HANDBOOK OF
NATURAL GAS
TRANSMISSION AND
PROCESSING
PRINCIPLES AND PRACTICES
THIRD EDITION
SAEID MOKHATAB, WILLIAM A. POE, AND JOHN Y. MAK
Handbook of Natural Gas Transmission and Processing
Disclaimer

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This book is dedicated to all professionals that preceded us; researchers, scientists, engineers, operators, and educators in the natural gas industry, who inspire us to assemble their knowledge and experience, manifesting this book that is much needed in today’s changing landscape of natural gas.
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Contents

About the Authors ........................................................................................................ xxiii
Foreword ...................................................................................................................... xxv
Preface to the Third Edition ...................................................................................... xxvii
Endorsements for the Third Edition ......................................................................... xxix

CHAPTER 1  Natural Gas Fundamentals ................................................................. 1
  1.1  Introduction ..................................................................................................... 1
  1.2  Natural gas history ....................................................................................... 1
  1.3  Natural gas origin and sources ..................................................................... 2
  1.4  Natural gas composition and classification ............................................... 3
  1.5  Natural gas phase behavior .......................................................................... 4
  1.6  Natural gas properties .................................................................................. 6
      1.6.1  Chemical and physical properties .............................................. 6
      1.6.2  Thermodynamic properties ..................................................... 15
  1.7  Natural gas reserves ..................................................................................... 16
  1.8  Natural gas exploration and production ..................................................... 16
      1.8.1  Conventional gas ................................................................. 17
      1.8.2  Unconventional gas ............................................................ 20
      1.8.3  Well deliverability .............................................................. 22
  1.9  Natural gas transportation .......................................................................... 24
      1.9.1  Pipelines .................................................................................. 25
      1.9.2  Liquefied natural gas .......................................................... 25
      1.9.3  Compressed natural gas ....................................................... 26
      1.9.4  Gas-to-liquids ....................................................................... 27
      1.9.5  Gas-to-solid ........................................................................ 28
      1.9.6  Gas-to-wire ......................................................................... 29
      1.9.7  Comparison of various methods ......................................... 30
  1.10 Natural gas processing .................................................................................. 33
  1.11 Sales gas transmission .................................................................................. 33
  1.12 Underground gas storage .......................................................................... 33
      1.12.1  Depleted reservoirs ............................................................ 34
      1.12.2  Aquifers ............................................................................. 34
      1.12.3  Salt caverns ....................................................................... 34
  References ............................................................................................................ 35

CHAPTER 2  Raw Gas Transmission ..................................................................... 37
  2.1  Introduction .................................................................................................. 37
  2.2  Multiphase flow terminology ..................................................................... 37
2.2.1 Superficial velocity .......................................................38
2.2.2 Mixture velocity ...........................................................38
2.2.3 Holdup ..........................................................................38
2.2.4 Phase velocity ...............................................................39
2.2.5 Slip ................................................................................39
2.2.6 Mixture density .............................................................40
2.2.7 Mixture viscosity ..........................................................40
2.2.8 Mixture pressure drop ..................................................41
2.2.9 Mixture enthalpy ...........................................................41

2.3 Multiphase flow regimes ........................................................42
2.3.1 Two-phase flow regimes ...............................................42
2.3.2 Three-phase flow regimes ............................................48
2.3.3 Gas-condensate flow regimes .........................................49

2.4 Determining multiphase flow design parameters ....................49
2.4.1 Steady-state two-phase flow .........................................50
2.4.2 Steady-state three-phase flow .......................................56
2.4.3 Transient multiphase flow ............................................57
2.4.4 Multiphase gas and condensate flow ............................59

2.5 Predicting temperature profile of multiphase pipeline ............60
2.6 Velocity criteria for sizing multiphase pipelines ..................64
2.7 Multiphase pipeline operations .........................................65
2.7.1 Leak detection ................................................................65
2.7.2 Pipeline depressurization .............................................66
2.7.3 Pigging .........................................................................66

2.8 Multiphase flow assurance .................................................68
2.8.1 Gas hydrates .................................................................68
2.8.2 Corrosion ......................................................................86
2.8.3 Wax .............................................................................90
2.8.4 Slugging .....................................................................102
2.8.5 Flow assurance risk management .................................111

References ............................................................................114

CHAPTER 3  Basic Concepts of Natural Gas Processing ...........123

3.1 Introduction .......................................................................123
3.2 Natural gas processing objectives ......................................123
3.3 Gas processing plant configurations ...................................124
3.3.1 Gas plant with hydrocarbon dewpointing ....................125
3.3.2 Gas plant for NGL production .....................................128
3.4 Finding the best gas processing route .................................131
3.5 Support systems ...............................................................132
3.5.1 Utility and off-site ........................................................132
3.5.2 Process control .............................................................132
3.5.3 Safety systems .............................................................133
CHAPTER 4  Phase Separation .................................................................137
  4.1  Introduction.............................................................................137
  4.2  Gravity separators.................................................................137
    4.2.1  General description.......................................................138
    4.2.2  Separator selection.......................................................140
    4.2.3  Gravity separation theory.............................................142
    4.2.4  Design considerations.................................................144
  4.3  Multistage separation.............................................................145
  4.4  Centrifugal separators...........................................................145
  4.5  Twister supersonic separator ................................................146
  4.6  Slug catchers.........................................................................148
  4.7  High-efficiency liquid/gas coalescers....................................150
    4.7.1  Aerosols.........................................................................150
    4.7.2  Coalescer construction/operation principles...............151
    4.7.3  Modeling the liquid/gas coalescer...............................154
    4.7.4  Coalescer performance/operational limits .................157
    4.7.5  Liquid/gas coalescer applications...............................157
  4.8  High-efficiency liquid–liquid coalescers................................157
    4.8.1  Emulsions.......................................................................157
    4.8.2  Coalescer principles and materials of construction.........158
    4.8.3  Coalescer mechanism of operation...............................158
    4.8.4  Liquid–liquid coalescer performance...........................161
    4.8.5  Limitations of using coalescers....................................162
    4.8.6  Applications..................................................................162
  4.9  Practical design of separation systems.................................162
    References................................................................................165

CHAPTER 5  Condensate Production......................................................169
  5.1  Introduction.............................................................................169
  5.2  Condensate stabilization........................................................170
    5.2.1  Stabilization by cascade flash separation.....................170
    5.2.2  Stabilization by distillation............................................171
    5.2.3  Design considerations....................................................173
    5.2.4  Operating problems.......................................................174
  5.3  Condensate hydrotreating.......................................................174
  5.4  Effluent treatment.................................................................176
5.4.1 MEG regeneration and reclaiming ......................... 176
5.4.2 Sour water stripping ............................................. 177
5.5 Condensate storage .................................................. 178
  5.5.1 Tank design considerations ................................... 178
  5.5.2 Tank emission control .......................................... 180

CHAPTER 6 Natural Gas Treating ........................................ 181
  6.1 Introduction .......................................................... 181
  6.2 Gas treating specifications ......................................... 181
  6.3 Gas treating processes ............................................. 182
  6.4 Chemical absorption processes ................................. 183
    6.4.1 Alkanolamine solvents ...................................... 183
    6.4.2 The potassium carbonate solution ........................ 195
  6.5 Physical solvent processes ....................................... 196
    6.5.1 Propylene carbonate ......................................... 198
    6.5.2 Dimethyl ether of polyethylene glycol .................. 202
    6.5.3 Methanol (Rectisol process) ............................... 207
    6.5.4 N-Methyl-2-pyrrolidone ................................... 208
  6.6 Mixed physical and chemical absorption processes ......... 208
  6.7 Solid bed absorption processes ................................. 210
    6.7.1 Iron sponge process .......................................... 210
    6.7.2 Zinc oxide process ........................................... 211
    6.7.3 PURASPEC™ ..................................................... 211
    6.7.4 Slurry processes .............................................. 213
  6.8 Solid bed adsorption process .................................... 213
  6.9 Membrane .................................................................. 214
  6.10 Cryogenic fractionation ........................................... 218
  6.11 Microbiological treatment processes ......................... 218
  6.12 Selecting the gas treating process ............................. 219
References ................................................................. 220

CHAPTER 7 Natural Gas Dehydration ................................. 223
  7.1 Introduction .......................................................... 223
  7.2 Water content determination ..................................... 224
  7.3 Glycol dehydration .................................................. 226
    7.3.1 Conventional TEG dehydration process ................. 227
    7.3.2 Enhanced TEG dehydration process ...................... 229
    7.3.3 Glycol injection process ..................................... 230
    7.3.4 TEG unit design considerations ........................... 232
    7.3.5 Operational problems ......................................... 235
    7.3.6 Future technology developments ......................... 237
  7.4 Solid-bed dehydration .............................................. 237
    7.4.1 Adsorption capacity ........................................... 238
    7.4.2 Adsorbent selection ........................................... 238
CHAPTER 7  Gas Dehydration Processes ........................................... 258
7.1 Introduction ............................................................................ 258
7.2 Dehydration ........................................................................... 258
7.2.1 Adsorption ......................................................................... 259
7.2.2 Operation of solid-bed dehydrator .................................... 259
7.2.3 Dehydration unit design considerations ............................ 261
7.2.4 Operational problems ......................................................... 263
7.3 Other gas dehydration processes ............................................. 264
7.4 Gas dehydration process selection .......................................... 265
7.4.1 Introduction ........................................................................ 265
7.4.2 Adsorption technology ...................................................... 266
7.4.3 Operation of solid-bed dehydrator .................................... 266
7.4.4 Dehydration unit design considerations ............................ 268
7.4.5 Operational problems ......................................................... 268
7.4.6 Other gas dehydration processes ........................................ 269
7.5 Gas dehydration process selection .......................................... 270
7.5.1 Introduction ........................................................................ 270
7.5.2 Adsorption technology ...................................................... 271
7.5.3 Operation of solid-bed dehydrator .................................... 271
7.5.4 Dehydration unit design considerations ............................ 273
7.5.5 Operational problems ......................................................... 273
7.6 Mercury removal.................................................................... 274
7.6.1 Nonregenerative mercury sorbents .................................... 274
7.6.2 Regenerative mercury adsorbents ..................................... 276
7.6.3 Mercury recovery ............................................................... 278
7.7 Process selection considerations ............................................ 278
7.8 References ............................................................................. 280

CHAPTER 8  Natural Gas Liquids Recovery .................................... 265
8.1 Introduction ............................................................................ 265
8.2 Refrigeration processes.......................................................... 266
8.2.1 Propane refrigeration ......................................................... 266
8.2.2 Cascade refrigeration ......................................................... 267
8.2.3 Mixed refrigerants ............................................................. 270
8.3 Liquid recovery processes ....................................................... 271
8.3.1 Hydrocarbon dew pointing with Joule–Thomson cooling .... 271
8.3.2 Hydrocarbon dew pointing with propane refrigeration .... 272
8.3.3 Deep hydrocarbon dew pointing ....................................... 273
8.3.4 Turboexpander NGL recovery processes ......................... 274
8.3.5 Lean oil absorption ........................................................... 279
8.3.6 Modern NGL recovery processes ..................................... 281
8.3.7 Other hydrocarbons removal processes ......................... 287
8.4 Selection of NGL recovery process ........................................ 289
8.5 NGL recovery technology development ............................... 290
8.6 NGL recovery unit design considerations ......................... 290
8.7 NGL recovery unit operating problems ................................. 290
8.8 NGL fractionation ................................................................. 291
8.9 Liquid product processing ..................................................... 293
8.9.1 NGL contaminant treating .............................................. 293
8.9.2 Dehydration ................................................................. 297
8.9.3 Liquid product processing .............................................. 298
8.9.4 Other hydrocarbons removal processes ......................... 301

CHAPTER 9  Sulfur Recovery and Handling .................................... 301
9.1 Introduction ............................................................................ 301
9.2 Sulfur properties ................................................................. 301
9.3 Sulfur recovery ................................................................. 302
9.3.1 Modified Claus process .................................................... 303
9.3.2 Direct oxidation processes .............................................. 309
9.3.3 Small- and medium-scale processes .................311
9.3.4 Microbiological treatment processes ..................314
9.4 Tail gas cleanup .................................................314
  9.4.1 Reduction processes .......................................314
  9.4.2 SO₂ scrubbing processes ...............................317
  9.4.3 Catalytic oxidation .........................................318
  9.4.4 Other tail gas treating configurations ...............318
9.5 Sulfur degassing .................................................320
  9.5.1 D’GAASS process .........................................321
  9.5.2 Aquisulf process ..........................................322
9.6 Sulfur storage and handling ..................................322
  9.6.1 Molten sulfur handling system .........................323
  9.6.2 Sulfur forming .............................................323
  9.6.3 Conveying formed sulfur ...............................324
  9.6.4 Storage of formed sulfur ...............................324
9.7 SRU design considerations ....................................324
  9.7.1 Piping ......................................................325
  9.7.2 Acid gas feed drums .......................................325
  9.7.3 Combustion air blowers ..................................326
  9.7.4 Main burner and reaction furnace .....................326
  9.7.5 Waste heat boiler .........................................326
  9.7.6 Sulfur condensers ........................................327
  9.7.7 Sulfur pit ..................................................327
9.8 SRU operation problems .......................................327
  9.8.1 Proper air ratio ............................................327
  9.8.2 Reactor activity ............................................328
  9.8.3 Excessive COS and CS₂ .................................329
  9.8.4 Leakage of reheat exchanger .........................329
  9.8.5 Reactor pressure drop ..................................329
  9.8.6 Carbon deposits .........................................329
  9.8.7 Catalyst support screens ..............................330
  9.8.8 Water vapor and carbon dioxide .....................330
  9.8.9 Steam heater .............................................330
  9.8.10 Combustion air control ..............................330
9.9 Selecting the sulfur recovery process ......................331
9.10 Sulfur disposal by acid gas injection .....................332
References ..................................................................333

CHAPTER 10  Nitrogen Rejection .................................335
10.1 Introduction ....................................................335
10.2 Nitrogen rejection options ..................................335
  10.2.1 Cryogenic processes ....................................335
  10.2.2 Noncryogenic processes ..............................336
10.3 Nitrogen rejection unit integration .......................336
10.4 Cryogenic nitrogen rejection ............................................. 338
  10.4.1 Classical single-column design .................................... 338
  10.4.2 Modified single-column design .................................. 339
  10.4.3 Double-column design ............................................ 340
  10.4.4 Two-column design ............................................... 343
  10.4.5 Process selection ............................................... 344
10.5 Design considerations .................................................. 345
  10.5.1 Feed gas characteristics ........................................ 345
  10.5.2 Reflux and reboiler duties ...................................... 345
  10.5.3 Temperature and pressure control ............................ 346
  10.5.4 Insulation ....................................................... 346
  10.5.5 Reboiler hydraulics ............................................. 346
10.6 Operating problems ..................................................... 346
  10.6.1 Feed contaminants ............................................. 346
  10.6.2 Foaming ......................................................... 347
  10.6.3 High methane content in nitrogen vent .................... 347
  10.6.4 Hydrate formation ............................................. 347
  10.6.5 Nitrogen safety ............................................... 347
  10.6.6 BAHX failure .................................................. 347

CHAPTER 11 Natural Gas Compression ....................................... 349
11.1 Introduction ............................................................ 349
11.2 Reciprocating compressors .......................................... 350
11.3 Centrifugal compressors ............................................. 351
11.4 Comparison between compressors ................................ 353
11.5 Compressor selection ................................................. 354
11.6 Thermodynamics of gas compression ............................... 355
  11.6.1 Basic relations .................................................. 356
  11.6.2 Isentropic model ............................................... 357
  11.6.3 Polytropic model ............................................... 360
  11.6.4 Real gas behavior ............................................. 361
11.7 Compression ratio ..................................................... 362
11.8 Compressor design ..................................................... 364
  11.8.1 Determining number of compression stages ............... 364
  11.8.2 Compression power calculation ............................... 366
11.9 Compressor control ................................................... 367
  11.9.1 Reciprocating compressors ................................... 368
  11.9.2 Centrifugal compressors ...................................... 369
11.10 Compressor performance maps ..................................... 375
  11.10.1 Reciprocating compressors .................................. 375
  11.10.2 Centrifugal compressors ................................... 375
11.11 Example for operating a compressor in a pipeline system ............................................. 376
References ........................................................................ 380
CHAPTER 12 Sales Gas Transmission .............................................383
12.1 Introduction ........................................................................383
12.2 Gas flow fundamentals .......................................................383
   12.2.1 General flow equation ............................................384
   12.2.2 Friction factor correlations ....................................385
   12.2.3 Simplified flow equations ......................................388
12.3 Predicting gas temperature profile .....................................390
12.4 Transient flow in gas transmission pipelines .....................392
12.5 Compressor stations ...........................................................394
   12.5.1 Station facilities .....................................................395
   12.5.2 Compressors arrangements ....................................399
   12.5.3 Station control ........................................................399
   12.5.4 Acoustical treatment ..............................................400
   12.5.5 Reliability and availability .....................................401
12.6 Reduction and metering stations ........................................402
   12.6.1 Filters......................................................................402
   12.6.2 Heaters....................................................................403
   12.6.3 Pressure reduction and regulation system .............403
   12.6.4 Metering system.....................................................403
12.7 Design considerations of sales gas pipelines.....................403
   12.7.1 Line sizing criteria .................................................403
   12.7.2 Compressor station spacing ...................................404
   12.7.3 Compression power ................................................408
12.8 Pipeline operations .............................................................409
   References ..........................................................................410

CHAPTER 13 Gas Processing Plant Automation .......................413
13.1 Introduction ........................................................................413
13.2 Early methods of gas plant automation.............................413
13.3 Microprocessor-based automation .....................................414
   13.3.1 Programmable logic controllers.............................414
   13.3.2 Distributed control systems ...................................414
   13.3.3 Standards and protocols .........................................416
13.4 Control of equipment and process systems .......................417
   13.4.1 Gas gathering .........................................................417
   13.4.2 Gas treating ............................................................418
   13.4.3 Sulfur recovery .......................................................418
   13.4.4 Gas dehydration .....................................................419
   13.4.5 Liquids recovery ....................................................420
   13.4.6 NGL fractionation..................................................422
   13.4.7 Centrifugal compressors ........................................422
   13.4.8 Centrifugal pumps..................................................423
   13.4.9 Reciprocating pumps .............................................423
   13.4.10 Utilities .................................................................423
13.5 Automation applications ................................................................. 424
  13.5.1 Data historians ........................................................................ 424
  13.5.2 Asset and performance management ...................................... 425
  13.5.3 Statistical process control ...................................................... 426
  13.5.4 Advanced regulatory control ................................................. 427
  13.5.5 Multivariable predictive control ............................................. 427
  13.5.6 Optimization ..................................................................... 429
  13.5.7 Leveraging automation .......................................................... 431
13.6 Condensate stabilizer case study ............................................... 433
References ......................................................................................... 436

CHAPTER 14 Gas Processing Plant Operations ..................................... 437
  14.1 Introduction .............................................................................. 437
  14.2 Commissioning and start-up .................................................... 437
    14.2.1 Mechanical completion and precommissioning .................. 437
    14.2.2 Control systems testing .................................................... 438
    14.2.3 Initial start-up procedures ............................................... 440
    14.2.4 Process commissioning ................................................... 440
    14.2.5 Performance testing ........................................................ 442
  14.3 Control room management ...................................................... 442
    14.3.1 Roles and responsibilities .................................................. 443
    14.3.2 Process safety management .............................................. 443
    14.3.3 Hazard and operability study .......................................... 445
    14.3.4 Layer of protection analysis .......................................... 446
    14.3.5 Fatigue mitigation .......................................................... 448
    14.3.6 Alarm management ........................................................ 449
    14.3.7 Training ....................................................................... 453
  14.4 Maintenance .......................................................................... 454
    14.4.1 Types of maintenance ....................................................... 455
    14.4.2 Enterprise asset management systems ............................ 457
    14.4.3 Reliability centered maintenance ..................................... 457
  14.5 Troubleshooting ...................................................................... 458
    14.5.1 Troubleshooting steps ....................................................... 459
    14.5.2 Troubleshooting documentation ...................................... 461
    14.5.3 Instrumentation ............................................................... 461
    14.5.4 Process troubleshooting .................................................. 463
  14.6 Turnarounds ............................................................................ 464
References .......................................................................................... 464

CHAPTER 15 Dynamic Simulation of Gas Processing Plants .................. 467
  15.1 Introduction .............................................................................. 467
  15.2 Areas of application of dynamic simulation .............................. 467
    15.2.1 Plant design ................................................................... 468
    15.2.2 Plant operation ............................................................... 471
15.3 Modeling considerations ....................................................473
  15.3.1 Level of detail in the model .........................................473
  15.3.2 Model speed .............................................................474
  15.3.3 Equipment specific considerations ................................474
15.4 Control of equipment and process systems .......................477
  15.4.1 Gas gathering and transmission .....................................477
  15.4.2 Gas treating .............................................................477
  15.4.3 Sulfur recovery .........................................................477
  15.4.4 Gas dehydration ........................................................477
  15.4.5 Liquids recovery, natural gas liquefaction .................478
  15.4.6 NGL fractionation ....................................................478
15.5 Case study I: Analysis of a fuel gas system start-up ........478
  15.5.1 Introduction ............................................................478
  15.5.2 Steady-state analysis ................................................479
  15.5.3 Dynamic analysis .....................................................480
  15.5.4 Conclusion ............................................................481
15.6 Case study II: Online dynamic model of a trunk line ..........481
References ........................................................................485

CHAPTER 16 Real-Time Optimization of Gas Processing Plants 487
16.1 Introduction ....................................................................487
16.2 Real-time optimization ..................................................487
  16.2.1 Physical properties ....................................................489
  16.2.2 Optimization models ...............................................491
  16.2.3 Plant model integration .............................................502
16.3 RTO project considerations ..............................................504
16.4 Example of RTO ...........................................................505
  16.4.1 Process description ....................................................506
  16.4.2 Plant operation ........................................................506
  16.4.3 Production objectives ...............................................509
  16.4.4 Project drivers ........................................................509
  16.4.5 Features of the optimization model .........................512
References ........................................................................516

CHAPTER 17 Maximizing Profitability of Gas Plant Assets 517
17.1 Introduction .....................................................................517
17.2 The performance strategy—integrated gas plant ..........518
17.3 Strategies for organizational behavior and information ....519
17.4 Organizational behavior model ........................................519
  17.4.1 Information quality ....................................................520
  17.4.2 Perception of information ..........................................522
  17.4.3 Capability to perform ...............................................524
  17.4.4 Organizational hierarchy of needs .........................526
17.5 The successful information strategy .............................................. 528
17.6 The impact of living with information technology .................. 529
17.7 Vision of the modern plant operation ........................................... 530
17.8 Operations strategy ................................................................. 531
17.9 Model-based asset management ................................................. 532
17.10 Optimization ......................................................................... 533
   17.10.1 Tools for optimization .................................................... 534
   17.10.2 Optimization alternatives ................................................ 534
17.11 Industrial relevance ............................................................... 536
17.12 The technology integration challenge .................................... 537
17.13 Scientific approach ............................................................... 537
17.14 Other miscellaneous initiatives .............................................. 539
17.15 Conclusion ........................................................................... 539
References .................................................................................. 541

CHAPTER 18 Gas Plant Project Management .................................. 543
18.1 Introduction ............................................................................ 543
18.2 Project management overview ................................................ 543
18.3 Industry perspective .............................................................. 544
18.4 The project management process .......................................... 545
   18.4.1 Defining business and project objectives ....................... 546
   18.4.2 Contracting strategy ...................................................... 548
   18.4.3 Conceptual estimates and schedules ............................ 549
   18.4.4 Project execution planning ........................................... 551
   18.4.5 Pre-project planning measurement ............................. 552
   18.4.6 The responsibility matrix ............................................. 552
18.5 Project controls ....................................................................... 554
   18.5.1 Project timeline ............................................................ 555
   18.5.2 Risk management ......................................................... 556
18.6 Quality assurance ................................................................. 564
18.7 Commissioning and start-up .................................................. 565
18.8 Operate and evaluate ............................................................ 566
18.9 Project closeout ..................................................................... 567
18.10 Conclusion ............................................................................ 567
References .................................................................................. 568

Appendix 1: Conversion Factors ..................................................... 571
Appendix 2: Standard Gas Conditions ............................................. 573
Appendix 3: Physical Properties of Fluids ........................................ 575
Index .......................................................................................... 581
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Saeid Mokhatab is one of the most recognizable names in the natural gas community through his contributions to advancing the technologies in the natural gas processing industry. He has worked in a variety of senior technical and managerial positions with major petroleum companies and has been actively involved in several large-scale gas-field development projects, concentrating on design, precommissioning, and start-up of processing plants. He has presented numerous invited lectures on gas processing technologies, and has authored or coauthored over 200 technical publications including two well-known Elsevier’s handbooks, which are considered by many as major references to be taken into account for any gas processing/LNG project in development. He founded the world’s first peer-reviewed journal devoted to the natural gas science and engineering (published by Elsevier, the United States); has held editorial positions in many scientific journals/book publishing companies for the hydrocarbon processing industry; and served as a member of technical committees for a number of professional societies and famous gas-processing conferences worldwide. As a result of his outstanding work in the natural gas industry, he has received a number of international awards/medals including the Einstein Gold Medal of Honor and Kapitsa Gold Medal of Honor; and his biography has been listed in highly prestigious directories.

William A. “Bill” Poe is a Senior Principal Technical Consultant at the Invensys Division of Schneider Electric, the United States. He has over 30 years of international business and industrial experience in design, operations, and project management of gas processing plants with a special focus on automation, multivariable predictive control (MPC), advanced process control (APC), optimization design and implementation, and real-time performance monitoring. Bill started his career at Shell Oil Company, the United States, in 1981, working over a decade in natural gas processing plants operations and engineering as well as management of multimillion-dollar projects. In 1993, he joined Continental Controls to lead the process engineering department in support of executing contracts with the Gas Research Institute, the United States, where he developed new multivariable control applications in the natural gas industry. After joining GE as part of the Continental Controls acquisition, he became vice president of this division of GE where his responsibilities included direction of product development, projects, technical sales support, and customer service for multivariable control and optimization applications in the natural gas industry. In 2001, Bill joined Invensys Process Systems, the United States, where he has developed APC and Optimization Master Plans for international companies such as Saudi Aramco, ADNOC, Statoil, and PDVSA, as well as automation and advanced process control feasibility studies for over 100 natural gas processing plants.
Bill is an Associate Editor of the *Journal of Natural Gas Science and Engineering*, has authored or coauthored more than 50 technical papers, and made numerous technical presentations at prestigious international conferences. He received the GE Innovators Award in 1999 and attained the Invensys Circle of Excellence in 2011.

**John Y. Mak** is a Senior Fellow and Technical Director at Fluor, the United States, and leads the technology and design development for the chemical and energy sectors at Fluor. He has been with Fluor for over 40 years and has been leading domestic and global oil and gas and refinery projects from conceptual design, feasibility study, FEED and detailed engineering to plant start-up and operation. John has made significant contributions to the technologies in natural gas treating, NGL recovery, LNG liquefaction and regasification, coal gasification, and carbon capture. He is the coauthor of the *Handbook of Liquefied Natural Gas*—first edition (2013) published by Elsevier, the United States, and has presented over 60 technical papers at the GPA (Gas Processors Association) and LRGCC (Laurance Reid Gas Conditioning Conference), China Coal Forum, and other technical conferences. John is the inventor of over 80 patent and patent-pending processes, which have been used in projects at Fluor. His patented technologies have been proven to improve energy efficiency, lower emissions, and reduce cost for many of his clients. John’s current focus is on liquid recovery for shale gas projects and treating of the difficult gases, such as the high-carbon dioxide and nitrogen content gases for offshore projects.
Foreword

Natural gas is an environment friendly, clean-burning fuel, offering important environmental benefits compared to other fossil fuels. It is also a remarkably safe source of energy when transported, stored, and used. The emergence of natural gas in the global markets further underscores the importance of gas transmission and processing. In this regard, it is fitting that technical books covering these subjects become required reading for technical specialists and project planners.

While several books exist for natural gas upstream engineering, the literature has lacked a comprehensive book that captures the midstream and downstream sectors of the natural-gas value chain. This book provides well-organized, theoretical, and technical background knowledge on natural gas transmission and processing; and covers subject areas missed by other books in these areas.

Because the authors have extensive experience in the design and operation of natural gas processing plants throughout the world, this book has attempted to concentrate upon what they perceive to be modern and proven practices. To keep things interesting for the experts, they have included some of the latest innovations and developments reported in the recent published works, although some material also reflects their own research involvement in problems related to natural gas transmission and processing. In summary, this book is a major contribution to the professional literature and will serve as a valuable desk reference for scientists, researchers, and engineers working in the upstream, midstream, and downstream sectors of natural gas industry. Literature references for in-depth study enhance the reference aspect of this comprehensive work. The organization of materials also permits flexibility in designing courses in the gas processing field for university students in chemical/petroleum engineering curricula.

Professor Kenneth R. Hall, PhD.,
Jack E. and Frances Brown, Chair and Professor, Department of Chemical Engineering, Texas A&M University, College Station, TX, USA;
Associate Dean, Research and Graduate Studies,
Texas A&M University at Qatar, Doha, Qatar.
Preface to the Third Edition

Many interesting and exciting developments in the natural gas industry have occurred since the publication of the second edition in 2012 that make it necessary to revise portions of the book and include additional materials that are becoming important in the gas processing industry. This realization has led us to embark on updating our work to make it a more complete, refreshed, and comprehensive reference for all major aspects of natural gas transmission and processing. The operating practices and guidelines for the gas processing units have been added and updated where we feel they are relevant. In this respect, we hope that the basic contents are useful as introductory materials in the chemical and petroleum engineering fields and the more detailed description of design and operation are useful for the engineers and plant operators.

The preparation of a book that covers such a broad subject requires many sources of information. There are many contributors of technical materials and we gratefully acknowledge our indebtedness to all of the individuals who contributed to the development of this book. An invaluable contribution to this edition is the insight by experts in their specialties and applications. Special thanks are due to friends and colleagues, who encouraged, assessed, and made this book possible. Among them are Dr Louis Mattar and Dr Mehran Pooladidarfish of the Fekete Associates Inc., Canada, who prepared the section on “Natural Gas Exploration and Production” in Chapter 1. We also appreciate Mr Cris Heijckers of Kranji Solutions Pte Ltd, Singapore, for preparing a section on “Practical Design of Separation Systems” in Chapter 4. We deeply acknowledge the greatest help of Dr Rainer Kurz of Solar Turbines Inc., the United States, in updating Chapter 11. We thank Mr Sidney P. Santos of At Work Rio Engineering and Consulting Ltd, Brazil, for his constructive comments and suggestions on Chapter 12. We also express our sincere thanks to Professor Kenneth R. Hall of the Texas A&M University, the United States, who accepted to write its foreword. Finally, we appreciate the editorial staff members of Elsevier who have been an excellent source of strong support during the preparation and publication of this book.

We know that this handbook required updating to reflect the current state of the art in the changing landscape of natural gas processing. In order to meet this requirement, the authors have endeavored to assemble the best available knowledge and expertise to make this work as current as possible, but by no means is this work perfect. The accelerated delivery of information contained in this handbook is important as we all desire to make use of the latest technology for a greener and safer environment through the effective use of natural gas. Our ultimate mission is to continuously follow progress in this exciting industry and
add innovations as well as expanded topics in future editions of this handbook while retaining the basic fundamental and practical materials already available in this edition.

Saeid Mokhatab
William A. Poe
John Y. Mak
Endorsements for the Third Edition

This handbook is a valuable reference that covers all aspects of the natural gas transmission and processing industries. It contains much needed design, operation and optimization information, all in a single source and does an excellent job of highlighting the key considerations for any gas processing project, as well as providing innovative solutions in natural gas liquids recovery and treating high nitrogen and carbon dioxide content gases in unconventional gas plants. It is a key addition to any gas processing professional’s library.

Jason Kraynek, Vice President of Business Transformation & Innovations, Fluor, USA.

This well-balanced handbook is the only book of its kind, covering all aspects of natural gas transmission and processing in more details. I believe it will serve as a valuable desk reference for practicing gas engineers and technologists, and as a text for graduate students in the gas engineering curriculum.

Dr J.C. Kuo, Senior Advisor of Gas Team, Process Engineering, Chevron ETC, USA.

This high quality, comprehensive book gives an accurate picture of where the natural gas transmission and processing industry stands today, and describes some relatively new technologies that could become important in the future. I recommend this book for any professional gas processing engineer and technologist.

David Messersmith, Bechtel Fellow and Manager of LNG Technology and Services Group, Bechtel OG&C, USA.

This is the only book that covers all technical and operational aspects of natural gas transmission and processing as well as subject areas missed by other similar references. I recommend that if you work in the natural gas industry, you have this unique reference available.

Dr Jaleel V. Valappil, Manager, Advanced Simulation/Process, Bechtel OG&C, USA.

This is a valuable handbook to both the experienced engineer and the graduate just commencing in natural gas engineering. It provides practical advice for design and operation based on sound engineering principles and established techniques as well as introducing process solutions based on new and emerging technologies.

Adrian Finn, Manager of Process Technology, Costain Natural Resources, UK.

This book does an excellent job of describing the fundamental handling of natural gas from the wellhead to the consumer. I particularly found the
segment on natural gas liquids recovery most informative. I plan to make this very informative publication available to my employees as an excellent training tool as well.

Randy Johnson, Operations Manager, LNG, Energy Transfer, USA.

Must have book for anyone in the Gas Processing Industry. The authors did an incredible job at covering the process dynamics and introducing control theories for optimal operation. Absolutely enjoyed the descriptive nature of the chapters without compromising technical details.

Cesar A. Felizzola, Process Control Engineer, South Region Engineering, DCP Midstream, USA.

This comprehensive book provides in-depth coverage of all technical aspects of natural gas transmission and processing, beyond those addressed in other books. This is a “must addition” to library of anyone working in the midstream and downstream sectors of natural gas utilization to achieve higher career goals. I commend the authors’ continuous effort to make it an excellent source book for all professionals, engineers, and scientists in the natural gas industry.

Dr Suresh C. Sharma, ONEOK Chair Professor and Director of Natural Gas Engineering and Management, University of Oklahoma, USA.

This comprehensive book provides a state-of-the-art treatment of the different aspects of natural gas transmission and processing from the fundamental principles to the latest technology developments. It is a unique reference for all professionals involved in natural gas industry and an excellent textbook for graduate programs on the subject.

Dr Valerio Cozzani, Professor of Chemical Engineering and Director of Post-Graduate Program on Oil & Gas Process Design, University of Bologna, Italy.

This unique handbook, written by internationally renowned gas-engineering experts, is a major contribution to the professional and scholarly literature, offering an excellent coverage of key topics in the natural gas supply chain. It addresses the principles, practices, advanced technologies, new issues and challenges related to the natural gas transmission and processing industry, which have not been addressed in depth in any existing books. I recommend it highly, as a reference and textbook.

Dr Brian F. Towler, Professor and Chair of Petroleum Engineering, University of Queensland, Australia.
1.1 Introduction

Natural gas is the most energy efficient fossil fuel—it offers important energy saving benefits when it is used instead of oil or coal. Although the primary use of natural gas is as a fuel, it is also a source of hydrocarbons for petrochemical feedstock and a major source of elemental sulfur, an important industrial chemical. Its popularity as an energy source is expected to grow substantially in the future because natural gas can help achieve two important energy goals for the twenty-first century—providing the sustainable energy supplies and services needed for social and economic development and reducing adverse impacts on global climate and the environment in general. Natural gas consumption and trade have been growing steadily over the past two decades and natural gas has strengthened its position in the world energy mix. Although natural gas demand declined in 2009, as a result of the economic slowdown, it is expected to resume growth in both emerging and traditional markets in the coming decades. Such increase in the near future will be driven because of additional demand in current uses, primarily power generation. There is yet little overlap between the use of natural gas and oil in all large markets. However, there are certain moves in the horizon, including the electrifying of transportation, which will push natural gas use to ever higher levels.

This chapter gives the reader an introduction to natural gas by describing the origin and composition of natural gas, gas sources, phase behavior and properties, and transportation methods.

1.2 Natural gas history

The discovery of natural gas dates from ancient times in the Middle East. It was considered by ancients to be a supernatural manifestation. Noticed only when ignited, it appeared as a mysterious fire bursting from fissures in the ground. Natural gas seeps were discovered in Iran between 6000 and 2000 BC. The practical use of natural gas dates back to the Chinese of 2500 years ago, who used bamboo pipes to collect it from natural seeps and convey it to gas-fired evaporators, where it was used to boil ocean water for the salt. Apparently, natural gas was unknown in Europe until its discovery in England in 1659. However, since manufactured gas (coal gas) was already commercially available, natural gas
remained unpopular. In 1815, natural gas was discovered in the United States during the digging of a salt-brine well in Charleston, West Virginia. In 1821, an American gunsmith named William Aaron Hart drilled the first natural gas well in the United States. It was covered with a large barrel, and the gas was directed through wooden pipes that were replaced a few years later with lead pipe. One of the earliest attempts of monetization occurred in 1824 in Fredonia, New York, which led to the formation of the first natural gas company in the United States, the Fredonia Gas Light Company, in 1858.

The nineteenth century is considered as the starting point of the gas industry. In the early 1900s, huge amounts of natural gas were found in Texas and Oklahoma, and in the 1920s modern seamless steel pipe was introduced. The strength of this new pipe, which could be welded into long sections, allowed gas to be carried under higher pressures and, thus, in greater quantities. As the technology to create seamless steel pipe and related equipment advanced, the size and length of pipelines increased, as did the volumes of gas that could be transported easily and safely over many miles. The first natural gas pipeline longer than 200 miles was built in 1925, from Louisiana to Texas.

Steady growth in the use of gas marked the early and mid-twentieth century. However, it was the shortages of crude oil in the late 1960s and early 1970s that forced major industrial nations to seek energy alternatives. Since those events, gas has become a central fossil fuel energy source. Today, natural gas has become extremely important as a concentrated, clean fuel for home heating and cooking and electrical power generation, and is sought after almost as much as oil.

1.3 Natural gas origin and sources

There are different theories as to the origins of fossil fuels. The most widely accepted theory of the origin of natural gas assumes that natural gas hydrocarbons come from organic matter (the remains of land and aquatic plants, animals and microorganisms) that was trapped within sediments as they were deposited and transformed over long periods of time into their present form. Two main mechanisms, namely, biogenic and thermogenic are responsible for the degradation of fossil organic material in sediments (Rojey et al., 1997). Biogenic gas is formed at shallow depths and low temperatures due to the action of bacteria on the organic debris accumulating in the sediments. In contrast, thermogenic gas is formed at deeper depths by degradation of organic matter, called kerogen, accumulated in fine-grained sediments, especially clays and shales. This degradation occurs through the combined effects of temperature and pressure. Thermogenic gas is believed to be produced by two mechanisms, namely, direct thermal cracking of sedimentary organic matter and secondary thermal cracking of oil that is formed in the first stage. The former is called the primary thermogenic gas that coexists with oil, while the latter is called secondary thermogenic gas that coexists with insoluble solid matter, called pyrobitumen. Both
mechanisms involve thermal cracking with some degree of sustained pressure, mainly through the weight of the sedimentary formation. Little information is available on the time required to generate thermogenic gas other than the general assumption that it is a long time.

Natural gas comes from both “conventional” (easier to produce) and “unconventional” (more difficult to produce) geological formations. Conventional gas is typically “free gas” trapped in multiple, relatively small, porous zones in various naturally occurring rock formations such as carbonates, sandstones, and siltstones. Conventional natural gas generally occurs in deep reservoirs, either associated with crude oil (associated gas1) or in reservoirs that contain little or no crude oil (nonassociated gas2). Natural gas from coal (also known as coal-bed methane), tight gas sands, gas shales, geopressurized aquifers, and gas hydrates3 are often referred to as unconventional gas resources. The common characteristic of the different types of unconventional gas resources is that they contain large quantities of natural gas, but it is usually more difficult to produce this gas as compared to conventional reservoir rocks. New technologies are continually being developed to allow more accurate estimations of the amount of gas in these unconventional reservoirs and to stimulate these rocks to produce the gas.

1.4 Natural gas composition and classification

Natural gas is a complex mixture of hydrocarbon and nonhydrocarbon constituents and exists as a gas under atmospheric conditions. Virtually hundreds of different compounds may be present in natural gas in varying amounts. Even two wells producing from the same reservoir may produce gases of different composition as the reservoir is depleted.

While natural gas is formed primarily of methane (CH₄), it can also include significant quantities of ethane (C₂H₆), propane (C₃H₈), butane (C₄H₁₀), pentane (C₅H₁₂), as well as traces of hexane (C₆H₁₄) and heavier hydrocarbons. Many natural gases often contain nitrogen (N₂), carbon dioxide (CO₂), hydrogen sulfide (H₂S), and other sulfur components such as mercaptans (R-SH), carbonyl sulfide (COS), and carbon disulfide (CS₂). Trace quantities of argon, hydrogen, and helium may also be present. Trace quantities of metallic substances are known to exist in natural gases including arsenic, selenium, mercury, and uranium.

According to the proportion of hydrocarbons heavier than methane, different types of natural gas (dry, wet, and condensate) can be considered. Natural gas is considered “dry” when it is almost pure methane, having had most of the other commonly associated hydrocarbons removed. When other hydrocarbons are

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1Associated gas is produced with the oil and separated at the casinghead or wellhead. Gas produced in this fashion is also referred to as casinghead gas or oil well gas.
2Nonassociated gas is sometimes referred to as gas-well gas.
3Gas hydrates are icelike structures of water and gas located under the permafrost.
4R signifies an organic group such as a methyl, ethyl, propyl, or other group.
present, the natural gas is “wet,” where it forms a liquid phase during production at surface conditions. “Condensate” gases have a high content of hydrocarbon liquids and form a liquid phase in the reservoir during production, during the depletion process.

Natural gases are commonly classified according to their liquid content as either lean or rich and according to their sulfur content as either sweet or sour. The lean and rich terms refer to the amount of potentially recoverable liquids. The term usually applies to ethane and heavier components but sometimes applies instead to propane and heavier components (if ethane is not regarded as a valuable liquid component). To quantify the liquids content of a natural gas mixture, the industry uses GPM, or gallons of liquids recoverable per 1000 standard cubic feet (Mscf) of gas. Lean natural gas has liquid content less than 2 GPM. Moderately rich natural gas has between 2 and 5 GPM, and very rich natural gas has greater than 5 GPM (Ewan et al., 1975).

The sweet and sour terms refer to the H2S content. Strictly speaking, “sweet” and “sour” refer to both acid gases (H2S and CO2) but are usually applied to H2S alone. A sweet gas contains negligible amounts of H2S, whereas a sour gas has unacceptable quantities of H2S. The terms are relative, but generally, sweet means the gas contains less than 4 ppmv of H2S. Carbon dioxide can be tolerated to much higher levels, say 3–4 mol%, as long as the heating value of the sales gas is satisfactory.

1.5 Natural gas phase behavior

Natural gas is a naturally occurring hydrocarbon mixture that is found underground and at elevated conditions of pressure and temperature. Therefore, there is an essential need to know a priori how the gas fluid will behave under a wide range of pressure and temperature conditions, particularly in terms of its volumetric and thermophysical properties that are required in simulating reservoirs, evaluating reserves, forecasting production, designing production facilities, and designing gathering and transportation systems. In fact, an accurate knowledge of hydrocarbon fluid phase behavior is crucial in designing and operating the gas-engineering processes efficiently and optimally. This means, having advanced predictive tools for the characterization of hydrocarbon phase behavior with the highest accuracy possible is the key to mastering the economics of natural gas systems.

The natural gas phase behavior is a plot of pressure versus temperature that determines whether the natural gas stream at a given pressure and temperature consists of a single gas phase or two phases, gas and liquid. The phase behavior for natural gas with a given composition is typically displayed on a phase diagram, an example of which is shown in Figure 1-1. The left-hand side of the curve

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5In the petroleum industry, 1000 is usually abbreviated by the Roman numeral “M.”
is the bubble point line and divides the single-phase liquid region from the two-phase gas–liquid region. The right-hand side of the curve is the dew point line and divides the two-phase gas–liquid region and the single-phase gas region. The bubble point and dew point lines intersect at the critical point, where the distinction between gas and liquid properties disappears. The maximum pressure at which liquids can form is called the cricondenbar (Pcc), and the maximum temperature at which liquids can form is called the cricondentherm (Tcc).

However, there is something very interesting going on within the region \( T_c < T < T_{cc} \), where we will be moving from a 0% liquid to another 0% liquid condition (both on the dew point curve) in an isothermal compression. This different behavior of a vapor under compression is called retrograde (contrary to expectation) condensation. It is also important to see that a similar behavior is to be expected within the region \( P_c < P < P_{cc} \). In this case, we talk about retrograde vaporization since we will be moving from a 100% liquid to another 100% liquid condition (both on the bubble point curve) in an isobaric heating.

The natural gas phase behavior is a function of the composition of the gas mixture and is strongly influenced by the concentration of the heavier hydrocarbons, especially hexane plus. The presence of heavier hydrocarbons will increase the phase envelope, and failure to include them in a phase calculation will underpredict the phase envelope. There is also an essential need for proper characterizing the heavy-ends. In fact, although some different fluid descriptions

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FIGURE 1-1
Pressure–Temperature diagram for a typical natural gas mixture.

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\(^6\)Hexane plus or \( C_6^+ \) is a terminology used in natural gas engineering which indicates a grouping of heavier hydrocarbons including the hexane and heavier hydrocarbons such as heptane, octane, nonane, etc.
match to some extent the behavior of the reservoir fluids at reservoir conditions, they exhibit larger variations once surface simulators are used and the fluids are subjected to process conditions.

1.6 Natural gas properties

1.6.1 Chemical and physical properties

Natural gas is colorless, odorless, tasteless, shapeless, and lighter than air (see Table 1-1). The natural gas after appropriate treatment for acid gas reduction, and moisture and hydrocarbon dew point adjustment, would then be sold within prescribed limits of pressure, heating value and possibly Wobbe Index (often referred to as the Wobbe Number).

One of the principal uses of natural gas is as a fuel, and consequently, pipeline gas is normally bought and sold on the basis of its heating value that can be produced by burning the gas. The heating value of natural gas is variable and depends on its accumulations which are influenced by the amount and types of gases they contain. The gas industry always uses the gross heating value (frequently called higher heating value) in custody transfer. Obviously, the numerical difference between the two net and gross heating values is the heat of condensation of water at the specified conditions. Heating values for custody transfer are determined either by direct measurement, in which calorimetry is used, or by computation of the value on the basis of gas analysis (Gas Processors Association, 1996). The heating value of natural gas is measured in British thermal unit (Btu). A British thermal unit is the energy required to raise the temperature of 1 pound of water by 1 °F. For larger industrial customers, the abbreviations MBtu (1000 Btu) or MMBtu (1 million Btu) are more commonly used. Since meters measure volume and not energy content, a conversion factor of 1000 Btu/ft³ is commonly used by gas companies to convert the volume of gas used to its heat equivalent, and thus calculate the actual energy use.

The Wobbe Index (defined as the gross heating value of the gas divided by the square root of the specific gravity) gives a measure of the heat input to an appliance through a given aperture at a given gas pressure. Using this as a vertical coordinate and the flame speed factor as the horizontal coordinate, a combustion diagram can be constructed for an appliance, or a whole range of appliances, with the aid of appropriate test gases. This diagram shows the area within which variations in the Wobbe Index and flame speed factor of gases may occur for the given range of appliances without resulting in either incomplete combustion, flame lift, or the lighting back of preaerated flames. This method of prediction of

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Since the heat energy of the natural gas is related to the relative proportion of “lighter” methane versus “heavier” ethane, propane, butane, pentane, and other components, heat energy is not a constant value between different natural gas sources (Chandra, 2006).
combustion characteristics is not sufficiently accurate to eliminate entirely the need for the practical testing of new gases.

Since natural gas as delivered to pipelines has practically no odor, the addition of an odorant is required by most regulations in order that the presence of the gas can be detected readily in case of accidents and leaks. This odorization is provided by the addition of trace amounts of some organic sulfur compounds to the gas before it reaches the consumer. The sulfur compound, a chemical odorant (a mercaptan also called a thiol with the general formula R-SH and the odor of rotten eggs) is added to natural gas so that it can be smelled if there is a gas leak. The standard requirement is that a user will be able to detect the presence of the gas by odor when the concentration reaches 1% of gas in air. Since the lower limit of flammability of natural gas is approximately 5%, this requirement is equivalent to one-fifth the lower limit of flammability. The combustion of these trace amounts of odorant does not create any serious problems of sulfur content or toxicity.

In the following section, we discuss important gas properties including specific gravity, compressibility factor, formation volume factor, density, isothermal compressibility, and viscosity.

<table>
<thead>
<tr>
<th>Table 1-1 Properties of Natural Gas</th>
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</thead>
<tbody>
<tr>
<td><strong>Properties</strong></td>
</tr>
<tr>
<td>Relative molar mass</td>
</tr>
<tr>
<td>Carbon content, weight %</td>
</tr>
<tr>
<td>Hydrogen content, weight %</td>
</tr>
<tr>
<td>Oxygen content, weight %</td>
</tr>
<tr>
<td>Hydrogen/carbon atomic ratio</td>
</tr>
<tr>
<td>Relative density, 15 °C</td>
</tr>
<tr>
<td>Boiling point, °C</td>
</tr>
<tr>
<td>Autoignition temperature, °C</td>
</tr>
<tr>
<td>Octane number</td>
</tr>
<tr>
<td>Methane number</td>
</tr>
<tr>
<td>Stoichiometric air/fuel ratio, weight</td>
</tr>
<tr>
<td>Vapor flammability limits, volume %</td>
</tr>
<tr>
<td>Flammability limits</td>
</tr>
<tr>
<td>Lower heating/calorific value, MJ/kg</td>
</tr>
<tr>
<td>Stoichiometric lower heating value, MJ/kg</td>
</tr>
<tr>
<td>Methane concentration, volume %</td>
</tr>
<tr>
<td>Ethane concentration, volume %</td>
</tr>
<tr>
<td>Nitrogen concentration, volume %</td>
</tr>
<tr>
<td>Carbon dioxide concentration, volume %</td>
</tr>
<tr>
<td>Sulfur concentration, weight % ppm</td>
</tr>
<tr>
<td>Specific CO₂ formation, g/MJ</td>
</tr>
</tbody>
</table>
1.6.1.1 Gas specific gravity

Specific gravity of a natural gas is defined as the ratio of gas density to the density of air, both defined at the same pressure and temperature. These densities are usually defined at standard conditions (14.7 psia and 60 °F). Therefore, specific gravity of gas is defined as:

\[ \gamma_g = \frac{M}{M_{\text{air}}} \]  

(1-1)

where \( M \) is the molecular weight of gas, and \( M_{\text{air}} \) is the molecular weight of air that is equal to 29. Once we can calculate the value of molecular weight of mixture, we can calculate the specific gravity of mixture. For a gas mixture, we can calculate the molecular weight as:

\[ M = \sum_{i=1}^{n} y_i M_i \]  

(1-2)

where \( M_i \) is the molecular weight of component \( i \), \( y_i \) is the mole fraction of component \( i \), and \( n \) is the total number of components.

Various gas properties, including the molecular weights for pure components, are given in Table 1-2.

1.6.1.2 Gas compressibility factor

The volume of a real gas is usually less than what the volume of an ideal gas would be, and hence a real gas is said to be super-compressible. The ratio of the real volume to the ideal volume, which is a measure of the amount the gas deviates from perfect behavior, is called the super-compressibility factor, sometimes shortened to the compressibility factor. It is also called the gas deviation factor, and given the symbol “\( Z \).” The gas deviation factor is by definition the ratio of the volume actually occupied by a gas at a given pressure and temperature to the volume it would occupy if it behaved ideally.

The real gas equation of state (EOS) is then written as:

\[ PV = ZnRT \]  

(1-3)

where \( P \) is the pressure, \( V \) is the volume, \( T \) is the absolute temperature, \( Z \) is the compressibility factor, \( n \) is the number of kilomoles of the gas, and \( R \) is the gas constant.

The gas deviation factor, \( Z \), is close to 1 at low pressures and high temperatures which means the gas behaves as an ideal gas at these conditions. At standard or atmospheric conditions, the gas \( Z \)-factor is always approximately 1.

Empirical correlations for \( Z \)-factor for natural gases were developed before the advent of digital computers. Although their use is in decline, they can still be used for rapid estimates of the \( Z \)-factor. Chart lookup is another means of determining \( Z \)-factor of natural gas mixtures. These methods are invariably based on some type of corresponding states development. According to the theory of
<table>
<thead>
<tr>
<th>Compound</th>
<th>Formula</th>
<th>Molecular Weight</th>
<th>$P_c$ psia (kPa)</th>
<th>$T_c$ °R (°K)</th>
<th>$V_c$ ft³/lb (m³/kg)</th>
<th>$Z_c$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane</td>
<td>CH₄</td>
<td>16.043</td>
<td>667.8 (4604)</td>
<td>343 (190.6)</td>
<td>0.0991 (0.0062)</td>
<td>0.2884</td>
</tr>
<tr>
<td>Ethane</td>
<td>C₂H₆</td>
<td>30.070</td>
<td>707.8 (4880)</td>
<td>549.8 (305.4)</td>
<td>0.0788 (0.00492)</td>
<td>0.2843</td>
</tr>
<tr>
<td>Propane</td>
<td>C₃H₈</td>
<td>44.097</td>
<td>616.3 (4249)</td>
<td>665.7 (369.8)</td>
<td>0.0737 (0.0046)</td>
<td>0.2804</td>
</tr>
<tr>
<td>n-Butane</td>
<td>C₄H₁₀</td>
<td>58.124</td>
<td>550.7 (3797)</td>
<td>765.3 (425.2)</td>
<td>0.0702 (0.00438)</td>
<td>0.2736</td>
</tr>
<tr>
<td>Isobutane</td>
<td>C₄H₁₀</td>
<td>58.124</td>
<td>529.1 (3648)</td>
<td>734.7 (408.2)</td>
<td>0.0724 (0.00452)</td>
<td>0.2824</td>
</tr>
<tr>
<td>n-Pentane</td>
<td>C₅H₁₂</td>
<td>72.151</td>
<td>488.6 (3369)</td>
<td>845.4 (469.7)</td>
<td>0.0675 (0.00422)</td>
<td>0.2623</td>
</tr>
<tr>
<td>Isopentane</td>
<td>C₅H₁₂</td>
<td>72.151</td>
<td>490.4 (3381)</td>
<td>828.8 (460.4)</td>
<td>0.0679 (0.00424)</td>
<td>0.2701</td>
</tr>
<tr>
<td>Neopentane</td>
<td>C₅H₁₂</td>
<td>72.151</td>
<td>464.0 (3199)</td>
<td>781.11</td>
<td>0.0674 (0.00421)</td>
<td>0.2537</td>
</tr>
<tr>
<td>n-Hexane</td>
<td>C₆H₁₄</td>
<td>86.178</td>
<td>436.9 (3012)</td>
<td>913.4 (507.4)</td>
<td>0.0688 (0.0043)</td>
<td>0.2643</td>
</tr>
<tr>
<td>n-Heptane</td>
<td>C₇H₁₆</td>
<td>100.205</td>
<td>396.8 (2736)</td>
<td>972.5 (540.3)</td>
<td>0.0691 (0.00432)</td>
<td>0.2633</td>
</tr>
<tr>
<td>n-Octane</td>
<td>C₈H₁₈</td>
<td>114.232</td>
<td>360.6 (2486)</td>
<td>1023.9 (568.8)</td>
<td>0.0690 (0.0043)</td>
<td>0.2587</td>
</tr>
<tr>
<td>n-Nonane</td>
<td>C₉H₂₀</td>
<td>128.30</td>
<td>332 (2289)</td>
<td>1070.3 (594.6)</td>
<td>0.0684 (0.00427)</td>
<td>0.2536</td>
</tr>
<tr>
<td>n-Decane</td>
<td>C₁₀H₂₂</td>
<td>142.30</td>
<td>304 (2096)</td>
<td>1111.8 (617.7)</td>
<td>0.0679 (0.00424)</td>
<td>0.2462</td>
</tr>
<tr>
<td>Ethylene</td>
<td>C₂H₄</td>
<td>28.054</td>
<td>729.8 (5032)</td>
<td>508.6 (282.6)</td>
<td>0.0737 (0.0046)</td>
<td>0.2765</td>
</tr>
<tr>
<td>Propene</td>
<td>C₃H₆</td>
<td>42.081</td>
<td>669 (4613)</td>
<td>656.9 (364.9)</td>
<td>0.0689 (0.0043)</td>
<td>0.2752</td>
</tr>
<tr>
<td>Acetylene</td>
<td>C₂H₂</td>
<td>26.038</td>
<td>890.4 (6139)</td>
<td>555.3 (308.5)</td>
<td>0.0695 (0.00434)</td>
<td>0.2704</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>CO₂</td>
<td>44.010</td>
<td>1071 (7382)</td>
<td>547.6 (304.2)</td>
<td>0.0342 (0.00214)</td>
<td>0.2742</td>
</tr>
<tr>
<td>Hydrogen sulfide</td>
<td>H₂S</td>
<td>34.076</td>
<td>1306 (9005)</td>
<td>672.4 (373.6)</td>
<td>0.0459 (0.00287)</td>
<td>0.2831</td>
</tr>
<tr>
<td>Sulfur dioxide</td>
<td>SO₂</td>
<td>64.059</td>
<td>1145. (7894)</td>
<td>775.5 (430.8)</td>
<td>0.0306 (0.00191)</td>
<td>0.2697</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>N₂</td>
<td>28.013</td>
<td>493 (3399)</td>
<td>227.3 (126.3)</td>
<td>0.0514 (0.00321)</td>
<td>0.2916</td>
</tr>
<tr>
<td>Water</td>
<td>H₂O</td>
<td>18.015</td>
<td>3208. (22,105)</td>
<td>1165.0 (647.2)</td>
<td>0.0500 (0.00312)</td>
<td>0.2350</td>
</tr>
</tbody>
</table>
corresponding states, substances at corresponding states will exhibit the same behavior. The theory of corresponding states dictates that the Z-factor can be uniquely defined as a function of reduced pressure and reduced temperature. The reduced pressure and reduced temperature are defined as:

$$P_r = \frac{P}{P_c} \quad \text{and} \quad T_r = \frac{T}{T_c}$$

(1-4)

where $P_r$ and $T_r$ are reduced pressure and reduced temperature, respectively; and $P_c$ and $T_c$ are critical pressure and critical temperature of the gas, respectively. The values of critical pressure and critical temperature can be estimated from the following equations if the composition of the gas and the critical properties of the individual components are known.

$$P_c = \sum_{i}^n P_{ci} y_i \quad \text{and} \quad T_c = \sum_{i}^n T_{ci} y_i$$

(1-5)

where $P_{ci}$ and $T_{ci}$ are the critical pressure and critical temperature of component $i$, respectively; and $y_i$ is the mole fraction of component $i$.

Once critical properties of the mixture are calculated as stated in Eqn (1-5), we can use Eqn (1-4) to calculate the reduced properties of the mixture.

The values of critical pressure and critical temperature can be estimated from its specific gravity if the composition of the gas and the critical properties of the individual components are not known. The method uses a correlation to estimate pseudo-critical temperature and pseudo-critical pressure values from the specific gravity. There are several different correlations available. The most common is the one proposed by Sutton (1985), which is based on the basis of 264 different gas samples. Sutton (1985) used regression analysis on the raw data to obtain the following second order fits for the pseudo-critical properties.

$$P_{pc} = 756.8 - 131.07 \gamma_g - 3.6 \gamma_g^2$$

(1-6)

$$T_{pc} = 169.2 + 349.5 \gamma_g - 74.0 \gamma_g^2$$

(1-7)

These equations are valid over the range of specific gas gravities with which Sutton (1985) worked $0.57 < \gamma_g < 1.68$.

The most commonly used method to estimate the Z-factor is the chart provided by Standing and Katz (1942). The Z-factor chart is shown in Figure 1-2. The chart covers the range of reduced pressure from 0 to 15, and the range of reduced temperature from 1.05 to 3. This chart is generally reliable for sweet natural gases with minor amounts of nonhydrocarbons. It was developed using data for binary mixtures of methane with propane, ethane, butane, and natural gases having a wide range of composition. None of the gas mixtures had molecular weight in greater than 40. For low-molecular weight gases, it was found that the Z-factor estimated from Standing and Katz (1942) chart has error in the order of 2–3%. However, for gas mixtures whose components greatly differ
in molecular weight from 40, this chart provides inaccurate Z-factors (Elsharkawy et al., 2001).

Wichert and Aziz (1972) have developed a correlation to account for inaccuracies in the Standing and Katz chart when the gas contains significant fractions of acid gases, specifically carbon dioxide (CO₂) and hydrogen sulfide (H₂S). The Wichert and Aziz (1972) correlation modifies the values of the

FIGURE 1-2
Compressibility of natural gases as a function of reduced pressure and temperature (Standing and Katz, 1942).
pseudo-critical temperature and pressure of the gas. Once the modified pseudo-critical properties are obtained, they are used to calculate pseudo-reduced properties and the Z-factor is determined from Figure 1-2. The Wichert and Aziz (1972) correlation first calculates a deviation parameter \( \varepsilon \):

\[
\varepsilon = 120 \left( A^{0.9} - A^{1.6} \right) + 15 \left( B^{0.5} - B^4 \right)
\]  

(1-8)

where:

- \( A \) is the sum of the mole fractions of CO\(_2\) and H\(_2\)S in the gas mixture
- \( B \) is the mole fraction of H\(_2\)S in the gas mixture.

Then, \( \varepsilon \) is used to determine the modified pseudo-critical properties as follows:

\[
T_{pc}' = T_{pc} - \varepsilon
\]  

(1-9)

\[
P_{pc}' = \frac{P_{pc} T_{pc}}{T_{pc} - B (1 - B) \varepsilon}
\]  

(1-10)

The correlation is applicable to concentrations of CO\(_2\) < 54.4 mol % and H\(_2\)S < 73.8 mol%. Wichert and Aziz (1972) found their correlation to have an average absolute error of 0.97% over the following ranges of data: 154 psia < \( P \) < 7026 psia and 40 °F < \( T \) < 300 °F.

Methods of direct calculation using corresponding states have also been developed, ranging from correlations of chart values to sophisticated equation sets based on theoretical developments (Elsharkawy et al., 2001; Heidaryan et al., 2010). However, the use of EOSs to determine Z-factors has grown in popularity as computing capabilities have improved. EOSs represent not only the most complex method of calculating Z-factor, but also the most accurate.

### 1.6.1.3 Gas formation volume factor

The formation volume factor for gas is defined as the ratio of volume of 1 mol of gas at reservoir conditions to the volume of 1 mol of gas at standard conditions \( (P_s \text{ and } T_s) \). Using the real gas law and assuming that the Z-factor at standard conditions is 1, the equation for formation volume factor \( (B_g) \) can be written as:

\[
B_g = \frac{V_R}{V_s} = \frac{nZRT}{P} \frac{P_s}{nZsRT_s} = \frac{P_sZT_s}{T_sP}
\]  

(1-11)

When \( P_s \) is 1 atm (14.6959 psia or 101.325 kPa) and \( T_s \) is 60 °F (519.67 °R or 288.71 K), this equation can be written in three well-known standard forms:

\[
B_g = 0.0283 \frac{ZT}{P}
\]  

(1-11-1)

where \( B_g \) is in ft\(^3\)/SCF, \( P \) is in psia, and \( T \) is in °R. Alternately,

\[
B_g = 0.3507 \frac{ZT}{P}
\]  

(1-11-2)

where \( B_g \) is in m\(^3\)/Sm\(^3\), \( P \) is in kPa, and \( T \) is in °K.
In some cases, it is more convenient to define the value of $B_g$ in bbl/SCF. The equation can be written as:

$$B_g = 0.005 \frac{ZT}{P}$$  \hspace{1cm} (1-11-3)

where $T$ is in °R and $P$ is in psia.

### 1.6.1.4 Gas density

The gas density is defined as mass per unit volume and so can also be derived and calculated from the real gas law:

$$\rho_g = \frac{m}{V} = \frac{PM}{ZRT}$$  \hspace{1cm} (1-12)

Knowing that molecular weight of gas is the product of specific gravity and molecular weight of air, and the value of $R$ is 10.73 in field units (8.314 in SI units), we can write the equation for density as:

$$\rho_g = 2.7 \frac{P\gamma_g}{ZT}$$  \hspace{1cm} (1-12-1)

where $\rho_g$ is in lbm/ft$^3$, $P$ is in psia, and $T$ is in °R. Alternately,

$$\rho_g = 3.49 \frac{P\gamma_g}{ZT}$$  \hspace{1cm} (1-12-2)

where $\rho_g$ is in kg/m$^3$, $P$ is in kPa, and $T$ is in °K.

The density can also be written as:

$$\rho_g = 0.0764 \frac{\gamma_g}{B_g}$$  \hspace{1cm} (1-12-3)

where $\rho_g$ is in lbm/ft$^3$, and $B_g$ is in ft$^3$/SCF. Alternately,

$$\rho_g = 1.224 \frac{\gamma_g}{B_g}$$  \hspace{1cm} (1-12-4)

where $\rho_g$ is in kg/m$^3$, $B_g$ is in m$^3$/Sm$^3$.

### 1.6.1.5 Isothermal compressibility of gases

The isothermal gas compressibility, which is given by the symbol $c_g$, is a useful concept that will be used extensively in determining the compressible properties of the reservoir. The isothermal compressibility is also called the bulk modulus of elasticity. Gas usually is the most compressible medium in the reservoir. However care should be taken so that it is not confused with the gas deviation factor, $Z$, which is sometimes called the super-compressibility factor:

$$c_g = -\frac{1}{V_g} \left( \frac{\partial V_g}{\partial P} \right)_T$$  \hspace{1cm} (1-13)
where $V$ and $P$ are volume and pressure, respectively, and $T$ is the absolute temperature. For ideal gas, we can define the compressibility as:

$$c_g = \frac{1}{P} \quad (1-13-1)$$

Whereas, for nonideal gas, compressibility is defined as:

$$c_g = \frac{1}{P} - \frac{1}{Z} \left( \frac{\partial Z}{\partial P} \right)_T \quad (1-13-2)$$

If the relationship between the Z-factor and pressure at a given temperature is known, we can calculate the compressibility of gas. Since we already know the relationship between $Z$ and $P$, we can numerically calculate the derivative and, hence, the value of the compressibility.

### 1.6.1.6 Gas viscosity

A number of methods have been developed to calculate gas viscosity. The method of Lee et al. (1966) is a simple relation which gives quite accurate results for typical natural gas mixtures with low nonhydrocarbon content. The Lee et al. (1966) correlation was evaluated by Jeje and Mattar (2004) and has the form shown below:

$$\mu_g = 1.10^{-4} K \exp \left[ X \left( \frac{\rho_g}{62.4} \right)^Y \right] \quad (1-14)$$

where:

$$K = \frac{(9.4 + 0.02M_g)T^{1.5}}{209 + 19M_g + T} \quad (1-14-1)$$

$$X = 3.5 + \frac{986}{T} + 0.01M_g \quad (1-14-2)$$

$$Y = 2.4 - 0.2X \quad (1-14-3)$$

In this expression, temperature is given in (${}^\circ$R), the density of the gas in lbm/ft$^3$ (calculated at the pressure and temperature of the system), and the resulting viscosity is expressed in centipoises (c_p).

The viscosity of gas mixtures at 1 atm and reservoir temperature can be determined from the gas mixture composition:

$$\mu_{ga} = \frac{\sum_{i=1}^{N} y_i \mu_i \sqrt{M_{gi}}}{\sum_{i=1}^{N} y_i \sqrt{M_{gi}}} \quad (1-15)$$

where $\mu_{ga}$ is the viscosity of the gas mixture at the desired temperature and atmospheric pressure, $y_i$ is the mole fraction of the $i$th component, $\mu_{ga}$ is the viscosity of the $i$th component of the gas mixture at the desired temperature and
atmospheric pressure, \(M_{gi}\) is the molecular weight of the \(i\)th component of the gas mixture, and \(N\) is the number of components in the gas mixture.

This viscosity is then multiplied by the viscosity ratio to obtain the viscosity at reservoir temperature and pressure.

### 1.6.2 Thermodynamic properties

The principles of thermodynamics find very wide application in predicting the properties of hydrocarbons. The properties of greatest interest are specific heats of natural gas and its Joule–Thomson coefficient.

#### 1.6.2.1 Specific heat

Specific heat is defined as the amount of heat required raising the temperature of a unit mass of a substance through unity. It is an intensive property of a substance. It can be measured at constant pressure (\(C_p\)), or at constant volume (\(C_v\)), resulting in two distinct specific heat values. In terms of basic thermodynamics quantities:

\[
C_p = \left(\frac{\partial H}{\partial T}\right)_p \quad (1-16)
\]

\[
C_v = \left(\frac{\partial U}{\partial T}\right)_v \quad (1-17)
\]

where \(H\) is the molal enthalpy (BTU/lbmole), \(U\) is the molal internal energy (Btu/lbmole), \(C_p\) is the molal specific heat at constant pressure (BTU/lbmole-\(°\)R), and \(C_v\) is the molal specific heat at constant volume (Btu/lbmole-\(°\)R).

Both heat capacities are thermodynamically related. It can be proved that this relationship is controlled by the P-V-T behavior of the substances through the below relationship:

\[
C_p - C_v = T \left(\frac{\partial V}{\partial T}\right)_p \left(\frac{\partial P}{\partial T}\right)_v \quad (1-18)
\]

For ideal gases:

\[
C_p - C_v = R \quad (1-19)
\]

where \(R\) is the universal gas constant.

#### 1.6.2.2 Joule–Thomson coefficient

When a nonideal gas suddenly expands from a high pressure to a low pressure, there is often a temperature change. Note that this is far from a reversible effect. It is however an adiabatic effect due to the fact that the pressure change occurs too quickly for significant heat transfer to occur. Thermodynamically, the Joule–Thomson coefficient is defined as:

\[
\eta = \left(\frac{\partial T}{\partial P}\right)_H \quad (1-20)
\]
Using thermodynamic relationships, alternative expressions can be written. For example, using the cycling rule we may write

\[ \eta = \frac{RT^2}{PC_p} \left( \frac{\partial Z}{\partial T} \right)_p \]  

(1-21)

An interesting observation from the above expressions for the Joule–Thompson coefficient is that the Joule–Thompson coefficient of an ideal gas is identically equal to zero. However, real fluids take positive or negative Joule–Thompson values.

1.7 Natural gas reserves

As natural gas is essentially irreplaceable, it is important to have an idea of how much natural gas is left in the ground to use. Measuring natural gas in the ground involves a great deal of inference and estimation. There is no single way that every industry player uses to quantify estimates of natural gas. Therefore, it is important to delve into the assumptions and methodology behind each study to gain a complete understanding of the estimate itself. With new technologies, these estimates are becoming more and more reliable; however, they are still subject to revision. The Energy Information Administration (EIA) estimates world proved\(^8\) natural gas reserves to be around 5210.8 Tcf. The biggest reserves are located in the Middle East with 1836.2 Tcf (or 34% of the world total), and Europe and the Former USSR with 2158.7 (or 42% of total world reserves), but very significant reserves exist also in other regions. Note should be made that the most recent data on the world proved natural gas reserves can be found by visiting the EIA Web site (www.eia.doe.gov).

Natural gas reserves have grown rapidly in recent years. However, the share of natural gas reserves located onshore, easy to produce, and close to consumers is decreasing and the share of reserves located offshore and in hostile environments appears to be increasing. The availability of these reserves for the end-user is therefore hampered by production and transportation costs which can exceed the price at which the gas can be sold. In such cases, innovative technical options are required for reducing the costs and providing new outlets for natural gas.

1.8 Natural gas exploration and production

There are four major activities involved when developing a natural gas resource. They are exploration, drilling, completion, and production. These activities vary, depending on whether the natural gas is “conventional” or “unconventional,” and accordingly, the discussion will follow these two broad categories. In the

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\(^{8}\)Proved reserves are those estimated quantities of natural gas that can be commercially recoverable from a given date forward, from known reservoirs and under current economic conditions.
discussion of production, the emphasis will be on the flow characteristics within
the reservoir. However, as the gas needs to travel from the bottom of the well to
the wellhead, the wellbore’s delivery capacity will also be reviewed.

1.8.1 Conventional gas

1.8.1.1 Exploration

Gas is contained in porous rocks at different depths in different locations. So, the
first step is to locate where the gas is likely be found. This involves the two
disciplines of geophysics and geology.

A seismic survey (geophysics) is conducted to determine the layering of the
different rock strata. This helps determine where an accumulation of gas (or oil)
may occur, and the potential extent of this accumulation. A seismic survey
consists of a wave generator (an explosive device or a vibrator) and a series of
geophones, all located at the surface. The geophones detect reflections of the
wave from the various layers. The geophysicist interprets the seismic data to
generate a 2-D or 3-D picture of the layers in the Earth (see Figure 1-3).

The seismic survey only defines the structure of the rock layers. It does not
identify what is contained within the rock. However, it indicates where the highs
and lows of the structure are, or if there are any faults. Because of buoyancy
effects, the hydrocarbons are more likely to be found in the highs of the structure.
In combination with the geoscientist, who brings regional experience and
knowledge of the hydrocarbon-bearing potential of these layers, a potential
drilling location is determined. The chances of success vary significantly,
depending on the degree of oil and gas development in the region. Where there is
little knowledge (wildcat exploration) the chances of a successful well are
15–30%, but where there has already been a lot of successful drilling (develop-
ment wells), the chance of success increases to 75–100%.

![Seismic Schematic](image1)

**FIGURE 1-3**

Seismic survey used in exploration.
1.8.1.2 Drilling

The mechanics of drilling are complex, and are significantly more so for an offshore well than for one on land. The well depth can vary from 1000 to 20,000 ft. The well may encounter several layers of gas-, oil-, or water-bearing rocks, but usually, there is one particular layer that is the primary target of interest. While drilling, the wellbore is filled with drilling mud. The hydrostatic pressure of the mud counterbalances the reservoir’s pressure, and avoids blowouts during drilling.

Once the intended depth is reached, a suite of electric logs is run from the bottom to the top of the well, to determine the porosity, shale content, fluid saturations, and thickness of the rock layers (Figure 1-4). If a potential gas-bearing reservoir is found, flow tests can be performed to estimate the permeability of the reservoir. If no productive reservoirs are found, the well is abandoned. If one or more productive layers are found, a steel pipe (casing) is

![Electric logs used to determine the presence of hydrocarbons.](image)
placed in the wellbore and cemented. A “wellhead,” which is an assembly of control valves, is placed on top of the well to control the flow from the well.

### 1.8.1.3 Completion

The casing is perforated where the gas reservoir is known to exist. The perforations penetrate both the steel casing and the surrounding cement sheath, thus creating a flow path, allowing the gas to flow from the reservoir into the well (Figure 1-5). When the reservoir rock has a low permeability, the deliverability of the well can be improved significantly by acidizing or hydraulic fracturing. Acidizing increases the permeability in the vicinity of the well. Fracturing increases the contact area between the well and the reservoir. To fracture a well, a liquid (usually gelled water with a proppant) is injected at high rates and pressures until the rock cracks. After the completion treatment, flow and shut-in tests are conducted to determine the deliverability potential of the well and to estimate the permeability of the reservoir and the effectiveness of the completion.

### 1.8.1.4 Production

The rate of production of a gas well depends, principally, on two fundamental equations, namely Darcy’s law and the material balance equation (Dake, 1978). Darcy’s law relates the flow rate in a reservoir to the driving force (pressure difference) and the resistance (permeability). Material balance relates the decrease in the average reservoir pressure to the size of the reservoir and production volumes (hence to time). The principal factors that determine the deliverability of a well are, therefore, reservoir pressure, flowing pressure, permeability, completion effectiveness, original gas-in-place, and wellbore configuration.

For conventional gas wells, the deliverability equation can, for all practical purposes, be expressed as (ERCB, 1975):

\[
q = \frac{C}{C_0} \left( \frac{P_R^2 - P_{wf}^2}{P_{wf}^2} \right)^n
\]  

(1-22)
where \( q \) is the flow rate; \( C \) is a constant that encompasses permeability and completion effectiveness and other gas properties such as gravity, viscosity, temperature; \( P_R \) is the reservoir pressure; \( P_{wf} \) is the pressure at the well; and \( n \) is an exponent (between 0.5 and 1) that accounts for Darcy and non-Darcy flow in the reservoir (commonly referred to as laminar and turbulent flow).

The constants \( C \) and \( n \) are determined from a flow test (often called absolute open flow). The \( P_{wf} \) depends on surface facility restrictions, and usually reflects the pressure at which the gas is to be delivered to the transportation system or to the gas plant.

The material balance equation relates the reservoir pressure to the amount of gas produced:

\[
\frac{P_R}{Z} = \frac{P_i}{Z_i} \left( 1 - \frac{G_p}{G} \right)
\]

where \( Z \) is the gas deviation factor (compressibility factor), \( G_p \) is the gas produced, and \( G \) is the original gas-in-place. Subscript “R” represents current reservoir pressure, and subscript “i” represents initial reservoir pressure.

The above two equations appear to be independent of time. However, that is not so. In Eqn (1-23), \( G_p \) changes with time. Hence, the reservoir pressure changes with time. This, in turn, causes the flow rate calculated in Eqn (1-22) to decrease as the reservoir depletes.

Combining Eqns (1-22) and (1-23), and superimposing operating restrictions such as contract rates, facilities limitations (compression/pipeline/processing) result in a forecast of production rate versus time. Production stops when the flow rate becomes uneconomic. The total gas that will have been produced at the time of abandonment is called the expected ultimate recovery (EUR).

The original gas-in-place (\( G \)) is determined from geological estimates of the areal extent and thickness of the reservoir, as well as its porosity and saturation, or from analysis of production data.

The recovery factor is the ratio of the EUR to the original gas-in-place (EUR/G). It depends primarily on the reservoir pressure at the time of abandonment. The abandonment pressure itself is directly related to the delivery pressure and to the distance of transmission. As a consequence, the recovery factor can range from 50% (for remote locations) to 90% for wells adjacent to fully developed transportation systems.

1.8.2 Unconventional gas

1.8.2.1 Exploration

Unlike conventional gas, the location of unconventional resources is generally known, often as an indirect result of conventional mining or oil field activity. These unconventional gas resources are generally laterally extensive, and therefore, less exploration is required to locate them.
1.8.2.2 Drilling

Whereas most conventional gas wells are vertical, unconventional gas wells are either horizontal wells or multiple-well pads (Figure 1-6). There are two principal reasons for this:

1. The productivity of the vertical well is too low which is improved by using a horizontal well (single or multilateral).
2. The area of the reservoir that is drained by a vertical well is so small that it would require a large number of wells to effectively drain the whole reservoir. This could be very intrusive on the surface land activity and is resolved by drilling multiple directional wells from a single pad, which significantly minimizes the surface footprint.

1.8.2.3 Completion

Typically, the permeability of unconventional gas reservoirs is extremely small. As a consequence, with conventional completions, the deliverability per well can be uneconomic, in spite of the known presence of large quantities of gas. The best way to improve productivity per well is to increase the area of contact between the well and the reservoir. This is what underlies the success of unconventional gas completions. The single most significant contribution to the commercialization of “shale” gas and “tight” gas has been the multistage hydraulic fracturing of horizontal wells to create this increased contact area (Figure 1-7).

1.8.2.4 Production

The deliverability of gas from unconventional gas wells is governed by a combination of Darcy’s law and material balance (similar to conventional reservoirs). However the Eqns (1-22) and (1-23) are not very useful, because they are only applicable after the flow from the reservoir has stabilized. When a

![Figure 1-6](image.png)

**Different drilling patterns.**
well is first opened to flow, a pressure transient travels through the reservoir (transient flow). It is only when this pressure transient reaches the boundaries of the reservoir (boundary-dominated flow) that the flow is considered to be stabilized and Eqns (1-22) and (1-23) become applicable. The time to reach stabilization is a function of permeability, and in conventional reservoirs, this time is, for all practical purposes, relatively short (<3 months). However, in unconventional reservoirs, the permeability is so low that it can take many years before Eqn (1-22) becomes applicable. Accordingly, more complex equations applicable during transient flow are needed. Such equations will usually incorporate time explicitly, and typically require computer programs. The form of such an equation is: 

\[ q = f(p_i, p_{wf}, \text{permeability, well type, completion type, fluid properties, reservoir geometry, and time}) \]

Unconventional gas reservoirs can be extensive, but because of the very low permeability, the effective drainage area per well is very limited. Accordingly, the original gas-in-place is often an “assigned” value, directly linked to the well spacing.

The recovery factor is the ratio of the EUR to the original gas-in-place (EUR/G). As a consequence, the reported recovery factor depends on the spacing of the wells and the assigned drainage area. Accordingly, it can range from 20% to 70%.

1.8.3 Well deliverability

The flow of gas from the reservoir into the wellbore, described above, is often called the inflow performance relationship (IPR). It is usually represented as a graph of flowing bottom-hole pressure versus gas rate (see Figure 1-8(a)). This graph demonstrates that the flow rate depends on the flowing pressure.

From a production perspective, the flowing pressure is controlled at the surface. The bottom-hole (sand-face) flowing pressure that corresponds to the wellhead flowing pressure depends on the flow rate, the pressure level, the gas composition, the pipe diameter, and the depth of the well. For a given wellbore
configuration and a specified wellhead pressure, the flowing bottom-hole pressure can be calculated at various arbitrary flow rates. This is shown in Figure 1-8(b), and is called a “tubing performance curve” (TPC).

Since the reservoir and the wellbore must interact, the “operating point” is obtained by overlaying the TPC and the IPR curves (see Figure 1-8(c)). The point of intersection is the production rate from the well at the specified conditions. Several TPCs can be generated by varying the specifications of the wellhead conditions or the tubulars. These are overlain on the reservoir’s IPR and different operating points will result. The choice of operating conditions is based on the economic factors associated with each TPC.

Calculation of the TPC is based on the friction and hydrostatic pressure drop calculations outlined in Chapter 2. In general, for single-phase flow of gas, the bigger the tubing diameter the larger the operating point flow rate, and the choice of tubing diameter is straightforward and is directly linked to the cost of tubing. However, many gas wells produce gas and liquids (water or condensate). Multiphase calculations are complex and often show that the operating point rate can be increased by “reducing” the tubing diameter.

When a gas well produces liquids, care must be taken to efficiently remove all the liquids; otherwise, they accumulate in the wellbore, and eventually the increasing hydrostatic back pressure “kills” the well. Various mechanisms exist for removing the liquids. These range from “siphon strings” (small-diameter tubing) to “plunger lift,” to reducing the wellhead pressure, and to bottom-hole pumps. Many gas wells stop producing (die), even though there is still a lot of gas left in the reservoir. They die simply because it is not economically viable to remove the liquids from the wellbore.

The production rate of a gas well decreases with time, because the reservoir pressure depletes. Eventually, the flow rate becomes uneconomic and the well is abandoned. Typically, to maintain a gas supply contract, additional wells are drilled over time to supplement the decreasing deliverability of the wells.
1.9 Natural gas transportation

Transportation is an essential aspect of the gas business, since gas reserves are often quite distant from their main markets. For almost a century, natural gas has been transported safely, reliably, and economically via pipelines which are used to bring the gas supply in various production wells to the metropolis. Pipeline system has been providing the stability and long-term security by balancing the supply and demand markets. Now, in the twenty-first century, the vast majority of the large, easy to produce, and pipeline gas reserves plays have been already tapped, and attention is shifting to stranded reservoirs that were previously thought too small, too remote, or technically too difficult to develop. The liquefied natural gas (LNG) industry has commercialized many large remote gas fields over the past three decades and developed gas markets commercially unreachable by pipeline.

Over the last two decades, several technologies have been also evaluated and proposed for monetizing hitherto remote gas reserves (see Figure 1-9). These include: a number of technologies converting natural gas into a range of easily transported and marketed hydrocarbons (grouped under the generic term “gas-to-liquids” (GTL) technologies); using gas to produce electricity at the producing field and then transporting it by high-voltage direct current (HVDC) transmission lines over long distances (generically referred to as “gas-to-wire” (GTW) technologies); compressed natural gas (CNG) that avoids the cost of gas liquefaction; converting gas, particularly associated gas, into solids or slurries formed of gas hydrates for storage and transportation (generically referred to as “gas-to-solids”)

![Figure 1-9: Technologies available to transport natural gas over long distances](Wood and Mokhatab, 2008).

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Alternative Methods of Moving Gas to Market

- Fully Commercial
- Pipeline
- LNG
- GTL
- Gas to Wire - HVDC
- CNG
- Gas to Solids - Hydrates
- R&D - No Commercial Projects Yet Sanctioned

David Wood & Associates

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Technologies available to transport natural gas over long distances (Wood and Mokhatab, 2008).
(GTS) technologies). The latter two technologies are still in the research and development stage and, although much work has been done to verify the potential of these options in the past decade, no commercial projects exploiting them have yet been sanctioned (Mokhatab and Wood, 2007).

The following section examines some of these technical methods by which natural gas energy can be transported, and covers many of the essential points needed to enter the discussion.

1.9.1 Pipelines

Pipeline transportation has been employed to deliver gas to markets for a long time. However, supply of natural gas to markets via gas pipelines is often faced with technical, economic, and political uncertainties.

Pipelines are a very convenient method of transport but are not flexible. If the pipeline has to be shut down, the production and receiving facilities and the processing plant often also have to be shut down. The downstream facilities can continue to operate for only a short time by depleting the inventory in the gas lines.

For onshore and near-shore gas, pipeline is an economical option for transporting natural gas to the consumers. However, as transportation distance increases, particularly in offshore and deepwater exploration, development of these projects is challenging and expensive, and requires large-diameter long-distance pipe network. It requires both large, high-value markets and substantial proven reserves to be economically viable. On the other hand, gas-processing technology has advanced significantly in the past two decades, overcoming challenges of upstream gas conditioning, hydrate control, and subsea pipeline design that makes the gas transportation economical. Raw (unprocessed) gas transmission by means of pipelines is discussed in great details in Chapter 2.

Intercontinental pipelines usually entail the crossing of a number of countries and borders, which means that several governments may have jurisdiction over the pipelines, increasing the operational complexity of project. This may also increase the probability of supply interruptions through diversion of gas volumes, terrorist attack on infrastructure, or pipeline shutdown during political turmoil.

1.9.2 Liquefied natural gas

LNG technology has been proved to be effective over the last 30 years. When natural gas is cooled to approximately $-260 \, ^\circ\text{F}$ ($-162 \, ^\circ\text{C}$) at atmospheric pressure, it results in a condensed liquid form called LNG. The volume reduction is about 1/600th the volume of natural gas at the burner tip. The physical properties of LNG allow for its long-distance transport by ship across oceans to markets and for its local distribution by truck onshore. Occasionally, liquefaction of natural gas also provides the opportunity to store the fuel for use during high-consumption periods close to demand centers, as well as in areas where geologic conditions are not suitable for developing underground storage facilities.
The much lower volume of LNG relative to gaseous natural gas can reduce transportation costs by allowing delivery using cargo ships or transport trucks instead of pipelines. However, the costs of building an LNG plant are typically high relative to comparable energy projects for a number of reasons, including remote locations, strict design and safety standards, large amounts of cryogenic material required, and a historic tendency to overdesign to ensure supply security. Costs throughout the value chain have been declining in the LNG industry in recent years. Further advances in LNG technology can be expected in liquefaction and shipping, which could lead to lower overall project costs (Cornot-Gandolphe et al., 2003; Mokhatab et al., 2014). In the longer term, floating LNG (FLNG) concept, where processing and storage facilities are based on a vessel moored offshore in the vicinity of the producing fields, could reduce costs even more and make the development of some small and remote gas reserves or deep offshore gas feasible. This technology can reduce costs by minimizing the cost of offshore platforms and pipelines, eliminating the need for port facilities and reducing the time needed to build the plant. Vessel construction can be also carried out in a low-cost location. However, there are potentially many commercial and technical challenges that need to be addressed during the development of FLNG projects. The key for the industry now is to delineate these challenges, and discuss means to resolve these problems (Chiu and Quillen, 2006; Mokhatab et al., 2014).

1.9.3 Compressed natural gas

Gas can be transported in containers at high pressures, typically 1800 psig for a rich gas (significant amounts of ethane, propane, etc.) to roughly 3600 psig for a lean gas (mainly methane). Gas at these pressures is termed CNG. CNG offers proven technology that has the potential to provide an early-to-market, economic solution for remote offshore gas developments with small to medium reserves or for associated gas reserves in large oil field developments. They could work where subsea pipelines are not viable because of distance, ocean topography, limited reserves, modest demand, or environmental factors, but where LNG is also not economical due to its high cost of liquefaction and regasification facilities, or feasible due to community or safety issues. In addition to being a cost-effective solution for regional gas projects, CNG transport projects also offer several unique and valuable solutions in terms of flexibility and risk mitigation compared to the LNG projects (i.e., easy supply and market access, redeploy-ability, scalability).

9Safety and security are contentious issues for CNG, but by locating the offloading facilities tens of kilometers offshore it is possible to remove many of the community-related safety and environmental issues that have dogged the permitting for LNG receiving terminals located onshore United States in populous areas in recent years.
Energy consumed in operating a CNG project is about 40% of that of an LNG project. The threshold volume of gas reserves is relatively low for commercial viability, provided shipping costs can be kept low and work effectively. Greater than 85% of a CNG project cost is likely to be associated with the ships, which are based upon conventional bulk carriers with at least four competing certified containment designs (Hatt, 2003): EnerSea (US) VOTRANS carbon steel pressure cells; TransOceanGas (Canada) fiber-reinforced plastic (FRP) covering high-density polyethylene cells; TransCanada CPV steel liner cell overwrapped with a glass fiber laminate; Sea NG (Canada) patented Coselle of coiled X70 high-strength steel pipe wound into a cylindrical storage container. EnerSea’s VOTRANS™ (Volume Optimized Transport and Storage) containment system is the most cost-effective CNG solution in the marketplace due to greater gas storage/delivery efficiency. Certainly with most of the capital costs of the technology in the vessels, it is important to have a large and experienced shipping company such as Teekay at the helm (Wood and Mokhatab, 2008).

1.9.4 Gas-to-liquids

In GTL transportation processes, the natural gas is converted to a liquid and transported as such. The technology of converting natural gas to liquids is not new. In the first step, methane is mixed with steam and converted to syngas or synthetic gas (mixtures of carbon monoxide and hydrogen) by one of a number of routes using suitable new catalyst technology (Keshav and Basu, 2007). The syngas is then converted to a liquid using a Fischer–Tropsch (FT) process (in the presence of a catalyst) or an oxygenation method (mixing syngas with oxygen in the presence of a suitable catalyst). The produced liquid can be a fuel, usually a clean burning motor fuel (syncrude) or lubricant, or ammonia, methanol, or some precursor for plastics manufacture, e.g., urea, dimethyl ether (DME), which is also used as a transportation fuel, LPG substitute, or power generation fuel as well as a chemical feedstock (Knott, 1997; Skrebowski, 1998; Thomas, 1998; Apanel, 2005). The environmental benefits of GTL products have been well demonstrated and have much potential in improving air quality in cities polluted with transportation fuel emission. However, on a full-cycle environmental analysis, GTL fuels do not significantly outperform refinery fuels because the GTL plants and their components involve substantial emissions. The problem is primarily with the low energy efficiency of syngas generation and the low carbon efficiencies of the conversion processes such as FT (O’Rear and Goede, 2007). Technology breakthroughs are required to improve these efficiencies, which will lead to both cost and environmental benefits.

The FT-GTL method is an application of the basic FT process, where synthesis gas (or syngas) is reacted in the presence of an iron or cobalt catalyst. End products are determined by the length of the hydrocarbon chain which, in turn, is determined by catalyst selectivity and reaction conditions. FT-GTL is essentially a three-stage process (syngas generation, FT transformation, and product
upgradation). Possible end products include kerosine, naphtha, methanol, DME, alcohols, waxes, synthetic diesel, and gasoline, with water or carbon dioxide produced as a by-product. FT-GTL has reached a key milestone as an industry in that worldscale capacity plants are in the process of being planned, constructed, and commissioned. However, FT-GTL has a lot left to prove about its technical, economic, and environmental efficiency before it can be considered as a rival to LNG for developing and monetizing stranded gas reserves on a large scale (Fleisch et al., 2003; Al-Saadoon, 2005; Apanel, 2005). Hundreds of modifications and patents have been applied to this complex, energy-intensive process, and further developments continue. Most recent technical advances are focused on lowering capital expenditures, establishing economies of scale, and improving operating and energy efficiencies for large-scale FT-GTL plants. Efforts to reduce the footprint, energy consumption, and efficient and safe use of oxygen in syngas reforming plants have also received attention in recent years. Much research is still ongoing with respect to other GTL processes, but none has yet reached large-scale commercialization, although further developments, particularly in methanol to gasoline plants, should be expected in this high-price gasoline market (Wood and Mokhatab, 2008; Wood et al., 2008).

GTL technology development has reached a stage where its marinization, enabling smaller footprints and more compact plants than other technologies, may be considered for Floating Production, Storage and Offloading (FPSO) application. Floating GTL will provide new opportunities for companies to produce, transport, and market gas reserves that would otherwise remain stranded. Also, floating GTL plants could be moved from location to location, enabling access to small fields that would otherwise not justify building a dedicated GTL facility. GTL FPSO concept, which is still in the development phase, could greatly improve potential project economics (Loenhout et al., 2006).

### 1.9.5 Gas-to-solid

Gas can be transported as a solid, with the solid being gas hydrate (Børrehaug and Gudmundsson, 1996). Natural gas hydrates (NGHs), which are essentially natural gas in a “frozen” state, form when water and natural gas combine at low temperatures and high pressures. Gas hydrates are clathrates, where the guest gas molecules are occluded in a lattice of host water molecules. They are most commonly encountered in the industry as a production problem in pipelines to be avoided as part of flow assurance.

GTS involves three stages: production, transportation, and regasification. Conceptually, hydrate slurry production is simply mixing chilled water and gas. In practice, processed gas is fed to a hydrate production plant, where a series of reactors convert it into hydrate slurry. Each reactor further concentrates the hydrate slurry. It is then stored and eventually off-loaded onto a transport vessel (insulated to near-adiabatic conditions). At the receiving terminal, the hydrate is dissociated and the gas can be used as desired. The
water can be used at the destination if there is water shortage, or returned as ballast to the hydrate generator and, since it is saturated with gas, will not take more gas into solution.

Gas storage in hydrate form becomes especially efficient at relatively low pressures where substantially more gas per unit volume is contained in the hydrate than in the free state or in the compressed state when the pressure has dropped. When compared to the transportation of natural gas by pipeline or as LNG, the hydrate concept has lower capital and operating costs for the movement of quantities of natural gas over adverse conditions. NGHs are chemically stable at about \(-20\,^\circ C\) compared with \(-162\,^\circ C\) for LNG, reducing the costs of transportation and storage. One cubic meter of NGH contains approximately 160 m\(^3\) of natural gas, while one cubic meter of LNG contains 600 m\(^3\) of natural gas, thus limiting the quantity of gas that can be transported with the NGH technology. This is a considerable volume penalty (and hence transport cost) if considered in isolation, with the cheaper ships for hydrate transport the process could be economic.

GTS options for transporting and storing stranded gas to market have been extensively researched and laboratory tested for more than a decade. BG Group, Marathon, NTNU (Norway) and others have worked on a range of GTS transportation technologies (Fischer, 2001) and have taken them to small-scale pilot plants. Concepts include storage and transport of gas either as atmospheric hydrates or as a paste in pressurized insulated containers or as frozen hydrates are mixed with refrigerated crude oil for atmospheric pressure transport. Even though these studies have proved the concept of storing natural gas in hydrates technically feasible, applications have not progressed beyond the laboratory stage because of complexities of the process, slow hydrate formation rates, and costs. No projects are close to commercialization, but technological advances continue to be made, e.g., forming the hydrates from a surfactant solution (Rogers et al., 2005), that may make this possible in the long term.

1.9.6 Gas-to-wire

Currently, much of the transported gas is used as a fuel for electricity generation. Electricity generation at or near the reservoir source and transportation by cable to the destination(s) (GTW) is possible. Thus, for instance, offshore or isolated gas could be used to fuel an offshore power plant (may be sited in less hostile waters), which would generate electricity for sale onshore or to other offshore customers.

HVDC transmission lines offer the most technically viable solution to moving large quantities of electric power over large distances (up to about 1500 km) keeping line losses less than 10%. However, HVDC is capital intensive and requires costly converter stations at either end of the transmission line. Additional costs for installing and then operating and maintaining gas turbines at the remote site would be incurred.
While HVDC cables are now being used more widely to transport electricity tens of kilometers, no projects to develop remote gas fields in this way have yet been sanctioned. Distance, cost, and efficiency of remote generation make other options currently more attractive. There are other practical considerations to note such as if the gas is associated gas, then if there is a generator shutdown and no other gas outlet, the whole oil production facility might also have to be shut down, or the gas released to flare. Also, if there are operational problems within the generation plant, the generators must be able to shut down quickly (in around 60 s) to keep a small incident from escalating. Additionally, the shutdown system itself must be safe so that any plant that has complicated processes that requires a purge cycle or a cool-down cycle before it can shut down is clearly unsuitable (Ballard, 1965). Finally, if the plant cannot shut down easily and/or be able to start up again quickly (perhaps in an hour), operators will be hesitant to ever shut down the process, for fear of financial retribution from the power distributors.

1.9.7 Comparison of various methods

As discussed above, there are a number of options of exporting natural gas energy from oil and gas fields to market. However, distance to market and potential production volumes (dependent on the reserve size of the gas resource) influence the technologies that might viably be used to exploit remote gas fields (see Figure 1-10).

Any gas energy export route requires a huge investment in infrastructure, and long-term “fail-proof” contracts, covering perhaps 20 years or more. However,
improvements in technology, economies of scale, and synergies will doubtless lower capital costs and further improve the project economics over the next few years. Most of these technologies have reached a stage where commercialization on a wide scale is only a few years away. However, their use carries risks in terms of technology, credit worthiness, revenue security, and market competition risks, each of which needs to be appropriately mitigated.

Gas is currently transported to markets primarily via two long-established methods (pipeline and LNG technology). Alternative technologies, which have been refined and developed in recent years, are yet to make serious inroads into the challenges of developing remote gas fields. Figure 1-11 shows that as the distance over which natural gas must be transported increases, usage of LNG has economic advantages over usage of pipelines. As can be seen, for short distances, pipelines—where feasible—are usually more economical. LNG is more competitive for long-distance routes particularly those crossing oceans or long stretches of water since overall costs are less affected (Economides and Mokhatab, 2007).

LNG, an effective long-distance delivery method, constitutes only 25% of global gas movement. However, LNG can offer flexibility, diversity, and security of supply advantages over pipeline alternatives. LNG projects require large investments along with substantial natural gas reserves and are economically viable for transporting natural gas in offshore pipelines for distances of more than 700 miles or in onshore pipelines for distances greater than 2200 miles (Mokhatab and Purewal, 2006). However, further improvements in

![Gas Transportation Costs](image)

**FIGURE 1-11**
Comparison of the cost of transporting gas via pipeline and LNG; for 1 tcf/year and including regasification costs (Economides and Mokhatab, 2007).
LNG technology can be expected, which could lead to lower overall project costs.

Note, CNG is an economic alternative to monetize stranded gas reserves, and creates new market where pipelines and LNG deliveries are not practical. CNG technology provides an effective way for smaller-volumes and shorter-distance transport of gas. The results show that for distances up to 2200 miles, natural gas can be transported as CNG at prices ranging from $0.93 to $2.23 per MMBTU compared to LNG, which can cost anywhere from $1.5 to $2.5 per MMBTU depending on the actual distance. At distances above 2200 miles, the cost of delivering gas as CNG becomes higher than the cost for LNG because of the disparity in the volumes of gas transported with the two technologies (Economides et al., 2005). Also, unit gas transportation comparisons for pipeline, LNG, and CNG (see Figure 1-12) suggest that pipelines will match the CNG costs at transportation volumes of about 750 MMcf/d or greater over intermediate distances. At lower volumes below this, CNG will offer lower unit transportation costs. LNG will match CNG unit transportation costs at larger volumes or longer distances. Therefore, CNG technology may have the potential to challenge LNG transportation for some niche markets. CNG can start with much smaller initial volumes than either a pipeline or LNG, and can grow incrementally to meet demand by simply adding ships. It should be noted that, while pipeline and LNG are proven either as concepts or technologies, CNG is still awaiting its first commercial application. Nevertheless, reflecting on the system’s simplicity, the level of engineering efforts invested and developed reliable cost estimates, it is believed that CNG technology is ready for commercialization (Economides and Mokhatab, 2007).

Further to CNG technology, there are numerous technology alternatives to the currently commercial pipeline and LNG options for transporting natural gas to

![Figure 1-12](image)

**FIGURE 1-12**
Transportation tariffs (Economides et al., 2005).
markets on a wide range of scales, which have been described before. Unfortu-
nately, few of them are yet receiving the levels of investment and commitment 
required to bring them into widespread development. High prevailing oil and gas 
prices around the world could change this lack of enthusiasm for embracing novel 
technologies.

The transport options preferred by governments and companies must not only 
take the economic risks into account but also consider the negative effects of 
possible terrorist activity, political changes, and trade embargos over long periods 
of time. Thomas and Dawe (2003) cover many of the essential technical points 
and broad economic pointers needed to enter the discussion of gas-rich states 
which do not need the gas for domestic use, but wish to monetize their reserves 
by export.

1.10 Natural gas processing

As discussed in Section 1-4, natural gas as it leaves the reservoir contains water 
vapor and all the impurities. Therefore, raw natural gas produced from wells must 
be processed and treated to meet the quality standards specified by pipeline 
companies, which must be compatible with the pipeline design and the cus-
tomer’s requirements. Off-spec natural gas will cause operational problems 
associated with corrosion and plugging, which may result in unsafe operation.

If the raw gas contamination levels and acid gas content are low, the gas can 
be treated and dried at the wellhead and sent directly to the sales gas pipeline. 
However, if the contaminant levels are high and the gas is sour, then it is typically 
collected and piped to a gas-processing complex where the gas is treated, 
conditioned, and dried to meet pipeline specifications. The design and operation 
of the different processing units are discussed in following chapters.

1.11 Sales gas transmission

After the raw gas has been treated and conditioned to meet pipeline specifica-
tions, the treated (sales) gas is transported via a pipeline network to the industrial 
and residential consumers. A number of compressor stations (containing one or 
more compressor units) are located along the pipeline system to maintain the 
pipeline pressure and overcome the pressure drop in the pipeline. Sales gas 
transmission will be discussed in detail in Chapter 12.

1.12 Underground gas storage

Strategically, underground gas storage provides security of supply in case there 
are disruptions to production and transmission. This could be due to commercial 
reasons, such as sales gas price negotiation, political reasons, or an outage. It is a 
means to balance seasonal variations in consumption; the winter demand is
typically greater than the summer usage. It can also enhance the effectiveness of gas transportation and production, where it can be stored locally to where it is being used. Commercially, natural gas is a commodity for trading.

There are three principal types of underground storage methods: depleted oil or gas reservoirs, aquifers, and salt cavern formations.

1.12.1 Depleted reservoirs

The most common form of underground storage is depleted gas reservoirs. They are formations that have already been exhausted of most of their recoverable oil and natural gas. The empty reservoirs can be used to hold natural gas. Typically, an extensive pipeline network is located close to the reservoir sites for the convenience of the injection and withdrawal operation. Of the three types of underground storage, depleted reservoirs are the cheapest and the quickest to develop, operate, and maintain.

Two of the most important characteristics of an underground storage reservoir are the capability to hold natural gas and the rates at which natural gas inventory can be injected and withdrawn. Note that most of these depleted oil fields had been injected with nitrogen for enhanced oil recovery, and consequently, the nitrogen content in these fields would vary during withdrawal. In some reservoirs, the nitrogen content in the reservoir during the withdrawal mode can vary from 3 mol% to 30%. In these facilities, a nitrogen rejection unit must be installed to remove the nitrogen content to meet pipeline specification.

1.12.2 Aquifers

In the United States, 16% of storage capacity is in aquifers. Aquifers are underground porous, permeable rock formations that act as natural water reservoirs. An aquifer is suitable for natural gas storage if the water-bearing sedimentary rock formation is overlaid with an impermeable cap rock. While the geology of aquifers is similar to the depleted production fields, they are more expensive to develop for natural gas storage, and consequently, there are limited numbers of aquifer natural gas storage, found only in areas where there are no other alternatives.

1.12.3 Salt caverns

Salt caverns are formed out of existing salt bed deposits. Most of the large salt caverns are located in the salt domes along the Gulf Coast in the United States. Salt caverns in Northeastern, Midwestern, and Western states are also available but the applications are limited by the lack of suitable geology.

The cavern is artificial by drilling a well down into the formation, and pumping water through the completed well to dissolve the salt which returns to the surface as brine. The walls of the cavern are every resilient against reservoir degradation.
As the salt cavern is an open vessel, it offers very high deliverability. Flow rates can be high and they can be brought on stream and ramped to full flow quickly. They are best for peak loads and short-term trading rather than long-term seasonal storage. Peak load can be provided by salt caverns, where the deliverability is higher, turnovers will be higher and facilities are smaller. Salt caverns turnover can be daily or weekly, entirely dictated by commercial trading.

References
Ewan, D.N., Lawrence, J.B., Rambo, C.L., Tonne, R.R., March 10–12, 1975. Why cryogenic processing (investigating the feasibility of a cryogenic turbo-expander plant). Paper Presented at the 54th GPA Annual Convention, Houston, TX, USA.


Wood, D., Mokhatab, S., Economides, M.J., March 2–5, 2008. Technology options for securing markets for remote gas. Paper Presented at the 87th GPA Annual Convention, Grapevine, TX, USA.
2.1 Introduction

Natural gas is often found in places where there is no local market, such as in the many offshore fields around the world. For natural gas to be available to the market, it must be gathered, processed, and transported. Quite often, collected natural gas (raw gas) must be transported over a substantial distance in pipelines of different sizes, due to the drive for reduced field processing facilities particularly for offshore fields. These pipelines vary in length between hundreds of feet to hundreds of miles, across undulating terrain with varying temperature conditions. Liquid condensation in pipelines commonly occurs because of the multicomponent nature of the transmitted natural gas and its associated phase behavior to the inevitable temperature and pressure changes that occur along the pipeline. Condensation subjects the raw gas transmission pipeline to multiphase, gas-condensate-water, flow transport.

Multiphase transportation technology has become increasingly important for developing marginal fields, where the trend is to economically transport unprocessed well fluids via existing infrastructures, maximizing the rate of return and minimizing both capital expenditure and operational expenditure. In fact, by transporting multiphase well fluid in a single pipeline, separate pipelines and receiving facilities for separate phases, costing both money and space, are eliminated that reduces capital expenditure. On the other hand, phase separation and reinjection of water and gas save both capital expenditure and operating expenditure, by reducing the size of the fluid transport/handling facilities and the maintenance required for the pipeline operation. Given the savings that can be available to the operators using multiphase technology, the market for multiphase flow transportation is an expanding one. Hence, it is necessary to predict multiphase flow behavior and other design variables of gas-condensate pipelines as accurately as possible so that pipelines and downstream processing plants may be designed optimally. This chapter covers all the important concepts of multiphase gas-condensate transmission from a fundamental perspective.

2.2 Multiphase flow terminology

This section defines the variables commonly used to describe multiphase flow. Definitions for these variables are described based on Figure 2-1, which is the ideal flow of three fluids. It is assumed that the water is heavier than the oil and flows at the bottom, while the oil flows in the middle and the gas is the top layer.
2.2.1 Superficial velocity

The superficial velocity is the velocity of one phase of a multiphase flow, assuming that the phase occupies the whole cross-section of pipe by itself. It is defined for each phase as follows:

\[ V_{SW} = \frac{Q_W}{A} \]  \\
\[ V_{SO} = \frac{Q_O}{A} \]  \\
\[ V_{SG} = \frac{Q_G}{A} \]

where

\[ A = A_W + A_O + A_G \]

The parameter \( A \) is the total cross-sectional area of pipe, \( Q \) is volumetric flow rate, \( V \) is velocity, and the subscripts are \( W \) for water, \( O \) for oil, \( G \) for gas, and \( S \) for superficial term.

2.2.2 Mixture velocity

The fluid mixture velocity is defined as the sum of the superficial velocities:

\[ V_M = V_{SW} + V_{SO} + V_{SG} \]

where, \( V_M \) is the multiphase mixture velocity.

2.2.3 Holdup

Holdup is the cross-sectional area, which locally occupied by one of the phases of a multiphase flow, relative to the cross-sectional area of the pipe at the same local position.
For the liquid phase:

\[ H_L = \frac{A_L}{A} = \frac{A_W + A_O}{A} = H_W + H_O \quad (2-6) \]

For the gas phase:

\[ H_G = \frac{A_G}{A} \quad (2-7) \]

where the parameter H is the phase-holdup and the subscripts are L for the liquid and G for the gas phase.

Although “holdup” can be defined as the fraction of the pipe volume occupied by a given phase, holdup is usually defined as the in situ liquid volume fraction, while the term “void fraction” is used for the in situ gas volume fraction.

### 2.2.4 Phase velocity

Phase velocity (in situ velocity) is the velocity of a phase of a multiphase flow based on the area of the pipe occupied by that phase. It may also be defined for each phase as follows:

\[ V_L = \frac{V_{SL}}{H_L} = \frac{V_{SW} + V_{SO}}{H_L} \quad (2-8) \]

\[ V_G = \frac{V_{SG}}{H_G} \quad (2-9) \]

### 2.2.5 Slip

Slip is the term used to describe the flow condition that exists when the phases have different phase velocities. In most two-phase flow pipelines, the gas travels faster than the liquid. Under this condition, there is said to be slippage between the phases.

The slip velocity is defined as the difference between the actual gas and liquid velocities, as follows:

\[ V_S = V_G - V_L \quad (2-10) \]

If there is no slip between the phases, \( V_L = V_G \), and by applying the no-slip assumption to the liquid holdup definition, it can be shown that

\[ H_{L,\text{no-slip}} = \lambda_L = \frac{V_{SL}}{V_M} \quad (2-11) \]

Investigators have observed that the no-slip assumption is not often applicable. For certain flow patterns in horizontal and upward inclined pipes; gas tends to flow faster than the liquid (positive slip). For some flow regimes in downward flow, liquid can flow faster than the gas (negative slip).
2.2.6 Mixture density

The equations for two-phase gas-liquid density used by various investigators are as follows:

\[ \rho_S = \rho_L H_L + \rho_G H_G \]  
(2-12)

\[ \rho_{nS} = \rho_L \lambda_L + \rho_G \lambda_G \]  
(2-13)

where the subscripts “S” and “nS” represent the slip and the no-slip conditions respectively.

In these equations, the total liquid density can be determined from the oil and water densities and flow rates if no slippage between these liquid phases is assumed:

\[ \rho_L = \rho_O f_O + \rho_W f_W \]  
(2-14)

where

\[ f_O = \frac{Q_O}{Q_O + Q_W} = 1 - f_W \]  
(2-15)

where the parameter f is the volume fraction of each phase.

2.2.7 Mixture viscosity

To determine the mixture viscosity, three approaches have been proposed (Brill and Beggs, 1991):

\[ \mu_S = \mu_L H_L + \mu_G H_G \]  
(2-16)

\[ \mu_S = \mu_L H_L + \mu_G H_G \]  
(2-17)

\[ \mu_{nS} = \mu_L \lambda_L + \mu_G \lambda_G \]  
(2-18)

where \( \mu_L \) and \( \mu_G \) are the liquid and gas viscosity, respectively.

The liquid viscosity can be the viscosity of water, oil, or water–oil mixture. Normally, the water-oil mixture viscosity can be calculated as

\[ \mu_L = \mu_O f_O + \mu_W f_W \]  
(2-19)

Studies have shown that Eqn (2-19) often is not valid for the viscosity of two immiscible liquids, such as oil and water. For some oil/water systems, the emulsion viscosity can be many times higher than the individual phase viscosities. Peak viscosities (over 30 times the oil phase value) typically occur near the inversion point where the emulsion reverts from a water-in-oil dispersion to an oil-in-water dispersion. The inversion point usually prevails at water cuts in the range 20–50% (Arirachakaran, 1983).

Over the years, quite extensive research has been conducted to develop simplified correlations for the water-oil emulsion viscosity. But since there are so many parameters that affect the emulsion viscosity, none of these correlations...
can be universally applied to engineering calculations. Instead, the best way to
determine water-oil emulsion viscosity is to perform lab measurements of
emulsions of different water cut at elevated pressure and temperature
conditions.

2.2.8 Mixture pressure drop

The general pressure-drop equation for multiphase (two- and three-phase) flow
is similar to that for single-phase flow except some of the variables are
replaced with equivalent variables which consider the effect of multiphase. The
general pressure-drop equation for multiphase flow is as follows (Brill and
Beggs, 1991):

\[
\frac{dP}{dx}_{\text{tot}} = \frac{dP}{dx}_{\text{ele}} + \frac{dP}{dx}_{\text{fri}} + \frac{dP}{dx}_{\text{acc}}
\]  
\tag{2-20}

where

\[
\frac{dP}{dx}_{\text{ele}} = \rho_\text{tp} \left( \frac{g}{g_c} \right) \sin \theta \]  
\tag{2-20-1}

\[
\frac{dP}{dx}_{\text{fri}} = \rho_\text{tp} f_\text{tp} V_{\text{tp}}^2 \]  
\tag{2-20-2}

\[
\frac{dP}{dx}_{\text{acc}} = \rho_\text{tp} f_\text{tp} \left( \frac{dV}{dx} \right) \]  
\tag{2-20-3}

where \(dP/dx\) is flow pressure gradient, \(x\) is pipe length, \(\rho\) is flow density, \(V\) is flow
velocity, \(f\) is friction coefficient of flow, \(D\) is internal diameter of pipeline, \(\theta\)
is inclination angle of pipeline, \(g\) is gravitational acceleration, and \(g_c\) is
gravitational constant. The subscripts are “tot” for total, “ele” for elevation, “fri”
for friction loss, “acc” for acceleration change terms, and “tp” for two- and/or
three-phase flow.

The pressure-drop component caused by acceleration is normally negligible
and is considered only for cases of high flow velocities.

Many methods have been developed to predict multiphase, flowing pressure
gradients. They differ in the manner used to calculate the three components of
the total pressure gradient. Section 2-4 describes these methods.

2.2.9 Mixture enthalpy

When performing temperature change calculations for multiphase flow in
pipelines, it is necessary to predict the enthalpy of the multiphase mixture. If
enthalpies of the gas and liquid phases are expressed per unit mass, the enthalpy
of a multiphase mixture can be calculated from

\[
h_M = H_L h_L + (1 - H_L) h_G
\]  
\tag{2-21}
2.3 Multiphase flow regimes

Multiphase flow is characterized by the existence of interfaces between the phases and discontinuities of associated properties. The flow structures are rather classified in “flow regimes” or “flow patterns,” whose precise characteristics depend on a number of parameters. Flow regimes vary depending on operating conditions, fluid properties, flow rates, and the orientation and geometry of the pipe through which the fluids flow. The transition between different flow regimes may be a gradual process. Due to the highly nonlinear nature of the forces that rule the flow regime transitions, the prediction is near impossible. In the laboratory, the flow regime may be studied by direct visual observation using a length of transparent piping. However, the most utilized approach is to identify the actual flow regime from signal analysis of sensors whose fluctuations are related to the flow regime structure. This approach is generally based on average cross-sectional quantities, such as pressure drop or cross-sectional liquid holdup.

2.3.1 Two-phase flow regimes

The description of two-phase flow can be simplified by classifying types of “flow regimes” or “flow patterns.” The distribution of the fluid phases in space and time differs for the various flow regimes and is usually not under the control of the pipeline designer or operator.

Hubbard and Dukler (1966) suggested three basic flow patterns: separated, intermittent, and distributed flow. In separated flow patterns, both phases are continuous and some droplets or bubbles of one phase in the other may or may not exist. In the intermittent flow patterns, at least one phase is discontinuous. In dispersed flow patterns, the liquid phase is continuous, while the gas phase is discontinuous.

Due to multitude of flow patterns and the various interpretations accorded to them by different investigators, the general state of knowledge on flow patterns is unsatisfactory and no uniform procedure exists at present for describing and classifying them. In this section, the basic flow patterns in gas-liquid flow in horizontal, vertical, and inclined pipes are introduced.

2.3.1.1 Horizontal flow regimes

Two-phase, gas-liquid flow regimes for horizontal flow are shown in Figure 2-2. These horizontal flow regimes are defined as follows.

Dispersed bubble flow
At high liquid flow rates and for a wide range of gas flow rates small gas bubbles are dispersed throughout a continuous liquid phase. Due to the effect of buoyancy these bubbles tend to accumulate in the upper part of the pipe.

Plug (elongated bubble) flow
At relatively low gas flow rates, as the liquid flow rate is reduced, the smaller bubbles of dispersed bubble flow coalesce to form larger bullet-shaped bubbles that move along the top of the pipe.
Stratified (smooth and wavy) flow
At low liquid and gas flow rates gravitational effects cause total separation of the two phases. This results in the liquid flowing along the bottom of the pipe and the gas flowing along the top, where the gas-liquid surface is smooth. As the gas velocity is increased in stratified smooth flow the interfacial shear forces increase, rippling the liquid surface and producing a wavy interface.

 Slug flow
 As the gas and liquid flow rates are increased further, the stratified liquid level grows and becomes progressively more wavy until eventually the whole cross-section of the pipe is blocked by a wave. The resultant “piston” of liquid is then accelerated by the gas flow; surging along the pipe, and scooping up the liquid film in front as it progresses. This “piston” is followed by a region
containing an elongated gas bubble moving over a thin liquid film. Hence an intermittent regime develops in which elongated gas bubbles and liquid slugs alternately surge along the pipe. The major difference between elongated bubble flow and slug flow is that in elongated bubble flow there are no entrained gas bubbles in the liquid slugs.

Annular flow
When gas flow rates increase, annular (also referred to as annular mist) flow occurs. During annular flow, the liquid phase flows largely as an annular film on the wall with gas flowing as a central core. Some of the liquid is entrained as droplets in this gas core. The annular liquid film is thicker at the bottom than at the top of the pipe because of the effect of gravity and, except at very low liquid rates, the liquid film is covered with large waves.

2.3.1.2 Vertical flow regimes
Flow regimes frequently encountered in upward vertical two-phase flow are shown in Figure 2-3. These flow regimes tend to be somewhat more simpler than those in horizontal flow. This results from the symmetry in the flow induced by the gravitational force acting parallel to it. A brief description of the manner in which the fluids are distributed in the pipe for upward vertical two-phase flow is as follows. It is worth noting that vertical flows are not so common in raw gas systems (i.e., wells normally have some deviation and many risers are also inclined to some extent).

![Figure 2-3](image)

**FIGURE 2-3**
Upward vertical two-phase flow regimes.
Bubble flow
At very low liquid and gas velocities, the liquid phase is continuous and the gas phase travels as dispersed bubbles. This flow regime is called bubble flow. As the liquid flow rate increases, the bubbles may increase in size via coalescence.

Based on the presence or absence of slippage between the two phases, bubble flow is further classified into bubbly and dispersed bubble flows. In bubbly flow, relatively fewer and larger bubbles move faster than the liquid phase because of slippage. In dispersed bubble flow, numerous tiny bubbles are transported by the liquid phase, causing no relative motion between the two phases.

Slug flow
As the gas velocity increases, the gas bubbles start coalescing, eventually forming large enough bubbles (Taylor bubbles) which occupy almost the entire cross-sectional area. This flow regime is called slug flow. Taylor bubbles move uniformly upward and are separated by slugs of continuous liquid that bridge the pipe and contain small gas bubbles. Typically, the liquid in the film around the Taylor bubbles may move downward at low velocities although the net flow of liquid can be upward. The gas bubble velocity is greater than that of the liquid.

Churn (transition) flow
If a change from a continuous liquid phase to a continuous gas phase occurs, the continuity of the liquid in the slug between successive Taylor bubbles is repeatedly destroyed by a high local gas concentration in the slug. This oscillatory flow of the liquid is typical of churn (froth) flow. It may not occur in small diameter pipes. The gas bubbles may join and liquid may be entrained in the bubbles. In this flow regime, the falling film of the liquid surrounding the gas plugs cannot be observed.

Annular flow
As the gas velocity increases even further, the transition occurs and the gas phase becomes a continuous phase in the pipe core. The liquid phase moves upward partly as a thin film (adhering to the pipe wall) and partly in the form of dispersed droplets in the gas core. This flow regime is called an annular flow or an annular-mist flow.

Although downward vertical two-phase flow is less common than upward flow, it does occur in steam injection wells and downcomer pipes from offshore production platforms. Hence a general vertical two-phase flow pattern is required that can be applied to all flow situations. Reliable models for downward multiphase flow are currently unavailable and the design codes are deficient in this area.

2.3.1.3 Inclined flow regimes
The effect of pipeline inclination on the gas-liquid two-phase flow regimes is of a major interest in hilly terrain pipelines that consist almost entirely of uphill and
downhill inclined sections. Pipe inclination angles have a very strong influence on flow pattern transitions. Generally, the flow regime in a near horizontal pipe remains segregated for downward inclinations and changes to intermittent flow regime for upward inclinations. An intermittent flow regime remains intermittent when tilted upward and tends to segregated flow pattern when inclined downward. The inclination should not significantly affect the distributed flow regime (Scott et al., 1987).

2.3.1.4 Flow pattern maps

In order to obtain optimal design parameters and operating conditions, it is necessary to clearly understand multiphase flow regimes and the boundaries between them, where the hydrodynamics of the flow as well as the flow mechanisms change significantly from one flow regime to another. If an undesirable flow regime is not anticipated in the design, the resulting flow pattern can cause system pressure fluctuation and system vibration, and even mechanical failures of piping components.

Most early attempts to predict the occurrence of the various flow patterns in pipes were based on conducting experimental tests in small diameter pipes at low pressures with air and water. The results of experimental studies were presented as a flow pattern map. The respective pattern was represented as areas on a plot, the coordinates of which were the dimensional variables (i.e., superficial phase velocities), or dimensionless parameters containing these velocities. For horizontal flows, the classical flow pattern map is that of Mandhane et al. (1974) as shown in Figure 2-4. This particular map is based on air-water data at atmospheric pressure in 0.5- to 6.5-inch pipes.

![Flow Pattern Map](image)

**FIGURE 2-4**
Mandhane et al. (1974) flow regime map for horizontal flow.
Figure 2-5 shows the flow pattern map developed by Aziz et al. (1972) for vertical upward flow. Equations (2-22) and (2-23) define the coordinates.

\[
N_x = V_{SG} \left( \frac{\rho_G}{0.0764} \right)^{\frac{1}{3}} \left[ \left( \frac{72}{\sigma_L} \right) \left( \frac{\rho_L}{62.4} \right) \right]^{\frac{1}{4}}
\]

\[
N_y = V_{SL} \left[ \left( \frac{72}{\sigma_L} \right) \left( \frac{\rho_L}{62.4} \right) \right]^{\frac{1}{4}}
\]

The bracketed terms in the above equations are attempts to validate the flow pattern map for fluids other than air and water.

Equations (2-24) to (2-26) represent the flow pattern transitions in Figure 2-5.

\[
N_1 = 0.51 \left( 100N_y \right)^{0.172}
\]

\[
N_2 = 8.6 + 3.8N_y
\]

\[
N_3 = 70 \left( 100N_y \right)^{-0.152}
\]

where superficial velocities are in feet per second, liquid and gas densities are in pounds per cubic feet, and liquid surface tension\(^1\) (\(\sigma_L\)) is in dynes per centimeter.

\(^1\)Assuming no slippage, the liquid surface tension is calculated from \(\sigma_L = \sigma_{O}f_{O} + \sigma_{W}f_{W}\).
Empirically based flow maps are not particularly accurate for systems where the fluid properties, pipe size, and inclination are different from those for which the flow map was originally produced. In addition, in these flow maps a change in flow rate can lead to a change of flow pattern between the old and new steady states. However, an interesting and important effect that frequently occurs is the existence of temporary flow pattern changes between the respective steady states (Shaha, 1999). A more flexible method, which overcomes this difficulty, is to examine each transition individually and derive a criterion valid for that particular transition. There have been some attempts to evaluate the basic mechanisms of flow pattern transitions and thus to provide a mechanistic flow pattern map for estimating their occurrence (Taitel and Dukler, 1976; Barnea, 1987; Taitel et al., 1980; Petalas and Aziz, 2000). In these transition models, the effects of system parameters are incorporated; hence they can be applied over a range of conditions. However, most of them are somewhat complex and they do require the use of a predetermined sequence to determine the dominant flow pattern.

2.3.2 Three-phase flow regimes

The main difference between two-phase (gas/liquid) flows and three-phase (gas/liquid/liquid) flows is the behavior of the liquid phases, where in three-phase systems the presence of two liquids gives rise to a rich variety of flow patterns (Hall, 1997). Basically, depending on the local conditions, the liquid phases appear in a separated or dispersed form (Brauner and Maron, 1992). In the case of separated flow, distinct layers of oil and water can be discerned, though there may be some interentrainment of one liquid phase into the other. In dispersed flow, one liquid phase is completely dispersed as droplets in the other, resulting in two possible situations, namely an oil continuous phase and a water continuous phase (Chen and Guo, 1999). The transition from one liquid continuous phase to the other is known as phase inversion. If the liquid phases are interdispersed, then prediction of phase inversion is an important item. For this purpose, Decarre and Fabre (1997) developed a phase inversion model that can be used to determine which of two liquids is continuous.

Due to the many possible transport properties of three-phase fluid mixtures, the quantification of three-phase flow pattern boundaries is a difficult and challenging task. Acikgoz et al. (1992) observed a very complex array of flow patterns and described 10 different flow regimes. In their work, the pipe diameter was only 0.748 inch and stratification was seldom achieved. In contrast, the Lee et al. (1993) experiments were carried out in a 4-inch diameter pipe. They observed and classified seven flow patterns, which were similar to those in of two-phase flows: (1) smooth stratified, (2) wavy stratified, (3) rolling wave, (4) plug flow, (5) slug flow, (6) pseudo slug, and (7) annular flow. The first three flow patterns can be classified as stratified flow regime and they noted that the oil and the water are generally segregated, with water flowing as a liquid layer at the bottom of the
pipe and oil flowing on top. Even for plug flow, the water remained at the bottom, because the agitation of the liquids was not sufficient to mix the oil and water phases. Note, turbulence, which naturally exists in a pipeline, can be sufficient to provide adequate mixing of the water and oil phase. However, the minimum natural turbulent energy for adequate mixing depends on the oil and water flow rates, pipe diameter and inclination, water concentration, viscosity, density, and interfacial tension. Dahl et al. (2001) provided more detailed information on the prediction methods that can be used to determine whether a water-in-oil mixture in the pipe is homogeneous or not.

2.3.3 Gas-condensate flow regimes

Gas-condensate flow is a multiphase flow phenomenon commonly encountered in raw gas transportation. However, the multiphase flow that takes place in gas-condensate transmission lines differs in certain respects from the general multiphase flow in pipelines. In fact, in gas-condensate flow systems, there is always interphase mass transfer from the gas phase to the liquid phase because of the temperature and pressure variations. This leads to compositional changes and associated fluid property changes (Ayala and Adewumi, 2003). In addition to that, the amount of liquid in such systems is assumed to be small, and the gas flow rate gives a sufficiently high Reynolds number that the fluid flow regime for a nearly horizontal pipe can be expected to be annular-mist flow and/or stratified flow (Asante, 2002). For other inclined cases, even with small quantities of liquids, slug-type regimes may be developed if liquids start accumulating at the pipe lower section.

In water/wet gas-condensate systems it can normally be assumed that the condensate and water phases are well mixed. However, when the water loading is a substantial proportion of the total liquid loading, the water and condensate can separate out at low flow rates leading to the accumulation of stagnant water at low points in the line. The presence of stagnant water can lead to the production of water slugs during normal line operation or flow rate changes.

2.4 Determining multiphase flow design parameters

The hydraulic design of a multiphase flow pipeline is a two-step process. The first step is the determination of the multiphase flow regimes, because many pressure-drop calculation methods rely on the type of flow regime present in the pipe. The second step is the calculation of flow parameters, such as pressure drop and liquid holdup, to size pipelines and field processing equipment such as slug catchers.

The analysis of multiphase flow phenomena in pipeline systems is usually classified into steady-state and transient approaches. In steady-state flow, there are no major changes transgressing the pipeline network. In transient or dynamic flows, the flow behavior is changing on a regular and significant basis. Both steady-state and transient analyses shall be performed to study the pipeline
system performance under different operating conditions. In preliminary engineering, a design engineer often uses steady-state analysis. However, the dynamic response of the system to start-up, shutdown and flow rate changes must also be considered. For example, liquid slugs are often produced during flow rate increase or line depressurization. The volume of these transient slugs must be determined to ensure that they can be handled by the downstream separation and processing facilities. If there is insufficient capacity, operating procedures can be introduced to limit the rate of change of flow rate and therefore the size of any produced slugs.

2.4.1 Steady-state two-phase flow

The techniques most commonly used in the design of two-phase flow pipeline can be classified into three categories: the single-phase flow approaches, the homogeneous flow approaches, and the mechanistic models. Within each of these groups are subcategories that are based on the general characteristics of the models used to perform design calculations.

2.4.1.1 Single-phase flow approaches

In this method, the two-phase flow is assumed to be a single-phase flow having pseudoproperties arrived at by suitably weighting the properties of the individual phases. These approaches basically rely solely on the well-established design equations for single-phase gas flow in pipes. Two-phase flow is treated as a simple extension by use of a multiplier; a safety factor to account for the higher pressure drop generally encountered in two-phase flow. This heuristic approach was widely used and generally resulted in inaccurate pipeline design (Schweikert, 1986).

In the past, single-phase flow approaches were commonly used for the design of wet-gas pipelines. When the amount of the condensed liquid is negligibly small, the use of such methods could at best prevent under design, but more often than not, the quantity of the condensed liquid is significant enough that the single-phase flow approach grossly over predicts the pressure drop (Ullah, 1987). Hence, this method is not covered in-depth here but can be reviewed elsewhere (Uhl, 1965).

2.4.1.2 Homogeneous flow approaches

The inadequacy of the single-phase flow approaches spurred researchers to develop better design procedures and predictive models for two-phase flow systems. This effort led to the development of homogeneous flow approaches to describe these rather complex flows. The homogeneous approach, also known as the friction factor model, is similar to that of the single-phase flow approach except that mixture fluid properties are used in the determination of the friction factor. Therefore, the appropriate definitions of the fluid properties are critical to the accuracy of the model. The mixture properties are expressed empirically as a
function of the gas and liquid properties as well as their respective holdups. Many of these correlations are based on flow regime correlations that determine the two-phase (gas/liquid) flow friction factor, which is then used to estimate pressure drop. While some of the correlations predict pressure-drop reasonably well, their range of applicability is generally limited, making their use as a scale-up tool marginal. This limitation is understandable because the database used in developing these correlations is usually limited and based on laboratory-scale experiments. Extrapolation of these data sets to larger lines and hydrocarbon systems is questionable at best (Cindric et al., 1987).

Historically, empirical correlations have been successful in terms of being relatively easy to employ as design tools. Brill and Beggs (1991) and Collier and Thome (1996) provide in-depth comparative analyses of the available correlations. However, two of the better known, more widely applicable correlations are outlined here.

Lockhart and Martinelli method
The method was developed by correlating experimental data generated in horizontal isothermal two-phase flow of two-component systems (air-oil and air-water) at low pressures (close to atmospheric) in a 1-inch diameter pipe. Lockhart and Martinelli (1949) separated the data into four sets dependent on whether the phases flowed in laminar or turbulent flow, if the phases were flowing alone in the same pipe. In this method, a definite portion of the flow area is assigned to each phase and it is assumed that the single-phase pressure-drop equations can be used independently for each phase. The two-phase frictional pressure drop is calculated by multiplying by a correction factor for each phase, as follows:

$$\frac{dP}{dx} = \varphi_G^2 \left( \frac{dP}{dx} \right)_G = \varphi_L^2 \left( \frac{dP}{dx} \right)_L$$  \hspace{1cm} (2-27)

where

$$\left( \frac{dP}{dx} \right)_G = \left( f_G \rho_G V_{SG}^2 \right) \frac{2gCD}{\mu_G}$$  \hspace{1cm} (2-27-1)

$$\left( \frac{dP}{dx} \right)_L = \left( f_L \rho_L V_{SL}^2 \right) \frac{2gCD}{\mu_L}$$  \hspace{1cm} (2-27-2)

The friction factors $f_G$ and $f_L$ are determined from Moody (1944) diagram using the following values of the Reynolds number:

$$N_{Re,G} = \frac{\rho_G V_{SG}D}{\mu_G}$$  \hspace{1cm} (2-28)

$$N_{Re,L} = \frac{\rho_L V_{SL}D}{\mu_L}$$  \hspace{1cm} (2-29)
The two-phase flow correction factors \( \Phi_G, \Phi_L \) are determined from Eqns (2-30) and (2-31) (Lockhart and Martinelli, 1949; Chisholm and Sutherland, 1969):

\[
\Phi_G^2 = 1 + CX + CX^2 \tag{2-30}
\]

\[
\Phi_L^2 = 1 + \frac{C}{X} + \frac{1}{X^2} \tag{2-31}
\]

where

\[
X = \left[ \frac{(dP/dx)_L}{(dP/dx)_G} \right]^{0.5} \tag{2-32}
\]

and parameter \( C \) is given in Table 2-1.

Note, the laminar flow regime for a phase occurs when the Reynolds number for that phase is less than 1000.

In this method, the correlation between liquid holdup and Martinelli parameter, \( X \), is independent of the flow regime and can be expressed as follows (Collier and Thome, 1996):

\[
H_L^2 = 1 + \frac{20}{X} + \frac{1}{X^2} \tag{2-33}
\]

The fluid acceleration pressure drop was ignored in this method. Although different modifications of this method have been proposed, but the original method is believed to be generally the most reliable (Collier and Thome, 1996).

**Beggs and Brill method**

This method was developed from 584 experimental data sets generated on a laboratory scale test facility using an air-water system. The facility consisted of 90 ft of 1- or 1.5-inch diameter acrylic (smooth) pipe, which could be inclined at any angle (Beggs and Brill, 1973). The pipe angle was varied between horizontal to vertical and the liquid holdup and the pressure were measured. For each pipe size in the horizontal position, flow rates of two phases were varied to achieve all flow regimes. Beggs and Brill (1973) developed correlations for the liquid holdup for each of three horizontal flow regimes and then corrected these for the pipe inclination/angle.

<table>
<thead>
<tr>
<th>Table 2-1 “C” Parameter</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid Phase</td>
</tr>
<tr>
<td>----------------</td>
</tr>
<tr>
<td>Turbulent</td>
</tr>
<tr>
<td>Laminar</td>
</tr>
<tr>
<td>Turbulent</td>
</tr>
<tr>
<td>Laminar</td>
</tr>
</tbody>
</table>
The following parameters are used for determination of all horizontal flow regimes (Brill and Beggs, 1991):

\[
N_{Fr} = \frac{V^2}{gD} \quad (2-34)
\]

\[
L_1 = 316\lambda_L^{0.302} \quad (2-35)
\]

\[
L_2 = 0.0009252\lambda_L^{-2.4684} \quad (2-36)
\]

\[
L_3 = 0.10\lambda_L^{-1.4516} \quad (2-37)
\]

\[
L_4 = 0.5\lambda_L^{-6.738} \quad (2-38)
\]

And the flow regimes limits are

**Segregated**

\[
N_{Fr} < L_1 \quad \text{for} \quad \lambda_L < 0.01
\]

\[
N_{Fr} < L_2 \quad \text{for} \quad \lambda_L \geq 0.01
\]

**Transition**

\[
L_2 \leq N_{Fr} \leq L_3 \quad \text{for} \quad \lambda_L \geq 0.01
\]

**Intermittent**

\[
L_3 \leq N_{Fr} \leq L_1 \quad \text{for} \quad 0.01 < \lambda_L \leq 0.4
\]

\[
L_3 \leq N_{Fr} \leq L_4 \quad \text{for} \quad \lambda_L \geq 0.01
\]

**Distributed**

\[
N_{Fr} \geq L_1 \quad \text{for} \quad \lambda_L < 0.4
\]

\[
N_{Fr} > L_4 \quad \text{for} \quad \lambda_L \geq 0.4
\]

Also, the horizontal liquid holdup, \(H_L(0)\), is calculated using the following equation:

\[
H_L(0) = \frac{a \lambda_L^b}{N_{Fr}^c} \quad (2-39)
\]

where parameters a, b, and c are determined for each flow regime and are given in Table 2-2.
With the constraint that \( H_L(0) \geq \lambda_L \).

When the flow is in the transition region, the liquid holdup is calculated by interpolating between the segregated and intermittent flow regimes as follows:

\[
H_L(0)_\text{transition} = A \ H_L(0)_\text{segregated} + (1 - A)H_L(0)_\text{intermittent} \tag{2-40}
\]

where

\[
A = \left(\frac{L_3 - N_{Fr}}{L_3 - L_2}\right) \tag{2-41}
\]

The amount of liquid holdup in an inclined pipe, \( H_L(\theta) \), is determined by multiplying an inclination factor (\( \psi \)) by the calculated liquid holdup for the horizontal conditions:

\[
H_L(\theta) = H_L(0) \times \psi \tag{2-42}
\]

where

\[
\psi = 1 + \alpha \left[ \sin \left(1.8\theta\right) - 0.333 \sin^3 \left(1.8\theta\right) \right] \tag{2-42-1}
\]

where \( \theta \) is the pipe angle and \( \alpha \) is calculated as follows:

\[
\alpha = (1 - \lambda_L) \ln \left[ d \ \dfrac{\lambda_c N_{LV}^f N_{Fr}^g}{g} \right] \tag{2-42-2}
\]

where

\[
N_{LV} = VSL \left( \frac{\rho_L}{g \ \sigma} \right)^{0.25} \tag{2-42-3}
\]

The equation parameters are determined from each flow regime using the numbers from Table 2-3.

<table>
<thead>
<tr>
<th>Flow Regime</th>
<th>d</th>
<th>e</th>
<th>f</th>
<th>g</th>
</tr>
</thead>
<tbody>
<tr>
<td>Segregated uphill</td>
<td>0.011</td>
<td>-3.768</td>
<td>3.539</td>
<td>-1.614</td>
</tr>
<tr>
<td>Intermittent uphill</td>
<td>2.96</td>
<td>0.305</td>
<td>-0.4473</td>
<td>0.0978</td>
</tr>
<tr>
<td>Distributed uphill</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>All flow regimes downhill</td>
<td>4.7</td>
<td>-0.3692</td>
<td>0.1244</td>
<td>-0.5056</td>
</tr>
</tbody>
</table>

\[ a = \frac{L_3 - N_{Fr}}{L_3 - L_2} \]
The two-phase pressure gradient due to pipeline elevation can be determined as follows:

\[ \frac{dP}{dx}_{\text{ele}} = \frac{g}{g_c} \left\{ \rho_L H_L(\theta) + \rho_G [1 - H_L(\theta)] \right\} \]  \hspace{1cm} (2-43)

Also, the two-phase frictional pressure drop is calculated as follows:

\[ \frac{dP}{dx}_{\text{fri}} = \frac{f_p \rho_n V_M^2}{2 g_c D} \]  \hspace{1cm} (2-44)

where

\[ f_p = f_n \exp(\beta) \]  \hspace{1cm} (2-44-1)

where \( f_n \) is the no slip friction factor determined from the smooth pipe curve of the Moody diagram using no slip viscosity and the density in calculating the two-phase flow Reynolds number. In other words:

\[ f_n = \frac{1}{\left[ 2 \log\left( \frac{N_{Re,n}}{4.5223 \log N_{Re,n} - 3.8215} \right) \right]^2} \]  \hspace{1cm} (2-44-1-1)

where

\[ N_{Re,n} = \frac{\rho_n V_M D}{\mu_n} = \left[ \frac{\left( \rho_L \lambda_L + \rho_G (1 - \lambda_L) \right) V_M D}{\mu_H \lambda_L + \mu_G (1 - \lambda_L)} \right] \]  \hspace{1cm} (2-44-1-2)

The exponent \( \beta \) is given by Eqn (2-44-2):

\[ \beta = \frac{[\ln Y]}{-0.0523 + 3.182 \ln Y - 0.8725 (\ln Y)^2 + 0.01853 (\ln Y)^4} \]  \hspace{1cm} (2-44-2)

where

\[ Y = \frac{\lambda_L}{[H_L(\theta)]^2} \]  \hspace{1cm} (2-44-3)

The pressure drop due to acceleration is only significant in gas transmission pipelines at high gas flow rates. However, it can be included for completeness as follows:

\[ \left( \frac{dP}{dx} \right)_{\text{acc}} = \left[ \frac{\rho_S V_M V_{SG}}{g_c \rho} \right] \left( \frac{dP}{dx} \right)_{\text{tot}} \]  \hspace{1cm} (2-45)

The Beggs and Brill (1973) method can be used for horizontal and vertical pipelines, although its wide acceptance is mainly due to its usefulness for inclined pipe pressure-drop calculation (Brill and Beggs, 1991). However, the proposed approach has limited applications dictated by the database on which the correlations are derived.
Note, the Beggs and Brill (1973) correlation is not usually employed for the design of wet gas pipelines. Specifically the holdup characteristic (holdup versus flow rate) is not well predicted by this method. This makes design difficult because one is unable to reliably quantify the retention of liquid in the line during turndown conditions.

2.4.1.3 Mechanistic models
While it is indeed remarkable that some of the present correlations can adequately handle noncondensing two-phase flow, they give erroneous results for gas-condensate flow in large diameter lines at high pressures (Battara et al., 1985). This shortcoming of existing methods has led to the development of mechanistic models that are based on fundamental laws and thus can offer more accurate modeling of the pipe geometric and fluid property variations. All of these models firstly predict stable flow pattern under the specified conditions and then use the momentum balance equations to calculate liquid holdup, pressure drop and other two-phase flow parameters with a greater degree of confidence than that possible by purely empirical correlations (Collier and Thome, 1996; Holt et al., 1999). The mechanistic models presented in the literature are either incomplete in that they only consider flow pattern determination, or are limited in their applicability to only some pipe inclinations or small-diameter, low-pressure two-phase flow lines (Taitel and Dukler, 1976; Barnea, 1987; Xiao et al., 1990; Ansari et al., 1994; Taitel et al., 1995; Wilkens, 1997). However, new mechanistic models presented by Petalas and Aziz (2000) and Zhang et al. (2003) have proven to be more robust than previous models, although further investigations and testing of these models are needed with high-quality field and laboratory data in larger diameter, high-pressure systems. Readers are referred to the original references for a detailed treatment of these models.

2.4.2 Steady-state three-phase flow
Compared to numerous investigations of two-phase flow in the literature, there are only limited works on three-phase flow of gas-liquid-liquid mixtures. In fact, the complex nature of such flows makes prediction very difficult. In an early study, Tek (1961) treated the two immiscible liquid phases as a single fluid with mixture properties, thus a two-phase flow correlation could be used for pressure loss calculations. On the other hand, recent studies by Pan (1996) have shown that the classical two-phase flow correlations for gas-liquid two-phase flow can be used as the basis in the determination of three-phase flow parameters; however, the generality of such empirical approaches is obviously questionable. Hence, an appropriate model is required to describe the flow of one gas and two liquid phases. One of the most fundamental approaches used to model such systems is the two-fluid model (TFM), where the presented approach can be used by combining the two liquid phases as one pseudoliquid phase and modeling the three-phase flow as a two-phase flow. However, for more accurate results,
three-fluid models should be used to account for the effect of liquid-liquid interactions on flow characteristics, especially at low flow rates. Several three-fluid models were found in the literature. All of these models are developed from the three-phase momentum equations with few changes from one model to the other. The most obvious model has been developed by Barnea and Taitel (1996), however, such a model introduces much additional complexity and demands much more in computer resources compared with the TFM for two-phase flow (Bonizzi and Issa, 2003).

2.4.3 Transient multiphase flow

Steady-state operation is the exception rather than the rule in multiphase systems due to a number of factors such as the nature of the flow regimes, interactions with the pipeline profile, and changes in supply or demand. As a result, detailed information of the transient flow behavior is necessary for the designer and the operator of the system to construct and operate the pipeline economically and safely.

Transient multiphase flow is traditionally modeled by one-dimensional averaged conservation laws, yielding a set of partial differential equations. In this section, two models, of particular industrial interest, are described:

- The TFM, consisting of a separate momentum equation for each phase.
- The drift-flux model (DFM), consisting of a momentum equation and an algebraic slip relation for the phase velocities.

The TFM is structurally simpler, but involves an extra differential equation when compared to the DFM. They do yield somewhat different transient results, though the differences are often small (Masella et al., 1998).

The following major assumptions have been made in the formulation of the differential equations:

1. Two immiscible liquid phases (oil and water) that are assumed to be a single fluid with mixture properties.
2. Flow is one-dimensional in the axial direction of the pipeline.
3. Flow temperature is constant at wall, and no mass transfer occurs between the gas and liquid phase. Note, most commercial codes allow phase change.
4. The physical properties of multiphase flow are determined at the average temperature and pressure of flow in each segment of the pipeline.

2.4.3.1 Two-fluid model

The TFM is governed by a set of four partial differential equations, the first two of which express mass conservation for gas and liquid phases respectively:

\[
\frac{\partial}{\partial t} [\rho_G H_G] + \frac{\partial}{\partial x} [\rho_G H_G V_G] = 0 \tag{2-46}
\]

\[
\frac{\partial}{\partial t} [\rho_L H_L] + \frac{\partial}{\partial x} [\rho_L H_L V_L] = 0 \tag{2-47}
\]
and the last two equations represent momentum balance for the gas and liquid phases respectively:

\[
\frac{\partial}{\partial t}[\rho_G H_G V_G] + \frac{\partial}{\partial x}\left[\rho_G H_G V_G^2 + H_G \Delta P_G\right] + H_G \frac{\partial P}{\partial x} = \tau_G + \tau_i - \rho_G g \sin \theta
\]  
(2-48)

\[
\frac{\partial}{\partial t}[\rho_L H_L V_L] + \frac{\partial}{\partial x}\left[\rho_L H_L V_L^2 + H_L \Delta P_L\right] + H_L \frac{\partial P}{\partial x} = \tau_L - \tau_i - \rho_L g \sin \theta
\]  
(2-49)

In Eqns (2-48) and (2-49), parameter P denotes the interface pressure, while \(V_k\), \(\rho_k\), and \(H_k\) are the velocity, the density, and the volume fraction of phase \(k \in \{G, L\}\) respectively. The variables \(\tau_i\) and \(\tau_k\) are the interfacial and wall momentum transfer terms. The quantities \(\Delta P_G\) and \(\Delta P_L\) correspond to the static head around the interface (De Henau and Raithby, 1995), defined as follows:

\[
\Delta P_G = P_G - P = -\rho_G \left[\frac{1}{2} \cos \left(\frac{\omega}{2}\right) + \frac{1}{3\pi H_G} \sin^3 \left(\frac{\omega}{2}\right)\right] g D \cos \theta
\]  
(2-50)

\[
\Delta P_L = P_L - P = -\rho_L \left[\frac{1}{2} \cos \left(\frac{\omega}{2}\right) + \frac{1}{3\pi H_L} \sin^3 \left(\frac{\omega}{2}\right)\right] g D \cos \theta
\]  
(2-51)

where \(\omega\) is the wetted angle.

The detailed description of the solution algorithm for the above equations is based on a finite volume method (Masella et al., 1998).

One major limitation of this type of model is the treatment of the interfacial coupling. Whilst this is relatively easy for separated flows (stratified and annular), this treatment is intrinsically flawed for intermittent flows.

### 2.4.3.2 Drift-flux model

The DFM is derived from the TFM by neglecting the static head terms \(\Delta P_G\) and \(\Delta P_L\) in Eqns (2-48) and (2-49) and replacing the two momentum equations by their sum. This leads in the DFM to a new model that consists of three partial differential equations, i.e., Eqns (2-46, 2-47 and (2-52).

\[
\frac{\partial}{\partial t}\left[\rho_G H_G V_G + \rho_L H_L V_L\right] + \frac{\partial}{\partial x}\left[\rho_G H_G V_G^2 + \rho_L H_L V_L^2 + P\right] = \tau_G + \tau_L - (\rho_G H_G + \rho_L H_L) \sin \theta
\]  
(2-52)

The detailed description of the numerical solution for the relevant equations can be found in Faille and Heintze (1996).

The main advantages of this three-equation model are as follows:

- The equations are in conservative form, which makes their solution by the finite volume methods less onerous.
- The interfacial shear term, \(\tau_i\), is cancelled out in the momentum equations, although it appears in an additional algebraic relation called the slip law.
- The model is well posed and does not exhibit complex characteristic.
Note, the drift-flux approach is best applied to closely coupled flows such as bubbly flow. Its application to stratified flows is, at best, artificial (Banerjee, 1986).

### 2.4.4 Multiphase gas and condensate flow

In order to achieve optimal design of gas-condensate pipelines and downstream processing facilities, one needs a description of the relative amount of condensate and the flow regime taking place along the pipelines, where fluid flowing in pipelines may traverse the fluid phase envelope such that the fluid phase changes from single-phase to two-phase or vice versa. Hence, compositional single-phase/multiphase hydrodynamic modeling, which couples the hydrodynamic model with the natural gas phase behavior model, is necessary to predict fluid dynamic behavior in gas-condensate transmission lines. The hydrodynamic model is required to obtain flow parameters along the pipeline, and the phase behavior model is required for determining the phase condition at any point in the pipe, the mass transfer between the flowing phases, and the fluid properties.

Despite the importance of gas flow with low liquid loading for the operation of gas pipelines, few attempts have been made to study flow parameters in gas-condensate transmission lines. While, the single-phase flow approaches have been previously applied to gas-condensate systems, only a few attempts have been reported for the use of TFM for this purpose. Some of them have attempted to make basic assumptions (e.g., no mass transfer between gas and liquid phases, etc.) in their formulation and others simply assume one flow regime for the entire pipe length. However, a compositional hydrodynamic model that describes the steady-state behavior of multiphase flow in gas-condensate pipelines has been recently presented by Ayala and Adewumi (2003). The model couples a phase behavior model, based on the Peng and Robinson (1976) equation of state (EOS), and a hydrodynamic model, based on the TFM. The proposed model is a numerical approach, which can be used as an appropriate tool for engineering design of multiphase pipelines transporting gas and condensate. However, the complexity of this model precludes further discussion here.

Note, the presence of liquid (condensates), besides reducing deliverability, creates several operational problems in gas-condensate transmission lines. Periodic removal of the liquid from the pipeline is thus desirable. To remove liquid accumulation in the lower portions of pipeline, pigging operations are performed. These operations keep the pipeline free of liquid, reducing the overall pressure drop and by the same increase the pipeline flow efficiency. However, the pigging process associates with transient flow behavior in the pipeline. Thus, it is imperative to have a means of predicting transient behavior encountered in multiphase, gas/gas-condensate pipelines. Until recently, most available commercial codes are based on the TFM. However, the model needs many modifications to be suitable for
simulating multiphase transient flow in gas/gas-condensate transmission lines. For example, the liquid and gas continuity equations need to be modified to account for the mass transfer between phases. So far, several codes have been reported for this purpose, where three main commercial transient codes are OLGA (Bendiksen et al., 1991), PROFES (Black et al., 1990), and TACITE (Pauchon et al., 1993). Detailed discussion of these codes is beyond the scope of this book, readers are referred to the original papers for further information.

2.5 Predicting temperature profile of multiphase pipeline

Predicting the flow temperature and pressure changes has become increasingly important for use in both the design and operation of flow transmission pipelines. It is therefore imperative to develop appropriate methods capable of predicting these parameters for multiphase pipelines. A simplified flow chart of a suitable computing algorithm is shown in Figure 2-6. This algorithm calculates pressure and temperature along the pipeline by iteratively converging on pressure and temperature for each sequential “segment” of the pipeline. The algorithm converges on temperature in the outer loop and pressure in the inner loop because robustness and computational speed are obtained when converging on the least sensitive variable first (Brill and Beggs, 1991).

The pipeline segment length should be chosen such that the fluid properties do not change significantly in the segment. More segments are recommended for accurate calculations for a system where fluid properties can change drastically over short distances. Often, best results are obtained when separate segments (with the maximum segment length less than about 10% of total line length) are used for up, down, and horizontal segments of the pipeline (Brill and Beggs, 1991).

Prediction of the pipeline temperature profile can be accomplished by coupling the pressure gradient and enthalpy gradient equations as follows (Brill and Beggs, 1991):

$$\Delta h = -\frac{V_M V_{SG} \Delta P}{(778)(32.17)(\bar{P})} + \frac{\Delta Z}{778} - \frac{U \pi D(T - T_a) \Delta L}{3600(M_M)}$$  

where $\Delta h$ is enthalpy change in the calculation segment, Btu/lbm; $V_M$ is velocity of the fluid, ft/s; $V_{SG}$ is superficial gas velocity, ft/s; $\Delta P$ is estimated change in pressure, psi; $\bar{P}$ is average pressure in calculation segment, psia; $\Delta Z$ is change in elevation, ft; $U$ is overall heat transfer coefficient, Btu/h-ft²°C; $D$ is reference diameter on which $U$ is based, ft; $\bar{T}$ is estimated average temperature in calculation segment, °F; $T_a$ is ambient temperature, °F; $\Delta L$ is change in segment length, ft; and $M_M$ is gas-liquid mixture mass flow rate, lbm/s.

As can be seen from Eqn (2-53), temperature and pressure are mutually dependent variables so that generating a very precise temperature profile
FIGURE 2-6
Pressure and temperature calculation procedure (Brill and Beggs, 1991).
requires numerous iterative calculations. The temperature and pressure of each pipe segment are calculated using a double-nested procedure in which for every downstream pressure iteration, convergence is obtained for the downstream temperature by property values evaluated at the average temperature and pressure of that section of pipe. Experience shows that it is essential to use a good pressure-drop model to assess the predominant parameter, i.e., pressure.

The overall heat transfer coefficient (U) can be determined by combination of several coefficients, which depend on the method of heat transfer and pipe configuration. Figure 2-7 is a cross-section of a pipe, including each “layer” through which heat must pass to be transferred from the fluid to the surroundings or vice versa. This series of layers has an overall resistance to heat transfer made up of the resistance of each layer.

In general, the overall heat transfer coefficient for a pipeline is the reciprocal of the sums of the individual resistances to heat transfer, where each resistance definition is given in Table 2-4.

The individual resistances are calculated from the given equations in Table 2-5.
Since fluid properties are key inputs into calculations such as pressure drop and heat transfer, the overall simulation accuracy depends on accurate property predictions of the flowing phases. Most of the required physical and thermodynamic properties of the fluids are derived from the EOS. However, empirical correlations are used for calculation of viscosity and surface tension. When the physical properties of the fluids are calculated for two-phase flow, the physical properties of the gas and liquid mixture can be calculated by taking the mole
fraction of these components into account. Similarly, when water is present in
the system, the properties of oil and water are combined into those of a
pseudoliquid phase. To produce the phase split for a given composition, press-
ure, and temperature; an equilibrium flash calculation utilizing an appropriate
EOS must be used.

2.6 Velocity criteria for sizing multiphase pipelines

Having selected a line size to meet throughput and pressure-drop constraints, it is
important to check whether acceptable flowing conditions exist within the
pipeline. However, a more useful means of assessing whether flowing conditions
are acceptable is to check whether fluid velocities lie within certain limits.

At the low end of the normal throughput range, the actual (not superficial)
liquid velocity should ideally be greater than 3 ft/s. This will ensure that sand and
water are continuously transported with the liquid phase and not allowed to drop
out and accumulate at the bottom of the pipe. However; at the maximum
throughput conditions the maximum mixture velocity in the multiphase pipeline
must be calculated in order to check that its value does not exceed the erosional
velocity limit.

The current industry standard method of determining the erosional velocity
limit is through use of the relationship given in API RP 14E (1991):

\[ V_e = \frac{C}{(\rho_M)^{0.5}} \]  

where \( V_e \) is the maximum acceptable mixture velocity to avoid excessive erosion,
ft/s; \( C \) is empirical constant; and \( \rho_M \) is the no-slip mixture density at flowing
conditions, lb/ft\(^3\).

Industry experience to date indicates that for solid-free fluids a C-factor of
100 for continuous service and a C-factor of 125 for intermittent service are
conservative.

The above simple criterion is specified for clean service (noncorrosive and
sand-free), and that the limits should be reduced if sand or corrosive conditions
are present. However, no guidelines are provided for these reductions.

If the erosional velocity limit is exceeded either the line diameter must be
increased or the production profile constrained to reduce the maximum mixture
velocity. Alternatively a higher grade of pipeline material could be used.

As well as the erosional velocity limit, API RP 14E (1991) also recommends
that a line velocity of 60 ft/s should not be exceeded to ensure that the level of
noise emitted by the flow is not excessive. In addition, in some pipeline systems
flowing velocities need to be limited by a requirement to avoid removal of
corrosion inhibitor film, ensuring a lubricating effect that shifts the erosion
velocity limit.
2.7 Multiphase pipeline operations

After a pipeline is installed, efficient operating procedures must be provided to maintain safe, long-term operation of the pipeline in the face of unexpected upsets, and to improve efficiency and economics of operation. Leak detection, depressurization, and pigging are typically important procedures.

2.7.1 Leak detection

Leaks in the pipelines originate from a variety of causes, which may include material-related damage, physical damage caused by construction in the right-of-way, etc. Because accidental product discharges cannot be entirely eliminated, one of the most effective methods of reducing the impact of spills is to quickly detect the leak and to act quickly to stop the discharge. Methods used to detect product leaks along the pipeline can be divided into two categories; external and internal leak detection systems (API, 1995a). Externally based methods include traditional procedures such as visual inspection, as well as technologies like hydrocarbon sensing using fiber optic or dielectric cables. Internally based methods, also known as computational pipeline monitoring, use instruments to monitor internal pipeline parameters (i.e., flow, pressure, temperature, and fluid properties), which are continuously input into a computer simulation software linked to a supervisory control and data acquisition (SCADA) system that mathematically or statistically analyzes the information. Leaks result in unexpected variations or well-defined deviation pattern between simulated and measured values. These patterns can be detected and assessed to determine if a leak is present (API, 1995b). The method of leak detection selected for a pipeline is dependent on a variety of factors including physical pipeline characteristics (length, diameter, thickness, etc.), product characteristics (density, viscosity, etc.), instrumentation and communications capabilities, and economics (Muhlbauer, 1996). Compared to other leak detection methods, the SCADA-based leak detection methods have the widest range of applicability to pipeline leak detection and are by far the most highly developed of the leak detection methods presented as it can rapidly detect large leaks and, over a period of time, detect smaller leaks as well (Jolly et al., 1992). However, for multiphase flow pipelines, the SCADA-based leak detection methods become much more difficult to apply, and the sensitivity to detecting leaks is reduced because metering multiphase lines is very difficult and inaccurate. Most condition the flow so has to create a measurable flow stream by making a uniform mist or separates the gas and liquid content. The big problem in multiphase leak detections is the varying percentage of gas and liquid content and most notably, the formation and expulsion of liquid slugs in the pipeline. Hence, for multiphase SCADA-based leak detection systems, accurate and reliable flow measurement instrumentation and transient flow computer models will need to be developed. Moreover, the capabilities and advantages/disadvantages of new technologies
should be analyzed with special consideration given to the possible application in subsea pipelines, for which the remoteness of these pipelines, coupled with a number of complex interactions between the released fluids and the subsea environment makes detection much more difficult.

### 2.7.2 Pipeline depressurization

Pipeline depressurization is generally used to refer to the controlled and relatively slow evacuation of a pipeline system. Depressuring is usually performed for process reasons/maintenance requirements. When pipelines transport a gas and liquid mixture a rupture can cause rapid depressuring and critical flow at the break, additionally, with gas-condensate lines the temperature can drop rapidly due to Joule–Thompson cooling. This can have implications on the brittle fracture of the pipeline and safety valves which may cool significantly before being actuated.

The simulation of the depressuring of a pipeline system is performed in order to determine the minimum temperature experienced, or to calculate the time required to depressure a system such that the temperature does not drop below the minimum specification value.\(^2\) The transient analysis of pipeline depressuring can become very complex depending on what is required from the analysis and to what accuracy. OLGA commercial software can simulate the depressuring of multiphase pipelines and is recommended for most cases.

### 2.7.3 Pigging

Pigging is a term used to describe a mechanical method for removing contaminants and deposits within the pipe or to clean accumulated liquids in the lower portions of hilly terrain pipelines, using a mechanized plunger or pigs. Because of the ability of the pig to remove both corrosion products and sludge from the line, it has been found to be a positive factor in the corrosion control of the offshore pipelines. Some specially instrumented pigs, known as “intelligent pigs,” may also be used intermittently for the purpose of pipeline integrity monitoring that includes detecting wall defects (e.g., corrosion, weld defects, and cracks).

Pipeline pigs fit the inside diameter of the pipe and scrape the pipe walls as they are pushed along the pipeline by the flowing fluid. For offshore platforms the pig is launched from offshore and received in the onshore pig catcher. The receiver is in a direct line with the sealine and can be isolated from it to allow the pigs to be removed. Pigs are available in various shapes (Figure 2-8) and are made of different materials, depending on the pigging task to be accomplished. Some have spring-loaded steel knives, wire brushes, or abrasive grit surfaces for removal of adhering contaminants. Others are semirigid, nonmetallic spheres.

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\(^2\)The time taken to fully depressure a pipeline to atmospheric pressure will depend on several factors, not least of which will include, the size and type of pipeline inventory, the operating pressure at time of rupture, the rate of flow escaping, and the maximum vent rate at the end section.
During pigging operations there are often technical problems due to the lack of the reliable tools for the prediction of the many variables related to the motion of the pig through the pipeline. Hence, the pigging operation requires careful control and coordination. For example, overfrequent pigging causes production downtime or higher operation expenses and infrequent pigging results in less production reduction but increases the risk of pipeline blockages including sticking a pig. The pipeline operator must therefore give serious consideration to whether the pipeline really needs to be pigged and whether it is economical to do so. Since there is no commercially available tool to determine the optimum pigging frequency; operators must choose their pigging frequency using rules of thumb based on their field experience, which often involves a high degree of uncertainty.

In pigging operations where the pipe content is unloaded, the liquid holdup builds up as a slug ahead of the moving pig. The arrival of a slug at production or processing equipment is problematic. It causes both mechanical problems (due to high velocities and momentum) and process problems (increasing liquid levels, causing surges and trips). There are in general several means to reduce slugging in pipelines. In some cases operators can minimize liquid accumulation by managing fields and pipelines in such a way as to create a suitable fluid flow regime (i.e., mist flow regime) under which the gas velocity is high enough to keep liquids continuously dispersed. While it is desirable to design the lines to avoid slugging, in practice this can be difficult while maintaining the ability to turn down the pipeline flow rate. In these cases, consideration should be given to providing suitable process equipment to handle possible slugging (Xiao and Shoup, 1998).

The pigging operation in multiphase pipelines is a transient operation. Transient flow is observed not only during the pig running time, but also for a long period after the pig exits. This situation occurs even if the inlet gas and liquid flow rates and outlet pressures are kept constant (Minami, 1991). Analysis of such
Transient flow behavior in a pipeline is necessary not only for designing the downstream processing facilities, but also for establishing safe operating procedures. Hence, there is a definite need to develop reliable and comprehensive pigging models for better understanding of transient behavior of fluids during these operations. A simple model to simulate transient flow behavior in a two-phase flow pipeline under pigging operation has been presented by Minami (1991). In this model, Minami (1991) assumed that the gas phase can be considered to be flowing in a quasi-steady condition, and then coupled it with the Taitel et al. (1989) simplified transient TFM. The model, however, needs significant modifications in order to be used for simulating transient flow in a pipeline-riser system, where the quasi-steady-state approach is not suitable for such systems due to high accumulation of gas upstream of the pig (Yeung and Lima, 2002). For this purpose, a new transient TFM has been recently developed by Yeung and Lima (2002), which is appropriate for estimating the two-phase flow pigging hydraulics, especially in pipeline-riser systems.

### 2.8 Multiphase flow assurance

Flow assurance, which is a critical component in the design and operation of multiphase production facilities, refers to ensuring the flow of produced hydrocarbons from the reservoir to the downstream processing facilities. Flow assurance encompasses the thermal–hydraulic design and assessment of multiphase production/transportation systems as well as the prediction, prevention, and remediation of flow problems such as gas hydrate formation, wax and asphaltene deposition on walls, corrosion, erosion, scaling, emulsions, foaming, severeslugging, etc. In all cases, flow assurance designs must consider the capabilities and requirements for all parts of the system throughout the entire production life of the system to reach a successful solution for securing the production operations, minimizing the downtimes, and reducing the production/transportation costs.

This section addresses the major flow assurance issues for the raw gas transmission lines and summarizes the commonly used mitigation practices in industry for such purpose. It also provides innovative techniques and practical strategies to design the deepwater pipelines, where traditional approaches are inappropriate for deepwater development systems due to extreme distances, depths, temperature, or economic constraints (Wilkens, 2002).

#### 2.8.1 Gas hydrates

A gas hydrate is an icelike crystalline solid called a clathrate, which occurs when water molecules form a cage-like structure around small guest molecules at

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3Flow assurance is a term originally coined by Petrobras in the early 1990s. The term in Portuguese was “Garantia de Fluxo,” which translates literally to “Guarantee the Flow.”
certain temperatures and pressures. The most common guest molecules are methane, ethane, propane, isobutane, normal butane, nitrogen, carbon dioxide, and hydrogen sulfide, of which methane occurs most abundantly in natural hydrates. Several different hydrate structures are known. The two most common are structure I and structure II. Type I forms with smaller gas molecules such as methane, ethane, hydrogen sulfide, and carbon dioxide while structure II is a diamond lattice, formed by large molecules like propane and isobutane (Sloan, 2000). However, nitrogen, a relatively small molecule, also forms a Type II hydrate (Carroll, 2003). In addition, in the presence of free water, the temperature and pressure can also govern the type of hydrate structure, where the hydrate structure may change from structure II hydrate at low temperatures and pressures to structure I hydrate at high pressures and temperatures (GPSA, 2004). It should be noted that $n$-butane does form a hydrate, but is very unstable (Ng and Robinson, 1976; Kumar, 1987). However, it will form a stabilized hydrate in the presence of small “help” gases such as methane or nitrogen (Edmonds et al., 1998). It has been assumed that normal paraffin molecules larger than $n$-butane are nonhydrate formers (Kumar, 1987). Furthermore, the existence of another hydrate structure, Type H, has been described by Ripmeester et al. (1987). Some isoparaffins and cycloalkanes larger than pentane are known to form structure H hydrates (Mehta and Sloan, 1996). However, little is known about type H structures.

While many factors influence hydrate formation, the two major conditions that promote hydrate formations are (1) the gas being at the appropriate temperature and pressure, and (2) the gas being at or below its water dew point (Sloan, 2000). Note, free water is not necessary for hydrate formation, but it certainly enhances hydrate formation (Carroll, 2003). Other factors that affect, but are not necessary for hydrate formation include turbulence, nucleation sites, surface for crystal formation, agglomeration, and the salinity of the system. The exact temperature and pressure at which hydrates form, depends upon the composition of the gas and the water. For any particular composition of gas at a given pressure, there is a temperature below which hydrates will form and above which hydrates will not form. As the pressure increases, the hydrate formation temperature also increases. As a general rule, when the pressure of the gas stream increases or as the gas becomes colder, the tendency to form hydrates increases (GPSA, 2004). Hence, many gas-handling systems are at significant risk of forming hydrate plugs during shut-in and subsequent start-up (Wilkens, 2002).

Although gas hydrates may be of potential benefit both as an important source of hydrocarbon energy and as a means of storing and transmitting natural gas, they represent a severe operational problem as the hydrate crystals may deposit on the pipe wall and accumulate as large plugs that can completely block pipelines, shutting in production (Edmonds et al., 1998; Sloan, 2000). Acceleration of these plugs due to a pressure gradient can also cause considerable damage to production facilities. In addition, the remediation of hydrate blockages can
present significant technical difficulties with major cost implications (Mehta et al., 2001). Because of these problems, methods of preventing hydrate solids development in gas production/transportation systems have been of considerable interest for a number of years.

### 2.8.1.1 Hydrate locus for natural gas components

The thermodynamic stability of hydrates, with respect to temperature and pressure, may be represented by the hydrate curve. The hydrate curve represents the thermodynamic boundary between hydrate stability and dissociation. Conditions to the left of the curve represent situations in which hydrates are stable and “can” form. Operating under such conditions does not necessarily mean that hydrates will form, only that they are possible. Figure 2-9 shows the hydrate locus for natural gas components. The extension to mixtures is not obvious from this diagram. The hydrate curve for multicomponent gaseous mixtures may be generated by a series of laboratory experiments, or more commonly, is predicted using thermodynamic software based on the composition of the hydrocarbon and aqueous phases in the system.

The thermodynamic understanding of hydrates indicates the conditions of temperature, pressure, and composition that hydrates are stable. However, it does not indicate when hydrates will form and, more importantly, whether they will cause blockages in the system.

### 2.8.1.2 Prediction of hydrate formation conditions

There are numerous methods available for predicting hydrate formation conditions. Three popular methods for rapid estimation of hydrate formation conditions are discussed below. A detailed discussion of other methods that are, perhaps, beyond the scope of the present discussion, can be found in publications by Kumar (1987), Sloan (2000), and Carroll (2003).

**K-factor method**

This method has been originally developed by Carson and Katz (1942), although additional data and charts have been reproduced since then. In this method, the hydrate temperature can be predicted using vapor-solid (hydrate) equilibrium constants. The basic equation for this prediction is (Carson and Katz, 1942):

\[
\sum_{i=1}^{n} \left( \frac{y_i}{K_i} \right) = 1.0
\]  

(2-55)

where \( y_i \) is mole fraction of component \( i \) in gas on a water-free basis, \( K_i \) is vapor-solid equilibrium constant for component \( i \), and \( n \) is number of components.

The calculation is iterative and the incipient solid formation point will determine when the above equation is satisfied. This procedure is akin to a dew point calculation for multicomponent gas mixture.

The vapor-solid equilibrium constant is determined experimentally and is defined as the ratio of the mole fraction of the hydrocarbon component in gas on a...
water-free basis to the mole fraction of the hydrocarbon component in the solid on a water-free basis (Carson and Katz, 1942):

\[ K_i = \frac{y_i}{x_i} \]  

(2-56)

where \( x_i \) is mole fraction of component \( i \) in solid on a water-free basis.

Figures 2-10 to 2-14 provide the vapor-solid equilibrium constants at various temperatures and pressures. For nitrogen and components heavier than butane, the equilibrium constant is taken as infinity. It should be stressed that in the original method of Carson and Katz (1942), it is assumed that nitrogen is a
nonhydrate former and that \( n \)-butane if present in mole fractions less than 5%, has the same equilibrium constant as ethane. Theoretically, this assumption is not correct, but from a practical viewpoint, even using an equilibrium constant equal to infinity for both nitrogen and \( n \)-butane, provides acceptable engineering results (Campbell, 1992).

FIGURE 2-10
Vapor-solid equilibrium constants for methane (top) and ethane (bottom) (GPSA, 2004).
FIGURE 2-11
Vapor-solid equilibrium constant for propane (GPSA, 2004).
The Carson and Katz (1942) method gives reasonable results for sweet natural gases and has been proven to be appropriate up to about 1000 psia (GPSA, 2004). However, Mann et al. (1989) presented new K-value charts that cover a wide range of pressures and temperatures. These charts can be an alternate to the Carson and Katz (1942) K-value charts, which are not a function of structure or composition.

Poettmann et al. (1989) also developed vapor-solid equilibrium ratio charts for various natural gas systems using the Colorado School of Mines hydrate program. These charts cover a wide range of pressure, temperature, and composition. Polynomial regression analysis was used to curve fit the computer-generated K-values for various systems so that the resulting equations can be programmed and used to predict the hydrate-forming conditions for natural gases.

Baillie and Wichert method
The method presented by Baillie and Wichert (1987) is a chart method (Figure 2-15) that permits the estimation of the hydrate formation temperatures at

![Figure 2-12: Vapor-solid equilibrium constants for isobutane (GPSA, 2004).](image)
pressure in the range of 100–4000 psia for natural gas containing up to 50% hydrogen sulfide, and up to 10% propane (Carroll, 2003). The method may not apply to a sweet gas mixture containing CO₂, but is considered fairly accurate if the CO₂ is less than about 5 mol% (Carroll, 2004).

Gas gravity method

Until now, several other methods have been proposed for predicting hydrate-forming conditions in natural gas systems. The most reliable of these require a gas analysis. However, if the gas composition is not known, even the previous methods cannot be used to predict the hydrate formation conditions and the Katz (1945) gravity chart (Figure 2-16) can be used to predict the approximate pressure and temperature for hydrate formation, provided hydrates exist in the

**FIGURE 2-13**

Vapor-solid equilibrium constants for n-butane (GPSA, 2004).

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**CHAPTER 2 Raw Gas Transmission**
FIGURE 2-14
Vapor-solid equilibrium constants for carbon dioxide (top) and for hydrogen sulfide (bottom) (GPSA, 2004).
pressure–temperature region above the appropriate gravity curve. Therefore, as a first step to predict hydrate formation temperature one can develop an appropriate equation representing the Katz (1945) gravity chart. Such a correlation uses two coefficients that correlate temperature, pressure, and specific gravity of gas (Towler and Mokhatab, 2005):

\[ T_h = 13.47 \ln(P) + 34.27 \ln(SG) - 1.675[\ln(P) \times \ln(SG)] - 20.35 \] (2-57)

where \( T_h \) is gas flow temperature, °F; \( P \) is gas flow pressure, psia; and \( SG \) is specific gravity of gas (air = 1.0).

Note that while Eqn (2-57) is based on the GPSA chart but is only accurate up to 65 °F. Beyond that it overestimates the temperature slightly.
The Katz (1945) gravity chart was generated from a limited amount of experimental data and a more substantial amount of calculations based on the K-value method. The components used for the construction of this chart are methane, ethane, propane, butane, and normal pentane, and therefore using this chart for compositions other than those used to derive these curves will produce erroneous results (Sloan, 2000). In fact, this method is an appropriate method of estimating hydrate formation conditions for sweet natural gas mixtures. However, the Baillie and Wichert (1987) method is better than this chart when applied to sweet gas, because of the inclusion of a correction factor for propane (Carroll, 2003).

Commercial software programs
Hydrate formation conditions based on fluid compositions are normally predicted with the commercially available software programs. These programs are generally quite good, and so simple to use, they often require less time than the simplified methods presented. The bases of these computer programs are the statistical thermodynamic models, which use a predictive algorithm with additional experimental data included to modify or “tune” the mathematical predictions. Most commercially available softwares use algorithms developed by

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**FIGURE 2-16**
Katz’s gravity chart for predicting hydrate formation conditions (GPSA, 2004).
D.B. Robinson and Associates (EQUIPHASE), and by Infochem Computer Services Ltd (MULTIFLASH). A detailed discussion of the accuracy of these programs and other hydrate software programs can be found in publications by Sloan (2000) and Carroll (2003).

### 2.8.1.3 Hydrate prevention techniques

The multiphase fluid produced at the wellhead will normally be at a high pressure and a moderate temperature. As the fluid flows through the pipelines, it becomes colder which means such pipelines could experience hydrates at some point in their operating envelope. For this reason, the hydrate formation in gas transmission pipelines should be prevented effectively and economically to guarantee the pipelines operate normally. Control of hydrates relies upon keeping the system conditions out of the region in which hydrates are stable. It may be possible to keep the fluid warmer than the hydrate formation temperature (with the inclusion of a suitable margin for safety) or operate at a pressure less than the hydrate formation pressure.

Although there are several methods to avoid hydrate formation, depending on the possible location of a gas hydrate plug, some of the techniques to remediate onshore hydrates (e.g., installation of line heaters, and line depressurization) may not be practical in long and high pressure, subsea gas transmission pipelines (Wilkens, 2002).

There are few methods of preventing hydrate formation in offshore transmission systems. The permanent solution is removal of water prior to pipeline transportation, using a large offshore dehydration plant that is not often the most cost-effective solution. Pipeline system depressurization is also often used to melt the hydrate plug formed in a pipeline. However, the process is fairly slow and can take up to weeks or even months to completely melt a long hydrate plug. In general, at the wellsite two methods are applicable, namely thermal and chemical. These techniques are discussed in the following sections.

#### Thermal methods

Thermal methods use either the conservation or introduction of heat in order to maintain the flowing mixture outside the hydrate formation range. Heat conservation is common practice and is accomplished through insulation\(^4\). By using insulation, it is easy to keep fluid flowing temperature everywhere along the pipeline above the hydrate formation temperature. But no matter how much insulation is put on the pipeline, after a long pipeline shutdown, the fluid temperature will fall below the hydrate formation temperature and eventually cool down to the ambient temperature. Therefore, thermal insulation itself is not enough for hydrate mitigation for long pipeline shutdowns and other mitigation strategies will be needed (Guo et al., 2005).

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\(^4\)Burying the pipeline alone is not enough for thermal insulation and some extra insulation will be needed.
A number of different concepts are available for introducing additional heat to a pipeline. The simplest is an external hot-water jacket, either for a pipe-in-pipe system or for a bundle. Other methods use either conductive or inductive heat tracing. There is concern over the reliability of conductive systems. An electrical resistance heating system may be desirable for long offset systems, where available insulation is insufficient, or for shut-in conditions (Langner et al., 1999). The ability to heat during production depends on the specific electrical heating implementation. Such systems provide environmentally friendly fluid temperature control without flaring for pipeline depressurization. The effect is also an increase in production as there is no time lost by unnecessary depressurization, pigging, heating-medium circulation, or removal of hydrate blockage. But still it is difficult to persuade operators to install acting heating system.

Chemical inhibition

An alternative to the thermal processes is chemical inhibition. Chemical inhibitors are injected at the wellhead and prevent the hydrate formation by depressing the hydrate temperature below that of the pipeline operating temperature. Chemical injection systems for subsea lines have a rather high capital expenditure price tag associated with them, in addition to the often high operating cost of chemical treatment (Sloan, 2000). However, hydrate inhibition using chemical inhibitors is still the most widely used method for unprocessed gas streams, and the development of alternative, cost-effective and environmentally acceptable hydrate inhibitors is a technological challenge for the gas production industry.

Types of inhibitors. Traditionally, the most common chemical additives used to control hydrates in gas production systems have been methanol, ethylene glycol, or triethylene glycol at a high enough concentration (Sloan, 2000). These chemicals are called “thermodynamic inhibitors” and have the effect of shifting the hydrate curve to the left, which causes the hydrate stability point to be displaced to a lower temperature and/or a higher pressure. Figure 2-17 shows how the hydrate curve shifts with different amounts of methanol inhibition. Increasing salt content in the produced brine (by injecting electrolyte solutions such as sodium chloride, calcium chloride, and potassium chloride) can also provide some hydrate temperature suppression5 (see Figure 2-18), but normally this alone is not sufficient to avoid hydrates in the subsea gas production systems. In some cases, blended inhibitors of methanol/glycols and electrolyte(s) are preferred for subsea applications (Zuo and Zhang, 1999).

The inhibitor selection process often involves comparison of many factors including capital/operating cost, physical properties, safety, corrosion inhibition, gas dehydration capacity, etc. However, a primary factor in the selection process

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5Even though salt solution can be used for hydrate inhibition, too much salt can cause salt deposition and scale deposits in the system facilities. Salt solution is also corrosive and can cause corrosion problems to equipment.
is whether or not the spent chemical will be recovered, regenerated, and reinjected. Usually methanol is not regenerated because its use is intermittent (i.e., during start-up or shutdown). However, when it is injected continuously, as is often observed in gas system, then it is sometimes regenerated. But loses to the vapor phase can be prohibitive, in which case operators select monoethylene glycol (GPSA, 2004). Often when applying this inhibitor, there is a significant expense associated with the cost of “lost” methanol. However, since methanol is lower viscosity and lower surface tension it makes for an effective separation from the gas phase at cryogenic conditions (below \(-13 \, ^\circ\text{F}\)), and it is usually preferred (Esteban et al., 2000). In many cases hydrate plug formation is prevented through the addition of glycols (usually ethylene glycol because of its lower cost, lower viscosity and lower solubility in liquid hydrocarbons, and lower vapor pressure giving reduced gas-phase losses) to depress the hydrate
formation temperature. In order to be effective, glycols must be added at rates of up to 100% of the weight of water. Since glycols are expensive inhibitors, there is a definite need for extra, costly, and space consuming, onshore or offshore plants for their regeneration. Therefore, it would be useful to develop new hydrate depressants, which can be used at much lower concentrations, and therefore much lower cost. Recently, two new types of low-dosage inhibitors have been developed, which will enable the subsea gas transmission pipelines to handle increased gas volumes without additional glycol injection, or extra glycol recovery units (Bloys et al., 1995; Lederhos et al., 1996; Palermo et al., 2000). These new hydrate inhibitors can lead to very substantial cost savings, not only for the reduced cost of the new inhibitor but also in the size of the injection, pumping, and storage facilities (Mehta et al., 2002). These new hydrate inhibitors called “Low Dosage Hydrate Inhibitors” (LDHIs), which form the basis of a technique that does not operate by changing the thermodynamic conditions of the system. In fact, LDHIs act at the early stages of hydrate formation by modifying the rheological properties of the system (Sinquin et al., 2004). There are two types of LDHIs: the “kinetic hydrate inhibitors” (KHIs) and “anti-agglomerants” (AAs). Most commercial kinetic inhibitors are high molecular weight polymeric chemicals (i.e., poly [N-vinyl pyrrolidone] or poly [vinylmethylacetamide/vinylcaprolactam]), which are effective at concentrations typically 10–100 times less than thermodynamic inhibitors concentrations. KHIs may prevent crystal nucleation or growth during a sufficient delay compared to the residence time in the pipeline. The deeper a system operates in the hydrate region, the shorter the time during which KHIs can delay hydrate formation. The achievable delays range between weeks if the pipeline operates less than 42 °F in the hydrate region to hours if the pipeline operates at 50 °F in the hydrate region (Mokhatab et al., 2007b). Kinetic inhibitors are relatively insensitive to the hydrocarbon phase and may therefore turn out to be applicable to a wide range of hydrocarbon systems. However, the industrial application of kinetic inhibitors depends on the repeatability of multiphase pipeline testing results among laboratory, pilot plant and field, and the transferability among different plants (Kelland et al., 2000).

In contrast to other types of inhibitors, AAs, which are surface active chemicals (i.e., alkyl aromatic sulfonates or alkylphenylethoxylates) do not prevent the formation of hydrate crystals but keep the particles small and well dispersed so that fluid viscosity remains low, allowing the hydrates to be transported along with the produced fluids. AAs performance is relatively independent of time. In addition, AAs appear to be effective at more extreme conditions than KHIs, which makes these products of interest to operators looking for cost-effective hydrate control in deepwater fields. These additives are currently applied in the Gulf of Mexico, the North Sea and West Africa (Mehta et al., 2002; Frostman, 2003). However, they have mainly limitations in terms of water cut, where they require a continuous oil phase and therefore only applicable at lower water cuts. The maximum water cut is expected to be between 40% and 50%.
This limitation is caused by the rheological properties of suspensions with high solid fraction and may depend on flow regime conditions (Frostman, 2000; Mehta et al., 2002).

As stated earlier, the choice between inhibitor alternatives must be based on physical limitations as well as economics. However, operating conditions may also limit the number of available choices. For example, in a recent project carried out by Baker Petrolite it was shown that under severe conditions, the required dosage of an anti-agglomerator unlike thermodynamic and kinetic inhibitors does not increase as the degree of subcooling increases. Therefore, this method of treatment would be a cost-effective solution for the control of gas hydrates (Ramachandran et al., 2000).

**Prediction of inhibitor requirements.** The inhibitor must be present in a minimum concentration to avoid hydrate formation. Accurate prediction of this minimum inhibitor concentration is required for cost-effective design and operation of multiphase pipelines. Various empirical methods, charts, and computer programs have been developed for this purpose including the venerable Hammerschmidt’s empirical equation (1934), which is a relatively simple method that has been used to calculate the amount of inhibitor required in the water phase to lower the hydrate formation temperature:

\[
\Delta T = \frac{K \cdot W}{100(MW) - (MW)W}
\]

where \(\Delta T\) is depression of hydrate formation temperature, °F; MW is molecular weight of inhibitor; W is weight percent of inhibitor in final water phase; and K is constant, depending on the type of inhibitor.

Experimentally determined values of K and molecular weights of inhibitors are given in Table 2-6. To use this equation, the hydrate formation temperature in the gas without the inhibitor being present must be known. In fact, Eqn (2-58) only predicts the deviation from the hydrate formation temperature without an inhibitor present.

The Hammerschmidt equation is limited to inhibitor concentrations of about 20–25 wt% for methanol and 60–70 wt% for the glycols (GPSA, 2004). However,
for higher methanol concentrations, Nielsen and Bucklin (1983) have recommended the following equation:

$$\Delta T = -129.6 \ln(X_{H_2O})$$  \hspace{1cm} (2-59)

where $X_{H_2O}$ is the mole fraction of water in the aqueous phase.

They claim that this equation is accurate up to 90 wt% methanol, which gives the maximum suppression because methanol freezes at concentrations above 90 wt%. However, recent studies of GPSA (2004) only recommend the Nielsen and Bucklin (1983) equation for methanol concentrations ranging from 25 to 50 wt%. Equation (2-59) was developed for use with methanol, however, this equation is actually independent of the choice of inhibitor and therefore, theoretically, it can be used for any glycols.

Recently, Maddox et al. (1991) have described a graphical procedure for estimating the required inhibitor concentration for both methanol and glycol. This method is a trial and error approach, which can be used when the activity coefficients of water in methanol and glycol are available. Although this method provides better accuracy, no recommended applicable range is provided (Esteban et al., 2000).

If the produced water or seawater is in contact with the hydrocarbon fluid, the salinity of the water will itself inhibit hydrate formation. Therefore, it is important to be able to estimate the effect of the brine in the produced water on the hydrate formation temperature. For this purpose, McCain (1990) presented the following equation:

$$\Delta T = AS + BS^2 + CS^3$$  \hspace{1cm} (2-60)

where $\Delta T$ is temperature depression, °F; S is water salinity, wt%; and

$$A = 2.20919 - 10.5746 \left(\text{SG}\right) + 12.1601 \left(\text{SG}^2\right)$$  \hspace{1cm} (2-60-1)

$$B = -0.10605 + 0.72269 \left(\text{SG}\right) - 0.85093 \left(\text{SG}^2\right)$$  \hspace{1cm} (2-60-2)

$$C = 0.00347 - 0.01655 \left(\text{SG}\right) + 0.01976 \left(\text{SG}^2\right)$$  \hspace{1cm} (2-60-3)

Equation (2-60) is limited to salt concentrations of less than 20 wt% and for gas specific gravities ranging from 0.55 to 0.68.

All of these simple methods predict the depression of the hydrate formation temperature; they do not predict the actual hydrate formation conditions. However, several thermodynamic models have been proposed for predicting the hydrate formation conditions in aqueous solutions containing methanol/glycols and electrolytes (Edmonds et al., 1996; Nasrifar et al., 1998; Zuo and Zhang, 1999). These rigorous models can also account for the effect of pressure and the type of hydrate that are neglected in the simple mentioned methods. However, available models have limitations that include the types of liquid, compositions of fluids, and inhibitors used.

The amount of inhibitor required to treat the water phase, plus the amount of inhibitor lost to the vapor phase, and the amount that is a soluble in the
hydrocarbon liquid, equals the total amount required. Methanol vaporization loss can be estimated from Figure 2-19, while glycols vaporization losses are generally very small and typically can be ignored. More details on how to estimate the amount of methanol or glycols lost in vapor and condensate can be found by Sloan (2000).

In addition, prediction of inhibitor losses to the hydrocarbon liquid phase is difficult. However, many of the commercially available software programs include proper calculations to account for the loss of methanol and glycols to the hydrocarbon liquid phase.

**Design of injection systems.** Proper design of an inhibitor injection system is a complex task that involves optimum inhibitor selection, determination of the necessary injection rates, pump sizing, and pipeline diameters. Inhibitors for a subsea gas transmission system are selected before gas production is started on the facility. This makes inhibitor selection difficult as a large number of factors including brine composition; temperature, pressure, etc. that affect the performance of inhibitors are unknown (Ramachandran et al., 2000). Therefore, at this stage, an appropriate multiphase flow simulation package must be used to calculate some of unknown necessary variables, which are required for injection systems design.

To determine the appropriate injection rate, it is preferable to determine inhibitor requirements by field-testing. The inhibitor dosing requirements are
later used to determine the requirements for inhibitor storage, pumping capacities, and number of inhibitor lines in order to ensure that inhibitor can be delivered at the required rates for treating wells and flow lines during start-up and shut-in operations. In addition, points of injection should be chosen to provide maximum benefit in the pipeline system. The most effective position for injection is generally at the center of the pipe in the direction of the product flow. However, the injection rate and location of injection points will be a function of flow geometry, fluid properties, pressure–temperature relationships, etc., that will be encountered in the actual field application. For instance, hydrate inhibitors for a deepwater production system are often pumped through long umbilicals that are $\frac{1}{2}$–$\frac{3}{4}$ inch in diameter. The injection pump is a positive-displacement metering pump (Kumar, 1987) capable of generating sufficient injection line pressure (normally between 3000 and 4000 psi) to overcome the line operating pressure. Ideally the injection pressure should be 100 psi above line pressure. However, varied injection rates can be achieved by changing the pressure differential. Ramachandran et al. (2000) provide a discussion on proper design of deepwater injection systems and predicting systems behavior.

2.8.2 Corrosion

One of the common problems in the multiphase flow transmission pipelines is metal corrosion. Corrosion is defined as the deterioration of material, usually a metal, because of its reaction with the environment or handling media. This material degradation leads to the impairment of the intended function of the metal, environment, or the integrity of the system. This can be general corrosion (regular loss of metal on the exposed surface) or it can be localized corrosion where only a limited portion of the surface is in contact.

Corrosion can occur in different forms and can be caused by a variety of different reasons. The cause of corrosion can be directly attributed to the impurities found in the produced gas as well as the corrosive components which are by-produced. Because of the nonspecificity of the components produced from a production well, some or all of these components may be active to create a corrosive environment in the pipelines. Corrosion in multiphase systems is a complex phenomenon, including dependency on the partial pressure, temperature, flow regime/velocity, pH, and concentration of corrosive components. Consequently, corrosion prediction requires substantial understanding of the simultaneous interaction of many process variables that govern both the flow and corrosion conditions.

An important aspect of maintaining pipeline performance is adequate control of corrosion both internally (caused by the flow components and their

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6Corrosion can result in the loss of millions of dollars if a pipeline is not properly protected.
by-products) and externally (because of pipeline exposure to the soil and water) (Hartt and Chu, 2004). Pipeline corrosion can be inhibited by several means:

- choice of corrosion-resistant metals
- injection of corrosion inhibitors
- cathodic protection
- external and/or internal protective coatings.

While corrosion control can be achieved by selection of an appropriate corrosion-resistant metal, operating considerations usually dictate that a high-efficiency corrosion control system be used. Protecting pipelines from corrosion is achieved internally by injection of inhibitors to mitigate internal corrosion and externally by use of cathodic protection and/or combination of coatings and cathodic protection (for buried or subsea pipelines).

### 2.8.2.1 Choice of corrosion-resistant metals

Corrosion resistance is a basic property related to the ease with which materials react with a given environment. All metals have a tendency to return back to stable conditions. This tendency causes metals to be classified according to rising nobleness that again leads to classification of decreasing activity and increasing potential. When it is observed that the existing material of construction is prone to corrosive attack, generally a decision is made to change the materials of construction and select alternate material to suit the specific need. For specific recommendations concerning materials selection refer to the ANSI B31.3 and B31.8, API RP 14E (1991), and NACE (1975) MR-01-75. It should be noted that no single material can serve as a cure for all the corrosion evils. Therefore, a detailed study of process and operating conditions has to be carried out before selection of a new material.

Corrosion resistance is not the only property to be considered in the material selection process, and the final selection will generally be the result of several compromises between corrosion resistance and economic factors. Historically, carbon steel has been the most economical material for construction of long-distance, large diameter gas transmission pipelines. However, the inherent lack of corrosion resistance of these materials in subsea application requires a corrosion control system with a high degree of reliability (Hartt and Chu, 2004).

### 2.8.2.2 Corrosion inhibitors

Sometimes it is necessary to reduce the intensity of corrosive attack of the environment by adding certain chemicals known as “inhibitors” to reduce the aggressiveness of the media. Corrosion inhibition is necessary in gas-condensate

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7Corrosion resistant alloy (CRA) is often used to replace carbon steel for corrosive applications. But, CRAs are normally expensive than carbon steel and their usage depends upon overall economics.
pipelines when the required corrosion allowance for an uninhibited pipeline is excessive. If this is the case a corrosion allowance for the inhibited pipeline system can be calculated from the allowance for the uninhibited system and the inhibitor efficiency$^8$. If the required corrosion allowance is still excessive a higher grade material must be used for the pipeline.

Corrosion inhibitors are cationic surfactant chemicals such as filming amines, which, when added in a small concentration (between 10 and 1000 ppm), effectively reduce the corrosion rate of a metal$^9$. The injected inhibitor forms a protective layer on the internal surface of the pipe and the continuity and the quality of this layer are the controlling factor for effective corrosion protection. Corrosion inhibitors can be applied in batches resulting in a protective film, which may last for weeks or months. Enough inhibitor should be introduced to provide an appropriate wall coating thickness. Frequency of the treatment is governed by the remaining effectiveness of the inhibitor after a specified amount of commodity has been moved through the line. Inhibitors are often continuously injected in low concentrations where a thin film is laid down and maintained over time.

Corrosion inhibitors will only perform at their nominal efficiency if they remain in contact with the pipe wall at the correct concentration (Erickson et al., 1993). However, for some specific flow patterns in multiphase flows such as slug flow, inhibitor stripping can occur due to turbulence and high shear stress (Kang et al., 1996). But the slug can also help by distributing corrosion inhibitor and preventing top-of-line corrosion. As gas-condensate pipelines operate primarily in stratified flow regime, corrosion inhibitor is only deposited on the pipe’s top wall if there is a substantial fraction of the liquid entrained as droplets in the gas phase. However, at some locations, corrosion will still occur due to the condensation of water that dilutes the inhibitor concentration.

There are numerous corrosion inhibitor types and compositions. Corrosion inhibitor selection is a complex process that needs to balance a number of competing demands. The inhibitor must both reduce corrosion rates to an acceptable level and be compatible with system materials and other production chemicals such as hydrate inhibitors and biocides. The selected corrosion inhibitor must also not adversely affect hydrocarbon production processes by formation of foams or emulsions. Finally, the selected chemical must meet environmental regulations.

Although inhibitors can be used to great advantage to suppress the corrosion of metals in many environments, there are certain limitations of this type of corrosion prevention which should be recognized. First, it may not be possible to

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$^8$For smooth pipeline, the efficiency of corrosion inhibitors can be as high as 85–95%, but can drop if the shear stress increases drastically at locations such as fittings, valves, and bends (Guo et al., 2005).

$^9$In gas-condensate pipeline, adding the thermodynamic hydrate inhibitors can also reduce the corrosion rate. This is because, they absorb free water and make the water phase less corrosive.
add inhibitors to all corrosive systems because they may contaminate the environment. Further, inhibitors generally rapidly lose their effectiveness as the concentration and temperature of the environment increase. Finally, many inhibitors are toxic, and their application is limited. Therefore, new inhibitors are continuously being developed to handle more aggressive conditions and comply with more stringent environmental regulations.

### 2.8.2.3 Cathodic protection

Cathodic protection is the most successful method for reducing or eliminating corrosion for buried or submerged metallic structures that involves using electric voltage to prevent corrosion (Colson and Moriber, 1997). When two metals are electrically connected to each other in an electrolyte (e.g., seawater), electrons will flow from the more active metal (anode) to the other (cathode) due to the difference in the electrochemical potential. The anode supplies current, and it will gradually dissolve into ions in the electrolyte and at same time produce electrons that the cathode will receive through the metallic connection with the anode. The result is that the cathode will be negatively polarized and hence be protected against corrosion. The two methods of achieving cathodic protection are (1) the use of sacrificial or reactive anodes with a corrosion potential lower than the metal to be protected and (2) applying a direct current. The use of direct current system is less costly than sacrificial anodes, and provides a higher range of possible potential differences, although they may require greater maintenance during the lifetime of the operation (Uhlig and Revie, 1985).

### 2.8.2.4 Protective coatings

While cathodic protection has historically been employed as the sole corrosion control methodology for subsea gas production systems, the nature of multiphase pipelines is such that the combined use of protective coatings with cathodic protection is necessary to achieve the effective protection10 (Samant, 2003). In fact, protective coatings help control pipeline corrosion by providing a barrier against reactants such as oxygen and water. But, because all organic coatings are semipermeable to oxygen and water, coatings alone cannot prevent corrosion and so a combination of cathodic protection is often used (Varughese, 1993).

Pipelines are often internally coated to minimize corrosion when lying in pipeline dumps prior to construction, and to provide a smooth surface that reduces friction when fluids are in transit. It should be noted that the initial period in a production well’s life can be its most corrosive time due to the high partial pressure of the carbon dioxide. Therefore, it may be more economical to protect the multiphase pipelines from a young field. The coating must be compatible with the commodity and should have suitable resistance to attack by the commodity as well as any contaminants, corrosives, or inhibitor associated with it. Coatings

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10For some installations (e.g., deepwater), one might choose continuous inhibition over protective coating due to the implications of a coating failure.
such as epoxies, plastics, or metallic compounds can be used for selected applications. The oldest and still used approach consists of hot-applied bituminous material wrapped with an appropriate covering (Polignano, 1982). External coatings are used to reduce the value of the external current to an economic level by imposing a barrier of high electrical resistance between metal and its environment. Rhodes (1982) reports the use of many such coatings, including, polyurethane, phenolic resin, phenolic epoxy, and fusion-bonded epoxy powder.

2.8.2.5 Corrosion monitoring

Pipelines are susceptible to both internal and external corrosion. Therefore, both internal as well as external monitoring of pipeline is required for a complete assessment, thus providing the direction that will ensure proper utilization of materials and corrosion control methodologies. While traditional inspection and monitoring techniques (such as pigging and corrosion coupons) may represent effective solutions for assessment of the condition and integrity of a pipeline, the sensitivity and accuracy of these methods may be inadequate for monitoring inhibitor performance since they do not provide real-time information but rather focus on historical analysis. Therefore, traditional techniques of monitoring must be supplemented with online techniques that provide more timely real-time data. This allows operators to be more proactive in preventing corrosion-related pipeline failures.

2.8.3 Wax

Multiphase flow can be severely affected by deposition of organic solids, usually in the form of wax crystals, and their potential to disrupt production due to deposition in the production/transmission systems. The wax crystals reduce the effective cross-sectional area of the pipe and increase the pipeline roughness, which results in an increase in pressure drop. The deposits also cause subsurface and surface equipment plugging and malfunction, especially when oil mixtures are transported across Arctic regions or through cold oceans. Wax deposition leads to more frequent and risky pigging requirements in pipelines. If the wax deposits get too thick, they often reduce the capacity of the pipeline and cause the pigs to get stuck. Wax deposition in well tubings and process equipment may lead to more frequent shutdowns and operational problems.

2.8.3.1 Wax deposition

Precipitation of wax from petroleum fluids is considered to be a thermodynamic molecular saturation phenomenon. Paraffin wax molecules are initially dissolved in a chaotic molecular state in the fluid. At some thermodynamic state the fluid becomes saturated with the wax molecules, which then begin to precipitate. This thermodynamic state is called the onset of wax precipitation or solidification. It is analogous to the usual dew point or condensation phenomenon, except that in wax precipitation a solid is precipitating from a liquid, whereas in condensation a
liquid is precipitating from a vapor. In wax precipitation, resin and asphaltene micelles behave like heavy molecules. When their kinetic energy is sufficiently reduced due to cooling, they precipitate out of solution but they are not destroyed. If kinetic energy in the form of heat is supplied to the system, these micelles will desegregate and go back into stable suspension and Brownian motion.

**Wax deposition envelope**

Many reservoir fluids at some frequently encountered field conditions precipitate field waxes. It is very important to differentiate field waxes from paraffin waxes. Field waxes usually consist of a mixture of heavy hydrocarbons such as asphaltenes, resins, paraffins (or paraffin waxes), cycloparaffins, and heavy aromatics. Wax precipitation primarily depends on fluid temperature and composition and it is dominated by van der Waals or London dispersion type of molecular interactions. Pressure has a smaller effect on wax precipitation. As with asphaltenes, the fact that waxes precipitate at some and not at other thermodynamic states, for a given fluid, indicates that there is a portion of the thermodynamic space that is enclosed by some boundary within which waxes precipitate. This bounded thermodynamic space has been given the name wax deposition envelope, WDE (Leontaritis, 1995). A typical WDE is shown in Figure 2-20.

![Typical wax deposition envelope](image-url)

**FIGURE 2-20**

Typical wax deposition envelope (Leontaritis, 1995).
The upper WDE boundary can have either a positive slope or a negative slope. In most cases where the WDE has been obtained experimentally, the upper WDE boundary was very close to a vertical line.

The intersection of the WDE boundary with the bubble-point line is generally expected to be to the left of the onset of wax crystallization (cloud point) of the stock tank oil. This has been the case with many reservoir fluids whose WDE has been measured (Leontaritis, 1995). This is due to the fact that light ends, when pressured into oil, usually cause a suppression of the onset of wax crystallization temperature. The actual shape of the lower WDE boundary is primarily a function of the compositions of the intermediates and light ends of the reservoir fluid. The technology for measuring WDEs only recently has been perfected. Because the technology is new and the cost of measurements is relatively high, most companies do not find it economical to obtain the complete WDE of their reservoir fluid through laboratory measurements. Hence, obtaining only a few experimental data points and using them to fine-tune phase behavior models that are then used to calculate the remaining WDE more economically are preferred. It should be noted that most of these models have very little predictive capacity at this time and they are primarily used as combination correlational-predictive type of tools. One of the main reasons for the lack of pure predictive capability is improper or inadequate crude oil characterization (Leontaritis, 1997a).

Gas-condensate wax deposition envelope
Some gas condensates, especially rich gas condensates with yields in excess of 50 bbls/MMSCF, are known to contain high carbon number paraffins that sometimes crystallize and deposit in the production facilities. The obvious question is what is the shape of the thermodynamic envelope (i.e., P and T surface) of these gas condensates within which waxes crystallize? Or, in order to maintain the previous terminology, what is the WDE of gas condensates typically?

The shapes of the WDEs of two gas condensates in the Gulf of Mexico are presented here. The shapes of the above WDEs indicate potential wax deposition in those cases where the gas condensate contains very high carbon number paraffins that precipitate in solid state at reservoir temperature. In other words, the temperature of the reservoir may not be high enough to keep the precipitating waxes in liquid state. Hence, the gas condensate, which is a supercritical fluid, enters the WDE at the “dew point” pressure. This casts new insight into the conventional explanation that the productivity loss in gas condensate reservoirs, when the pressure near the wellbore reaches the dew point, is only due to relative permeability effects.

Figure 2-21 shows the vapor-liquid envelope (V-L envelope) of what one might call a typical Gulf of Mexico gas condensate. This gas condensate (called gas condensate “A” for our purposes here) was analyzed with PARA (paraffin-aromatic-resin-asphaltene) analysis (Leontaritis, 1997a) and found to contain normal paraffins with carbon numbers exceeding 45. The V-L envelope was simulated using the Peng and Robinson (1976) original EOS that had been
fine-tuned to PVT data obtained in a standard gas-condensate PVT study. The first question that was addressed in a wax study involving this fluid was what happens as the fluid is cooled at some constant supercritical pressure? What actually happened is shown in Figure 2-22.

Figure 2-22 shows several onset of wax crystallization data points obtained with the near-infrared (NIR) equipment (Leontaritis, 1997b) by cooling the gas condensate “A” at different constant pressures. It was evident from the NIR data that there was a thermodynamic envelope, similar to the one defined and obtained experimentally for oils, to the left of which (i.e., at lower temperatures) wax crystallization occurred. The complete wax deposition envelope shown in Figure 2-22 was calculated with a previously tuned wax phase behavior model (Narayanan et al., 1993). Despite the clarity of the WDE obtained for gas condensate “A” as shown in Figure 2-22, more data were needed to confirm the presence of WDE in other condensates and establish its existence as a standard thermodynamic diagram.

Figure 2-23 shows the V-L envelope of another typical Gulf of Mexico condensate. This condensate (called gas condensate “B” for our purposes here) also contains paraffins with carbon numbers exceeding 45, although the data show that gas condensate “B” is lighter than gas condensate “A.” The V-L envelope was again simulated using the Peng and Robinson (1976) original EOS after it had been tuned to PVT data obtained in a standard gas condensate PVT study.
Figure 2-24 shows the NIR onset data superimposed on the V-L envelope. It is evident again from the NIR data that there is a thermodynamic envelope to the left of which (i.e., at lower temperatures) wax crystallization occurs. Once again, the complete wax deposition envelope shown in Figure 2-24 was calculated with a previously tuned wax phase behavior model (Narayanan et al., 1993).

Data presented here confirm the presence of a WDE in gas condensates that contain high carbon number paraffin waxes (≥45). This WDE is similar to oil WDEs and as a result it should be considered a standard thermodynamic diagram. The shape of the WDE inside the V-L envelope seems to be consistent with existing information regarding the effect of light hydrocarbons on the onset of wax crystallization or wax appearance temperature. That is, as the pressure rises the WDE tilts to the left (negative slope) due to the ability of light hydrocarbons to depress wax crystallization. However, at the pressure where retrograde condensation begins the WDE turns forward thus acquiring a positive slope. This is because the light ends begin to vaporize and the waxes remaining in the liquid phase begin to concentrate. This is simply caused by the change in normal paraffin concentration which in turn is caused by retrograde concentration.
In most condensates the V-L envelope is fairly horizontal at the saturation line (dew point or bubble point). Hence, when this general pressure is reached the WDE seems to coincide with the V-L saturation line until the temperature becomes low enough for the waxes to begin crystallizing from the supercritical condensate. This is in agreement with prior observations that indicate a substantial increase in the solvent power of some fluids when they become supercritical (i.e., propane, CO$_2$, etc.). That is, supercritical hydrocarbon fluids are expected to require cooling to much lower temperatures before paraffin waxes begin to crystallize because of their increased solvent power.

2.8.3.2 Wax formation in multiphase gas-condensate pipelines

When the production pressure and temperature profile of a gas condensate crosses the WDE and the hydrate envelope (HE), waxes and hydrates may form. The formed hydrates can grow in mass and yield strength to the point that they restrict and finally stop the flow. If the waxes form while the liquid is in contact with the wall, some of them should attach to the wall thus resulting in wax deposition. The P–T diagram of a Gulf of Mexico gas condensate as shown in Figure 2-25 demonstrates the above situation.
The diagram indicates that there was not any live NIR done with this fluid. The only point on the WDE that was measured was the cloud point of the condensate, shown in the diagram at 95 °C at the bottom of the orange vertical line. The clean tubing and flow line of this system would deliver on 6/17/2002 about 14.5 MMSCFD. From the light blue P and T profile line and dark blue hydrate line, it was very clear at the outset that the main flow assurance issues in this system would be hydrate and wax formation and deposition. Hence, because of this illustrative important information, the effort to obtain the above diagram is obviously very worthwhile.

Sometimes oil and gas operators decide to design the facilities such that their operation is to the left of the WDE and HE (such as the example in Figure 2-25). In these situations, wax and hydrate formation takes place if the fluids are left untreated with chemicals. In the above example, methanol was injected to inhibit successfully the formation of hydrates formed due to production of reservoir equilibrium water. However, because the operator was not aware of the wax phase behavior of this condensate at start-up, no wax chemical was injected. Wax deposition was severe enough to cause the production rate to drop down to 1.8 MMSCFD on 3/14/2003. The estimated via simulation maximum production rate

![Wax deposition envelope, Gulf of Mexico gas condensate “B” (Leontaritis, 1998).](image)
on 3/14/2003 was 16.5 MMSCFD. The data in the plot show that the maximum friction loss was in the upper part of the tubing. Indeed, the yellow P and T line shows that the wax deposition and HEs were being crossed in the tubing at that time.

Identification of wax deposition problems
A rather simple chart that allows daily monitoring of wax deposition problems in multiphase gas-condensate pipelines is shown in Figure 2-26.

The above performance chart is for the gas condensate shown in Figure 2-25. The above chart was made by plotting the performance index (PI) versus time. The PI is calculated from the following equation:

\[ PI = \frac{\Delta P^2}{Q^{1/0.54}} \]  

where \( \Delta P \) is the pressure drop in the pipeline, psi; and \( Q \) is the flow rate, MMSCFD.
The PI should remain as a horizontal straight line during production if there is no restriction formed in the line. It is evident in Figure 2-26 that around 10/21/2002 a restriction was being formed in the line. At the beginning of January there was a very sharp loss of hydraulic capacity. This cannot be due to wax deposition because the rate of wax deposition does not change so sharply in a produced fluid. Indeed, the sharp rise in the PI was caused by hydrate formation that occurred on top of wax crystal formation and deposition. Several sharp rises in PI followed caused by hydrates until it was determined that the well started producing free water and required a much higher methanol injection rate. While all of this is happening with the advent of water and hydrate formation the wax deposition rate was increasing steadily, as shown on the chart.

It should be noted that the engineer needs to be very careful in attributing a rise in the PI only to wax and hydrates. One needs to make sure that other culprits such as fines, salts, returning drilling, and completion fluids, etc. are not present.

Wax deposition inhibition/prevention
It is not always clear to the engineer what would be the wax deposition mechanism and what type of chemical treatment would be the most effective in gas
condensate situations. In oil flow lines, wax deposition occurs by diffusion of wax molecules and crystals toward and attachment at the wall. Prevention or inhibition of wax deposition is mainly accomplished by injection of a special class of molecules that interact with paraffin molecules at temperatures above the cloud point and influence their crystallization process in a way that diminishes the attraction of the formed crystals toward the wall. The inhibited formed wax crystals are removed from the system by the shear forces caused by the flowing oil.

The above mechanism occurs when the wax crystal forms at the wall. This is the case with most liquid-filled bare ( uninsulated) lines. The two important requirements are fast heat transfer and fast diffusion of paraffin molecules toward the wall. This is not the case in many situations of flow lines carrying gas condensates. There is fast cooling but the majority of the gas cools down while in the main flow thus forming wax crystals with the main flow that would have a tendency to deposit or sediment by gravity in the liquid holdup. Also, the liquid that forms near the inlet of the bare line that has not reached its cloud point cools with a velocity that is almost standstill compared to the main flow. Slugging in this case would be beneficial in removing wax slush from the line. Additionally, these flow lines are essentially soaked in methanol to prevent hydrate formation. It just so happens that most ( if not all) chemicals with wax crystal modification properties, as discussed later, are incompatible with alcohols and glycols. Hence, the inhibited wax crystals accumulate in the flow line in the liquid hold-up thus adding to the wax accumulation. This accumulation forms what has been called “wax slush.” This wax slush is viscous but it moves and can flow given enough shear force.

The above theory has been documented several times from field data. In the year 2000 enough production and laboratory data and simulations were available to make the case for one field, but there was no direct evidence of wax slush from the field. During a shutdown caused by a hurricane, the flow line was depressurized at the tree, in an effort to dissolve any existing hydrates. At start-up, the well was started very fast to try to initiate movement and flow of any liquids ( via an induced slug) and the previously theorized existing “wax slush.” The fast flow did mobilize a phase that was loaded with wax. The wax slush flowed into the separators and plugged lines and equipment making them temporarily incapable of operation.

It is appropriate at this time to give a brief description of the main two chemical classes used to treat wax deposition. There are three types of wax crystals: plate crystals, needle crystals, and mal or amorphous crystals.

Paraffinic oils form plate or needle crystals. Asphaltenic oils form primarily mal or amorphous crystals. Asphaltenes act as nucleation sites for wax crystal growth into mal crystals.

Plate crystals look as their name implies like plates under the microscope. Needle crystals look like needles and mal crystals are amorphous and generally
like small round spheres. The interaction between crystals and pipe wall increases from mal to needle to plates. Thus, maintaining newly formed wax crystals small and round, i.e., like mal crystals, is desirable.

The behavior and properties, e.g., cloud point and pour point, of paraffin crystals precipitating from a hydrocarbon can be affected in three ways:

1. Crystal size modification: that is modification of the crystal from larger sizes to smaller sizes.
2. Nucleation inhibition: that is inhibition of the growth rate of the crystal and its ultimate size.
3. Crystal type or structure modification: that is modification of the crystal from one type to the other. For instance, modify a crystal from needle to mal type.

A wax crystal modifier works primarily to modify the crystal size. The plate crystals of an n-paraffin look much smaller under the microscope when precipitating from hydrocarbons inhibited with a wax crystal modifier. Smaller crystals have lower molecular weights and thus higher solubility in oil. Furthermore, smaller particles have smaller energy of interaction among themselves and the pipe wall. A crystal modifier interrupts the normal crystal growth of the n-paraffins by inserting itself in the crystal and thus stopping its growth.

It is noted that another name for a wax crystal modifier is pour point depressant or PPD. As the name implies, wax crystal modifiers are very effective at suppressing the pour points of crude oils because they suppress the wax crystal growth and thus minimize the strength of their interactions.

A wax dispersant may act to inhibit wax nucleation and change the type of wax crystals from plate or needle to mal or amorphous. Inhibited amorphous crystals are smaller and carried much easier by the hydrodynamic drag of the flowing fluid. The presence of asphaltenes and resins facilitates the effect of the dispersant. The dispersant interacts with the asphaltenes and resins and ties them up thus removing nucleation sites required by wax crystals to grow. A dispersant, when added to the oil at a temperature above the cloud point, has occasionally an additional benefit of suppressing the cloud point by interacting with and tying up the asphaltenes and resins that tend to come out first. A dispersant tends to usually disperse the wax particles at the water oil interface.

Both wax crystal modifiers and wax dispersants are useful chemicals that have the ability to diminish wax formation and deposition, although through different mechanisms. Wax dispersants are usually much smaller in molecular weight and size than wax crystal modifiers. Hence, their viscosity and flow properties in general are more favorable in cold applications. Also, some wax dispersants are soluble with alcohols and glycols thus making them compatible for simultaneous injection. The selection of the wax chemical should be made after careful consideration of the produced hydrocarbon and facilities.
Wax deposit remediation

On occasion, if a substantial amount of wax accumulates in the line, as evidenced by the PI chart such as the one shown in Figure 2-26, a temporary shutdown to do a chemical soak, with a potential modification to the chemical to give it more penetrating power at cold temperatures, and fast start-up might be necessary to cause the wax slush cough.

Pigging is an option, but only after very careful consideration of the system’s performance to understand the dynamics of the moving pig and continuous removal of the wax cuttings ahead of it. This is a very difficult job. Controlling the bypass of flow around the pig for such a purpose is difficult. Hence, many pigging operations end up in failure with stuck pigs. The pigging analysis and decision must be left to true experts. Starting a pigging program at the beginning of the life of the system has a better chance of success than at any other time. Even then, excellent monitoring of the system’s PI is a must. It is recommended that a short shutdown, chemical soak, and fast start-up be considered first, because it is the safest option.

Controlled production of wax deposits

From a technical standpoint, spending enough capital initially to design the facilities to operate outside the wax and hydrate forming conditions or to the right of the WDE and HE is obviously the best solution. However, very often in practice, controlling wax deposits during production is preferred because of the lower facilities cost. Economically marginal fields can only be produced under this scenario. Hence, in these cases, the following three options prevail:

1. pigging only
2. chemical injection only
3. combination of pigging and chemical injection.

While frequent pigging of the line clears up any wax deposits, it may be necessary to chemically inhibit wax deposition during times when pigging is unavailable. Pigging often is inadequate or uneconomical, unless used in conjunction with a chemical treatment program. This program is often performed into two stages: (1) removal of wax deposits in the production/transmission lines and (2) continuous chemical injection or periodic treatment (such as batch treatments, etc.) in order to ensure pipeline integrity.

Chemical injection is the safest of the three, if good technical support and testing is available. The approach is discussed as follows:

1. Inject a strong chemical dispersant/inhibitor down hole to keep the formed wax particles small and suspended in the flow line. The majority of them would be carried away with the gas-liquid flow.
2. The chemical dispersant/inhibitor must be compatible and soluble with methanol to prevent precipitation of the chemical itself in the line.
3. Monitor the hydraulics in the flow line and occasionally, if necessary, cause a “slug” or “cough” to cough up any accumulated wax.

4. On occasion, if a substantial amount of wax accumulates in the line, a temporary shutdown to do a chemical soak, with a modification to the chemical to give it more penetrating power at cold temperatures, and fast start-up might be necessary to cause the wax slush cough.

When properly implemented, the above approach should provide the lowest cost wax deposition control in gas-condensate transmission lines.

2.8.4 Slugging

Slugging, which refers to varying or irregular flows of gas and liquids (or liquid surges) in pipelines, is a major flow assurance challenge. Slug flow can pose serious problems to the designer and operator of two-phase flow systems. Large and fluctuating rates of gas and liquid can induce severe mechanical vibrations in the pipe, and severely reduce the production and in the worst case shut down or damage downstream equipment like separator vessels and compressors. As a result, prediction of slug characteristics is essential for the optimal, efficient, and safe and economically feasible design and operation of two-phase gas-liquid slug flow systems.

Slugging can take several forms (hydrodynamic slugging, terrain/riser-induced slugging and operationally induced slugging) that will be discussed here.

2.8.4.1 Hydrodynamic slugging

Hydrodynamic slugs are formed by growing of waves on the gas-liquid interface in the stratified flow under certain flowing conditions, where the amplitude of these waves becomes sufficiently large and they can bridge the pipe. Hydrodynamic (normal) slugs tend to be longer, the larger the pipe is and can, unfortunately, be thousands of feet long. Brill et al. (1981) noted from field tests at Prudhoe Bay that not all slugs are of the same length and that a log-normal distribution was observed.

When hydrodynamic slugs are numerous, the pipeline is said to be operating in the slug flow regime. Hydrodynamic slugs travel through a pipeline and can cause significant damage to, or operational problems for, downstream equipment. A usual design practice is to find the volume of the large slugs and design a separator or slug catcher able to handle this. There are two main methods that are typically used for slug size prediction: the simple correlation of field data and transient one-dimensional slug-tracking simulators, which use empirical correlations to compensate for the fact that their one-dimensional models cannot model these mechanisms directly. The simple correlations have been developed by correlating field slugging results to key parameters such as flow line diameter and fluid physical properties. However, these correlations ignore the important effect of line topography, and therefore give misleading results. Slug-tracking simulators give reasonable predictions when they will be tuned to field data.
However, their ability to make a priori predictions of slug flow (including slug size and frequency) remains limited. Therefore, good design practice still requires the application of healthy design margins (FEESA, 2003).

2.8.4.2 Terrain-induced slugging

In pipes across undulating terrain, a major cause of slugging is the topography. Liquid tends to build up and sit at the lowest points of the pipeline, until it is forced onward through the rest of the pipe by the pressure of the gas caught behind. Terrain slugging is most likely to occur at low flow rates, with a low pipeline pressure. When terrain-induced slugging is predicted, slugs will only form in upwardly inclined section of flow line. These slugs are unlikely to persist throughout the length of the line. Instead they will steadily decay and then collapse in horizontal or downwardly inclined sections (Bendiksen et al., 1986).

Terrain-induced slugging is highly undesirable due to the long-duration instabilities and the related oscillating momentum which can damage process equipment and necessitate major slug catchers. A mechanistic model or a transient program can provide a more realistic estimate of the terrain slug characteristics (McGuinness and McKibbin, 2002). During production, the length and velocity of a slug can be also estimated by monitoring the pressure variation at the outlet of the pipeline.

2.8.4.3 Riser-induced (severe) slugging

Terrain-induced slugging at a pipeline-riser system is denoted severe slugging. For this system, liquid will accumulate in the riser and the pipeline, blocking the flow passage for gas flow. This results in a compression and pressure buildup in the gas phase that will eventually push the liquid slug up the riser and a large liquid volume will be produced into the separator that might cause possible overflow and shut down of the separator. The severe slugging phenomenon is very undesirable due to pressure and flow rate fluctuations, resulting in unwanted flaring and reducing the operating capacity of the separation and compression units (Schmidt et al., 1980). Figures 2-27 and 2-28 show example time traces for the riser outlet liquid and gas flow rates during severe slugging, respectively. These figures show the large surges in liquid and gas flow rates accompanying the severe slugging phenomenon. Clearly such large transient variations could present difficulties for topsides facilities unless they are designed to accommodate them.

Usually, the operators try not to operate in the severe slugging region. But, the inlet conditions of a production pipeline are linked to the number and the capacity of the producing wells; the availability of wells and also to some undesirable operation such as shutdown or restart. The natural trend when dimensioning a production line is to do whatever is possible to avoid critical flooding of the separator, and therefore to overdimension the separator unit. But in offshore production, overdimensioning the installation is very costly and not always possible. So, the design engineers require more accurate dynamic simulations to
FIGURE 2-27
Example time trace of the riser outlet liquid flow rate during severe slugging (Mokhatab, 2007).

FIGURE 2-28
Example time trace of the riser outlet gas flow rate during severe slugging (Mokhatab, 2007).
correctly design and dimension their production schemes (which are more sen-
sitive to transients occurring when slug flow conditions build up, and require
high-performance control systems to maintain the plant within the preset oper-
ating ranges), and to be able to propose new concepts suitable to every situation
they can be faced to (Sagatun, 2004). The prediction of pipeline simulation
respect to slug dynamics will be strongly affected by the specification of boundary conditions. Hence, a tight integration of the hydrodynamic model of pipeline-riser system with the dynamic model of the receiving plant is very
important for achieving correct simulation results, especially for studying severe
slugging dynamics (Mokhatab and Towler, 2007a).

Given these potential problems of severe slugging in such systems, an
understanding of how, and when severe slugging will form, as well as providing
significant information on the best method to prevent and control this phenom-
enon is highly necessary undertaking.

Although the research community has undertaken a thorough exposition of
severe slugging phenomenon in flexible risers, however, there is a lack of suffi-
ciently wide-ranging and openly available transient code testing and full-scale
data on flexible risers for better understanding of the physics and characteris-
tics of severe slugging and carrying out more vigorous verification (Mokhatab
and Towler, 2007b). Unless such a vigorous link with reality is maintained,
designers and operators will not have the confidence to expand the use of flexible
risers in more critical applications.

Severe slugging mechanism
The process of severe slugging in a pipeline-riser system is considered to consist
of four steps: (1) slug formation, (2) slug production, (3) bubble penetration, and
(4) gas blow down. This phenomenon had been previously identified by Schmidt
et al. (1980) as a cyclic flow rate variation, resulting in periods both of no flow
and very high flow rates substantially greater than the time average. Figure 2-29
illustrates the stages of a severe slugging cycle.

The first step, slug formation, corresponds to an increase of the pressure in
bottom of the riser. The liquid level does not reach the top of the riser. During this
period, the liquid is no longer supported by the gas and begins to fall, resulting in
the riser entrance blockage and the pipeline pressure buildup until the liquid level
in the riser reaches to the top. During the second step, slug production, the liquid
level reaches the riser outlet, and the liquid slug begins to be produced until the
gas reaches the riser base. In third step, bubble penetration, gas is again supplied
to the riser, so the hydrostatic pressure decreases. As a result, the gas flow rate
increases. The fourth step corresponds to gas blow down. When the gas produced
at the riser bottom reaches the top, the pressure is minimal and the liquid is no
longer gas-lifted. The liquid level falls and a new cycle begins (Fabre et al.,
1990). This cyclic process becomes steady state when the rate of penetration of
the gas into the riser is always positive. However, it is also possible that the
penetration of the gas into the riser becomes zero. In this case, liquid blocks the bottom of the riser. This is followed by a movement of the liquid interface into the pipeline and blocking of the gas passage into the riser until the liquid interface reaches the bottom of the riser. At this point, penetration of gas into the riser starts and a new cycle begins again.

When liquid penetrates into the pipeline, the gas in the riser propagates to the top until all of the gas in the riser disappears. When the liquid input is very low, the propagation of the gas toward the top of the riser causes accumulation of all the gas at the top as the liquid falls back. This process is termed cyclic process with fallback, while the former case is termed cyclic process without fallback. In summary, three different possibilities that can occur as a result of penetration of gas into a liquid column in a quasi-steady severe slugging process are identified (Jansen et al., 1996):

1. Penetration of the gas that leads to oscillation, ending in a stable steady-state flow.
2. Penetration of the gas that leads to a cyclic operation without fallback of liquid.
3. Penetration of the gas that leads to a cyclic operation with fallback of liquid.

Severe slugging in a pipeline-riser system can be considered as a special case of flow in low-velocity hilly terrain pipelines, which are often encountered in offshore field. This is a simple case of only one downward inclined section (pipeline), one riser, and constant separator pressure. For this reason, severe slugging has been termed “terrain-induced slugging.” This phenomenon has also various names in the industry, including “riser-base slugging,” and “riser-induced slugging.”

FIGURE 2-29
Description of severe slugging in pipeline-riser systems (Fabre et al., 1990).
Stability analysis
The flow characteristics of multiphase flow in a pipeline-riser system are divided into two main regions, stable (steady flow) region, and unstable (pressure cycling) region, in which the stability line on the flow pattern map separates the two regions. Steady region includes acceptable slugging, annular-mist and bubble flows, while the pressure cycling region includes the severe slugging and transitional flows. Figure 2-30 shows a typical flow map for a pipeline-riser system developed by Griffith and Wallis (1961) featuring regions of stable and unstable behavior. In the above-mentioned figure, $N_{Fr}$ and $\lambda_G$ are the Froude number of two-phase gas-liquid flow and no-slip gas holdup, defined previously. As it can be seen from the figure, at low Froude numbers, bubble flow prevails and fluids will flow through riser pipes without slug formation. However, as Froude number increases the slug flow range is entered.

The stability analysis predicts the boundary between stable and unstable regions, where the resultant stability map helps the engineer to design systems that operate well into the stable zone thus offering an adequate margin of safety. The stability analysis seek to model a particular process required for severe slugging and hence predicts the likelihood of severe slugging, as such these stability models are termed criteria for severe slugging. The severe slugging process was first modeled by Schmidt et al. (1980); however, their model formed the basis of much of the early works for the stability of severe slugging. A review of the existing stability criteria for predicting severe slugging in a pipeline-riser system can be found in Mokhatab (2010).

Prevention and control of severe slugging
The flow and pressure oscillations due to severe slugging phenomenon have several undesirable effects on the downstream topside facilities unless they are designed to accommodate them. However, designing the topside facilities to

![Figure 2-30](image-url)

**FIGURE 2-30**
Griffith and Wallis flow pattern map with Yocum (1973) transition band (Brill and Beggs, 1991).
accept these transients may dictate large and expensive slug catchers with compression systems equipped with fast-responding control systems. This may not be cost-effective and it may be more prudent to design the system to operate in a stable manner (Sarica and Tengesdal, 2000). While lowering production rates (slowing fluid velocity) can minimize severe slugging, operators are investigating alternatives that would allow for maximum production rates without the interruptions caused by slugs (Furlow, 2000). Mokhatab et al. (2007a) reference a combination of industrial experience and information from the literature to compile a list of methods of remediating the problems associated with severe slugging in pipeline-riser systems. In general, the severe slugging prevention and elimination strategies seek three following approaches.

**Riser base gas injection.** This method provides artificial lift for the liquids, moving them steadily through the riser. This technique can alleviate the problem of severe slugging by changing the flow regime from slug flow to annular or dispersed flow, but does not help with transient slugging in which the liquid column is already formed before it reaches the riser base. It is one of the most frequently used methods for the current applications. However, for deepwater systems increased frictional pressure loss and Joule–Thomson cooling are potential problems resulting from high injection gas flow rates.

Riser base gas injection method was first used to control hydrodynamic slugging in vertical risers. However, Schmidt et al. (1980) dismissed it as not being economically feasible due to the cost of a compressor or pressurizing the gas for injection, and piping required to transport the gas to the base of the riser. Pots et al. (1985) investigated the application of the method to control severe slugging. They concluded that the severity of the cycle was considerably lower for riser injection of about 50% inlet gas flow. Hill (1990) described the riser base gas injection tests performed in the S.E. Forties field to eliminate severe slugging. The gas injection was shown to reduce the extent of the severe slugging.

The riser base gas lift method may cause additional problems due to Joule–Thomson cooling of the injected gas, where the lift gas will cause cooling and make the flow conditions more susceptible for the wax precipitation and hydrate formation. Hence, Johal et al. (1997) proposed an alternative technique “multiphase riser base lift” that requires nearby high capacity multiphase lines diverted to the pipeline-riser system that alleviate the severe slugging problem without exposing the system to other potential problems.

Sarica and Tengesdal (2000) proposed a new technique for sourcing riser base gas lift. The principle of the proposed technique is to connect the riser to the downward inclined segment of the pipeline with a small diameter conduit, where the conduit will transfer the gas from the downward inclined segment to the riser at points above the riser base (multiposition gas injection). The transfer process reduces both the hydrostatic head in the riser and the pressure in the pipeline consequently lessening or eliminating the severe slugging occurrence. This method can be considered as self-gas lifting. Sarica and Tengesdal (2000) claimed that the proposed technique is expected to increase production since it
does not impose additional backpressure to the production system. The cost of implementing and operation of the proposed systems in the field application is also expected to be low compared to other elimination methods. **Topside choking.** This method induces bubble flow or normal slug flow in the riser by increasing the effective backpressure at the riser outlet. While, a topsides choke can keep liquids from overwhelming the system, it cannot provide required control of the gas surges that might be difficult for the downstream system to manage. This is a low-cost slug-mitigation option, but its application might be associated with considerable production deferment.

Topside choking was one of the first methods proposed for the control of severe slugging phenomenon (Yocum, 1973). Yocum observed that increased backpressure could eliminate severe slugging but would severely reduce the flow capacity. Contrary to Yocum’s claim, Schmidt et al. (1980) noted that the severe slugging in a pipeline-riser system could be eliminated or minimized by choking at the riser top, causing little or no changes in flow rates and pipeline pressure. Taitel (1986) provided with a theoretical explanation for the success of choking to stabilize the flow as described by Schmidt et al. (1980).

Jansen et al. (1996) investigated different elimination methods such as gas lifting, choking, and gas lifting and choking combination. They proposed the stability, and the quasi-equilibrium models for the analysis of the above elimination methods. They experimentally made three observations: (1) large amounts of injected gas were needed to stabilize the flow with gas-lifting technique, (2) careful choking was needed to stabilize the flow with minimal backpressure increase, (3) gas-lifting and choking combination were the best elimination method reducing the amount of injected gas and the degree of choking to stabilize the flow. **Control methods.** Control methods (feed forward control, slug choking, active feedback control) for slug handling are characterized by the use of process and/or pipeline information to adjust available degrees of freedom (pipeline chokes, pressure, and levels) to reduce or eliminate the effect of slugs in the downstream separation and compression unit. Control-based strategies are designed based on simulations using rigorous multiphase simulators, process knowledge, and iterative procedures. To design efficient control systems, it is therefore advantageous to have an accurate model of the process (Bjune et al., 2002).

The feed-forwarded control aims to detect the buildup of slugs and, accordingly, prepares the separators to receive them, e.g., via feed-forwarded control to the separator level and pressure control loops. The aim of slug choking is to avoid overloading the process facilities with liquid or gas. This method makes use of a topside pipeline choke by reducing its opening in the presence of a slug, and thereby protecting the downstream equipment (Courbot, 1996). Like slug choking, active feedback control makes use of a topside choke. However, with dynamic feedback control, the approach is to solve the slug problem by stabilizing the multiphase flow. Using feedback control to prevent severe slugging has been proposed by Hedne and Linga (1990), and by other researchers (Molyneux...
et al., 2000; Havre and Dalsmo, 2001; Bjune et al., 2002). The use of feedback control to stabilize an unstable operating point has several advantages. Most importantly, one is able to operate with even, nonoscillatory flow at a pressure drop that would otherwise give severe slugging. Figure 2-31 shows a typical application of an active feedback control approach on a production flow line/pipeline system, and illustrates how the system uses pressure and temperature measurements (PT and TT) at the pipeline inlet and outlet to adjust the choke valve. If the pipeline flow measurements (FT) are also available, these can be used to adjust the nominal operating point and tuning parameters of the controller.

Note, the response times of large multiphase chokes are usually too long for such a system to be practical. The slug suppression system (S3) developed by Shell has avoided this problem by separating the fluids into a gas and liquid stream, controlling the liquid level in the separator by throttling the liquid stream and controlling the total volumetric flow rate by throttling the gas stream. Hence, the gas control valve backpressures the separator to suppress surges and as it is a gas choke, it is smaller and therefore more responsive than a multiphase choke.

The S3 is a small separator with dynamically controlled valves at the gas and liquid outlets, positioned between the pipeline outlet and the production separator. The outlet’s valves are regulated by the control system using signals calculated from locally measured parameters, including pressure and liquid level in the S3 vessel and gas and liquid flow rates. The objective is to maintain constant total volumetric outflow. The system is designed to suppress severe slugging and decelerate transient slugs so that associated fluids can be produced at controlled rates. In fact, implementation of the S3 results in a stabilized gas and liquid production approximating the ideal production system. Installing S3 is a

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**FIGURE 2-31**

Typical configuration of feedback control technique in flow line/riser systems (Bjune et al., 2002).
cost-effective modification and has lower capital costs than other slug catchers on production platforms. The slug suppression technology also has two advantages over other slug-mitigation solutions, where unlike a topside choke, the S3 does not cause production deferment and controls gas production, and the S3 controller uses locally measured variables as input variables and is independent of downstream facilities (Kovalev et al., 2003).

The design of stable pipeline-riser systems is particularly important in deepwater fields, since the propensity toward severe slugging is likely to be greater and the associated surges more pronounced at greater water depths. Therefore, system design and methodology used to control or eliminate severe slugging phenomenon become very crucial when considering the safety of the operation and the limited available space on the platform. Currently, there are three basic elimination methods that have been already proposed. However, the applicability of current elimination methods to deepwater systems is very much in question. Anticipating this problem, different techniques should be developed to be suitable for different types of problems and production systems (Mokhatab et al., 2007a).

2.8.4.4 Operationally induced slugging

These types of slugging are initiated by pipeline operations, which may result in large production losses. Pigging of a pipeline causes most of the liquid inventory to be pushed from the line as a liquid slug ahead of the pig. When a pipeline is shut down, the liquid will drain to the low points in the line, and when the pipeline is restarted, the accumulated liquid can exit the pipeline as a slug. Slugs may also form during depressurization due to high gas velocities. In addition to these mechanisms, slugs may be produced as a result of flow rate changes. Increasing or decreasing the flow rate of either gas or liquid leads to a change in liquid holdup. This can come out in the form of a slug, depending on the flow rate.

A transient multiphase flow program must be used to estimate the characteristics of these types of slugs. The volume of these transient slugs must be determined to ensure that they can be handled by the downstream separation and processing facilities. If there is insufficient capacity, operating procedures can be introduced to limit the size of any produced slugs.

2.8.5 Flow assurance risk management

Flow assurance risk management for a new pipeline project is usually performed during the different project engineering and design phases. The better that these flow assurance risks are defined during the project phases, the fewer operational problems that are likely to be encountered. Similarly, the sooner that the flow assurance risks are identified, the earlier their impacts on the project economics and the design of the pipeline system can be established.

Six phases exist for consideration of the flow assurance risk management for pipelines: assessing the risks, defining the mitigation strategies, defining flow
operability, finalizing the operating procedures, optimizing system performance, and real-time monitoring. A brief description of each of these phases of flow assurance management is given below (Song, 2008).

**Phase I: Assessing Flow Assurance Risks.** In the first Phase I flow assurance risks are assessed from the sample analysis of the reservoir fluid, which is one of the most critical phases in flow assurance management. No matter how accurate the laboratory measurements and interpretations are, if the fluids do not represent the real production fluids, erroneous conclusions may be drawn.

From the collected samples, producers then consider which tests are critical to enabling the goal of properly evaluating the flow characteristics of the fluid and designing a pipeline system. Any inaccurate fluid properties that are reported are likely to lead to inappropriate flow assurance mitigation strategies that will not work properly and also lead to the pipeline system risks encountering severe operational impacts.

Water sampling to determine composition extent is also very critical in establishing flow assurance risks and may be very challenging to perform. Analysis of hydrate risk, corrosion prediction, and scaling tendency are dependent upon an assessment of the salinity and composition of the water produced with oil or gas. However, no water sample may be available because either the exploration wells never reach the aquifer zones or the water samples are contaminated by drilling muds (Guo et al., 2005).

Also many oil fields start production with little associated water initially present, but water content gradually increases over the production life. In such cases, the flow assurance specialist must often be content with estimates of salinity and compositions based upon analogs, such as samples taken from nearby fields or adjacent reservoir zones that have already been put into production. Consequently, all these potential water sampling issues may result in high uncertainty in the development of the flow assurance mitigation strategies (Wilkens, 2002).

**Phase II: Defining Flow Assurance Mitigation Strategies.** How all the flow assurance risks will be mitigated is studied in this Phase II and a high level mitigation strategy is then developed. The flow assurance mitigation strategy is best developed with a good understanding of pipeline operating issues. The mitigation strategy details are also very closely related to the configuration of the pipeline system. The system configuration will dictate how many different ways the flow assurance must be mitigated and, similarly, critical flow assurance mitigation requirements will also drive some specific configurations being developed. For example, different hydrate mitigation strategies must be developed if the system configuration consists of single flow line or dual flow lines. On the other hand, if chemical inhibitors will be required to mitigate the hydrate risks, then a separate, additional service line or a flow line could be required in the pipeline system to deliver the chemicals.

**Phase III: Defining Flow Operability.** Phase III involves defining flow operability. Operability is the set of design provisions and operating strategies that
ensure the pipeline system can be started, operated, and shut down under all possible operating conditions (planned and unplanned) throughout the system life cycle. Operability is intrinsic to the flow assurance/pipeline system design process. In fact, after defining the high-level flow assurance mitigation plans in phase II, good operating procedures (applying dynamic multiphase flow modeling of the system) will be defined that enable the pipeline system to perform with minimal flow assurance risks under a specified range of operating conditions.

**Phase IV: Finalizing the Pipeline Operating Procedures.** During this next Phase IV, the pipeline system components (e.g., flow line connectors, insulation system, flow line joints, and valves) are changed from the original design for various reasons, such as different vendors, different materials, and/or different properties, etc. Due to these changes, the corresponding flow assurance mitigation techniques and operating procedures may need to be modified to reflect the changes. Also, even if the components are not modified; the actual manufactured ones may have different thermal–hydraulic properties from the designed ones based upon performance tests conducted on selected components. At the end of this phase, the flow assurance mitigation procedures and pipeline operating procedures will be finalized and used for system commissioning, start-up, and daily operations.

**Phase V: Optimizing System Performance.** In this Phase V, the procedures will be modified based upon the actual recorded performance data from the pipeline system. Analysis of such data may identify some requirements that can be beneficially adjusted to optimize the system performance.

**Phase VI: Real-Time Flow Assurance Monitoring.** The last Phase VI describes real-time flow assurance monitoring. In most cases flow assurance problems cannot be completely eliminated due to unpredictable system component failures, unsuitability for operating conditions, faulty operational procedures for some situations, or operator failures/human errors that can occur in real-time operations. Significant efforts are therefore necessary to minimize the occurrence and impact of such failures. In this regard, real-time flow assurance monitoring systems can provide optimum asset management. Continuously monitoring the operating conditions of the pipeline for anomalous readings may indicate possible restrictions leading to potential blockage (i.e., observation of erratic pressure fluctuations is usually indicative of hydrate formation or wax deposition).

In fact, reliable, real-time data provided quickly and continuously throughout the life cycle of the pipeline can be tied into a software simulation program which can enable the data to be analyzed and modeled in a timely fashion. When a process interruption or operational upset occurs, the software can predict where problems are most likely to occur (some can recommend the best corrective action sequence). This feature allows faster detection and diagnosis of problems decreasing the potential additional cost of multiphase production, and substantially reducing the risk of environmental disasters due to the failure of unmonitored pipelines.
Continuous monitoring systems can provide wide bandwidth capability via fiber-optic-distributed sensors. Real-time data can pass at optimal transmission rates for analysis, ensuring that the data is moved reliably and appropriately referenced at each interval along the pipeline (Brower et al., 2005). However, for deepwater installations, the need for accurate and advanced distributed sensor systems is more critical. Such systems present greater opportunities for the occurrence of flow assurance problems with potentially significant adverse consequences.

References

API, 1995a. Computational Pipeline Monitoring. API Publication, 1130, 17, American Petroleum Institute, TX.
API, 1995b. Evaluation Methodology for Software Based Leak Detection Systems. API Publication, 1155, 93, American Petroleum Institute (API), TX.
Arirachakaran, S.J., 1983. An Experimental Study of Two-Phase Oil-Water Flow in Horizontal Pipes (MSc thesis). University of Tulsa, Tulsa, OK, USA.


Frostman, L.M., October 1–4, 2000. Anti-aggolomerant hydrate inhibitors for prevention of hydrate plugs in deepwater systems. Paper Presented at the SPE Annual Technical Conference and Exhibition, Dallas, TX, USA.
CHAPTER 2 Raw Gas Transmission


Langner, et al., 1999. Direct impedance heating of deepwater flowlines, OTC 11037. Paper Presented at the Offshore Technology Conference (OTC), Houston, TX, USA.


Leontaritis, K.J., February 18–21, 1997a. PARA-Based (Paraffin-Aromatic-Resin-Asphaltene) reservoir oil characterization. SPE 37252, Paper Presented at the SPE International Symposium on Oilfield Chemistry, Houston, TX, USA.


Leontaritis, K.J., May 4–7, 1998. Wax deposition envelope of gas condensates. OTC 8776, Paper Presented at the Offshore Technology Conference (OTC), Houston, TX, USA.


CHAPTER 2 Raw Gas Transmission


Sarica, C., Tengesdal, J.Q., October 1–4, 2000. A new technique to eliminate severe slugging in pipeline/riser systems. SPE 63185, Paper Presented at the 75th SPE Annual Technical Conference & Exhibition, Dallas, TX, USA.


3.1 Introduction

Raw natural gas from production wells contains a full range of hydrocarbons, carbon dioxide, hydrogen sulfide, nitrogen, water, and other impurities. They can originate from any one of three types of production wells: crude oil wells, gas wells, and condensate wells. Natural gas that comes from crude oil wells is the associated gas. Gas can exist as a gas cap over the crude oil reservoirs, or can be dissolved by the crude oil under pressure. Natural gas from gas wells or from condensate wells, in which there is little or no crude oil, is the nonassociated gas. Gas wells generally produce sour natural gas, while condensate wells produce natural gas condensate or gasoline that is saturated with natural gas.

Raw natural gas can also come from unconventional sources. Natural gas can be found in coal seam, as the coal-bed methane gas. It can be the associated gas from shale oil production. Natural gas from offshore wells is another source. The unconventional sources typically contain a high level of CO₂ and nitrogen, which can be found in different parts of the world. These gas reservoirs are believed to be much larger than previous prediction. As technology in exploration and drilling advances, the unconventional reserves are becoming economically attractive and will become an important source of energy supply.

The objective of a natural gas processing plant is to produce a methane-rich gas by removing the acid gases, heavy hydrocarbons, nitrogen, water, and other impurities. This chapter gives an overview of the design and function of the different process units within a natural gas processing plant.

3.2 Natural gas processing objectives

Raw natural gas stream must be treated to comply with emissions regulations and pipeline gas specifications. Typical pipeline gas specifications are shown in Table 3-1. The specifications are to ensure gas qualities and provide a clean and safe fuel gas to the consumers. The product gas must meet the heating values or Wobbe Indexes specifications, which are required to ensure optimum operation of gas turbines and combustion equipment to minimize emissions. Pipeline operators also require the product gas to be interchangeable and similar in properties with existing pipeline gas.
When the gas is high in heavy hydrocarbon contents, they must be removed to meet the heating value specification. The removed liquid can typically command a higher value than natural gas for the same heating value. Propane and butane can be sold as liquefied petroleum gas (LPG). Ethane can be used as feedstock to petrochemical plants. The C$_5^{+}$ components can be exported to refineries as a blending stock for gasoline. The characteristics of various types of natural gas liquid (NGL) products can be found in GPSA Engineering Data Book (2004).

It is important to recognize the definitions and specifications of the treated gas and product specifications. The product specifications of condensate and NGL are described in the later chapters. Sometimes a slight change on the product specifications may have significant impacts on the treatment and processing options, which will affect the cost and complexity of the gas plant.

### Table 3-1 Typical Pipeline Gas Specifications

<table>
<thead>
<tr>
<th>Characteristic</th>
<th>Specification</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water content</td>
<td>4–7 lbm H$_2$O/MMscf of gas</td>
</tr>
<tr>
<td>Hydrogen sulfide content</td>
<td>0.25–1.0 grain/100 scf</td>
</tr>
<tr>
<td>Gross heating value</td>
<td>950–1200 Btu/scf</td>
</tr>
<tr>
<td>Hydrocarbon dewpoint</td>
<td>14–40 °F at specified pressure</td>
</tr>
<tr>
<td>Mercaptans content</td>
<td>0.25–1.0 grain/100 scf</td>
</tr>
<tr>
<td>Total sulfur content</td>
<td>0.5–20 grain/100 scf</td>
</tr>
<tr>
<td>Carbon dioxide content</td>
<td>2–4 mol%</td>
</tr>
<tr>
<td>Oxygen content</td>
<td>0.01 mol% (max)</td>
</tr>
<tr>
<td>Nitrogen content</td>
<td>4–5 mol%</td>
</tr>
<tr>
<td>Total inerts content (N$_2$ + CO$_2$)</td>
<td>4–5 mol%</td>
</tr>
<tr>
<td>Sand, dust, gums, and free liquid</td>
<td>None</td>
</tr>
<tr>
<td>Typical delivery temperature</td>
<td>Ambient</td>
</tr>
<tr>
<td>Typical delivery pressure</td>
<td>400–1200 psig</td>
</tr>
</tbody>
</table>

3.3 Gas processing plant configurations

The gas processing plant configuration and complexity depend upon the feed gas compositions and the levels of treating and processing required in meeting product specifications and emission limits. Liquid values can also be the drivers for process complexity, which determines the levels of NGL components to be recovered. Figure 3-1 shows two simplified gas processing plant schematics. The first scheme is to remove condensate, sulfur, and the heavier components to meet sales gas specifications. The second scheme is to process the feed gas for recovery of the NGL components to increase plant revenues. The residue gas is typically recompressed to a sales gas pipeline. It can also be sent to a natural gas.
liquefaction plant for liquefied natural gas (LNG) production, used as fuel gas to power plants, or as a feedstock to petrochemical plants.

The gas processing plant must be a “fit-for-purpose” design, meeting the project economics and environmental requirements. While contaminants and sulfur must be removed to meet emissions requirements as shown in the first scheme, the extent of processing in the second schemes is project specific. It depends on the commercial agreements between upstream producers and downstream product distributors and buyers. More details on the contractual terms are discussed later in this chapter.

3.3.1 Gas plant with hydrocarbon dewpointing

Raw gas to a gas processing plant can be relatively lean, that is, containing a small amount of C2+ hydrocarbons. This lean gas can be processed by the process units as shown in Figure 3-2. The main process units consist of acid gas removal, gas dehydration, and hydrocarbon dewpoint control. There are other off-site support systems such as the sulfur recovery, and tail gas treating and sulfur production, which are necessary to meet environmental requirements. If the gas contains liquid condensate, a condensate stabilization unit is required. If the gas contains high levels of nitrogen, nitrogen rejection is required. Other units such as gas compression unit may also be required.

3.3.1.1 Inlet facility

Feed gas from production wells arriving at the inlet facility is first separated in the inlet facility, which typically includes slug catchers, separation equipment, and pressure protection system. Typically a highly reliability safety system is installed at the inlet to protect the gas plant from a block discharged emergency condition.

The slug catcher is designed with adequate surge volume to hold the maximum liquid slug during the piping pigging operation and to provide to a steady flow to the stabilization unit.
The slug catcher is typically located at a safe distance from the gas plant, with adequate separation distance for safety reasons. The design is either a “vessel type” or a “finger type.” A vessel-type slug catcher consists of a phase separator for vapor, liquid, and water and a storage vessel to contain the liquid slug during pigging. The design is suitable to handle feed gas with high gas to liquid ratio and a steady flow. It is used where there is a space limitation, such as in offshore platform installation. The finger-type slug catcher is common in land-based plants where there are no space constraints. It is less costly than the vessel type, especially in high-pressure (HP) services. It consists of multiple long pieces of pipes, which provide the surge volume.

Produced gas from the slug catcher is directed to a HP separator, which protects the gas plant from upstream system upset. More details about the phase separation facilities are presented in Chapter 4.

The water content in the liquids from the slug catcher is separated and sent to a sour water stripper unit. The hydrocarbon liquids are processed in a condensate stabilization unit to reduce its vapor pressure for storage and transport. The produced water typically containing monoethylene glycol (MEG) used for hydrate control can be reclaimed in a MEG reclamation unit. Other contaminants, such as corrosion inhibitors and salts content, must be removed before MEG can be reused in the upstream unit.
3.3.1.2 Condensate stabilization
The condensate contains dissolved light hydrocarbons and H₂S, which must be removed to meet the export condensate specifications. A condensate stabilization unit is designed to produce a condensate with 4 ppm H₂S and RVP (Reid vapor pressure) specifications, ranging from 8 to 12 psi. The stabilizer overhead vapor is compressed and recycled to the HP separator. There are different design options in configuring the stabilization unit, which are discussed in Chapter 5.

The H₂S content in the condensate from the stabilization unit can typically meet 10H₂S ppmv specifications. However, it may contain higher levels of organic sulfurs such as COS and mercaptans. If condensate is exported as a product, the total sulfur content must be met, and a separate unit for removal of the mercaptans content may be necessary. If the condensate is sent to refineries, the condensate can be blended with the refinery feedstock and treated in the refinery units.

3.3.1.3 Acid gas removal unit
The acid gas removal unit is designed to remove the acidic components to meet sales gas sulfur and CO₂ specifications. H₂S must be removed to meet the sales gas specification of 4 ppmv, or ¼ grains per 100 scf of gas. In addition, COS, mercaptans, and other organic sulfur species must be removed. Considering today’s stringent emission regulations, acid gas removal unit alone may not be sufficient to meet the requirements. Treated gas from the acid gas removal unit may need to be further treated with additional units, such as molecular sieves or sulfur scavengers. Design details of the acid gas removal units are discussed in Chapter 6.

3.3.1.4 Sulfur recovery and handling unit
Acid gas from the amine regenerator contains concentrated H₂S, which cannot be vented for safety reasons or flared due to acid gas pollution. If reinjection wells are available, acid gas can be reinjected back to the reservoirs for sequestration, which would avoid the investment of a sulfur recovery unit. In most instances, where reinjection facility is not available, H₂S is processed in a sulfur recovery unit. The sulfur recovery unit can be coupled with a tail gas treating unit to achieve 99.9% sulfur removal to meet today’s emission target. The designs of the sulfur recovery units are discussed in Chapter 9.

3.3.1.5 Gas dehydration unit
Treated gas from the acid gas removal is fed to the gas dehydration unit to meet the water dewpoint specification for pipeline transmission, typically 7 lbs water/MMscf. In colder climate areas, the water dewpoint specification can be as low as −40 °F in order to avoid hydrate formation in the pipeline. Depending on the plant capacity and extent of drying, different types of dehydration methods are available, including glycol dehydration and solid desiccant (i.e., molecular sieves, silica gels, activated alumina) dehydration. Designs of gas dehydration system are discussed in Chapter 7.
3.3.1.6 Hydrocarbon dewpointing
The hydrocarbon dewpoint temperature must be reduced to a lower temperature that is below the coldest ambient temperature during transmission. This is to avoid hydrocarbon liquid condensation in the pipeline, which is a safety hazard. Depending on the phase envelop of the pipeline gas, the hydrocarbon dewpoint can actually increase when the pressure is lowered, which must be considered in the design of the unit. Details of the hydrocarbon dewpoint processes are discussed in Chapter 8.

3.3.1.7 Nitrogen rejection
Nitrogen content in natural gas varies depending on the gas reservoirs. Nitrogen can be naturally occurring in high concentration in some gas fields, such as in the South China Sea where 30–50% nitrogen content gas can be found. For onshore facilities where nitrogen injection is employed for enhanced oil recovery, nitrogen content can also be very high.

When nitrogen is present in high concentrations, it should be removed downstream of the NGL recovery unit. Heavy hydrocarbons (BTEX aromatics) must be removed in the NGL recovery unit, as these hydrocarbons will solidify in the cryogenic section of the nitrogen rejection unit. If the gas is used for LNG production, the nitrogen content must be removed to below 1 mol% to meet LNG specification. Nitrogen is not desirable in LNG product, as it will lower the liquefaction temperature, which would increase the power consumption.

Nitrogen removal by cryogenic separation is more efficient than other alternatives. Membrane separators and molecular sieves can be used for nitrogen rejection, but their processing capacity is relatively limited. They are suitable for bulk separation, and are not economical to meet stringent specifications. The rejected nitrogen would contain a significant amount of hydrocarbons, which may be an environmental issue. More details on the nitrogen rejection designs are discussed in Chapter 10.

3.3.1.8 Gas compression and transmission
Feed gas to the gas plant is typically reduced in pressure such that phase separation is feasible. Most often, recompression of the residual gas to the pipeline pressure is necessary. Design details of the natural gas compression and transmission systems are discussed in Chapters 11 and 12, respectively.

3.3.2 Gas plant for NGL production
When the feed gas contains a significant amount of liquids, the C$_3$+ hydrocarbons, there are economic incentives to produce the LPG and sometimes ethane liquid as byproducts. The liquid facility typically includes storage, pipeline, metering, and custody transfer, and must include safety system to protect against liquid leakage or spillage. This type of plant is complex and costs more than the simple hydrocarbon dewpoint plant.
Figure 3-3 illustrates a block flow diagram of a gas plant for NGL production. The following sections describe the units that are unique to NGL production. The balance of plant is similar to the hydrocarbon dewpointing plant.

3.3.2.1 CO₂ removal
CO₂ removal is required to meet the sales gas CO₂ specification, typically limited to 2–3 mol%. CO₂ may need to be removed to even a lower level to avoid CO₂ freezing in the cold section of the NGL recovery unit. Typically, in propane recovery process, 2 mol% of CO₂, can be tolerated as the NGL column operates at a warmer temperature. Deep CO₂ removal may not be required, unless the gas is sent to a liquefaction unit, which, in this case, would require CO₂ to be removed down to 50 ppmv.

If the NGL recovery unit is used for ethane recovery, the demethanizer column would operate at a much lower temperature, which is prone to CO₂ freeze. Even CO₂ freezing may not be a problem, a good portion of the CO₂ will condense with the ethane product, which may not meet the CO₂ specification in the ethane product (typical Y-Grade NGL limits CO₂ content in ethane to 500 ppmv). When ethane recovery is required, design must ensure that CO₂ in the feed gas to be removed sufficiently to avoid CO₂ freezing as well as meeting the CO₂ specification of the ethane product.
3.3.2.2 Dehydration
For NGL recovery, the deethanizer or demethanizer must operate at low temperatures. This would require sufficient water to be removed to avoid hydrate formation in the columns. If only propane recovery is considered, the column operates at a warmer temperature, at about $-60 \, ^\circ F$. In this case, the use of DRIZO triethylene glycol (TEG) dehydration may be sufficient. TEG unit is more compact than molecular sieve unit and is more suitable for offshore design. If ethane recovery is required, then molecular sieve dehydration is necessary. More details of the dehydration system designs are discussed in Chapter 7.

3.3.2.3 Mercury removal
Mercury is invariably present in the feed gas in ppm levels. Mercury removal using carbon beds or catalytic beds is required to avoid the risks of mercury attack on the brazed aluminum heat exchangers. Brazed aluminum exchangers are commonly used in NGL recovery process to achieve high NGL recovery. Aluminum material is very reactive with mercury and can be corroded quickly resulting in failure of the exchangers.

The mercury removal bed is typically designed with mercury removal to below 0.01 $\mu g/Nm^3$. The mercury removal step can be positioned upstream of the acid gas removal unit or downstream of the dehydration unit. Installing the mercury removal unit upstream can prevent mercury migration to downstream units, ensuring both the sales gas and the liquid products are mercury free. More details of the mercury removal designs are discussed in Chapter 7.

3.3.2.4 NGL recovery
There are numerous NGL patents and proprietary technologies for NGL recovery. Unlike acid gas removal unit, the NGL recovery process selection must be evaluated based on meeting the NGL recovery levels, feed gas pressure, temperature, gas compositions, and product specifications, as well as NGL recovery flexibility. The NGL process must be designed to handle a rich gas case and a lean gas case. If nitrogen injection is considered in the same formation, the higher nitrogen feed gas case must be considered in the design.

Typically, the design of the front section of the NGL recovery unit is controlled by the rich gas case operating in the summer, as the refrigeration duty is higher. However, the lean gas and the high nitrogen gas would control the design of the demethanizer and the residue gas compressor as the gas flow is higher.

The NGL recovery unit can be designed for propane recovery or ethane recovery. For operating flexibility, the NGL process can be designed for ethane recovery that can be operated on ethane rejection when ethane margins are low. Another alternative is to design the unit for propane recovery that can be operated on ethane recovery. More details of the various NGL recovery unit designs are discussed in Chapter 8.
NGL product compositions depend on the desirable propane and ethane recovery. NGL product pump and NGL pipeline must be designed for the ethane recovery operation as the NGL flow rate and the pipeline pressure are higher. NGL is typically transported to an NGL fractionation center, which would produce the final liquid products. The NGL products may need additional treatment and processing to meet the product specifications. Details on NGL fractionation and products treating are discussed in Chapter 8.

### 3.4 Finding the best gas processing route

There are various technologies, conventional or proprietary, for configuring the gas processing plant. The best process must be selected based on meeting the project requirements, economic targets, and environmental requirements. The design must be “fit-for-purpose” meeting the immediate requirements but must consider modification of the process for future gas compositions and NGL recovery requirements. More importantly, the designs must be reviewed by the plant operators and have their buy-in (Mokhatab and Meyer, 2009; Mokhatab, 2010).

The design of the front section of the gas plant, such as slug catcher and condensate stabilization unit, is conventional. Designs of the gas treating unit and sulfur recovery/handling system are more involved, and are dependent on the feed gas’ acid gas contents. Design of the NGL recovery unit is dependent on the levels of NGL recovery, energy efficiency, and operation flexibility. In the past, most NGL recovery units are designed for very high propane recovery (typically, 99% propane), as NGL is valued significantly higher than natural gas. High NGL recovery unit requires the use of turboexpander, feed gas, and residue gas compressors, and propane refrigeration, which is very costly compared to the simple hydrocarbon dewpointing plants. However, with the shale gas, remote gas and distributed gas playing a vital role in the global energy economy, there is no shortage of NGL; the economic driver for maximizing NGL production is becoming less attractive.

Today’s NGL design trend tends to focus on minimizing capital investment. Ethane demand is driven by incremental demands from new petrochemical plants. To minimize capital investment, prefabricated modular equipment is used to reduce field construction time and lower capital and operating costs. These modular units can be prefabricated as packaged units in various capacities, typically 60–100 MMscfd, and can be as high as 200 MMscfd. This modular design when properly executed can generate significant value for gas producers and processors.

In summary, there is no single best approach for the gas processing plant. NGL design continues to evolve around the economics of NGL markets. NGL designs that are optimized today may not work well in the future when feed gas compositions changes. An optimized gas plant design must be flexible and be suitable for revamp to meet future requirements and project economics, while preserving most of the key equipment.
3.5 Support systems

3.5.1 Utility and off-site

Operating the gas plants requires support from the utility and off-site systems. These typically include power generation, steam generation, heat medium supply, cooling water, instrument and plant air, nitrogen inert gas, fuel gas system, relief and flare, potable water, and wastewater treater system.

Small gas plants typically purchase electrical power off the local power grid. Larger gas plants or plants in remote areas would generate their own power. For uninterrupted supply, power can be supplied from the power grid as a backup or using an emergency power generator. A diesel generation is typically required for plant black start. A battery power supply is required for life support during emergency.

Cogeneration plants are energy efficient and can be attractive options for reduction of operating costs, especially where the gas turbine exhaust can be used for heating or steam generation. Most large compressors in gas plants use gas engines or gas turbine drivers, which can be integrated with the steam system. Steam is used in gas plants for solvent regeneration and reboilers in NGL recovery and fractionation units. These units require cooling; and where available, cooling water is the most economical. In desert areas where water supply is not available, air coolers are the only choice of cooling.

A demineralizer unit is required to supply boiler feed water makeup to steam generators and makeup to amine treating units. Wastewater must be treated in a wastewater treater before it can be discharged from the facility.

A reliable source of instrument air is critical to the gas plant operation. Typically, instrument air is supplied at about 100–120 psig, and must be dried to meet $-40^\circ\text{F}$ water dewpoint. The air must be clean and dried to avoid water condensation and corrosion in the instrument air piping. One or multiple backup air compressors are required to ensure plant safety and reliability. Plant air can also be supplied from the same air compressors. Air receivers are sized to dampen out fluctuation in demand, and to provide uninterrupted supply of instrument air.

Nitrogen is required for plant purging during start-up. It is also used for seal gas and for blanketing of storage tanks. Nitrogen consumption is typically very low in gas plant operation. Nitrogen can be produced by membrane separator or purchased from outside suppliers. In small gas plants, nitrogen can be supplied by nitrogen bottles.

A HP fuel gas and low-pressure fuel gas are required to meet the fuel gas demands by steam boilers, hot oil furnaces, and gas turbine drivers. Fuel gas must be treated and conditioned to meet the sulfur and heating value specifications.

3.5.2 Process control

Process control has always played a role in gas plants but has become more important over the years as companies try to reduce labor costs. Most plants use
distributed control system (DCS) for individual units to provide both process control and operation history. Advanced process control (APC) systems, which are integrated to the DCS systems, provide sophisticated plant control for plant optimization. APC uses multivariable algorithms that are programmed to perform online optimization. Another commonly used process control is supervisory control and data acquisition (SCADA) system. One important use of SCADA is the monitoring of field operations, with the capability of controlling the process unit equipment, flow valves, and compressor stations from the gas plant. Automation requires accurate input data to make the proper control decisions. These temperature, pressure, flow sensors, and gas analyzers require constant maintenance and tuning to ensure accurate data are transmitted to the computer system.

3.5.3 Safety systems

A relief and flare system must be provided to ensure plant safety. Multiple flares are typically installed to allow one flare to be maintained. The flare must be designed to protect against feed gas block discharge, as well as relief loads from power failure and cooling water failure and other failure modes. Proper sizing of relief valves, rupture disks is critical to ensure that process unit is protected during emergency. During normal operations, the gas plant is designed to avoid venting and flaring. If the plant is flaring mostly methane, the flame is bright but smokeless. If heavier hydrocarbons are flared, the flame will smoke and is noticeable. Smoke would create environmental problems. Steam can be injected to the flare system to maintain a smokeless flare. If the fuel has a low Btu content, fuel gas can be added to ensure complete combustion.

3.6 Contractual agreements

The gas processing industry performs the processing steps required to make the product gas marketable. These functions may include gas gathering, gas compression, and pipeline transmission. The gas qualities must be conditioned to meet the sales gas specification. In addition to gas purification functions, the gas processors may elect to add discretionary activities to recover NGLs, as liquid revenue can sometimes generate higher revenues.

When new ventures are being evaluated, there are contractual agreements, which must be used to establish the legal framework that defines the gas plant economics. This is a balance between capital expenditures and operating costs. The type of contract is dependent on the risks and potential rewards. There is no straightforward answer to the type of agreements, as different companies may take a different approach, pending on their willingness to take risks. Most facilities may have a combination of agreements to hedge the risks.

There are four major types of agreements, which are discussed below.
3.6.1 **Keep-whole contracts**

In the keep-whole contracts, the gas processor agrees to process or condition the producer’s gas to meet sales gas specifications and to return to the producer 100% of the Btu value of the raw gas in exchange for retaining ownership of liquids extracted from the gas. The processor usually retains the liability for all processing costs and energy costs. The gas price is usually indexed on a spot gas price or an average of spot gas prices.

One of the variations to the keep-whole agreement is to have a Btu cap or crude cap clause. The gas price is capped at a maximum price (Btu cap) or a maximum price relative to the price of crude oil (crude cap).

These contracts are more complex, more favorable to the producer, and more risky to the processor. The producer, in essence, sells the whole hydrocarbon stream, at the price of natural gas on a Btu basis, to the processor. The processor makes or loses money, depending on the price difference between natural gas and the NGL, which the processor sells.

Most contracts contain penalties for variations from contracted liquid content, impurities, and delivery pressure. Contracts may be set to allow for incremental variations from base composition. Contracts are commonly a combination of two or more of the above basic types.

3.6.2 **Flat fees contracts**

In fee-based contracts, the producer pays the gas processor a flat fee based on the gas produced. The processor may charge fees for additional services, such as gas gathering, gas compression, and pipeline transmission. In these contracts, the processors income is independent of fluctuation of gas and NGL prices.

The gas processor takes custody of the gas and then bears all of the gas processing risk. The processor has the benefits of liquid upgrade, along with the risk on volatility, but has to carry all operating costs, such as fuel, power, and transmission. Some producers/processors would include offset terms to mitigate the risks of keep-whole contracts or flat-fee contracts.

3.6.3 **Percentage of proceeds contracts**

In percentage of proceeds contracts, the two parties agree to the percentage of the proceeds from the sale of the gas and liquids. Typically, the producer retains more than 70% of the proceeds from the sale of all products. In the case of multiple producers, each has a percentage share of the proceeds, allocated on the contribution to the proceeds.

Allocations can be based on the Btu content of the gas delivered at the wellhead for a producer divided by the sum of the Btu content of the gas from all producers. Both the producers and processors share the effect of gas and NGL price fluctuations, the cost of gathering and production, and the sales price of the gas and liquids production.
3.6.4 Processing fee contracts

This is a risk-free agreement from a processor’s standpoint. The producer pays a fee based on the gas volume processed, which is generally a flat fee based on the types of gas conditioning and processing requirements. If only gas gathering and compression are required, the fee will be lower than that required to sweeten, drying, and NGL recoveries.

Depending on the processing fees, the liquids may be retained by the producer. In most contracts, the processor will take title to a portion of the liquids. This arrangement is common in small isolated fields where there is limited competition.

In summary, the processing costs are shared between producer and processor for capital projects, and the terms depend on the nature of the project and the potential risks. Capital items that benefit both parties may be cost shared. However, maintenance, replacements, and costs of environment and emissions compliance are borne by the processor as a cost of staying in business. Situations arise in which costs are too high for the processor to absorb alone, and producers must decide to either share the costs or cease production.

References


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4.1 Introduction

Separation of oil and gas is a critical field processing operation. As producing pressure is increased and lighter condensates are produced, efficient separation has become more critical than ever. Moreover, some of the new concepts in separation technology have been applied to advantage on old lease producing oil at moderate or low pressures. As gas transmission lines raise their standards, separation becomes a part of the overall field processing necessary to condition the gas. Several technologies are available to remove liquids and solids from gases. However, selecting gas–liquid separation technologies requires not only knowledge of the process conditions, but also knowledge of the characteristics of the liquid contaminants. Selection should be made based on the droplet size, concentration, and whether the liquid has waxing or fouling tendencies (Brown et al., 1994). Before evaluating specific technologies, it is important to understand the mechanisms used to remove liquids and solids from gases. Three principles used to achieve physical separation of gases and liquids or solids are momentum, gravity settling, and coalescing. Any separator may employ one or more of these principles; however, the fluid phases must be immiscible and have different densities for separation to occur. Momentum force is utilized by changing the direction of flow and is usually employed for bulk separation of the fluid phases. The gravitational force is utilized by reducing velocity so the liquid droplets can settle out in the space provided. Gravity segregation is the main force that accomplishes the separation, which means the heaviest fluid settles to the bottom and the lightest fluid rises to the top. However, very small droplets such as mist cannot be separated practically by gravity. These droplets can be coalesced to form larger droplets that will settle by gravity.

The purpose of this chapter is to review the principles governing the basic separation process and to describe the commonly used separation facilities in the gas processing industry.

4.2 Gravity separators

Gravity separators are pressure vessels that separate a mixed-phase stream into gas and liquid phases that are relatively free of each other. In a gravity separator, gravitational forces control separation, and the efficiency of the gas–liquid separation is increased by lowering the gas velocity. Because of the large vessel size required to achieve settling, gravity separators are rarely designed to remove droplets smaller than 250 μm (Talavera, 1990). However, an analysis of this type
of separator is given because it is useful to help understanding the settling mechanism of other separators.

Gravity separators are often classified by their geometrical configuration (vertical, horizontal) and by their function (two-phase–three-phase separator). In other words, gravity separators are classified as “two-phase” if they separate gas from the total liquid stream and “three-phase” if they also separate the liquid stream into its crude oil- and water-rich phases. Additionally, separators can be categorized according to their operating pressure. Low-pressure units handle pressures of 10–180 psi. Medium-pressure separators operate from 230 to 700 psi. High-pressure units handle pressures of 975–1500 psi.

Separators are sometimes called “scrubbers” when the ratio of gas rate to liquid rate is very high. These vessels usually have a small liquid collection section and are recommended only for the following items:

- Secondary separation to remove carryover fluids from process equipment such as absorbers and liquid dust scrubbers.
- Gas line separation downstream from a separator and where flow lines are not long.
- Miscellaneous separation where the gas–liquid ratio is extremely high.

In any case, the processing equipment should have the same configuration and are sized in accordance with the same procedure of separators.

4.2.1 General description

All gravity separators normally have the following components or features (API Spec 12J, 1989):

- A primary gas–liquid separation section with an inlet diverter to remove the bulk of the liquid from the gas.
- A gravity settling section providing adequate retention time so that proper settling may take place.
- A mist extractor at the gas outlet to capture entrained droplets or those too small to settle by gravity.
- Proper pressure and liquid-level controls.

Gravity separators are designed as either horizontal or vertical pressure vessels. Figure 4-1 is a typical scheme of a three-phase horizontal separator. The fluid enters the separator and hits an inlet diverter. This sudden change in momentum generates the initial bulk separation of liquid and gas. In most designs, the inlet diverter contains a downcomer that directs the liquid flow below the oil–water interface. This forces the inlet mixture of oil and water to mix with the water continuous phase in the bottom of the vessel and rise through the oil–water interface. This process is called “water-washing” and it promotes the coalescence of water droplets that are entrained in the oil continuous phase. The inlet diverter
assures that little gas is carried with the liquid, and the water wash assures that
the liquid does not fall on top of the gas–oil or oil–water interface, mixing the
liquid retained in the vessel and making control of the oil–water interface difficult.
The liquid collecting section of the vessel provides sufficient time so that the oil
and emulsion form a layer or “oil pad” at the top. The free water settles to the
bottom. The produced water flows from a nozzle in the vessel located upstream of
the oil weir. An interface level controller senses the height of the oil–water
interface. The controller sends a signal to the water dump valve thus allowing the
correct amount of water to leave the vessel so that the oil–water interface is
maintained at the designed height.

The gas flows horizontally and out through a mist extractor (normally known
as a demisting device) to a pressure control valve that maintains constant vessel
pressure. The level of the gas–oil interface can vary from half the diameter to 75%
of the diameter depending on the relative importance of liquid–gas separation,
and what purpose the separator has. For example, degassers and produced water
flash drums have very high liquid–gas interfaces. However, the most common
configuration is half full.

Figure 4-2 shows a typical configuration for a vertical three-phase separator.
In the vertical separator, the flow enters the vessel through the side as in the
horizontal separator and the inlet diverter separates the bulk of the gas. The gas
moves upward, usually passing through a mist extractor to remove suspended
mist, and then the dry gas flows out. A downcomer is required to transmit the
liquid collected through the oil–gas interface so as to not disturb the oil skimming
action taking place. A chimney is needed to equalize gas pressure between the
lower section and the gas section. The spreader or downcomer outlet is located at
the oil–water interface. From this point, as the oil rises any free water trapped within the oil phase separates out. The water droplets flow countercurrent to the oil. Similarly, the water flows downward and oil droplets trapped in the water phase tend to rise countercurrent to the water flow. It should be clear that the principles of operation (such as oil–water interface level controlling) of three-phase vertical separators are the same as three-phase horizontal separators those described above. Essentially, the only difference is that horizontal separators have separation acting tangentially to flow, whereas vertical separators have separation acting parallel to flow. In the vertical separator, level control is not also critical, where the liquid level can fluctuate several inches without affecting operating efficiency (GPSA, 2004). However, it can affect the pressure drop for downcomer pipe (from the demister) and can therefore reduce demisting device drainage.

4.2.2 Separator selection

There are no simple rules for separator selection. Sometimes, both configurations should be evaluated to decide which is more economical. The relative merits and common applications of vertical and horizontal separators are summarized by Manning and Thompson (1995).
4.2.2.1 Horizontal separators
Horizontal separators are most commonly used in the following conditions:

- Large volumes of gases and/or liquids
- High-to-medium gas–oil ratio (GOR) streams
- Foaming crudes
- Three-phase separation.

The advantages and disadvantages of these separators are given below.

Advantages

- Require smaller diameter for similar gas capacity as compared to vertical vessels.
- No counterflow (gas flow does not oppose drainage of mist extractor).
- Large liquid surface area for foam dispersion generally reduces turbulence.
- Larger surge volume capacity.

Disadvantages

- Only part of shell available for passage of gas.
- Occupies more space unless “stack” mounted.
- Liquid-level control is more critical.
- More difficult to clean produced sand, mud, wax, paraffin, etc.

4.2.2.2 Vertical separators
These separators are used in the following conditions:

- Small flow rates of gases and/or liquids.
- Very high GOR streams or when the total gas volumes are low.
- Plot space is limited.
- Ease of level control is desired.

The advantages and disadvantages of these separators are given below.

Advantages

- Liquid-level control is not so critical,
- Have good bottom-drain and clean-out facilities,
- Can handle more sand, mud, paraffin, wax without plugging,
- Lower tendency for re-entrainment,
- Has full diameter for gas flow at top and oil flow at bottom,
- Occupies smaller plot area.

Disadvantages

- Require larger diameter for a given gas capacity, therefore, most competitive for very low GOR, very high GOR, or scrubber applications.
It is not recommended when there is a large slug potential.

More difficult to reach and service top-mounted instruments and safety devices.

4.2.3 Gravity separation theory

Vapor–liquid separation is usually accomplished in three stages. The first stage, primary separation, uses an inlet diverter\(^1\) to cause the largest droplets to impinge by momentum and then drop by gravity. The next stage, secondary separation, is gravity separation of smaller droplets as the vapor flows through the disengagement area. Gravity separation can be aided by utilizing distribution baffles that create an even velocity distribution in the fluid thus allowing enhanced separation. The final stage is mist elimination, where the smallest droplets are coalesced on an impingement device, such as a mist pad or vane pack, followed by gravity settling of the larger formed droplets.

In the gravity settling section of the separators, the liquid drops will settle at a velocity determined by equating the gravity force \(F_B\) on the drop with the drag force \(F_D\) caused by its motion relative to the vapor continuous phase. When the drag force is equal to the buoyancy \(F_B\) force, the droplets acceleration is zero so that it moves at a constant velocity. This velocity is the terminal or free settling velocity, which is determined with respect to the following equations:

\[
F_B = \left(\frac{\pi}{6}\right) D_d^3 (\rho_L - \rho_V) \left(\frac{g}{g_c}\right) \tag{4-1}
\]

where \(D_d\) is drop diameter, ft; \(\rho_L\) is liquid density, lbm/ft\(^3\); \(\rho_V\) is vapor density, lbm/ft\(^3\); \(g\) is gravitational constant, 32.174 ft/s\(^2\); and \(g_c\) is conversion factor, 32.174 lbm-ft/s\(^2\)-lbf.

Also, the drag force on the droplet is given by:

\[
F_D = C_D (A_p) \left(\frac{\rho_V V_d^2}{2 g_c}\right) \tag{4-2}
\]

where

\[
C_D = \text{drag coefficient, dimensionless,}
\]

\[
A_p = \text{projected drop area, ft}^2
\]

\[= (\pi/4) D_d^2 \text{ (area of circle, not sphere)}\]

\[V_d = \text{drop velocity, ft/s} \]

\(^1\)Inlet diverters are very old technology now and are very rarely used. Revamps in the North Sea replace these inlet devices all the time, especially where the asset is producing more throughput than originally expected. If incorrectly sized, inlet devices can cause serious separation issues due to droplet shatter. Schoepentoeters became the popular inlet device for a while, but these were designed for vertical separators and are therefore not always applicable. Cyclonic and distribution baffle inlet devices are more common nowadays.
Therefore, the terminal settling velocity of the liquid droplets ($V_t$) can be determined by equating the Eqns (4-1) and (4-2) as follows.

$$V_t = \sqrt{\frac{4/3}{D_d (\rho_L - \rho_V) g/(C_D \rho_V)}}$$  \hspace{1cm} (4-3)

The drag coefficient can also be calculated as follows (Svrcek & Monnery, 1993):

$$C_D = \frac{5.0074}{\ln(x)} + \frac{40.927}{\sqrt{X}} + \frac{484.07}{X}$$ \hspace{1cm} (4-4)

and

$$X = \frac{0.95 \times 10^8 D_d^3 \rho_V (\rho_L - \rho_V)}{\mu_v^2}$$ \hspace{1cm} (4-5)

where $D_d$ is in ft, densities are in lb/ft$^3$, and viscosity is in cP.

The droplet-settling velocity equation considers the escape of a drop from the continuous phase (e.g., the escape of an oil drop from the gas phase). For this purpose, the droplet-settling velocity must be greater than the superficial upward bulk vapor velocity, $V_V$. Typically, the allowable vapor velocity is set between 0.75 $V_t$ and $V_t$ (Svrcek & Monnery, 1993). Equation (4-3) can be rearranged as a Souders and Brown (1934)-type equation as follows:

$$V_t = K_{SB} \sqrt{(\rho_L - \rho_V)/\rho_V}$$ \hspace{1cm} (4-6)

where

$$K_{SB} = \sqrt{\frac{4gD_d}{3C_D}}$$ \hspace{1cm} (4-7)

In practice, the Souders and Brown design coefficient ($K_{SB}$), depends primarily on the type of mist extractor present, separator geometry, flow rates, and fluid properties. Therefore, $K_{SB}$ is usually determined from experiments. A well-known source of empirical $K_{SB}$ factors for mist pads is the GPSA (2004) Engineering Data Book. The GPSA’s $K_{SB}$ factors have been curve fitted and are given as:

$$K_{SB} = 0.35 - 0.0001(P - 100)$$ \hspace{1cm} (4-8)

where $K_{SB}$ is in ft/s and $P$ is the separator pressure in psig.

Also, the factor $K_{SB}$ should be adjusted as follows:

1. For most vapors under vacuum, $K_{SB} = 0.20$.
2. For glycols and amine solutions, multiply $K_{SB}$ values by 0.6–0.8.
3. For compressor suction scrubbers and expander inlet separators, multiply $K_{SB}$ by 0.7–0.8.

Maximum terminal velocities calculated using the $K_{SB}$ factors are for separators normally having a wire-mesh mist extractor and should allow all liquid droplets larger than 10 $\mu$m to settle out of the gas. If no mist extractor is present, multiply $K_{SB}$ by 0.5.
It is often necessary to separate two immiscible liquids, the light and heavy phases, and a vapor. A typical example in the petroleum industry is the separation of water, and a hydrocarbon liquid and vapor. For this system, the flow of rising light droplets in the heavy liquid phase or settling heavy droplets in the light liquid phase is considered laminar and is governed by Stokes law (Monnery & Svrcek, 1994):

\[ V_t = \frac{K_s (\rho_H - \rho_L)}{\mu} \]  
\[ K_s = 2.06151 \times 10^{-5} D_d^2 \]  

where, \( V_t \) is in inch/min, densities of light and heavy liquid phases (\( \rho_L, \rho_H \)) are in lb/ft\(^3\), viscosity is in cP, and \( D_d \) is in microns (1 \( \mu \)m \( = \) 3.28084 \( \times \) 10\(^{-6}\) ft).

As can be seen from Eqn (4-9), the settling velocity of a droplet is inversely proportional to the viscosity of the continuous phase. Hence, it requires more time for the droplets to settle out of the continuous phase with greater viscosity. In practice, \( V_t \) is typically limited to 10 inch/min.

### 4.2.4 Design considerations

The following factors must be determined before beginning separator design:

- Gas and liquid flow rates (minimum, average, and peak),
- Operating and design pressures and temperatures,
- Surging or slugging tendencies of the feed streams,
- Physical properties of the fluids such as density, viscosity, and compressibility,
- Designed degree of separation (e.g., removing 100% of particles greater than 10 \( \mu \)m).

Consideration for the future life of the field should also be included. For example, most North Sea separators were designed for high oil cuts, but are now operating on high water cuts, which produces a lot of nozzle problems.

The initial design and calculation of gravity separators are discussed in many books and basic references (Kumar, 1987; Campbell, 1992; Arnold & Stewart, 1998). However, a more accurate and detailed sizing of two- and three-phase separators can be performed by using the design methods developed by Svrcek and Monnery (1993), and Monnery and Svrcek (1994) which have been well received by the industry worldwide. These procedures are a result of a review of literature sources and accepted industrial design guidelines and allow the production facility engineer to choose the detailed sizing parameters of the two- and three-phase separators. However, it is reasonably common nowadays for vessel sizing to be subcontracted out either directly to the vessel supplier or sometimes to the internals supplier.
4.3 Multistage separation

To achieve good separation between the gas and liquid phases and maximize the hydrocarbon liquid recovery, it is necessary to use several separation stages at decreasing pressures in which the well stream is passed through two or more separators that are arranged in series. The operating pressures are sequentially reduced, hence the highest pressure is found at the first separator and the lowest pressure at the final separator. In practice, the number of stages normally ranges between two and four, which depends on the GOR and the well stream pressure. Two-stage separation is usually used for low GOR and low well stream pressure, three-stage separation is used for medium to high GOR and intermediate inlet pressure, and four-stage separation is used for high GOR and high-pressure well stream. Note, three-stage separation usually represents the economic optimum, where it allows 2–12% higher liquid recovery in comparison with two-stage separation and, in some cases, recoveries up to 25% higher (Rojey et al., 1997). To recover the gas fractions produced in the separators operating at medium pressure and low pressure, it is necessary to recompress them to the pressure of high-pressure separator. However, for an associated gas, recompression is sometimes considered too costly; hence the gas produced from the low-pressure separator may be flared.

It should be noted that the main objective of stages separation is to provide maximum stabilization to the resultant phases (gas and liquid) leaving the final separator, which means the considerable amounts of gas or liquid will not evolve from the final liquid and gas phases, respectively. The quantities of gas and liquid recovered at a given pressure are determined by equilibrium flash calculations using an appropriate equation of state. This helps to optimize the value of pressure that is set for each separator. The pressures are often staged so that the ratio of the pressures in each stage is constant. Therefore, if the pressure in the first separator (which is normally fixed by specification or economics) and the final separator (which will be near the atmospheric pressure) are known, the pressure in each stage can be determined.

4.4 Centrifugal separators

In centrifugal or cyclone separators, centrifugal forces act on a droplet at a force several times greater than gravity, as it enters a cylindrical separator (Figure 4-3). This centrifugal force can range from 5 times gravity in large, low velocity units, to 2000 times gravity in small, high pressure units. Generally, centrifugal separators are used for removing droplets greater than 100 μm in diameter and a properly sized centrifugal separator can have a reasonable removal efficiency of droplet size as low as 10 μm.

Centrifugal separators are also extremely useful for gas streams with high particulate loading (Talavera, 1990). Such equipment has already been studied; however, a simple, compact, and lightweight gas–liquid cylindrical cyclone
(GLCC) separator has been developed by Tulsa University that requires little maintenance and is easy to install and operate. The compact dimensions, smaller footprint, and lower weight of the GLCC have a potential for cost savings to the industry, especially in offshore applications. Also, the GLCC reduces the inventory of hydrocarbons significantly, which is critical to the environmental and safety considerations. The GLCC separator, used mainly for bulk gas–liquid separation, can be designed for various levels of expected performance. Typical performance levels from the GLCC separator are 0.5–2.0 gallons of liquid per MMscf in the gas outlet and 0–5% gas in the liquid outlet (NATCO, 2002). More information on the design, control system studies, experimental investigations, and field applications of GLCC separators is discussed in greater details by Gomez et al. (2000), Mohan and Shoham (2003), and Wang et al. (2003).

### 4.5 Twister supersonic separator

The Twister™ supersonic separator is a unique combination of the expansion, cyclonic gas–liquid separation, and recompression process steps in a compact, static, tubular device to condense and separate water and heavy hydrocarbons from natural gas. Significant potential has been identified for future of Twister technology for other gas processing applications including deep LPG extraction, bulk removal of CO₂ and H₂S, mercury removal, and subsea gas processing.
Twister has thermodynamics similar to a turbo-expander, which transforms pressure to shaft power. Therefore, Twister achieves a similar temperature drop by transforming pressure to kinetic energy (i.e., supersonic velocity). The tube is the heart of the Twister system that combines adiabatic cooling, in which no heat enters or leaves the system, with cyclonic separation in a single, compact device. Figure 4-4 details the cross-section and key components of the tube design. As can be seen, adiabatic cooling is achieved in a Laval nozzle—an aerodynamically shaped venturi tube—which is used to expand the saturated feed gas to supersonic velocity, resulting in a low temperature and pressure. This results in the formation of a mist of water and hydrocarbon condensation droplets. The fine, dispersed liquids formed during the adiabatic expansion are separated using a cyclonic separator as a result of the centrifugal forces exerted by the strong swirling flow, and removed from the dry flow with significantly high separation efficiency. The liquid stream contains slip gas, which will be removed in a compact liquid degassing vessel and recombined with the dry gas stream (see Figure 4-5). Twister is a low-temperature separation process, for which performance can be optimized by inlet cooling. The inlet separator upstream of the Twister tubes is designed to remove produced liquids and prevent carry-over of slugs and solids.

The residence time inside the Twister supersonic separator is only milliseconds, allowing hydrates no time to form and avoiding the requirement for hydrate inhibition chemicals. The simplicity and reliability of a static device, with no rotating parts which operates without chemicals, ensures a simple and safe facility with a high availability, suitable for unmanned operation in harsh and/or offshore environments.

Twister technology was developed in 1998 for hydrocarbon dewpoint control for offshore application. Current commercial experience is two Twister separation units installed on the Petronas/Sarawak platform in Malaysia. Recently, Twister BV introduced the Twister SWIRL valve, which claims to have higher separation efficiency. UOP has also joined the development effort, applying their multiple valves switching technology to dampen out gas flow fluctuation required by the Twister separator operation.

**FIGURE 4-4**

Cross-section of a Twister tube (Courtesy of Twister BV).
4.6 Slug catchers

Slug catchers are used at the terminus of large gas-condensate transmission pipelines to catch large slugs of liquid in pipelines, to hold these slugs temporarily, and then to allow them to follow into downstream equipment and facilities at a rate at which the liquid can be properly handled. In fact, the slug catcher provides temporary storage for any surges (slugs) in liquid flows, and roughly separates the gas from the liquids. The gas then exits the top of the slug catcher and flows to the plant inlet separator via a pressure control valve, which reduces the pressure of the gas and further condenses water and some of the heavier hydrocarbons.

Slug catchers may be either of the vessel or of the manifolded pipe type. A vessel-type slug catcher is essentially a knockout vessel. This type is simple in design and maintenance. A pipe-type slug catcher consists of several long pieces of pipes (fingers), which together form the buffer volume to store the largest slugs expected from the upstream system. Vessel-type slug catchers can only be used if the incoming liquid volume is small. When larger liquid volumes have to be accommodated, say of more than 100 m³ (3531 ft³), the pipe-type slug catcher should be used (Shell DEP 31.40.10.12-Gen, 1998). Pipe-type slug catchers are frequently less expensive than vessel-type slug catchers of the same capacity due to thinner wall requirements of smaller pipe diameter. The manifold nature of multiple pipe-type slug catchers also makes possible the later addition of additional capacity by laying more parallel pipes. As the pipe-type slug catcher is defined as a piping configuration rather than a pressure vessel, it is not constrained to the same requirements as a normal vessel. However, due to its

FIGURE 4-5
Process flow diagram for a typical Twister system (Courtesy of Twister BV).
large size, it will contain the majority of high-pressure hydrocarbon gas on the site. It is therefore recommended that the slug catcher be automatically depressurized (for prevention of fire) as quickly as possible without imposing unusually high flow rates on the flare system.

A schematic of a pipe-type slug catcher appears in Figure 4-6. The general configuration consists of the following parts:

- Fingers with dual slope and three distinct sections: gas–liquid separation, intermediate, and storage sections.
- Gas risers connected to each finger at the transition zone between the separation and intermediate sections.
- Gas equalization lines located on each finger. These lines are located within the slug storage section.
- Liquid header collecting liquid from each finger. This header will not be sloped and is configured perpendicular to the fingers.

Note, it has been assumed that all liquids (condensate and water) are collected and sent to an inlet three-phase separator although it is possible to separate condensate and water at the fingers directly. When doing condensate–water separation at the slug catcher itself, we have to allow separately for the maximum condensate slug and the maximum water slug in order to ensure continuous level control.

**FIGURE 4-6**

Three dimensional rendering of finger-type slug catcher.
Separation of gas and liquid phases is achieved in the first section of the fingers. The length of this section will promote a stratified flow pattern and permit primary separation to occur. Ideally liquid droplets, 600 μm and below, will be removed from the gas disengaged into the gas risers, which are located at the end of this section. The length of the intermediate section is minimal such that there is no liquid level beneath the gas riser when the slug catcher is full, i.e., storage section completely full. This section comprises of a change in elevation between the gas risers and the storage section that allows a clear distinction between liquid and gas phase. The length of the storage section ensures that the maximum slug volume can be retained without liquid carryover in the gas outlet. During normal operations, the normal liquid level is kept at around the top of the riser from each finger into the main liquid collection header, which is equivalent to approximately 5-min operation of the condensate stabilization units at maximum capacity.

Because the slug catcher is the first element in the gas processing plant, determining its proper size is vital to the operation of the entire plant. In fact, if more liquid is brought in than the slug catcher can handle, the plant normally shuts down. Therefore, slug sizing results should always be treated with caution and slug catchers should be designed with an ample design margin (Burke & Kashou, 1996; Xiao & Shoup, 1998). Shell DEP 31.40.10.12-Gen (1998) specifies requirements and gives recommendations for the design of multiple-pipe slug catchers. However, vendors should be requested to provide detailed design guidelines.

4.7 High-efficiency liquid/gas coalescers

Aerosols in gas streams can often be less than 5 μm in size and requires the use of special separation equipment. High-efficiency liquid/gas coalescers have been applied effectively for the removal of fine aerosols in gas production, processing, and transmission. Coalescers are typically constructed as cartridges that use pleated glass fiber media supported by a metal core. The coalescer cartridges are then placed in a housing that controls the inlet/outlet gas velocities to ensure good separation and prevent any re-entrainment of liquids. The coalescer media contains a much finer pore structure and larger surface area as compared to traditional separators that often use mesh pads or vane pack internals.

4.7.1 Aerosols

Aerosols are formed by three mechanisms: condensation, atomization, and entrainment. The relative sizes produced by these formation mechanisms are given in Figure 4-7.

Aerosols formed by condensation of a vapor into a liquid are the smallest and the most difficult to remove contaminants having a size distribution in the
range of 0.2–5 µm. Atomization creates aerosol drops by breaking up larger liquid drops through mechanical shear such as passing through a constriction in a valve under a high velocity. Atomization forms aerosols in the size range of 10–200 µm. Entrainment involves the movement of liquid slugs along pipelines and here the liquid drop sizes are very large from 500 to 5000 µm. All three types of aerosol liquids are commonly found in gas systems. High-efficiency liquid/gas coalescers can effectively remove the fine aerosols created by the condensation mechanism.

4.7.2 Coalescer construction/operation principles

High-efficiency liquid/gas coalescers are generally constructed from glass fibers since this material allows for a fine porous structure with fiber diameter of a few microns. The small pore size is needed to achieve greater capture and separation of these small aerosols. The primary rational for the use of high-efficiency coalescers is that significant aerosol contaminant exists in the plants that are in the sub-micron and low-micron size range (Brown et al., 1994).

This type of liquid/gas coalescer can operate at significantly lower flow rates than the initial design flow rate and therefore can tolerate a high turndown ratio. This is due to the separation mechanisms for coalescing that are primarily diffusion and direct interception unlike vane separators and mesh pads that rely heavily on inertial separation principles. This allows the high-efficiency liquid/gas coalescer systems a greater degree of flexibility and they can operate at peak performance even for high turndown ratios (reduced flow rates) which can occur during commonly encountered partial plant shutdowns and upset conditions.

The use of surface treatment (Miller et al., 1988) on high-performance vertical liquid/gas coalescer cartridge systems has been proven to significantly enhance
performance by allowing higher flow rates or smaller housing diameters compared to untreated coalescers. The surface treatment alters the properties of the coalescer medium so that it will not wet out with oil or water-based fluids. The treatment has also been found to extend the service of the coalescer by reducing fouling and also to lower the saturated pressure drop. A Pall vertical high-efficiency liquid/gas coalescer system is depicted in Figure 4-8.

The inlet gas with liquid aerosol contamination first enters at the bottom of the housing into a first-stage knockout section. Here any slugs or large-size droplets (approximately >300 μm) are removed by gravitational settling. The gas then travels upward through a tube sheet and flows radially from the inside of the cartridges through the coalescer medium to the annulus. The inlet aerosol distribution is in the size range of 0.1–300 μm and after passing through the coalescer medium is transformed to enlarged coalesced droplets in the size range of 0.5–2.2 mm. The advantage of flowing from the inside to outside of the coalescer cartridge is that the gas velocity can be more easily adjusted in the annulus by selecting the optimum housing diameter to prevent re-entrainment of coalesced droplets.

As the gas leaves the coalescer cartridge and travels upward in the annulus it contributes to the total flow, thereby increasing the annular velocity. The annular velocity is modeled as a linear function with vertical distance, and the annular velocity is zero at the bottom of the cartridge and increased to a maximum value at the top of the cartridge.

FIGURE 4-8
Once the coalesced droplets are formed, they immediately drain vertically downward in the coalescer medium pack. The surface treatment greatly enhances this drainage and as a direct consequence of the treatment, the coalesced droplets are shielded from the upward gas flow in the annulus in the upper 2/3 section of the coalescer cartridge. The coalesced droplets are first exposed to the annular gas flow when they appear on the external face of the coalescer medium pack at the bottom third of the coalescer cartridge (see Figure 4-9). Once the coalesced droplets are released to the annular space they are subject to the force of the upward flowing gas. The trajectory of the coalesced droplets is modeled on a force balance between gravity settling and the drag force created by the gas flow past the droplets. This analysis leads to the calculation of a critical annular velocity for re-entrainment.

Due to the surface treatment, there are minimal coalesced droplets present in the annulus above the drainage point at the bottom third of the coalescer cartridge. For a coalescer cartridge that is not specially surface treated, the coalesced liquids are present throughout the length of the coalescer in the annulus space and the critical annular velocity for re-entrainment is given for the top of the element (see Figure 4-9). For the treated coalescer, it is allowable to have annular velocities greater than the critical value for re-entrainment in the portion of the

![Figure 4-9](image-url) Effect of surface treatment on coalescer drainage (Wines, 2004).
annulus space where there are no liquids present. This allows the maximum
annular velocity at the top of the coalescer cartridge to be about three times the
critical re-entrainment value needed at the vertical position of the lower one-third
of the cartridge height where liquids are present.

Therefore, the maximum annular velocity at the top of the coalescer cartridge
is found to be about three times greater than the value for an untreated coalescer.
The annular area is determined using the maximum allowable annular velocity
and designed to be of sufficient size to prevent re-entrainment and as small as
possible to minimize the housing diameter.

### 4.7.3 Modeling the liquid/gas coalescer

The modeling of the liquid/gas coalescer system can be divided into two basic
aspects for performance: media velocity and annular velocity. The other
consideration to be taken into account is pressure drop. The pressure drop for a
given system can be decreased by using more coalescer elements.

#### 4.7.3.1 Media velocity

The media velocity ($V_{\text{med}}$) is defined as the actual flow rate divided by the
coalescer filter area (Wines, 2004):

$$V_{\text{med}} = \frac{Q_a}{N A_{\text{med}}} \tag{4-11}$$

where $N$ is number of coalescers, $A_{\text{med}}$ is media area for one coalescer, and $Q_a$ is
actual system flow rate at system conditions that is obtained from the standard
system flow rate ($Q_s$) as:

$$Q_a = \frac{Q_s(SG) \rho_{\text{Air, stp}}}{\rho_G} \tag{4-12}$$

where $SG$ is gas specific gravity, $\rho_{\text{Air, stp}}$ is density of air at standard temperature
and pressure, and $\rho_G$ is density of gas at system conditions.

The media velocity is not the actual velocity through the open pores of the
media, but rather an average by convention over the combined pore area and solid
matrix area in the spatial plane normal to the flow direction. The maximum media
velocity for a coalescer construction is related to a number of factors intrinsic to
the particular coalescer design and to the physical properties of the system. Four
steps have been identified with the mechanism of the formation and removal of
droplets in the coalescer medium: capture, coalescing, release, and drainage.

The formation of the coalesced droplets first involves the capture of the small
aerosols onto the fibers of the coalescer medium. The actual coalescing or
merging of the fine droplets is believed to take place on the fibers and especially
at fiber intersections. The coalesced droplets are then released from the fiber due
to the drag force of the gas flow exceeding the adsorption energy. This process is
repeated through the depth of the coalescer medium until the coalescing process is completed and the largest possible stable droplet size is achieved. During the coalescing stages, the growing droplets are also draining downward inside the media pack due to the force of gravity.

The surface treatment allows the release and drainage process to proceed at a faster rate which in turn frees up more coalescing sites on the fibers and allows the coalescer to process higher inlet liquid aerosol concentrations than the untreated coalescer medium.

4.7.3.2 Effect of system conditions on media velocity
The ability of the coalescer medium to perform effectively will also depend on the system environment. While different coalescer constructions will exhibit quantitative differences, they will follow the same qualitative behavior. The media velocity has been determined to depend on system parameters such as inlet aerosol concentration, aerosol density, gas density, and gas viscosity.

At low aerosol concentrations, the maximum media velocity is constant and is unaffected by aerosol levels. Under these conditions the media is limited by the capture mechanism and is not affected by drainage. At higher levels of aerosol concentration, the coalescer medium becomes limited by drainage and is inversely proportional to the aerosol concentration. The effect of the surface treatment on this process is to enhance the drainage and allow for higher maximum media velocities under the same aerosol loading when limited by drainage.

4.7.3.3 Annular velocity
The annular velocity \( V_{ann} \) is defined as the actual flow rate divided by the annulus area (Wines, 2004):

\[
V_{ann} = \frac{Q_h}{A_{ann}} \tag{4-13}
\]

where, \( A_{ann} \) is the cross-sectional annular area defined as the cross-sectional area of the housing without coalescers minus the area of the coalescer end caps:

\[
A_{ann} = \pi R_h^2 - N \pi R_c^2 \tag{4-14}
\]

where \( R_h \) is radius of the housing, \( R_c \) is radius of coalescer end cap, and \( N \) is number of coalescers.

The enlarged droplets leaving the coalescer media pack can be assumed to be as large as possible for the given flow conditions when complete coalescence has occurred. Therefore, the coalesced droplet diameter will be the same for any specific design of the coalescer cartridge as long as complete coalescence has been achieved. If complete coalescence is not achieved, the calculation of the coalesced droplets must take into account the degree of coalescence.

In most industrial applications, the coalesced droplets will range in size from 0.5 to 2.2 mm and will be mostly influenced by the interfacial tension which is
significantly affected by the liquid density, system temperature, and system pressure. As the pressure is increased, the gas density will increase while the liquid density is only slightly affected. The solubility of the gas in the liquid is enhanced with increasing pressure. This leads to a substantial decrease in interfacial tension with increasing pressure and consequently to significantly smaller coalesced droplets at the higher pressures.

Once the coalesced droplet size has been estimated, the next step is to determine the maximum annular velocity that can be sustained without re-entrainment. In general, the coalesced droplets will produce Reynolds numbers (Re) outside of the creeping flow regime (<0.1) and Stokes law. Instead, a force balance is used between the liquid droplets settling by gravity and the drag force of the gas flowing upwards in the opposite direction.

### 4.7.3.4 Determination of minimum housing diameter

The housing diameter is determined from the area of the annulus and the area of the coalescer end caps. The maximum annular velocity at the top of the coalescer cartridges is used to determine the annular area required. The value of the maximum annular velocity \(V_{\text{ann,max}}\) at the top of the coalescer cartridges is dependent on the critical annular velocity for re-entrainment \(V_c\) and the vertical location at which the coalesced droplets are present in the free annulus space. This relationship can be described as follows (Wines, 2004):

\[
V_{\text{ann,max}} = k_a V_c
\]

where, \(k_a\) is the annular velocity enhancement factor due to drainage.

For the untreated coalescer medium, the coalescer cartridge is completely wetted and coalesced droplets are present in the annulus space up to the top of the annulus where the annular velocity is highest. There is no drainage enhancement and \(k_a = 1\). The maximum annular velocity to prevent re-entrainment is then equal to the critical value for re-entrainment:

\[
\text{Untreated Coalescer: } V_{\text{ann,max}} = V_c
\]

The effect of the surface treatment is to greatly increase the drainage and the annular velocity at the top of the coalescer cartridge can now be significantly higher than the critical value since there are no coalesced droplets present in the annulus except in the bottom third of the cartridge. The maximum annular velocity is now determined with \(k_a = 3.1\) as follows:

\[
\text{Surface Treated Coalescer: } V_{\text{ann,max}} = 3.1V_c
\]

Convincing evidence for the enhanced maximum annular velocity given by Eqn (4-16) has been demonstrated by laboratory tests (Williamson et al., 1988; Murphy, 1984) and is presented in Figure 4-9. Visual observations during these tests also confirm that liquids are present on the outside of the coalescer pack only at the bottom third for the surface-treated coalescer and are present throughout the length of the wetted untreated coalescer.
4.7.4 Coalescer performance/operational limits

Generally, the high-efficiency liquid/gas coalescers are used for inlet aerosol concentrations of a few thousand ppmw or less and are placed downstream of other bulk removal separators as the final purification stage. Under these conditions, typical service life for liquid/gas coalescers is 1–2 years. Coalescer systems are usually sized for a clean differential pressure (DP) of 2–5 psi and when this DP reaches 15 psi, they are replaced with new elements. Outlet concentrations for these high-efficiency liquid/gas coalescers are as low as 0.003 ppmw (Murphy, 1984; Williamson et al., 1988).

4.7.5 Liquid/gas coalescer applications

The separation of liquid aerosol contamination with high-performance liquid/gas coalescer cartridge systems has found widespread acceptance in gas plants in recent years for a number of applications such as protection of compressors, turbo equipment, low NOx burner nozzles, amine and glycol contactors, and molecular sieve beds. This has largely been the result of traditional separation approaches including knockout vessels, centrifugal separators, mesh pads or vane separators not meeting the end user’s requirements for aerosol reduction. A brief description of some of the main applications is given by Mokhatab and Poe (2012).

4.8 High-efficiency liquid–liquid coalescers

Liquid–liquid separations may require the use of special equipment when the drop sizes are small typically in the range of 1–50 μm in size. These fluid systems are classified as stable emulsions and often conventional bulk separators with mist pads or plate type internals will not be effective. High-efficiency liquid–liquid coalescers have been developed to break these emulsions and provide improved separation.

4.8.1 Emulsions

Emulsions consist of the three components: oil (representing hydrocarbon or organic liquids), water (including any aqueous mixtures), and surfactants. Depending on the ratio of these components, oil-in-water emulsions or water-in-oil emulsions can exist. The structure of the oil-in-water or water-in-oil emulsions is well defined with spherical droplets of the dispersed phase surrounded by a bulk continuous phase and surfactant sheathing the droplets. Surfactants contain

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2Surfactants consist of polar or hydrophilic groups joined to nonpolar or hydrophobic hydrocarbon chains. The ability of surfactants to aid in the emulsification process will depend on the ratio of the polar to nonpolar groups, the charge density and size of the polar group, and the volume occupied by the nonpolar groups (branching and length).
both hydrophilic (water loving) and hydrophobic (water fearing) portions in the same molecule. This unique structure allows them to associate at water–oil interfaces and helps them to stabilize the droplet shape. A spherical drop shape is formed to minimize the surface area between the oil and water and this also minimizes the free energy required to make this surface. In order to make an emulsion the system must be subjected to shear or mixing to allow the three components to break up into the droplet structure.

Emulsions are inherently unstable and will spontaneously separate out into two bulk phases. This process requires that the small droplets merge together or coalesce repeatedly until they form increasing drop sizes that eventually merge with a bulk phase until all of the drops are gone. Depending on the nature of the emulsion, the separation can occur in a matter of seconds or months. Many of the same factors that affect emulsion stability also influence coalescer performance.

### 4.8.2 Coalescer principles and materials of construction

Coalescers are typically manufactured as either pads or cartridge filters that have been designed especially to take small droplets in an emulsion and grow them into large drops that are more easily separated. This process is accelerated over natural coalescing by the fibers present in coalescer media that force the contact of small droplets thereby promoting the coalescing process. The pore gradient of coalescer medium is constructed so that the inlet medium has fine pore sizes that increase in size with the flow direction (see Figure 4-10).

Coalescers have been primarily constructed with glass fiber media until recently when polymer and fluoropolymer materials were adopted. Glass fiber works adequately for emulsions with interfacial tensions >20 dyne/cm. It is known to disarm and lose efficiency in the presence of surfactants (Hughes, 1997). These coalescers are widely used to dewater jet fuel for the aviation industry.

High-efficiency liquid–liquid coalescers are the newest generation of coalescers, incorporating the latest in coalescer technology. They are constructed from polymer and fluoropolymer materials that have been optimized to separate the most difficult emulsions with interfacial tensions as low as 0.5 dyne/cm. This coalescer can be used with a broad range of applications. It can process aggressive chemicals and handle demanding operating conditions while providing the highest level of performance.

### 4.8.3 Coalescer mechanism of operation

The liquid–liquid coalescing system operates in three stages: separation of solids/preconditioning, coalescence, and separation of coalesced drops.

#### 4.8.3.1 Separation of solids/preconditioning of the fluid

Solids can increase the stability of an emulsion and removing solids can make coalescing easier. Generally, this step can be achieved by a separate cartridge...
filter system or by a regenerable backwash filter system for high levels of solids. In addition, the filtration stage protects the coalescer and increases service life. This step also initiates the coalescence of the hydrocarbon droplets, thereby enhancing the separation capabilities of the system.

4.8.3.2 Coalescence

The next step in the process is the primary coalescence. In this stage, the pore dimensions begin with a very fine structure and then become more open to allow for void space for the coalescing droplets. In the primary coalescence zone, the inlet dispersion containing fine droplets in the size range of 0.2–50 μm is transformed into a suspension of enlarged droplets in the size range of 500–5000 μm.

The coalescence mechanism can be described by the following steps:

1. Droplet adsorption to fiber,
2. Translation of droplets to fiber intersections by bulk flow,
3. Coalescence of two droplets to form one larger droplet,
4. Repeated coalescence of small droplets into larger droplets at fiber intersections,
5. Release of droplets from fiber intersections due to increased drag on adsorbed droplets caused by bulk flow,
6. Repeat of steps 1–5 with progressively larger droplet sizes and more open media porosity.

Based on this mechanism, we can predict that a number of factors will influence the coalescence performance. The specific surface properties of the coalescer fibers are critical in influencing the adsorption of droplets as well as the ultimate release after coalescing. There is a balancing act between increasing the attraction or adsorption characteristics of the fibers against the release mechanism.
which strong adsorption would inhibit. The necessary condition that droplet-fiber adsorption occur for coalescing has been supported by a number of sources (Jeater et al., 1980; Basu, 1993).

### 4.8.3.3 Separation of coalesced droplets

Once the droplets have been coalesced they are now as large as possible for the given flow conditions. The separation stage can be achieved in one of two ways:

**Horizontal configuration:** The coalescer housing contains a settling zone that relies on the difference in densities between the coalesced droplets and the bulk fluid (see Figure 4-11). This configuration can be used for both hydrocarbon from water and water from hydrocarbon separation, but the location of the collection sump and outlet nozzle will need to be reversed. For the case of the removal of hydrocarbon from water, a collection sump is located at the top of the housing and the purified water leaves at the bottom outlet nozzle. The sump can be manually drained on a periodic basis or equipped with an automatic level control and drain system. Estimation of the coalesced drop size and required settling zone are best determined through pilot-scale tests at field conditions.

**Vertical Configuration:** Once the droplets have been coalesced, they are now as large as possible for the given flow conditions, in the range of 0.5–2 mm in diameter. The separation stage is achieved using hydrophobic separator cartridges that provide an effective barrier to aqueous coalesced drops, but allow hydrocarbon to pass through them. The separator cartridges can be stacked below the coalescers for the most efficient utilization of the separator medium. This

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**FIGURE 4-11**

Horizontal liquid–liquid coalescer configuration (Katona et al., 2001).
configuration only applies to the separation of water or aqueous contaminants from hydrocarbons (see Figure 4-12).

After leaving the coalescing stage, the large aqueous coalesced drops and hydrocarbon then flow axially in a downward direction and the flow direction is from the outside of the separator to the inside. The large coalesced drops are repelled by the separators and are collected in the bottom sump. The purified hydrocarbon passes through the separators and exits at the bottom of the housing. The aqueous phase in the collection sump can be manually drained on a periodic basis or equipped with an automatic level control and drain system.

**4.8.4 Liquid–liquid coalescer performance**

Properly designed and sized high-efficiency coalescer systems can process inlet discontinuous phase concentrations up to 10% and reduce them to ppm levels in the outlet for interfacial tensions as low as 0.5 dyne/cm. For water from hydrocarbon separations, coalescer outlets below 15 ppmv per the AquaGlo (registered trademark of Gammon Corp.) method (ASTM D3240) can be achieved and for hydrocarbons from water, concentrations below 20 ppmw per the Oil and Grease method have been demonstrated.

The use of polymers and fluoropolymers in the coalescer materials of construction allows for expanded use of coalescers over earlier conventional types so
that they can withstand an array of aggressive chemical applications over a wide range of temperatures from $-40 \degree F$ up to $300 \degree F$.

### 4.8.5 Limitations of using coalescers

While liquid–liquid coalescers have many benefits in breaking tough emulsions, there are some limitations to consider. Solids can become problematic at higher concentrations and lead to excessive change out of disposable prefilters. Generally, the solids range that liquid–liquid coalescers can operate economically with disposable filters is less than 10 ppm. Above this level of solids, further pretreatment will be required such as backwash cartridge filters, mixed media packed beds, or hydrocyclones for solids removal.

The operational limits of the coalescer for removal of free liquids must also be understood. If the clarified stream leaving the coalescer is then cooled, condensation of previously dissolved contaminant can occur leading to a hazy fluid at the lower temperature. The coalescer will not be able to remove contaminants that are dissolved in solution and therefore, the location to place the coalescer and any subsequent change in process conditions after the coalescer must be considered carefully.

The coalescers typically will have a service life of 1–2 years when protected adequately by prefiltration. Despite the long life, the coalescers will eventually require disposal and replacement; however this will be minimal given the low frequency of replacement.

For liquid–liquid coalescers constructed from glass fiber medium, the problem of surfactant disarming must also be considered and for low interfacial tension emulsion systems ($<20$ dyne/cm), they will not operate efficiently for separation. For these conditions nondisarming fluoropolymer or polymer coalescers should be considered. These types of materials also have wider compatibility for chemical streams and a wider temperature range.

### 4.8.6 Applications

High-efficiency liquid–liquid coalescers are finding increasing applications in industry where problematic emulsions exist. They are used to protect equipment, recover valuable streams, and to meet environmental discharge limits. Some examples are given by Mokhatab and Poe (2012).

### 4.9 Practical design of separation systems

Designing a separation system capable of separating oil, water, gas, and any solids, traditionally relies upon the application of standard calculation methods using certain values for design criteria (such as inlet momentum, gas K-value, and liquid separation residence time), that are based on empirical information and
tested configurations. As such, these classic methods do not consider matters such as droplet size distribution and mist fraction in the inlet piping, droplet capture efficiencies, and separation performance of each separation step in the separator. As a result, the carryover rate associated with the separator design and operating conditions is commonly not known at the design stage. Therefore, designs based on these classic criteria-based methods have an inherent risk of malperformance and operating problems upon start-up. Fortunately, recent developments have made advanced separator design and rating softwares available in the market. These softwares include the complex calculations associated with droplet size distributions and internal efficiencies based upon the design and operating conditions, thus providing the designer with a prediction of the separator’s performance and carryover. With this key information, the designer can adjust the separator design to suit its required performance in the system. This significantly reduces the risk of operating problems in the field.

The classic design methods also assume that the nature of flow in the separators is uniform whereas in reality it will be strongly three-dimensional and highly turbulent. Therefore, problems occur frequently when commissioning new separators or existing separators under unusual loading conditions. Modeling the complex three-dimensional flow within the separation equipment (piping, vessels, and internals) can provide a visual indication of the weaknesses in the process design. This will help to optimize the design of new separators and solve operational problems with existing designs. Computational fluid dynamics (CFD) is a key tool for modeling and predicting fluid flow behavior inside the separation equipment under real operating conditions. CFD enables the nature of the flow in the separator to be illustrated so that the interaction between the equipment and the different flow components can be understood. Qualitative comparisons of the performance of different models can then be made to identify the optimal solution. The optimized CFD model can be verified with on-site test capabilities if this is required. The verified model can then be used in fine-tuning designs prior to separator manufacturing, resulting in a significant saving in time and money, and allows operational problems to be identified and solved before they even manifest themselves. In conclusion, many practical design and redesign applications for separators may be performed by CFD modeling and simulations.

4.9.1 Case study

The following case study demonstrates the power of using CFD to analyze and find the root causes of operational issues in separation facilities. Using this tool would reveal the areas in the separation facilities where special attention is required to mitigate or prevent operating issues. With these insights, the design engineer can address the potential issues at the design stage, which is far less costly than rectifying malperforming designs after they have been built and put into operation.
4.9.2 The situation

An operator has been experiencing problems with one of their gas scrubbers. Large amounts of carryover were found in equipment downstream of a gas scrubber. To reduce the carryover and associated operational issues, the operator had to reduce the production rate through the scrubber from 100 to <70 MMSCFD, resulting in a considerable loss of revenues.

The gas scrubber, with an inside diameter of 1.5 m and a Tan–Tan length of 2.6 m, was equipped with a deflector plate at the inlet and a horizontal flow vane pack acting as the demisting device. The operating pressure was 35 barg.

4.9.3 The objective and methodology

A first-pass analytical analysis showed that the root cause of the issue was most likely within the gas scrubber. To investigate this further, a CFD model of the existing vessel with its internals and the upstream piping was made (Figure 4-13). The actual operating conditions were imposed on the model.

4.9.4 The analysis

The CFD results revealed both analytical and visual insights. In Figure 4-13, it can be seen that the gas flow in the vessel is far from ideal. The configuration of the deflector and the horizontal flow vane pack causes severe maldistribution and “jetting” inside the vessel. The deflector causes high gas velocities over the liquid surface which may cause liquid re-entrainment. Furthermore, the CFD simulation also shows that the deflector baffle will cause significant droplet shattering due to high turbulent dissipation rates. The gas velocity profile on the face of the vane pack shows a nonuniform profile (i.e., maldistribution), with unacceptably high local velocities. The combination of high liquid load and the high local gas
velocities through the vane pack will cause the vane pack to fail. All these issues explain the high carryover observed during operation of this gas scrubber.

4.9.5 Modified situation

With the root causes of the malperformance identified, a solution to the problem was developed. The solution comprised of a redesign of the vessel internals, the adequacy of which was verified through CFD analysis. At the inlet nozzle, the inlet baffle was replaced with a proper vane-type inlet device. For the demisting section, the vane pack was replaced with a double mesh pad configuration; one acting as a mesh agglomerator and one as a mesh demister. The design and results of the CFD simulation are shown in Figure 4-14. The CFD simulation of the modified design showed that the gas distribution inside the vessel was improved drastically. The velocity over the liquid level became acceptable and the velocity profile on the face of the top mesh demister showed a proper uniform velocity profile. With the recommended internals, the operator can suitably operate at the required gas flow of 100 MMSCFD.

References

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5.1 Introduction

Production well fluids entering the inlet reception facility are first separated in the slug catcher. The hydrocarbon liquid stream contains mainly light hydrocarbons (methane and ethane in particular), water, salts, corrosion and hydrate inhibitors, acid gases, mercaptans, and other sulfur compounds. The condensate production unit is designed to separate these contaminants to produce a marketable hydrocarbon liquid (condensate) for export.

Condensate production involves three steps: water washing, condensate stabilization, and condensate treating. Depending upon the associated water quality, the condensate may require a water wash to remove salts and additives. After dewatering step, which requires a careful evaluation of the condensate/water separation technology, the condensate goes to the condensate stabilization unit where remaining lighter hydrocarbons are stripped and recombined with the gas that leaves the inlet reception facility. The process of increasing the amount of intermediate (C₃–C₅) and heavy (C₆+) components in the condensate is called condensate stabilization. This process is performed primarily in order to reduce the vapor pressure of the condensate so that a vapor phase is not produced upon flashing the liquid to atmospheric storage tanks. Stabilized condensate generally has a vapor pressure specification, typically identified by its Reid Vapor Pressure (RVP\(^1\)) or True Vapor Pressure (TVP). RVP is set by local emissions authorities to limit hydrocarbons emissions during storage and transport. Typical RVP specification ranges from 4 to 8 psia. For hydrocarbon storage in high elevation, the atmospheric pressure is lower, and consequently a low RVP as low as 4 psia may be necessary, allowing some safety margins. After the stabilization step, condensate must be treated to remove heavy mercaptans and other undesirable contaminants to very low levels in order to produce a liquid product that has specifications to be sold as “natural gasoline.” In addition to the above-mentioned RVP specification, other typical specifications of the commercial grade condensate are: water content (0.05 volume %), butanes (2 volume %), H₂S (10 ppmw), and total sulfur content as S (50 ppmw). Recent accidents of condensate rail cargoes and condensate trucks in North Dakota and Oklahoma areas prompt the transportation authority and

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\(^{1}\)Reid vapor pressure (RVP) is related to the vapor pressure of a petroleum product, which measures its inherent tendency to evaporate at 100 °F with vapor:liquid ratio of 4:1 (ASTM D323). RVP is a function of the hydrocarbon’s composition and is independent of operating temperature and pressure.
administrators to tighten the safety codes. It is expected that the condensate specifications will be more stringent: lower RVP and lower sulfur content.

The objective of this chapter is to review some of the basic processes for condensate stabilization, and the associated facilities such as condensate storage, condensate hydrotreating, monoethylene glycol (MEG) regeneration and reclaiming, and sour water treatment.

5.2 Condensate stabilization

There are two basic stabilization processes: cascade flash separation and distillation separation. Cascade flash separation, consisting of multiple separators and compressors, is common in offshore gas processing plants due to its simplicity, but its efficiency and condensate yields are lower. The distillation separation type is more complex, and more efficient, which is common in onshore gas processing plants.

5.2.1 Stabilization by cascade flash separation

The principle of the cascade flash separation is to remove the lighter components by flashing to lower pressures in several steps. The condensate can also be heated to promote removal of the light components. The process equipment can be very compact, which is advantageous in offshore gas installation where equipment weight and plot space must be minimized. However, the condensate yield is low, particularly when processing a lighter condensate to meet a low RVP requirement. The process is more suitable for processing crude oil or heavier condensate.

A typical cascade flash separation process is shown in Figure 5-1. The condensate is flashed and separated at three successive pressures at 500, 100, and

FIGURE 5-1
Condensate stabilization by cascade flash separation.
15 psig, respectively. The flashed vapor can be used as fuel gas in the facility or sent to the flare (during early production developments). Typically, the vapors are compressed back to the feed section for recovery. Vapor from compressor discharge is cooled and the condensed liquid is blended with the condensate product, which helps to reduce the condensate losses.

To achieve a low RVP condensate, the feed must be heated to higher temperatures, typically about 150–250 °F. In fact, the feed temperature must be sufficiently high to drive off its H₂S content to meet the H₂S specification in the product. Once the H₂S specification has been met, heating can be adjusted as needed to meet the vapor pressure specification.

It is impractical for the flash stabilization process to meet a very low RVP specification. If a low RVP is required, the only solution is the distillation type, which is discussed in the following sections.

5.2.2 Stabilization by distillation
Distillation process is an efficient method for separating the C₅+ from the lighter components, instead of using multiple flash stages. The distillation column can be a refluxed type or a nonrefluxed type (simple stripper). A nonrefluxed type is lower in capital cost, as the overhead reflux condenser system is not required. The drawback is the loss of the C₅+ components in the overhead. The C₅+ components can be partially recovered by recycling the condensate later, which in most cases, can be economically justified.

With the refluxed column, there are two design options. The first option is to produce a condensate product, with the butane and lighter components returned to the acid gas removal unit (AGRU). The second option is to produce a condensate and a liquefied petroleum gas (LPG) fraction. Both configurations are described in the following.

5.2.2.1 Condensate production only
The process flow schematic of a condensate stabilization unit is shown in Figure 5-2. The condensate is flashed and separated in an intermediate separator, with flashed vapors compressed and returned to the AGRU. The flashed liquid is preheated with the stabilizer bottom and routed to the stabilizer for separation. The stabilizer typically operates between 150 and 250 psia, and contains about 20–25 trays.

The stabilizer column is heated with medium pressure steam to meet the RVP specification. The overhead vapor is partially condensed, by air or cooling water, producing a reflux to the column and a butane and lighter vapor that is compressed to the AGRU. No liquid overhead product is produced in this configuration.

Since the condensate is saturated with water, water will be stripped and condensed in the upper column. Water is collected in the reflux drum as an aqueous phase and may be trapped in the upper section of the stabilizer column.
Any free water must be removed from the column or it will build up inside the column, resulting in column flooding. Typically, interface controllers are provided to allow withdrawal of water from the reflux drum and the top trays. The draw tray locations can be determined to detect three phase conditions with simulation software.

The advantage of the condensate only design is that it recycles the LPG portion to the gas processing plant. This avoids producing LPG that may not meet the color and water specifications for export. The disadvantage is that the gas plant has to be designed for a higher duty from the recycled LPG.

### 5.2.2.2 Condensate and LPG production

The condensate stabilization process can be configured to produce LPG from the stabilizer column. In order to meet the ethane content in the LPG product, a feed liquid stripper is required upstream of the stabilizer. This stripper is designed to remove all the acid gases and lighter components such that the LPG product from the downstream stabilizer would meet the H₂S and vapor pressure specifications. The process flow schematic for the feed liquid stripper and stabilizer is shown in Figure 5-3. The feed liquid stripper pressure is typically set at about 200 psia or at slightly higher pressure than the stabilizer such that pumping can be avoided. The advantage of this process is that it avoids recycling the LPG portion and minimizes the size of the gas processing plant. However, the LPG product may contain other contaminants that may not meet specifications for export.

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**FIGURE 5-2**

Condensate stabilization by distillation.
5.2.3 Design considerations

The condensate stabilization unit shall be designed to meet the process design basis, which must consider the ranges of condensate flow rates and compositions, inlet pressures, and temperatures. The design basis must be conservative, as it is difficult to predict the upstream operations. These design parameters will affect the sizing of the separators, stabilizer column, reboiler, heat exchangers, and compression system.

Two important design considerations of the stabilizer column are described as follows.

5.2.3.1 Stabilizer column pressure

For a given condensate composition (i.e., 0.1% C₄, 14.6% iC₅, 27.8% nC₅, and 57.5% C₆+), there is a relationship between the stabilizer bottom temperature and the product’s vapor pressure, and they vary with operating pressures, as shown in Figure 5-4.

For a given stabilizer pressure, the bottom temperature decreases with increasing RVP specification. Conversely, the bottom temperature increases when the RVP specification is lowered, as more butane and pentane components are stripped overhead. As can be expected, higher column pressure requires more stripping steam and tends to increase the equipment cost. However, this may be offset by the lower compression cost of the column overhead gas.

The column pressure should be selected by considering the operating costs such as utilities consumption (steam, cooling water, and power), capital costs, and incremental revenue from condensate production.
5.2.3.2 Stabilizer system control
The stabilizer operation is very sensitive to inlet feed gas changes. To maintain a stable column pressure, the slug catcher and the feed separator should be designed conservatively to dampen out flow fluctuation, which the plant will experience during pipeline pigging operation. Predictive methods can also be built into the control system, such as the use of upstream feed conditions, and column tray data to adjust the reboiler steam flow using a feed-forward controller. Dynamic simulation model can be setup to optimize the control variables that can respond to changes in feed composition and flow conditions.

5.2.4 Operating problems
The common operating problems of the condensate stabilization units are failure to meet product specifications and unstable column operation. These can be contributed by one of the following factors:

- Feed flow rate and conditions are significantly different than design
- Carryover of contaminants and injection chemicals from upstream operation
- High corrosion, resulting in failure of piping and heat exchanger equipment
- Column flooding and unstable operation
- Equipment fouling.

Many of these problems can be avoided in the FEED stage of the project by defining the ranges of operating conditions.

5.3 Condensate hydrotreating
While steam stripping in the stabilizer can be used to remove lighter hydrocarbon and acid gas components, it has minimal effect on the removal of mercaptans.
If the condensate contains the lower molecular weight mercaptans (methyl mercaptan), it can be treated by conventional liquid treating technologies, such as caustic wash, UOP’s Merox, molecular sieves, and catalyst solid beds.

If the condensate contains the higher molecular weight mercaptans, aromatic compounds, and other undesirable sulfur components, it must be processed with a hydrotreater, which is a common process in refinery to desulfurize high sulfur feedstock. The primary function is to use a hydrotreater catalyst to promote the following desulfurization reactions:

\[
R-S-H + H_2 = R-H + H_2S \quad (5-1)
\]

\[
R-S-R' + 2H_2 = R-H + R'-H + H_2S \quad (5-2)
\]

where, \(R\) and \(R'\) are hydrocarbon and alternate hydrocarbon chains, respectively.

A typical process flow scheme of a condensate hydrotreating unit is shown in Figure 5-5. The process consists of a high-pressure reactor loop where sulfur compounds are converted to \(H_2S\), and aromatic hydrocarbons are saturated and converted to paraffinic hydrocarbons. The reactor effluent is cooled; recycle gas and product liquids are separated. The \(H_2S\) content in the flashed gas is removed by an amine-treating unit.

The flashed liquid is letdown in pressure and fractionated in a debutanizer, which strips off the \(H_2S\), butane, and light components from the condensate to produce a sulfur-free hydrocarbon liquid. Hydrogen compression is necessary to maintain the hydrogen partial pressure in the hydrotreater reaction loop. A hydrogen generation unit is used to supply high-purity hydrogen to support the hydrotreater reaction.
5.4 Effluent treatment
5.4.1 MEG regeneration and reclaiming

The condensate may contain MEG solution, which is used for hydrate control in subsea pipelines. The glycol solution is contaminated with salts, corrosion inhibitors, and pipe scales, which must be removed to avoid fouling in the downstream process equipment.

Separation of condensate from the MEG solution requires special design attention. Phase separation is affected by operating temperatures. Temperatures lower than 40 °C/140 °F will result in poor separation, due to high viscosity of glycol and potential emulsion formation. Separation would be improved if the temperature were above 60 °C. For this reason, the condensate temperature is typically pre-heated to above 60 °C prior to the feed separator, which is necessary to avoid contamination of the condensate by the MEG solution.

Water separated from the MEG regeneration unit contains H₂S, CO₂, and ammonia, and may contain other undesirable pollutants such as phenols, cyanide and various salts, organic or inorganic acids. The type and quantity of these pollutants depend on the well properties and the injection chemicals. They must be removed to comply with environmental regulations.

A typical MEG regeneration/reclaiming unit is shown in Figure 5-6. As can be seen, MEG solution is flashed to a feed drum operating at a lower pressure, which removes the dissolved gas from MEG. The flashed gas is treated and sent to the fuel gas system. The flashed liquid is pumped and filtered, removing most of the solids and pipe scales before entering the regenerator.

**FIGURE 5-6**
MEG regeneration and reclaiming.
The MEG regenerator is stripped by medium pressure steam, producing an acid gas and sour water overhead, and a concentrated MEG solution. The acid gas is sent to the Claus burners in the sulfur recovery unit and the water is routed to a sour water stripper. Any entrained hydrocarbon will be skimmed off from the reflux drum.

The MEG regenerator bottom is concentrated in salt and other nonvolatile materials, which is purified before recycling to the upstream facility. MEG can be reclaimed by vacuum stripping to separate MEG from the salty water. There are commercial designs that can be used to reclaim the MEG. Licensed units, from Prosernat or Cameron Process System, are skid-mount units that can remove salts, water and other solids and produce a high-purity glycol.

5.4.2 Sour water stripping

The main function of a sour water stripping unit, as shown in Figure 5-7, is to remove H₂S and ammonia from sour water before sending it to a wastewater treatment unit or recycling back to the process units. The overhead gas from the sour water stripper is routed to the acid gas burners in the sulfur recovery unit for destruction. In some designs, in order to minimize acid gas corrosion, a pump-around system with cooler is used in the top section, instead of the conventional overhead condenser.

Sour water stripper typically contains about 30 trays that are required to meet the ammonia specification of 10 ppmw for emissions compliance. In some installations, there are traces of acidic compounds in the sour water, which would fix the ammonia in solution. In these cases, addition of caustic may be necessary to neutralize the acidic compounds.

Low-pressure steam can be used to supply stripping requirement by the reboiler. The stripper bottom temperature is typically maintained at about 230 °F.
or higher temperatures sufficient to meet the ammonia specification. The overhead temperature is controlled at 180 °C (356 °F) or higher to avoid formation of ammonia bisulfide. The formation of the ammonia salt will cause fouling and corrosion in the overhead system. The overhead piping and equipment are constructed with specialty alloy steel and must be electric or steam traced to stay above the bisulfide formation temperature.

5.5 Condensate storage
The stabilized condensate products are stored in condensate storage tanks before they are transported to the refineries for further processing. The storage tanks must be designed to avoid flashing of the condensate under different operations. A vapor recovery system should be used to minimize any hydrocarbon emissions as described in the following sections.

5.5.1 Tank design considerations
Condensates are commonly stored in cylindrical steel tanks at atmospheric pressure. The tanks are flat bottomed and are provided with a roof, which is of conical or domed shape.

The storage tank is typically designed to operate at a slightly positive and a slightly negative atmospheric pressure to provide flexibilities during operations. The tank’s design temperature should be based on the minimum storage fluid temperature or the lowest ambient air temperature. The exterior of the condensate tanks are typically coated or painted for corrosion protection and designed with a corrosion allowance per code and regulation.

The two standards commonly used for tank design are British Standard BS 14015 “Standards on specification for manufacture of vertical steel welded non-refrigerated storage tanks with butt-welded shells for the petroleum industry,” and the American Petroleum Institute Standard API 650 “Venting Atmospheric and Low-Pressure Storage Tanks.” API standards provide design guidelines for the determination of venting requirements that may be used under normal tank operations and emergency conditions. Specially designed pressure/vacuum vent valves should be provided to protect the tank against overpressure or vacuum conditions as follows:

**Vacuum conditions:**

- Withdrawal of product from the tank
- Contraction of vapors caused by a drop in atmospheric temperature.

**Pressure buildup conditions:**

- Vapor generation when condensate is letdown from a high-pressure source
- Vapor displacement by product into the tank, product evaporation and heat of mixing with dissimilar liquid in the tank
Expansion and evaporation caused by a rise in atmospheric temperature

Fire exposure.

Tanks should be configured to operate with a nitrogen or fuel-gas blanketing system that maintains the tank at positive pressures under all operating conditions.

Tank vent piping should include flame arrestors, which protect the tank against ignition of the vent gases due to lightning strike or a discharge of static electricity at the vent location. Where the vent piping is routed to a flare system, a constant bleed of purge gas into the vent is required in addition to a flame arrestor. More complex flow devices, such as fluidic seals and molecular seals, are available from several manufacturers to minimize the amount of purge gas needed to avoid hazardous conditions occurring in the storage tank.

Direct acting pressure/vacuum relief valves are special types of relief valves, specifically designed for tank protection. They can be designed for pressure relief only, vacuum only, and combined pressure/vacuum. Pressure and vacuum protection levels are controlled with weighted pallets or springs and can be adjusted to provide the required pressure/vacuum settings. It is common to combine pallet and spring systems in one unit, i.e., pressure settings require a spring section, while the vacuum settings use the pallet method.

![Schematic of tank emission control system.](image-url)
5.5.2 Tank emission control

Emissions of hydrocarbons from condensate storage tanks are sources of greenhouse gas pollutants. The tank emissions can be minimized with a vapor recovery system, as shown in Figure 5-8.

Storage tank pressure is controlled by at least two sets of pressure-regulating valves within a set pressure range. When the tank pressure drops below the set pressure, such as during unloading of the condensate, the first pressure control valve is opened introducing vapor (nitrogen or fuel gas) to the tank. When the tank pressure goes up above the set point, such as introduction of condensate to the tank, the second pressure control valve will open, sending the excess vapor to the vapor recovery system. These two pressure control valves operate using split-range control logic to avoid simultaneous opening of both valves. To handle the displacement vapor during truck unloading, a balance line between the truck and the tank can be used to equalize the pressure.
Natural Gas Treating

6.1 Introduction

Natural gas, which consists of mainly methane and light hydrocarbons, also contains acid gases such as hydrogen sulfide (H₂S) and carbon dioxide (CO₂). In addition to acid gases, natural gas may contain other sulfur contaminants such as mercaptans (R–SH), carbonyl sulfide (COS), and carbon disulfide (CS₂). Natural gas with H₂S or other sulfur compounds is called sour gas, whereas gas with only CO₂ is called sweet gas. Sour gas can cause corrosion to natural gas processing equipment and pipelines. Combustion of sulfur compounds produces sulfur oxide air pollutants that must be limited to protect the environment and to prevent health-related problems. Carbon dioxide is an inert gas that does not have any heating value, but it is the main source of greenhouse gas responsible for climate changes.

The main challenge in today’s natural gas treating units is to remove the high concentration of CO₂ from sour gas and the sulfur components to meet stringent emission standards. A number of methods are available for the removal of acid gases. Some are suitable for bulk acid gas removal. Some are effective in total acid gas removal but ineffective in the removal of organic sulfur compounds such as mercaptans, disulfides, or carbonyl sulfide. These different methods are discussed in this chapter. The objective of this chapter is to review the basic concepts and discuss some of today’s gas treating problems, but it is not intended to cover every aspect of gas treating. For more in-depth knowledge, the authors recommend the book *Gas Purification* by Kohl and Nielsen (1997).

6.2 Gas treating specifications

When natural gas is used by consumers, the gas must be treated to meet pipeline gas specifications. Detailed specifications are given in Table 3-1 in Chapter 3.

For NGL recovery plants, CO₂ removal is required to avoid CO₂ from freezing in the cold sections of the plants. Natural gas liquefaction plant typically requires CO₂ content to be less than 50 ppmv. For coal gasification plants, carbon capture of 90–95% of the CO₂ content for sequestration may be necessary to meet the greenhouse gas mandate.
6.3 Gas treating processes

The natural gas treating processes can be grouped in several categories as shown in Figure 6-1. If the sulfur content is relatively low, direct conversion processes are an option. These processes use an alkaline solution for absorption of $\text{H}_2\text{S}$ and then a chelating agent and oxidation with air to form elemental sulfur. Details of this process are discussed in Chapter 9.

There are two types of indirect acid gas removal: adsorption and absorption. Adsorption is a physicochemical phenomenon, in which impurities of the gas are trapped and removed physically or chemically by the surface of a selective solid. Absorption is the removal of acid gases by physical absorption by physical solvent toward acid gases or by chemical reaction with the acid gases using a chemical alkaline solvent.

The dry-bed processes, which use a fixed bed of solid materials to remove acid gases through either ionic bonding or chemical reactions, can be divided into two categories: physical adsorption (i.e., molecular sieve process) and chemical absorption.

The gas treating processes shown in Figure 6-1 will be discussed in the following sections.
6.4 Chemical absorption processes

The chemical absorption processes can be broadly classified into two groups:

- Alkanolamine solution
- Potassium carbonate solution

In chemical solvent processes, absorption of the acid gases is achieved mainly by the use of alkanolamines or alkaline salts of various weak acids such as sodium and potassium salts of carbonate. Regeneration (desorption) can be brought about by use of reduced pressures and/or high temperatures, whereby the acid gases are stripped from the solvent. Chemical absorption processes chemically absorb the H₂S, CO₂, and to some extent COS. Organic sulfur components do not chemically react with the solvent. Chemical solvents are specifically suitable when contaminants at relatively low partial pressure have to be removed to very low concentrations. Chemical solvents will not remove mercaptans down to low levels due to the low solubility of these components. An advantage, however, is that there is minimum coabsorption of hydrocarbons.

6.4.1 Alkanolamine solvents

The chemical solvent process using the various alkanolamines is the common gas treating process. Amines are well suited for applications where the acid gas partial pressures are low and low acid gas contents are required on the treated gas.

The application of alkanolamines as solvents for acid gas removal began in 1930. In earlier applications, these amines were represented by monoethanolamine (MEA) and diethanolamine (DEA). Later, other amines like diglycolamine (DGA) and diisopropanolamine (DIPA) gained commercial recognition. In recent years, various methyldiethanolamine (MDEA) and specialty solvents blends were developed which take advantage of the low-energy characteristics of MDEA.

Amines are compounds formed from ammonia (NH₃) by replacing one or more of the hydrogen atoms with another hydrocarbon group. Amines are categorized as primary, secondary, and tertiary, depending upon the degree of substitution of the hydrogen atoms by organic groups. Replacement of a single hydrogen atom produces a primary amine (MEA, DGA), replacement of two hydrogen atoms produces a secondary amine (DEA, DIPA), and replacement of all three hydrogen atoms produces a tertiary amine (MDEA).

The chemical solvent processes are characterized by a relatively high heat of acid gas absorption. A substantial amount of heat is required for solvent regeneration. Amines are thermally stable, but are sensitive to degradation by oxidation and high temperatures, and may have various side reactions with CO₂, COS, and CS₂. The degraded amine can be recovered thermally or by ion exchange, as specified by the solvent suppliers.
Acid gas loading of amines is not very sensitive to changes in acid gas partial pressure as compared to physical solvents. While acid gas loading of a physical solvent increases with partial pressure of the acid gases, amine acid gas loading stays fairly constant regardless of the operating pressure and cannot take advantage of the high operating pressure.

Different amines have different reaction rates with respect to the various acid gases. In addition, different amines vary in their equilibrium absorption characteristics for the various acid gases and have different sensitivities with respect to solvent stability and corrosion characteristics. Some of the specific details for the amine processes are discussed in the followings.

**Monoethanolamine**: MEA is a primary amine, and is the most aggressive amine among the group. MEA is also the lowest cost solvent.

MEA can remove both H$_2$S and CO$_2$ from gas streams to meet sales gas specifications. MEA solvent is very corrosive at high acid gas loadings and high solution concentrations. It reacts irreversibly with carbonyl sulfide (COS) and carbon disulfide (CS$_2$), which can degrade the solvent and form heat-stable salts. MEA has a higher vapor pressure than other amines, and the high equilibrium losses can be a problem. Amine losses can typically be minimized using a water wash section in the upper section of the amine absorber.

**Diglycolamine**: DGA was developed in the late 1960s by Fluor Corporation and Jefferson Chemical Company, a predecessor to Texaco Chemical and later Huntsman Corporation. The advantage of DGA in gas treating is that it can operate with air cooling in hot climate locations, and was chosen as the sole treating solvent for the Saudi Aramco gas projects in the 1970s. Fluor patented this technology as the Fluor Econamine process that includes a side cooler on the amine absorber.

As a primary amine, DGA is similar in many respects to MEA except that it exhibits lower vapor pressure, lower solution corrosion tendency, and improved solution properties and can operate at higher concentrations than MEA. The higher operating concentration (up to 60 wt%) results in significantly lower circulation rates and energy consumption.

DGA can remove COS, CS$_2$, and (partially) mercaptans but also forms degradation products that are not reversible at normal amine regenerator temperature. Degradation reactions produce $N,N'$,bis-(hydroxyethoxyethyl) urea, and $N,N'$,bis-(hydroxyethoxyethyl)thiourea. To maintain the amine activity, thermal reclaiming of a slip stream of the circulating amine is required.

**Diethanolamine**: DEA is a secondary amine. Its amine reactivity and corrosivity are lower than those of primary amines. The vapor pressure and heat of reaction are also lower. DEA is a common amine used for H$_2$S and CO$_2$ removal in refineries due to its stabilities with contaminants in the refinery gas streams. DEA can partially remove COS and CS$_2$. However, the reaction rate of DEA with COS and CS$_2$ is lower than with MEA. The solution strength of DEA solvent is typically 25–30 wt%.
When used for high-pressure acid gas removal, the Societe Nationale des Petroles d’Aquitaine (SNPA)-DEA process (developed by SNPA, France, in the gas field at Lacq) has success in meeting the pipeline specifications (Kohl and Nielsen, 1997).

Diisopropanolamine: DIPA is a secondary amine and its reactivity with acid gases is similar to that of DEA. DIPA is more widely used in gas processing industries in Europe and Asia. As it also has some steric hindrance, it can be used selectively either to remove H₂S (at low pressures) or to remove both CO₂ and H₂S (at higher pressures). DIPA can be used to partially remove COS and CS₂. The solution strength of DIPA solvent is typically 20–40 wt%.

Methyldiethanolamine: MDEA, a tertiary amine, is the most widely used gas treating agent today. Unlike primary and secondary amines, MDEA cannot react with CO₂ by the carbamate reaction. It can only absorb CO₂ by the slow bicarbonate formation. This property allows MDEA to selectively remove H₂S when treating a gas stream containing both H₂S and CO₂. The slippage of CO₂ opens up more capacities in debottlenecking gas treating units. By reducing CO₂ absorption, more solvent is available for H₂S removal. Other advantages of MDEA solvent include low vapor pressure and solution losses, low energy for solvent regeneration, low corrosiveness, and resistance to degradation. However, generic MDEA process would require low absorption temperature to improve the amine–H₂S equilibrium for absorption in order to meet low H₂S specification. Alternatively, MDEA that is acidified with phosphoric, sulfuric, or other acids, can be regenerated more easily, and can be used to produce an ultralow H₂S content lean amine for treating. However, the level of acidification has a certain limit, as there is an impact on the pH value of the amine solution and the equilibrium curve of the acid gas solubility in the solvent (Vorberg et al., 2010).

Generic MDEA has minimal COS and CS₂ removal capability. The solution strength of MDEA solvents is 35–50 wt%.

To meet stringent emissions requirements, and to take advantage of the low energy consumption of MDEA, MDEA can be blended with chemical promoters to meet the treating requirements. Promoters work by a shuttle mechanism and affect thermodynamics, but more importantly, they can control the reactivity with CO₂ (Weiland et al., 2003). Varying the concentration of chemical activators such as piperazine offers one of the solutions to meet the acid gas specifications in H₂S and CO₂. With other proprietary additives, they can be used for removal of mercaptans and COS and other contaminants. Some of the proprietary formulated technology suppliers include:

- aMDEA (by BASF)
- ADIP X (by Shell)
- AdvAmine (by Prosernat)
- GAS/SPEC (by Ineos)
- Jefftreat (by Huntsman)
- UCARSOL (by Dow Chemical Company)
Sterically hindered amines: Sterically hindered amines are compounds in which the nitrogen atom of the amine molecule is partially shielded by neighboring groups so that larger molecules cannot easily approach and react with the nitrogen.

Hindered amine concept is based on the reaction rates of the acid gases with different amine molecules. In the case of CO₂ removal, the capacity of the solvent can be greatly enhanced if one of the intermediate reactions, i.e., the carbamate formation reaction, can be slowed down by providing steric hindrance to the reacting CO₂. In addition to slowing down the overall reaction, bulkier substitutes give rise to less stable carbamates. By making the amine carbamate unstable, one can theoretically double the capacity of the solvent (Chakma, 1994).

The molecular configuration of hindered amine dictates the amount of CO₂ slippage while maintaining high H₂S removal. Moderately hindered amines are characterized by high rates of CO₂ absorption and high capacities for CO₂. Moderately hindered amine may be used with physical solvents for simultaneous removal of CO₂ and H₂S from synthesis gas and natural gas. A severely hindered amine is characterized by a very low rate of CO₂ absorption, and exhibits high solvent capacity, which is typically used in tail gas unit.

ExxonMobil developed the FLEXSORB SE process, which is based on a family of proprietary sterically hindered amines. The FLEXSORB SE process removes H₂S selectively, or removes a group of acidic impurities (H₂S, CO₂, COS, CS₂, and mercaptans) from a variety of gas streams, depending on the selected solvent. One version of the solvent, FLEXSORB SE Plus, is used to selectively remove H₂S as well as organic sulfur impurities commonly found in natural gas. It is claimed that FLEXSORB SE Plus in these services requires lower circulation rate than other amines (Garrison et al., 2002).

6.4.1.1 Amine processes
The basic flow scheme of amine treating units has remained unchanged throughout the years. Amine unit requires steam for solvent regeneration and is a major steam consumer in a gas plant. In some designs, the amine regenerator is actually used as a heat rejection method to get rid of excess steam in the gas plant. A picture of a DGA unit is shown Figure 6-2.

A typical process flow diagram of the amine unit is shown in Figure 6-3. Typically, sour feed gas to the amine unit is filtered and cleaned of entrained liquids and pipe scales using an inlet filter separator. The separator gas is fed to the amine absorber where the acid gas content is removed by an aqueous amine solution, producing a rich amine at the bottom and a H₂S-depleted gas from the top. Depending on the applications some CO₂, as much as 2 mol%, can be left in the treated gas if the gas is used as pipeline gas. If the gas is sent to an NGL recovery unit or an LNG production plant, deeper CO₂ removal is required.

The rich amine is let down in pressure to the rich amine flash drum. The rich amine flash drum is a three-phase separator designed for hydrocarbon liquids and amine separation. Hydrocarbon liquids are removed from the flash drum using a skimming device to avoid buildup that would cause foaming. The flashed gas containing some H₂S is treated using a small packed section and recovered as fuel gas.
The flashed amine is preheated in the lean/rich exchanger, typically a plate and frame heat exchanger, to about 200–215 °F, before entering the regenerator. The regenerator typically operates at about 10–15 psig producing an acid gas to the sulfur recovery unit. The rich amine is heated in the regenerator and stripped with steam, producing a lean amine. The lean amine is cooled in the lean/rich exchanger and lean amine cooler and filtered prior to returning to the absorber. The filter system typically includes a particulate filter for removal of corrosion materials and a carbon filter for removal of heavy hydrocarbons. The filtration system is typically installed on the lean amine side, which is easier for maintenance as the system is free of H₂S. Alternatively, it can also be installed on the rich amine side, which has the advantage of protecting the amine regenerator from fouling. In this design, proper operating procedure must be exercised because of the H₂S content in the rich amine.

The amine storage tank acts as a surge for the amine system, which dampens out flow fluctuation, maintaining a constant circulation. The storage tank is typically designed to hold the volume of the entire amine inventory. The amine tank content should be periodically checked for active amine and purity, and should be adjusted by makeup as necessary.

If the feed gas contains ammonia, it will be stripped off in the overhead system, which will combine with acid gases to form ammonia salt, such as ammonia bisulfate. Excessive buildup of ammonia salt will cause severe corrosion, causing hydrogen blistering on piping and exchangers. The ammonia content in the overhead system should be maintained at below 1 wt% by purging the reflux water.

FIGURE 6-2
DGA unit (Courtesy of Fluor).
FIGURE 6-3

Typical flow schematic of an amine treating unit.
There are variations of the basic flow scheme, which may include lean and semilean solvent circuits, additional flash regeneration stages, multiple absorbers, and H₂S concentrator. The complexity depends on the solvent used, the acid gas concentration, operating pressure and temperature, and the product gas specifications. Some of these process variations are discussed in the following sections.

Two-stage absorption process

The two-stage amine absorption process is applicable to MDEA solvent, which can be partially regenerated by pressure letdown, reducing the overall steam consumption. The two-stage MDEA process has been used in treating high-CO₂-content synthesis gas in ammonia plants. Synthesis gas mainly consists of hydrogen, CO and CO₂, and H₂S and does not contain any hydrocarbon. For this reason, there is typically no foaming problem in synthesis gas plants. On the other hand, there are problems with a two-stage MDEA process in natural gas plants. This is typically due to adequate solvent regeneration, which results in buildup of hydrocarbons in the amine circuit, resulting in foaming and failure to meet specifications. For this reason, the two-stage amine unit is seldom used in natural gas processing plants.

A typical process configuration of a two-stage process is shown in Figure 6-4. The amine absorber consists of two absorption sections, a top lean amine section and a lower semilean solvent section. The lean amine is produced by steam

![FIGURE 6-4](image)

Flow schematic of the two-stage MDEA process. CW, Cooling Water.
stripping and the semilean solvent is produced by flash regeneration. The main purpose of the two-stage pressure is to reduce steam consumption using the semilean amine. The ratio of lean and semilean solvent flow is an important parameter in designing a two-stage MDEA system. Increasing the lean to semilean ratio will increase the steam consumption but would enhance stripping of hydrocarbons, and can produce a lean amine to meet a more stringent CO₂ specification. Lowering the ratio has the opposite effect. It would lower steam consumption but may lead to inadequate stripping, resulting in foaming and failure to meet specifications. The design must consider the different feed gas conditions and compositions, and contaminant levels. Operating experience is important in the design of a two-stage MDEA unit as the performance depends on the reaction kinetics of CO₂ absorption, which is difficult to accurately predict even with a rate-based simulator.

The two-stage MDEA process has a higher equipment count than a single-stage MDEA process. The higher equipment cost must be justified by the reduction in operating cost. In summary, the use of a two-stage process may be justified for a feed gas with low hydrocarbon content, and where steam supply is limited.

Double absorption process

The double absorption process is used for enriching the H₂S content in the feed gas to operate a straight-through Claus sulfur recovery unit.

When the gas plant is used to treat a high-CO₂-content feed gas, the H₂S content in the acid gas from the amine unit frequently falls below 10%, which makes operating a conventional Claus sulfur recovery unit difficult. The acid gas can be processed in an innovative method using a patented double absorption process as shown in Figure 6-5.

The process uses a selective solvent, such as MDEA, to process a dilute acid gas to produce two gas streams via selective absorption of H₂S. One gas stream is predominantly composed of CO₂, which can be sent to a thermal oxidizer or incinerator, prior to being discharged to the atmosphere. The other gas stream is enriched in H₂S, which can be processed in a Claus sulfur recovery unit.

The process uses two absorbers; a primary and a secondary absorber, to separately treat the dilute feed gas and the regenerator overhead recycle gas. The dilute acid gas, typically at slightly above atmosphere pressure, is first contacted with a lean MDEA in the primary absorber, which selectively remove H₂S, while rejecting CO₂ to the overhead. The overhead gas can meet 150 ppm H₂S specification, as required to meet sulfur emission requirements. The rich solvent is pumped to a higher pressure to a regenerator operating at 10–15 psig.

To enrich the acid gas to the Claus unit, a portion of the regenerator overhead is recycled back to the secondary absorber. Similar to the primary absorber, CO₂ is rejected from the secondary absorber, concentrating the H₂S content in the rich amine, which indirectly increases the H₂S content in the regenerator overhead.
gas. The H_2S content in overhead gas can be increased by increasing the recycle flow, and can reach over 60% as needed for optimum Claus unit operation.

This double absorption process can also be integrated with the tail gas treating unit to achieve over 99.9% sulfur recovery.

### 6.4.1.2 Special design considerations

General equipment design guidelines of an amine treating unit can be found in the GPSA Engineering Data Book, (2004). The following are the design options that are normally not covered in typical amine unit design.

**Water wash trays**: When the absorber operates at high ambient temperature, the vaporization loss of amine can be significant. To minimize amine losses, a water wash section can be installed on the upper section of the amine absorber. Typically, a water wash pump and three to four scrubber trays are required. The water wash rate is designed to meet the minimum tray liquid loading hydraulic requirement.

**Feed gas and amine temperatures**: To avoid hydrocarbon condensation, lean amine must be controlled at 10–15 °F above the feed temperature. However, if the plant is located in desert areas, feed gas temperature can be as high as 140 °F, which means that the lean amine temperatures must be kept at 150–155 °F. The high amine temperature has two undesirable effects. First, there will be an equilibrium pinch at the top of the absorber resulting in the treated gas not meeting the H_2S specification. Second, there will also be a pinch at the absorber bottom due to high operating temperature, which would require a higher solvent circulation to reduce the absorption temperature.
In the hot climate applications, it may be more cost effective to cool the feed gas first to reduce the inlet temperature, and then use a chilled lean amine to avoid the temperature pinches. Typically, a chilled water system is available for process and utility cooling.

**Lean amine feed locations:** Multiple lean amine feed points can be installed in the absorber to provide the flexibility for selective CO₂ removal. When producing a pipeline gas, the feed location can be lowered to minimize CO₂ absorption as the pipeline gas specification is only 2 mol% CO₂. This is particularly useful in tail gas treating unit operation, which requires maximum slippage of CO₂, and moving the feed location to a lower location will help to concentrate the H₂S content in the acid gas to the Claus sulfur recovery unit.

However, if stringent sulfur specification must be met, the lean amine must be raised to a higher tray location, which would add more contacting stages, but would also reduce the amount of CO₂ slippage.

**Absorber side cooler:** When amine is used to treat high-partial-pressure acid gases, especially CO₂, the heat of reaction would be significant, which would raise the temperature of the absorber. The amount of acid gas that can be absorbed by the amine is determined by the rich amine acid gas loading at the bottom, which is a function of the operating temperature. In a conventional amine unit design, the absorber bottom temperature is maintained at about 200 °F (to minimize corrosion) by increasing the amine circulation. Therefore to treat a high-CO₂-content feed gas, a high amine circulation is required.

In treating high-CO₂ gases, a more efficient design is to install a side cooler in the lower section of the absorber, as shown in Figure 6-6. The hot semi-rich amine is withdrawn using a chimney tray in the absorber, pumped, cooled, and

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**FIGURE 6-6**
Absorber side cooler schematic. TC, Temperature Controller.
returned to the absorber. With this configuration, the absorber can be maintained at a relatively low temperature without increasing amine circulation.

The temperature profile of the absorber with the side cooler is shown in Figure 6-7. The side cooler should be located where the peak absorption temperature occurs, typically four trays from the bottom. With the side cooler design, the absorber bottom temperature can be controlled using a bypass on the heat exchanger.

**Design guidelines:** The operating parameters of an amine process vary with the type of the amine as well as the feed gas conditions. The main differences are the amine strengths, which depend on the physicochemical properties and the corrosion characteristics. Table 6-1 summarizes some of the typical design parameters for the different amines. Note that this table is intended to be used as a general guide, and the users must also consider other parameters such as operating pressure and temperature, feed gas, and product gas specifications.

The amine circulation rate depends on the acid gas loading of the rich amine, which, in turn, is a function of the amine–acid gases equilibrium at the absorber bottom condition. A higher acid gas loading means that the amine can pick up more acid gases per gallon of solution, requiring a lower amine circulation. Typically, the absorber bottom is designed at no more than 80% approach to equilibrium. The absorber contains at least 20 absorption trays, and more trays may be required to meet the treated gas specifications, which must be determined by simulation of the amine unit.

Low-pressure steam must be supplied for solvent regeneration to produce a lean amine suitable for treating. Steam consumption is typically defined as the mole of steam per mole of acid gas (typically 1.0) or pounds of steam per gallon of amine circulation (1–1.2), as shown in Table 6-1. Steam consumption is higher for the primary amines (MEA and DGA) than the tertiary amine

![Absorber temperature profile with side cooler.](image)
MDEA), which is mainly due to the differences in heat of reaction with acid gases.

Amine concentrations are set by unit operating experience considering the solvent stability, foaming tendency, corrosion rates, and unit performance. Higher concentration can reduce amine circulation, but would also increase the absorption temperature and corrosion rate. The typical amine concentrations for the different amines are shown in Table 6-1. For example, DGA can be operated at high concentration as it is proven to be stable and in fact, the high temperature enhances DGA’s ability to remove COS.

### 6.4.1.3 Amine unit operating problems

Amine unit operating problems have been well documented and published. New amine units can typically meet the performance guarantees and have no operating problems in early operation. However, unit performance generally deteriorates over time, mainly due to the buildup of contaminants and impurities, and lack of maintenance. The end results are failure to meet product specifications and reduction in treating capacity.

Absorber and regenerator typically experience foaming problems, resulting in carryover of amine to downstream units. Foaming is caused by hydrocarbon entrainment in the feed gas. Hydrocarbon contamination can be mitigated by maintaining a temperature approach between the lean amine and the feed gas, skimming of hydrocarbons from the rich amine flash drum and reflux drum, and frequent change-out of particulate filters and carbon filters.

### Table 6-1 Design Parameters of Generic Amines

<table>
<thead>
<tr>
<th>Solvent</th>
<th>MEA</th>
<th>DGA</th>
<th>DEA</th>
<th>DIPA</th>
<th>MDEA</th>
</tr>
</thead>
<tbody>
<tr>
<td>Typical concentration, wt%</td>
<td>15–20</td>
<td>45–60</td>
<td>25–30</td>
<td>30–40</td>
<td>35–50</td>
</tr>
<tr>
<td>Typical lean loading, mole/mole</td>
<td>0.1–0.15</td>
<td>0.05–0.1</td>
<td>0.05–0.07</td>
<td>0.02–0.05</td>
<td>0.004–0.01</td>
</tr>
<tr>
<td>Typical rich loading, mole/mole</td>
<td>0.30–0.35</td>
<td>0.35–0.45</td>
<td>0.35–0.40</td>
<td>0.30–0.40</td>
<td>0.45–0.55</td>
</tr>
<tr>
<td>Typical steam usage, lb/gal</td>
<td>1.0–1.2</td>
<td>1.0–1.2</td>
<td>0.9–1.1</td>
<td>0.8–1.1</td>
<td>0.9–1.1</td>
</tr>
<tr>
<td>Heat of reaction with CO₂, Btu/lb</td>
<td>825</td>
<td>850</td>
<td>653</td>
<td>550</td>
<td>475</td>
</tr>
<tr>
<td>Heat of reaction with H₂S, Btu/lb</td>
<td>820</td>
<td>674</td>
<td>511</td>
<td>475</td>
<td>455</td>
</tr>
</tbody>
</table>
Foaming can also be caused by heat-stable salts, which are formed from amine degradation. There are different methods that can be used to control the levels of heat-stable salts, including purging, caustic addition, and reclamation. Continuous amine reclaiming is considered more effective in solving the heat-stable-salt-related problems. Formation of heat-stable salts depends on the feed gas compositions, particularly sulfur contaminants, and the H₂S to CO₂ ratio. Most amine units can operate without reclaiming for some time, especially with low H₂S content gases. Most amine suppliers are knowledgeable of the heat-stable salt problems and can provide the reclaiming equipment to maintain the amine purity. For more detailed discussion on the operating issues and troubleshooting methods, please refer to technical papers such as by Seagraves et al. (2010) and the Laurance Reid Gas Conditioning Conference (LRGCC) proceedings.

6.4.2 The potassium carbonate solution

The potassium carbonate solution is a vintage treating solvent. In the potassium carbonate process, which is referred to as the “hot pot” process, the process system operates at high temperature (in the range of 230–240 °F) and was very popular in CO₂ removal in ammonia plant. Potassium carbonate is known to cause equipment failure from excessive corrosion. All carbon steel must be stress relieved. A variety of corrosion inhibitors are also available to combat corrosion.

Potassium carbonate solvent absorbs CO₂ at a relatively slow rate. The reaction rates are increased through the use of catalytic promoters, such as DEA, arsenic trioxide, selenous acid, and tellurous acid. In applications for the removal of hydrogen sulfide, tripotassium phosphate may be added. These activators are claimed by the licensors to improve the potassium carbonate performance.

There are three potassium carbonate processes using different additives: Catacarb, Vetrocoke, and Benfield.

The Catacarb process uses amine borates additive and vanadium-based inhibitors to mitigate the corrosion, which are toxic materials.

The Vetrocoke process utilizes organic and inorganic additives to activate the solution. One of the common corrosion inhibitors is arsenic trioxide, which, however, is also a toxic material.

The Benfield Process, licensed by UOP, uses the less toxic DEA as an activator. A variant of the Benfield process is the DEA-hot carbonate process, which consists of bulk CO₂ removal combined with a downstream DEA polishing unit. The UOP Benfield ACT-1™ activator is another activator that is claimed to have better performance. A typical flow scheme for the UOP Benfield process is shown in Figure 6-8.

Potassium carbonate units are slowly replaced by promoted MDEA units for CO₂ removal. However, there are still many existing potassium carbonate units that may benefit using the less toxic activators.
6.5 Physical solvent processes

Physical solvent processes can remove acid gases and organic sulfur based on physical absorption, and no chemical reaction occurs.

Physical solvents have an advantage over chemical solvents when treating sour feed gas streams with high acid gas content. The physical solvent’s acid gas holding capacity increases with the partial pressure of acid gas (Henry’s law). Thus, they are more economical than chemical solvents in treating high-pressure high-CO₂-content gas.

Unlike chemical solvents, physical solvents operate at ambient to subambient temperature. They are noncorrosive and do not require stainless-steel material to combat corrosion. Unless operating at cryogenic temperatures, only carbon steel is required.

Physical solvents can be regenerated by reduction in solvent pressure. Compared to chemical solvents, physical solvents absorb more hydrocarbons, and have high hydrocarbon losses. For this reason, physical solvents are more common in treating synthesis gas, which has no hydrocarbons. When used in natural gas treating applications, flash gas compressor recycle is required to minimize the hydrocarbon losses.
Some of the more common physical solvents used in gas treating are:

- Propylene carbonate (PC)
- Dimethyl ether of polyethylene glycol (DEPG)
- Methanol (MeOH)
- N-Methyl-2-pyrrolidone (NMP)

These solvents are used to remove CO2 from high-pressure sour gases. During the absorption processes, they also coabsorb other components. The solubility of the various components in relationship to CO2 is shown in Table 6-2 (Burr and Lyddon, 2008). To minimize hydrocarbon losses, the solvent that has the lowest solubility with methane and ethane are preferred.

PC and DEPG solvents have been used for both natural gas and syngas treating units. Among all the physical solvents, PC exhibits the lowest hydrocarbon solubility, which means a lower hydrocarbon loss in the CO2 stream; hence PC is more suitable for treating natural gas streams than other physical solvents.

Methanol and NMP solvents are common in syngas treating units. They are seldom used in natural gas treating units mainly due to the high solubility of

### Table 6-2 Relative Solubility of Components in Physical Solvents (Burr and Lyddon, 2008)

<table>
<thead>
<tr>
<th>Gas Component</th>
<th>PC at 25 °C</th>
<th>DEPG at 25 °C</th>
<th>NMP at 25 °C</th>
<th>MeOH at –25 °C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon dioxide</td>
<td>1.0</td>
<td>1.0</td>
<td>1.0</td>
<td>1.0</td>
</tr>
<tr>
<td>Hydrogen sulfide</td>
<td>3.29</td>
<td>8.82</td>
<td>10.2</td>
<td>7.06</td>
</tr>
<tr>
<td>Carbonyl sulfide</td>
<td>1.88</td>
<td>2.30</td>
<td>2.72</td>
<td>3.92</td>
</tr>
<tr>
<td>Methyl mercaptans</td>
<td>27.2</td>
<td>22.4</td>
<td>34.0</td>
<td>–</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>0.0078</td>
<td>0.013</td>
<td>0.0064</td>
<td>0.0054</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.0084</td>
<td>0.020</td>
<td>–</td>
<td>0.012</td>
</tr>
<tr>
<td>Oxygen</td>
<td>0.026</td>
<td>–</td>
<td>0.035</td>
<td>0.020</td>
</tr>
<tr>
<td>Carbon monoxide</td>
<td>0.021</td>
<td>0.028</td>
<td>0.021</td>
<td>0.020</td>
</tr>
<tr>
<td>Methane</td>
<td>0.038</td>
<td>0.066</td>
<td>0.072</td>
<td>0.051</td>
</tr>
<tr>
<td>Ethane</td>
<td>0.17</td>
<td>0.42</td>
<td>0.38</td>
<td>0.42</td>
</tr>
<tr>
<td>Ethylene</td>
<td>0.35</td>
<td>0.47</td>
<td>0.55</td>
<td>0.46</td>
</tr>
<tr>
<td>Propane</td>
<td>0.51</td>
<td>1.01</td>
<td>1.07</td>
<td>2.35</td>
</tr>
<tr>
<td>i-Butane</td>
<td>1.13</td>
<td>1.84</td>
<td>2.21</td>
<td>–</td>
</tr>
<tr>
<td>n-Butane</td>
<td>1.75</td>
<td>2.37</td>
<td>3.48</td>
<td>–</td>
</tr>
<tr>
<td>i-Pentane</td>
<td>3.50</td>
<td>4.47</td>
<td>–</td>
<td>–</td>
</tr>
<tr>
<td>n-Pentane</td>
<td>5.0</td>
<td>5.46</td>
<td>–</td>
<td>–</td>
</tr>
<tr>
<td>n-Hexane</td>
<td>13.5</td>
<td>11.0</td>
<td>42.7</td>
<td>–</td>
</tr>
<tr>
<td>n-Heptane</td>
<td>29.2</td>
<td>23.7</td>
<td>50.0</td>
<td>–</td>
</tr>
</tbody>
</table>
hydrocarbons, as shown in Table 6-2. They will coabsorb most of the propane and heavier hydrocarbons, resulting in high product losses.

Unlike amine processes, physical solvent process configurations are much more complex. Physical solvent processes require proven design and engineering skill and unit operating data, coupled with accurate thermodynamic data and heat and mass transfer model that are only available from licensors. The process configurations must be customized for specific feed gas compositions, operating temperatures and pressures, and product specifications.

6.5.1 Propylene carbonate

PC, C₄H₆O₃, is a polar solvent. As discussed in the previous section, PC has the lowest solubility of hydrocarbons compared to other solvents. Among all the solvents, PC has the least H₂S selectivity, and is not as effective for H₂S removal. In terms of regeneration, PC requires the lowest regeneration energy, and it can be regenerated without the use of heat.

The FLUOR Solvent process that uses PC as the solvent was commercialized by Fluor in the early 1960s. The FLUOR Solvent process is a refrigerated solvent process, operating at low temperatures (10 to −20 °F), which has several advantages. At low temperatures, the PC unit can operate with a lower solvent circulation, requiring less power and a lower capital investment than competing processes. The freezing point of PC is −57 °F, which makes PC an ideal solvent for cold climate operation.

PC is often compared to formulated or promoted MDEA for bulk CO₂ removal. Bulk CO₂ removal is typically required to meet pipeline gas specification, typically 2 mol%. Both processes have been proven suitable in meeting pipeline specifications. However, the unit complexity, process performance, and design characteristics are very different, which can be compared in Table 6-3.

As can be seen in the comparison table, when used to treat a lean gas with low H₂S content, PC has a clear advantage over MDEA. It can produce a dry gas, and does not require the use of steam or heating. No water makeup is required and no solvent monitoring is necessary. These advantages are important in offshore applications where limited support and operating personnel are available. The use of PC has a competitive advantage over MDEA when the CO₂ partial pressure is greater than 60 psi.

The following sections describe the development of the FLUOR Solvent process from the original FLUOR Solvent unit to the new innovations that are applicable in today’s high CO₂ gas plants.

6.5.1.1 FLUOR solvent unit

The first FLUOR Solvent unit is located in Terrell County, Texas. The PC unit was built in 1960 and has been operating successfully for the past 50 years, using the same equipment, without any operating problems.
The unit was originally designed for 220 MMscfd of feed gas with 53 mol% CO₂ at about 900 psig pressure. Over the years, the feed gas pressure and CO₂ content have declined. The unit is now treating 120 MMscfd of feed gas with about 36 mol% CO₂. The Terrell County process flow schematic is shown in Figure 6-9. The unit picture is shown in Figure 6-10.

The unit feed gas is treated in the absorbers with about 3000 gpm of lean PC. The rich solvent from the absorber is cooled by the absorber overhead gas and the CO₂ vent stream. The cooled solvent is regenerated by successive letdown in pressure to 445 psig and 175 psig. The flash gas, typically containing about 65% methane, is compressed and recycled back to the absorber for hydrocarbon recovery. The rich solvent is letdown in pressure using the hydraulic turbines, which generate refrigeration for solvent cooling and shaft power to operate the circulation pumps.

To comply with emissions regulations, the flash vapor from the medium pressure stage is treated in a reabsorber, which recovers the hydrocarbon contents in the solvent. The methane content in the CO₂ vent is reduced to about 1 mol%, which is required to meet the local methane emission limit. The vacuum pressure in the last flash stage is maintained by a vacuum compressor.

The FLUOR Solvent unit produces sales gas with a CO₂ content of about 2 mol%. However, H₂S in the treated gas is 6 ppmv, which is slightly higher

<table>
<thead>
<tr>
<th>Gas Treating Process</th>
<th>FLUOR Solvent</th>
<th>Formulated MDEA</th>
</tr>
</thead>
<tbody>
<tr>
<td>Equipment count</td>
<td>Low</td>
<td>High</td>
</tr>
<tr>
<td>Operational complexity</td>
<td>Low</td>
<td>High</td>
</tr>
<tr>
<td>H₂S removal</td>
<td>Limited</td>
<td>No limit</td>
</tr>
<tr>
<td>Suitability for rich gas</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Stainless steel materials</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Stress relieving of carbon steel</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Fired heater/steam required</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Produces dry treated gas</td>
<td>Yes</td>
<td>No</td>
</tr>
<tr>
<td>Ability to handle higher level of CO₂</td>
<td>Yes</td>
<td>No</td>
</tr>
<tr>
<td>Vulnerable to solvent foaming</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Winterization required</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Solvent concentration monitoring</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Hazardous solvent</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Water consumption</td>
<td>No</td>
<td>Yes</td>
</tr>
<tr>
<td>Recycle gas compressor</td>
<td>Yes</td>
<td>No</td>
</tr>
<tr>
<td>Hydrocarbon loss in CO₂</td>
<td>1%</td>
<td>Less than 1%</td>
</tr>
</tbody>
</table>

Table 6-3 Comparison of FLUOR Solvent to Formulated MDEA Processes (Mak et al., 2003)
than the 4 ppmv pipeline specification. Presently, a sulfur scavenger bed (PURASPEC) is installed to treat about 20% of the absorber overhead gas to meet the pipeline specification. The water content of the treated gas is about 1 lb water/MMscf, which meets pipeline specification.

FIGURE 6-9
FLUOR solvent process (Mak et al., 2007).

FIGURE 6-10
Terrell County FLUOR solvent unit (Mak et al., 2007).
6.5.1.2 Innovations in FLUOR solvent process

Natural gas from unconventional gas, such as coal bed methane and offshore gas fields, is typically very lean, containing almost no H₂S and a very high CO₂ content. As an example, offshore gas from South China Sea typically contains 30–60% CO₂ with almost no H₂S, and is available at 1000 psig. These gases are ideal for the FLUOR Solvent process.

For other solvents and membrane separation processes, the processing requirement increases with the CO₂ content. If the feed gas CO₂ content increases over time, additional treating units or membrane units must be added. Typically, high CO₂ gas will require more investment.

The patented FLUOR Solvent process is designed different than conventional process. High CO₂ content is actually an asset to the process. In the FLUOR Solvent process, the potential energy of CO₂ is used to produce power and cooling for the CO₂ absorption process. The development is a result of many years of research and development in the physical solvent process. With advanced modeling and accurate thermodynamic data, the FLUOR Solvent unit can operate with almost no power input and refrigeration when used in treating high-pressure high-CO₂ content gases. The basic configuration is shown in Figure 6-11. Main design features of the process are described by Mak et al. (2004), Mak (2007), and Mak et al. (2007).

The innovation is the use of the refrigeration produced by flash regeneration process to cool the absorber. The use of hydraulic turbines produces power that further generates cooling as well as power to operate the circulation pumps. The cooling of the absorber is achieved by a side cooler that cools the semirich solvent withdrawn from the absorber. With such integration, the process can operate

![FLUOR process for treating high CO₂ gas](Mak et al., 2004, 2007; Mak, 2007).
without the use of external refrigeration as most of the cooling can be achieved by the flashing of CO₂. This typically occurs when the CO₂ content in the feed gas is greater than 30%.

Similar to the original design, the FLUOR Solvent process can operate at as low as −20 °F, and water in the feed gas should be removed to avoid hydrate formation in the absorber.

The rich solvent is successively flashed to lower pressures, producing vapors which are recycled back to the absorber. The methane content in the flash vapors is recovered in the absorber. The rich solvent is finally flashed at atmospheric pressure, producing a 95% CO₂ stream that can be compressed for reinjection for enhanced oil recovery. With the improved process, hydrocarbon loss in the CO₂ stream can be reduced to less than 1%.

The solvent can be regenerated using a gas stripper, which is very efficient to produce an ultralean solvent. If H₂S has been removed upstream, dried air or nitrogen stripping can be used and can be economical. With this stripping method, the treated gas can meet 500 ppmv or lower CO₂ specification.

### 6.5.2 Dimethyl ether of polyethylene glycol

DEPG is a mixture of dimethyl ethers of polyethylene glycol. Solvents containing DEPG are manufactured by several companies including Coastal Chemical Company, Dow, or Clariant. DEPG is marketed by Dow as the Selexol® solvent and by Clariant as the Genosorb® solvent. The DEPG solvent process can be licensed from UOP as the Selexol process or from Fluor as the EconoSolv™ process.

Similar to PC, the DEPG solvent can be regenerated by successive pressure letdown and/or stripping with air or nitrogen for deeper CO₂ removal. DEPG can dehydrate feed gas and remove hydrogen cyanide (HCN), mercaptans, and heavy hydrocarbons from natural gas.

While PC can operate at 10 to −20 °F temperature, operation of DEPG is limited to 32 °F. The difficulty with DEPG operating at low temperature is the viscosity problem. At lower temperatures, DEPG viscosity will greatly increase, which will impede the heat and mass transfer operation in heat exchangers, separators, and columns, making absorption more difficult.

The other difference between the PC process and the DEPG process is the solvent concentration. PC can operate as a pure component, while the DEPG process requires maintaining 95 wt% concentration. This is necessary to maintain the regenerator bottom temperature at about 300 °F in order to avoid solvent degradation.

Compared to other physical solvents, DEPG requires higher solvent circulation mainly due to the high operating temperature.

The advantage of DEPG is the high H₂S to CO₂ selectivity (see Table 6-2), only second to NMP. The high selectivity means that DEPG can remove H₂S preferentially to CO₂. Vapor pressure of DEPG is very low and its equilibrium
losses are almost negligible. DEPG is a stable solvent and is very resistant to degradation, even with oxygen. For this reason, DEPG is often used in gasification plants for gas cleanup for power generation and in recovery methane in landfill gas plants.

6.5.2.1 DEPG process

DEPG has been used in gasification projects to selectively remove CO₂ from the synthesis gas, to produce a clean fuel for integrated gasification combined cycle (IGCC) power plants. In these plants where the syngas feed to the DEPG process is unshifted (without converting CO to CO₂ and hydrogen), a single-stage DEPG unit can be used. A picture of a DEPG unit for IGCC plant is shown in Figure 6-12.

The basic DEPG process flow diagram is shown in Figure 6-13. The feed gas at about 600 psig pressure is first cooled by heat exchanged with the cold treated gas and enters the absorber where its H₂S content and a portion of the CO₂ content are absorbed by the lean DEPG. For H₂S selective removal, the absorber must contain sufficient trays, and the solvent flow must be minimized to avoid excessive coabsorption of CO₂. The absorber typically requires four or more sections of packing.

The rich solvent from the absorber is letdown in pressure to a flash drum. The flash gas containing the absorbed hydrogen is recycled back to the absorber for recovery. The flashed solvent is then heated in the lean/rich exchanger and then regenerated in a solvent regenerator, similar to the design of an amine unit.

Heavy hydrocarbons and sulfur compounds such as mercaptans are preferentially absorbed by DEPG. These components may be removed in the

FIGURE 6-12

DEPG unit for IGCC plant (Mak, 2011).
FIGURE 6-13

Single-stage DEPG process.
regenerator by steam stripping and end up in the reflux drum. However, the DEPG regenerator typically operates at 300°F, which may not be high enough to boil off the heavier compounds. If this is the case, the heavier compounds will accumulate in the system, which may cause foaming problems. Since DEPG is soluble in water, the heavy hydrocarbons can be removed by extraction with water by processing a slip stream of the rich solvent from the process.

6.5.2.2 DEPG carbon capture process

In meeting today’s environmental requirements, the syngas from gasification plant must be shifted to form CO₂ and hydrogen, which would allow CO₂ to be captured; consequently, only hydrogen is combusted in the power plant for power production, while CO₂ is reinjected for sequestration, thus eliminating all the greenhouse gas.

The DEPG process can be configured for carbon capture using a two-stage absorption process. The first stage is the H₂S absorption stage that is designed to remove essentially all the H₂S from the feed gas and produce an H₂S-enriched acid gas to a Claus sulfur recovery unit. The second stage is the CO₂ absorption stage that removes the CO₂ from the H₂S absorption stage for sequestration.

The coabsorption of CO₂ in the H₂S absorption stage makes the design challenging. The H₂S absorption process must be designed to be H₂S selective for two reasons. First, the Claus sulfur recovery unit requires the H₂S content to be kept at high concentration (see Chapter 9). Second, excessive slippage of CO₂ to the sulfur recovery unit will increase carbon emissions in the tail gas treating unit, defeating the purpose of carbon capture.

To increase the H₂S content in the rich solvent, CO₂ is stripped from the solvent in a CO₂ stripper, and recycled back to the process, as shown in Figure 6-14. The recycle of CO₂ would require a higher solvent circulation to meet H₂S specification. To avoid reabsorption of the recycled CO₂, CO₂ pre-loaded solvent from the CO₂ absorption stage is used. With a CO₂-saturated solvent, the amount of CO₂ absorption can be minimized.

Similar to the single-stage absorption process, the rich solvent from the first stage is regenerated using steam, producing a H₂S-free lean solvent. The lean solvent is routed to the CO₂ absorber. In this arrangement, the treated gas from the second stage is sulfur free.

In the CO₂ removal section, the rich solvent is regenerated by successive reduction in pressures. Steam stripping is typically not required in the second stage. The last stage flash can operate at atmospheric pressure or under a slight vacuum, producing a semilean solvent that is adequate for bulk CO₂ removal. The process can be configured to produce high-pressure CO₂, which can be fed to the CO₂ compressors, reducing the power consumption for CO₂ sequestration. The treated gas can be sent to a power plant for power production or used for ammonia or urea production.

This two-stage DEPG system can be used to capture over 95% of the carbon content in today’s gasification plants (Mak et al., 2009; Mak, 2011).
FIGURE 6-14
Two-stage DEPG process (Mak et al., 2009; Mak, 2011).
6.5.2.3 Mercaptan removal
As can be seen in Table 6-2, the solubility of methyl mercaptan is 340 times that of methane. When molecular sieves are used for dehydration, the spent regeneration gas from the molecular sieve beds typically contains high levels of mercaptans. The mercaptan content can be removed by the DEPG solvent, producing a sulfur-free treated gas.

6.5.2.4 Landfill gas
Landfill gas typically contains significant amounts of CO₂, some H₂S, O₂, and other contaminants, such as mercaptans. Depending on the sales gas requirements, it may be necessary to remove some or all of these contaminants. DEPG is proven to be highly oxygen tolerant and has demonstrated its ability to remove heavy metals and hydrocarbons.

6.5.2.5 Hydrocarbon dew point control
Typical hydrocarbon dew point specification for the pipeline gas is 14 °F. If a gas stream is near this specification, usually all that is required is removal of a small amount of heavy hydrocarbons. Heavy hydrocarbons are very soluble in DEPG. The use of DEPG to meet the hydrocarbon dew point specification can be effective and simple. This approach can be used to condition small gas streams that are used in gas turbines or engine drivers.

6.5.3 Methanol (Rectisol process)
Historically, methanol was the first commercial organic physical solvent and has been used for hydrate inhibition, natural gas treating, dehydration, and recovery of hydrocarbon liquids. The methanol process is also called the Rectisol process (licensed by Linde AG and Lurgi AG) and uses cold methanol at −40 °F to remove acid gases at high pressure, which is very popular in ammonia plants.

Operation of a physical solvent unit favors low temperatures. Physical solvent’s acid gas holding capacity increases with low temperatures, which would reduce the solvent circulation requirement. Methanol is an alcohol-based solvent that can operate at cryogenic temperatures without encountering viscosity problems. Low-temperature operation is necessary for the methanol process due to the high vapor pressure of methanol resulting in high methanol losses.

Operation of the Rectisol process is similar to the DEPG process. Methanol’s H₂S selectivity is close to DEPG (see Table 6-2) and can also be used to produce an acid gas with high H₂S content to the sulfur recovery unit and remove CO₂ for sequestration. Similar to the DEPG process methanol can be regenerated by steam, which accounts for its capability to produce very-high-purity products.

For chemical production, such as that of ammonia and urea, high-purity hydrogen is required, which is the main application for Rectisol process. If the product gas is used as fuel gas for power generation in IGCC plant, the Rectisol process may be an overkill. Because of the extensive use of stainless-steel
material and the high power requirement of the refrigeration unit, Rectisol process is typically more expensive than competing processes. The use of molecular sieve technology is an alternative to the Rectisol process for production of high-purity hydrogen and can be a lower cost option (Mak et al., 2004).

The process flow diagram of a typical Rectisol unit is depicted in Figure 6-15 (Mak et al., 2004), and the basic process is described in the following.

The syngas is scrubbed in the methanol scrubber, which consists of an upper section for bulk CO₂ removal and a lower section for H₂S removal. A portion of the CO₂-rich solvent is drawn from the CO₂ removal section and letdown in pressure to the CO₂ flash drum. The H₂S-rich solvent from the H₂S removal section is also letdown in pressure to the H₂S flash drum. The flash gas from both drums is recompressed and recycled to the inlet of the unit for hydrogen recovery.

The rich solvent is letdown in pressure to the CO₂ stripper that produces an overhead tail gas containing methane and light components, and a bottoms product containing mainly CO₂ and H₂S. The bottom stream is fractionated in the H₂S concentrator that produces a CO₂ overhead product, and a bottom solvent enriched in H₂S. Inert or treated gas can be used as a stripping gas in the H₂S concentrator. Acid gas is finally regenerated in the methanol regenerator, and sent to the sulfur recovery unit. To maintain methanol purity and avoid buildup of water, a small fractionating column can be used to process a slip stream of the circulating solvent.

6.5.4 N-Methyl-2-pyrrolidone

NMP is a high boiling liquid with very high solubility for H₂S. The use of NMP (licensed as the Purisol® process by Lurgi AG) is particularly suited for purification of high-pressure, high-CO₂-content synthesis gas in gasification applications.

The “Purisol” process typically operates at 5 °F to minimize solvent losses. NMP has a relatively high vapor pressure compared to DEPG or PC, and low operating temperature would help to minimize solvent losses. NMP cannot significantly remove COS, but COS can be hydrolyzed and then removed by the NMP solvent. The Purisol process is similar to the DEPG process.

6.6 Mixed physical and chemical absorption processes

Some chemical manufacturers have developed processes that take advantage of the benefits of physical and chemical solvents. These mixed solvents or “hybrid solvents” have different formulation of the solvent mixtures, and are trademarked by the solvent manufacturers. These solvents take advantage of physical solvent properties, such as high acid gas solubility at high pressure and high solubility of mercaptans and organic sulfur. It also has the acid gas reactivity of a chemical solvent to meet tight specifications.
FIGURE 6-15
Typical Rectisol units process flow scheme (Mak et al., 2004).
The use of hybrid solvents is more beneficial at high acid gas partial pressures. By using various amine combinations with the physical solvents, the hybrid solvents can be formulated to allow for complete CO₂ removal, bulk CO₂ removal, or CO₂ slip, while achieving H₂S removal comparable to alkanolamines. In hybrid systems, mercaptans and organic sulfur can be removed by the physical solvent portion. However, the hybrid solvent system also inherits some of the drawbacks of a physical solvent system, such as more absorber stages, high hydrocarbon losses, and ineffective treating at low pressures. Examples of hybrid solvents include Sulfinol from Shell and UCARSOL from Dow Chemical.

Shell offers the Sulfinol solvent to treat various levels of acid gases and different sulfur contaminants. The solvent contains sulfolane, which can be formulated according to feed gas compositions and product gas specifications. The formulation combines the chemical reaction properties of amines and the physical absorption properties of sulfolane. Three different solvents are available for the Sulfinol process:

- Sulfinol-X consists of sulfolane, MDEA, piperazine, and water
- Sulfinol-M consists of sulfolane, MDEA, and water
- Sulfinol-D consists of sulfolane, DIPA, and water

Dow Chemical also offers similar hybrid solvents, UCARSOL LE-701, 702, and 703.

### 6.7 Solid bed absorption processes

The solid fixed-bed absorption processes remove a small quantity of H₂S from gas stream to meet a tight sulfur specification. The absorption processes rely on reaction of a metal oxide with H₂S to form a metal sulfide compound. The metal oxides can be regenerated with oxygen or air, producing the sulfur waste product. The metal oxides used for the absorption processes are typically iron oxide and zinc oxide.

#### 6.7.1 Iron sponge process

The iron sponge absorption is the earliest sulfur removal process and remains one of the widely used processes today. Iron sponge is very selective for H₂S removal.

In this process, the inlet gas is fed at the top of the fixed-bed reactor filled with hydrated ferric oxide (Fe₂O₃). The basic reaction is:

\[
2\text{Fe}_2\text{O}_3 + 6\text{H}_2\text{S} \rightarrow 2\text{Fe}_2\text{S}_3 + 6\text{H}_2\text{O}
\] (6-1)

The reaction requires the presence of slightly alkaline water and a temperature below 110 °F. A pH level in the order of 8–10 should be maintained through the injection of caustic soda with the water.
The bed is regenerated by controlled oxidation as:

\[ 2\text{Fe}_2\text{S}_3 + 3\text{O}_2 \rightarrow 2\text{Fe}_2\text{O}_3 + 6\text{S} \] (6-2)

For the conversion process oxygen should be introduced slowly to react with the iron sulfide. If oxygen is introduced quickly, there is a danger of intensive heat of reaction that may ignite the bed.

\[ \text{S}_2 + 2\text{O}_2 \rightarrow 2\text{SO}_2 \] (6-3)

Generally, the iron oxide process is suitable for small to moderate quantities of hydrogen sulfide (300 ppm) operating at low to moderate pressures (50–500 psig). Removal of larger amounts of hydrogen sulfide requires a continuous process, such as the ferrox process or the Stretford process. The ferrox process is based on the same chemistry as the iron oxide process except that it is in a liquid state and the operation is continuous. The Stretford process employs a solution containing vanadium salts and anthraquinone disulfonic acid, which are known to be difficult to handle (Maddox, 1982).

### 6.7.2 Zinc oxide process

The zinc oxide process can also be used for \( \text{H}_2\text{S} \) removal from various gas streams. It uses a solid bed of granular zinc oxide to react with hydrogen sulfide:

\[ \text{ZnO} + \text{H}_2\text{S} \rightarrow \text{ZnS} + \text{H}_2\text{O} \] (6-4)

The zinc oxide process can produce a treated gas with \( \text{H}_2\text{S} \) concentration as low as 1 ppm but would require operating temperature as high as 550 °F. The use of this process has been decreasing in recent years due to the problems on disposing the zinc sulfide waste product, which is considered a hazardous material.

### 6.7.3 PURASPEC\textsuperscript{TM}

PURASPEC\textsuperscript{TM} technology was originally developed by Johnson Matthey based on their catalytic absorbents for the refining industries. It can be used for natural gas processing plants to remove impurities, such as \( \text{H}_2\text{S}, \text{COS}, \text{mercury}, \text{and chlorides} \), from liquid and gaseous hydrocarbon streams to extremely low levels. \( \text{H}_2\text{S} \) and \( \text{COS} \) are removed by irreversible reactions with metal oxides (MO).

\[ \text{MO} + \text{H}_2\text{S} \rightarrow \text{MS} + \text{H}_2\text{O} \] (6-5)

\[ \text{MO} + \text{COS} \rightarrow \text{MS} + \text{CO}_2 \] (6-6)

The absorption beds are typically designed in a lead–lag reactor configuration as shown in Figure 6-16, which would facilitate the catalyst bed change-out.

The PURASPEC\textsuperscript{TM} catalysts are proprietary formulation oxides than have higher capacities and selectivity than generic metal oxides. Their costs are also higher and they should only be used for polishing to meet stringent emissions specifications (Mak et al., 2012).

The application range of the PURASPEC\textsuperscript{TM} process is shown in Figure 6-17.
Typical lead–lag absorption reactors.

Selection guide for PURASPEC\textsuperscript{TM} process (Mak et al., 2012).
6.7.4 Slurry processes

Slurries of iron oxide have been used to selectively absorb H\textsubscript{2}S. Two different slurry processes are described below.

6.7.4.1 Chemsweet\textsuperscript{®} process

The Chemsweet\textsuperscript{®} process (developed by NATCO) is a zinc-oxide-based process for the removal of H\textsubscript{2}S from natural gas. The Chemsweet white powder is an aqueous mixture of zinc oxide, zinc acetate, and also a dispersant that keeps the zinc oxide particles in suspension. The process has no impact on CO\textsubscript{2}, as the pH of the solution is low enough to prevent any appreciable absorption, even at a high CO\textsubscript{2} to H\textsubscript{2}S ratio (Manning and Thompson, 1991). The following reactions are performed in this process:

Absorption: \[ \text{ZnAc}_2 + \text{H}_2\text{S} \leftrightarrow \text{ZnS} + 2\text{HAc} \quad (6-7) \]

Regeneration: \[ \text{ZnO} + 2\text{HAc} \leftrightarrow \text{ZnAc}_2 + \text{H}_2\text{O} \quad (6-8) \]

Overall: \[ \text{ZnO} + \text{H}_2\text{S} \leftrightarrow \text{ZnS} + \text{H}_2\text{O} \quad (6-9) \]

The Chemsweet\textsuperscript{®} process can treat gas streams with high H\textsubscript{2}S concentration between pressures of 90 and 1400 psia. The presence of mercaptans in gas streams can be a problem. Some of the mercaptans will react with the zinc oxide forming a sludge composed of zinc mercaptides, Zn(OH)RH, which potentially can cause foaming problems (GPSA Engineering Data Book, 2004).

6.7.4.2 Sulfa-check process

The Sulfa-Check\textsuperscript{®} process (marketed by NALCO) selectively removes H\textsubscript{2}S from natural gas, in the presence of CO\textsubscript{2} (Dobbs, 1986). This process is accomplished using an aqueous solution of sodium nitrite (NaNO\textsubscript{2}) that is buffered to stabilize the pH of the solution to above 8.

This process is generally operated at ambient temperature and produces a slurry of sulfur and sodium salts.

\[ \text{NaNO}_2 + 3\text{H}_2\text{S} \leftrightarrow \text{NaOH} + \text{NH}_3 + 3\text{S} + \text{H}_2\text{O} \quad (6-10) \]

CO\textsubscript{2} also reacts with the sodium hydroxide to form sodium carbonate and sodium bicarbonate. Sodium nitrite and sodium hydroxide are consumed in the processes. When the chemicals are completely reacted, the system must be shutdown to remove and replenish the chemicals.

6.8 Solid bed adsorption process

For physical adsorption, the holding forces of the adsorbate on the adsorbent are weaker than chemical bonding, and the adsorbate can be desorbed by raising the temperature or reducing the partial pressure. When an adsorbed component reacts chemically with the adsorbent, the process is termed...
chemisorption and desorption by pressure reduction is generally not possible. The main advantage of physical adsorption over chemical is lower energy consumption in regeneration, which can be achieved using a pressure swing or temperature swing cycle.

The most common adsorbents in the natural gas treating industries are molecular sieves, activated alumina, and silica gels. Molecular sieve adsorbents are zeolite-based compounds that are commonly used for removal of CO₂ and H₂S, mercaptans, and other sulfur compounds. This technology is used for treating low concentrations of sulfur components in natural gas stream where deep removal is required.

The typical molecular sieve unit consists of two or more fixed-bed adsorbers and a regeneration system. With two molecular sieve beds, one bed is treating the feed gas, while the other bed is being regenerated. Switching between the beds is automated using programmed timer and switching valves. During regeneration, the molecular sieve bed is heated up by a slip stream of the product gas, which desorbs the sour components from the molecular sieves. The hot regeneration gas is cooled and treated by an amine or physical solvent (depending on the mercaptan level), which is later regenerated and sent to the sulfur recovery unit. Details of the molecular sieve unit operation are discussed in Chapter 7.

6.9 Membrane

Membrane systems, which consist of semipermeable elements (polymeric membranes), separate gases by selective permeation of the gas constituents. The gases dissolve in the membrane materials and move across the membrane barrier under a partial pressure gradient, which is established by maintaining a high feed pressure on one side of the membrane while maintaining a low pressure on the permeate side.

Many different types of membranes have been developed for gas separations. For natural gas separation, the industry’s common standard is cellulose acetate. These membranes are of the solution-diffusion type, which consists of a thin layer of cellulose acetate built on top of a thicker layer of a porous support material. The membranes are thin in order to maximize mass transfer, which would require a supporting layer for mechanical strength (Baker, 2002).

Membranes separate by solution and diffusion. The smaller and more soluble components such as H₂O, H₂, He, CO₂, and H₂S permeate quicker, leaving behind the less soluble components such as CO, methane, and nitrogen. The driving force is the partial pressure differential of the permeate components. The preferable operating conditions are high feed gas pressure and low permeate pressure.

The performance of membrane depends on the permeability of the membrane elements for the difference compounds. The relative permeation of the different compounds for a typical cellulose membrane is illustrated in Figure 6-18.
Membrane separation is intended for bulk removal, which results in high hydrocarbon losses. More advanced membrane design such as the spiral-wound membrane element may reduce hydrocarbon losses.

6.9.1 Membrane process advantages

For bulk removal where stringent specifications are not required, membrane process offers the following advantages over solvent treating processes:

- Membranes are skid-mounted units. They can be installed in modules, and are relatively inexpensive to install.
- The membrane module is compact, requiring small plot space for installation; they are suitable for offshore applications.
- Membranes have no moving parts, require minimal operator attention, and low operating and maintenance cost.
- Membranes are self-supporting and do not require utility such as water and instrument air.

6.9.2 Membrane process disadvantages

The decision to go with membrane for separation must consider the following shortcomings:

- Losses of hydrocarbon products with the permeate. If the permeate is vented to atmosphere, it may create greenhouse gas impacts. In most applications, it is desirable to minimize hydrocarbon losses as it represents a revenue loss.
- Prone to fouling from contaminants. The elements are prone to degradation from heavy hydrocarbons and aromatics. An upstream pretreatment unit is required to prevent damage to the membranes. The pretreatment unit will increase the installation cost and space requirement, reducing the advantages of a membrane unit.
- Replacement cost of membranes can be expensive, which must be factored into the operating costs.
Limited membrane manufacturer suppliers are available, making the membrane less competitive.

There is no economy of scales compared to the solvent treating process. Additional membranes module must be added to handle higher flow. The number of membrane unit is directly proportional to the feed gas flow rate. Solvent process can accommodate the higher feed gas flow by increasing the solvent circulation.

6.9.3 Membrane processes

Membrane process can be configured as a single membrane unit (Figure 6-19), which may result in significant hydrocarbon losses (typically over 10% of the feed gas).

Hydrocarbon losses may be reduced by using a two-stage membrane system as shown in Figure 6-20. However, this arrangement would require the addition of

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**FIGURE 6-19**  
Single membrane unit.

**FIGURE 6-20**  
Two-stage membrane unit.
a recycle gas compressor and a second set of membrane unit. While this may improve the recovery to some extent, the economics of additional compressor and membrane unit are seldom justified, in terms of equipment cost and space requirement.

6.9.4 Membrane pretreatment system

To protect the membrane elements and minimizing operating difficulties with heavy hydrocarbons, particulates and contaminants, a pretreatment system is necessary to ensure a continuous operation and avoid elements degradation. The pretreatment system typically includes the following design provisions:

- Filter coalescer for removal of hydrocarbon liquids, carryover amine, chemicals, and solvents.
- Adsorbent beds for removal of heavy hydrocarbons, aromatics, and contaminants that are detrimental to the life of the membrane elements. Heavy hydrocarbons can also be removed using a refrigeration unit.
- Particulate filters for removal of particulates and carryover adsorbent dust to avoid plugging of the membranes.
- A regeneration system for the adsorbent beds that includes heater, cooler, separator, and blower.
- A heater to maintain superheated feed gas to avoid condensation across the membrane, which is detrimental to the membrane elements.

The pretreatment system can be assembled as a modular unit upstream of the membrane unit, such as UOP’s MemGuard system, as shown in Figure 6-21.
6.10 Cryogenic fractionation

Cryogenic separation involves cooling the acid gases to a very low temperature so that the CO₂ can be liquefied and separated. This technology requires substantial power to operate the refrigeration unit. It also needs pretreatment and dehydration of the feed gas to remove components that would result in hydrate formation and CO₂ freezing in the cold section of the fractionation equipment.

For typical natural gas plants, cryogenic fractionation of CO₂, H₂S, and hydrocarbons requires several fractionation steps, which would require extensive refrigeration and heating. The Ryan–Holmes process that was created for enhanced oil recovery for fractionation of high CO₂ gases is still the most expensive process for CO₂ separation.

There are other techniques that were developed for cryogenic fractionation with the intent to make it economical and avoid freezing problems. FZ™ technology, developed at ExxonMobil Upstream Research Company, is a cryogenic distillation process for the single-step removal of CO₂ and H₂S from natural gas involving the controlled freezing and remelting of CO₂. This process is conceptually sound but has not been commercially proven (Northrop and Valencia, 2009).

The application for acid gas fractionation that proves to be economical is in the area of helium production and enhanced oil recovery. These plants typically process a very lean gas with high CO₂ content, typically over 70%, and a very low H₂S content. Where CO₂ is used for enhancing oil recovery, it is more energy efficient to fractionate CO₂ in a fractionation column and pump the bottom CO₂ liquid to high pressure for reinjection. Pumping CO₂ liquid requires significantly less power than conventional vapor compression.

For helium recovery, a methanol wash column is used to polish the overhead gas from the CO₂ fractionation column to produce an overhead vapor that is recycled to extinction. The clean gas from the methanol wash column can be fractionated in a nitrogen rejection column, producing a bottom stream containing the methane and hydrocarbons, which can be sent to the sales gas pipeline. The nitrogen rejection column overhead, containing nitrogen and helium, is fractionated in a helium recovery unit into the helium overhead product and a nitrogen bottom. More details of the helium recovery process can be found in Chapter 10.

6.11 Microbiological treatment processes

Continual search for more economical, safer, and simpler removal methods has led to the development of microbiological solutions for H₂S removal. One of the recent technologies for biological treatment is the THIOPAQ™ process (co-developed between Shell Global Solutions and Paques B.V), which can potentially be applied to replace the acid gas removal unit, the sulfur recovery unit, and the tail gas unit.
The THIOPAQ™ process, as shown in Figure 6-22, removes H2S by absorption into a mild alkaline solution followed by oxidation of the absorbed sulfide to elemental sulfur under the following reaction equations:

- Absorption and hydrolysis of H2S: H2S + OH⁻ → HS⁻ + H2O
- Biological sulfur formation: HS⁻ + ½O₂ → S⁰ + OH⁻

The oxidation reaction requires the use of a consortium of naturally occurring sulfur treating bacteria.

To avoid buildup of the sulfate waste product, a purge stream from the bioreactor is required. For heat and material balance of the process, makeup of water, caustic, and nutrient addition is required. To reduce the sulfate waste, membrane filter can be used to separate the sulfate ions from the other ions, concentrating the sulfate content, and reducing the effluent quantity. Recycling of the sulfate effluent to a sulfate reducing reactor can also reduce the sulfate effluent.

The practical processing capacity for the THIOPAQ process is about 50 tons per day of sulfur.

### 6.12 Selecting the gas treating process

The gas treating process selection should be based on the feed gas compositions, acid gas contents and sulfur compound levels, as well as the treated gas product specifications. If the sulfur compounds are not removed, they will end up in the liquid products, which will require further treatment.

A number of variables must be considered and evaluated in the selection process such as:

- Variation in sour gas flow rate and composition, pressure, and temperature.
- Unit capacity, site location and ambient temperatures, and onshore or offshore units.
Treated gas, acid gas, and liquid products specifications.

Unit turnup and turndown requirements.

Environmental requirements on air emissions, liquid effluent, and solid waste discharges.

Capital and operating costs.

Sometimes a high-level screening of the above variables can determine the treating gas technologies. For example:

- Fixed-bed scavengers are applicable when the sulfur capacity is less than 0.1 ton/day. When tight sulfur emissions must be met, scavengers should be considered for polishing the treated gas from the acid gas removal unit.
- When the sulfur capacity is between 15 and 50 ton/day, direct conversion processes, such as the redox processes (see Chapter 9) and microbiological treatment processes, are applicable.
- Low partial pressure of acid gases favors amine treating. High partial pressure (greater than 50 psi) and lean gases favor the use of physical solvents, especially for bulk CO₂ removal. Both PC and DEPG are suitable for natural gas treating, while PC is more economical due to its lower operating temperature.
- If selective removal of H₂S and other sulfur contaminants and mercaptans is required, physical solvent or hybrid solvent should be considered.
- If the feed gas contains COS, DGA or formulated MDEA should be considered.
- For tail gas treating, formulated MDEA and FLEXSORB can meet the stringent sulfur emissions.
- For LNG production, DGA or activated MDEA can meet the 50 ppm CO₂ specification.

In summary, in treating today’s higher CO₂ gases, the traditional amine processes may no longer be adequate. The high acid gas contents will require a different gas treating approach to meet the energy and emissions targets. The gas treating technologies will require the use of several technologies, and will invariably require the use of physical solvent for bulk removal to minimize energy consumption, the use of amine solvents for sulfur removal, and/or sulfur scavenger to meet the sulfur emissions requirements.

References


7.1 Introduction

Natural gas stream from production wells is saturated with water vapor, which will condense and form gas hydrates if the gas is cooled below its hydrate formation temperature. Gas hydrates are solids that can agglomerate and plug pipelines and equipment, interrupting operations and stopping gas production. This may create an unsafe condition, especially if significant pressure differential occurs across the hydrate plug. Also, water vapor may condense in pipelines, resulting in erosion and corrosion. When accumulated in the pipelines, it might form liquid plug, reducing the pipeline flow capacity. To avoid these potential problems, the gas stream needs to be dried to lower its water dew point.

Pipeline specifications typically call for water content to be no more than 7 lb/MMscf for US pipeline systems, 4 lb/MMscf for Canadian pipeline systems, and even lower, 1–2 lb/MMscf for Alaskan environment. These values provide protection against water condensation and hydrate formation during winter. For cryogenic natural gas liquids (NGLs) processing and LNG feed pretreatment, water removal level is even more demanding, down to a few parts per million water and dew point of −150 °F and below.

There are several methods of dehydrating natural gas, including absorption, adsorption, and direct cooling of the wet gas. Absorption using liquid (glycol) and adsorption using solid desiccants are the most common, when low water dew point gas is required for pipeline gas or for NGL recovery. The direct cooling method by expansion or refrigeration, with injection of hydrate inhibitors, is common for less dew point depression in production of pipeline gas in mild weather regions. Several other advanced dehydration technologies (i.e., membranes and supersonic processes) offer some potential advantages, particularly for offshore applications due to their compact design. However, they have limited commercial experience. There are other solvents that can remove both heavy hydrocarbons and water, including Dimethyl ether of polyethylene glycol (DEPG) and methanol (see Chapter 6), but water removal by these processes is considered incidental and typically cannot be customized. The glycol absorption and solid desiccant adsorption methods are discussed in more details in this chapter.

Mercury is present in many natural gas streams, and the levels can vary among different fields. Low levels of mercury are health hazards and will damage aluminum heat exchangers and equipment that are susceptible to corrosion attack by mercury. While this chapter addresses dehydration design, mercury removal
unit can be integrated to the molecular sieves unit. The different methods and configurations of mercury removal are discussed at the end of this chapter.

7.2 Water content determination

The first step in evaluating and/or designing a gas dehydration system is to determine the water content of the gas. This data is most important when one designs sour gas dehydration facilities and estimates water production with sour gas in the plant inlet separator.

Determining the saturation water content of a gas is a standard but complex problem in thermodynamics. There are numerous methods available for determining water contents of natural gases. A detailed discussion of those methods that are perhaps beyond the scope of the present discussion can be found in Carroll (2002, 2009). In general, for acid gas concentrations less than about 30%, existing methods are satisfactory. For higher acid gas concentrations (above 50%), particularly at higher pressures, existing methods can lead to serious errors in estimating water contents. An appropriate method has been introduced by Wichert and Wichert (2003). It is chart based and provides good estimates of the equilibrium water vapor content of sour natural gas for a range of conditions, including H$_2$S contents of 3–38 mol% with CO$_2$ contents of 3–43 mol%, pressures from 290 to 10,153 psi, and temperatures from 50 to 347 °F. The overall average error of this method is less than 1%. However, a few points showed discrepancy of more than ±10%. In the method developed by Wichert and Wichert (2003), the water content of the sour gas is calculated by multiplication of correction factor, and the water content of sweet gas from the McKetta and Wehe (1958) chart, as below:

1. Determine the equilibrium water vapor content of sweet gas at the operating temperature and pressure conditions using the McKetta and Wehe (1958) chart (see Figure 7-1).
2. Determine the mole% H$_2$S equivalent concentration of the sour gas as:
   \[
   \text{Mole}% \ H_2S \ \text{equivalent} = \text{mole}% \ H_2S + 0.7 \times \text{mole}% \ CO_2 \quad (7-1)
   \]
3. From Figure 7-2 at the bottom left-hand temperature scale, move to the right to the mole % H$_2$S equivalent (interpolate between the lines if necessary).
4. From this point, move to the upper chart, to the pressure of interest. From the pressure point, move to the left, to the ratio scale.
5. Multiply the value from step 4 by the water content determined for sweet gas in step 1. The result is the estimate of the saturated water content of the sour gas at the pressure and temperature of interest.

Commercial process simulation software are available with the latest GPSA data and vapor–liquid equilibrium correlations for accurate calculation of water content of sour gas streams. AQUAlibrium software by FlowPhase Inc., developed based on a rigorous thermodynamic model rather than mere empiricism, is a
FIGURE 7-1

McKetta and Wehe (1958) chart (GSA, 2004).
useful tool for predicting equilibrium water content of pure acid gas components and acid–gas mixtures, and may produce more accurate results, particularly for high-pressure systems where no data exist (Carroll, 2009).

7.3 Glycol dehydration

Among the different natural gas dehydration processes, absorption is the most common technique, where the water vapor in the gas stream becomes absorbed in a liquid solvent stream. Although many liquids possess the ability to absorb water from gas, the liquid that is most desirable to use for commercial dehydration purposes should possess the following properties:

1. High absorption efficiency
2. Easy and economic regeneration
3. Noncorrosive and nontoxic
4. No operational problems, such as high viscosity when used in high concentrations

FIGURE 7-2
Water content ratio chart (Wichert and Wichert, 2003).

\[
\% \text{H}_2\text{S equivalent} = \text{mole}\% \text{H}_2\text{S} + 0.7 \times \text{mole}\% \text{CO}_2
\]
5. Minimum absorption of hydrocarbons in the gas, and no potential contamination by acid gases.

Glycols are the most widely used absorption liquids as they approximate the properties that meet the commercial application criteria. Several glycols have been found suitable for commercial application. The properties of commonly available glycols can be found in manufacturer’s Web site. Their pros and cons can be summarized as follows (Katz et al., 1959):

1. Monoethylene glycol: High vapor pressure and seldom used in contactor at ambient temperature due to high losses in the treated gas. Normally, it is used as hydrate inhibitor where it can be recovered from gas by separation at below ambient temperatures. It is used in glycol injection exchanger operating at $-20 \, ^{\circ}\text{F}$ to minimize losses.

2. Diethylene glycol: High vapor pressure leads to high losses in contactor. Low decomposition temperature requires low reconcentrator temperature ($315–340 \, ^{\circ}\text{F}$) and thus glycol purity is not high enough for most applications.

3. Triethylene glycol (TEG): Relatively low vapor pressure when operating at temperatures below $120 \, ^{\circ}\text{F}$. The glycol can be reconcentrated at $400 \, ^{\circ}\text{F}$ for high purity. Dew point depressions up to $150 \, ^{\circ}\text{F}$ can be achieved with enhanced glycol process like DRIZO®.

4. Tetraethylene glycol: More expensive than TEG but less glycol loss at high gas contact temperatures. Reconcentrate at $400–430 \, ^{\circ}\text{F}$.

TEG is the most common liquid desiccant used in natural gas dehydration.

In process design of the TEG dehydration unit, the upstream unit operation must be considered, as the TEG inlet temperature and water saturation will significantly impact the unit performance. For regions with hot climate, the feed gas should be cooled to the lowest possible temperature with cooling water (or chilled water). This is necessary to ensure that the feed gas temperature meets the TEG unit’s inlet maximum temperature.

### 7.3.1 Conventional TEG dehydration process

Figure 7-3 shows the scheme of a typical TEG dehydration unit. As can be seen, wet natural gas is processed in an inlet filter separator to remove liquid hydrocarbons and free water. The separator gas is then fed to the bottom chamber of an absorber where residual liquid is further removed. It should be cautioned that hydrocarbon liquids must be removed as any entrainments will result in fouling of the processing equipment and produce carbon emissions. The separator gas is then contacted countercurrently with TEG, typically in a packed column.

Typically, the liquid loading on the tray (GPM per square foot) is very low, due to the low liquid to gas ratio. To avoid liquid maldistribution, structured packing or bubble cap trays should be used.

TEG will absorb the water content, and the extent depends on the lean glycol concentration and flow rate. TEG will not absorb heavy hydrocarbons to any
degree; however, it will remove a significant portion (up to 20%) of the BTEX (benzene, toluene, ethyl benzene, and xylenes) components. BTEX is considered as a volatile organic compound (VOC), which must be incinerated to comply with emissions requirements.

Dry natural gas exiting the absorber passes through a demister, and sometimes through a filter coalescer to minimize TEG losses. Because of the relatively low TEG flow rate, there is not much sensible heat exchange, hence the dried gas temperature is almost the same as that of the feed gas.

The rich glycol is used to cool the TEG regenerator overhead, minimizing glycol entrainment and losses from the overhead gas. Rich glycol is further heated by the glycol heat exchanger and then flashed to a flash tank. The flash gas can be recovered as fuel gas to the facility.

The rich TEG is filtered with solid and carbon filters, heated, and fed to the regenerator. The filtration system would prevent pipe scales from plugging the column and hydrocarbons from coking and fouling the reboiler. The water content in the glycol is removed with a reboiler. Heat to the reboiler can be by a fire heater or an electrical heater. Electric heater is preferred as it would avoid emission problems, particularly in smaller units. The water vapor and desorbed natural gas are vented from the top of the regenerator.

The dried glycol is then cooled via cross-exchange with rich glycol; it is pumped and cooled in the gas/glycol heat exchanger and returned to the top of the absorber.

The glycol regeneration is typically designed as a skid mounted unit that can be prefabricated and shipped to the site. A three-dimensional rendition of the modular TEG dehydration unit is shown in Figure 7-4.
There are improved regeneration techniques that can produce higher glycol concentration to be used to further lower the treated gas’ water dew point beyond the conventional TEG dehydration process. By injecting dry (stripping) gas into the base of the glycol reboiler to further reduce the partial pressure of water, and provide agitation of the glycol in the reboiler, TEG concentration can be increased from 99.1% to 99.6% by weight. Typically, a packed column located below the reboiler section is used for TEG stripping. 

DRIZO™ process (under PROSERNAT license) can regenerate TEG to a higher purity than the conventional gas stripping process. Solvent stripping can produce much higher glycol purities than gas stripping and consequently allows the process to achieve a much larger water dew point depression: up to $-150 \, ^\circ F$ and even higher in some cases. The solvent required by the DRIZO™ process is usually obtained from the $C_6^+(BTEX)$ present in the natural gas itself and in most cases the process will produce some liquid hydrocarbons.

The main advantages of the DRIZO process are that all BTEX compounds are recovered from the regenerator before being sent to the atmosphere and no external stripping gas is required. The DRIZO technology may be adapted to existing dehydration units, which need to be upgraded for meeting requirements for higher glycol purity, or for better emission control of BTEX.
A typical process flow schematic for the DRIZO system is shown in Figure 7-5. The main difference from the conventional TEG stripping unit is the proprietary separation process in the regenerator overhead where the oil is separated from the aqueous phase. The aqueous phase, containing the entrained glycol, is refluxed to the regenerator. The hydrocarbon phase is removed, heated, filtered, and used as the stripping gas for TEG regeneration. Purity of the lean glycol can be controlled by adjusting the amount of stripping gas recycle and the regeneration temperature.

There are other solvent stripping processes that can be used to improve the glycol purity without the use of stripping gas. There are two basic processes that can be used for glycol regeneration. One is the vacuum-based process that uses vacuum pressure to reduce the partial pressure of water in the lean glycol. The other is the “Coldfinger” process that uses a condenser to condense and collect water/hydrocarbons from the reboiler vapor phase and remove them from the reboiler such that can be applied to a glycol regeneration system to accomplish higher glycol purity. The Coldfinger process can achieve a TEG concentration of approximately 99.96 wt%. The water depression performance of these processes can be compared in Table 7-1.

7.3.3 Glycol injection process

Wet gas can be chilled to meet pipeline water as well as hydrocarbon dew point specification using the glycol injection process, which includes propane
refrigeration for chilling. Glycol injection is preferred by most pipeline operators because of its simplicity. The process flow schematic for a glycol injection dehydration system is shown in Figure 7-6.

Glycol injection systems are lower cost options compared with TEG dehy-
dration unit or fixed-bed dehydration systems. The lowest temperature at which ethylene glycol injection units can be operated is \(-30^\circ\text{F}\). Below this temperature, the viscosity of glycol becomes too high for good phase separation. Beyond this temperature, methanol injection or other dehydration techniques should be considered.

Glycol has limited solubility in the gas phase, so it must be carefully distributed to each heat exchanger tube to ensure that an adequate amount is present in each tube to prevent hydrate formation. The spray system is critical in the design of heat exchangers with glycol injection. Potential problems include:

- Incomplete coverage of all heat exchanger tubes with glycol spray, resulting in hydrate formation and plugging in some tubes.
- Inadequate mist formation by the injection system, resulting in uneven glycol distribution.
- Plugging of the spray nozzles.

<table>
<thead>
<tr>
<th>Regeneration Process</th>
<th>TEG Purity, wt%</th>
<th>Water Dew Point</th>
</tr>
</thead>
<tbody>
<tr>
<td>Vacuum</td>
<td>99.2–99.9</td>
<td>100–150</td>
</tr>
<tr>
<td>COLDFINGER</td>
<td>99.96</td>
<td>100–150</td>
</tr>
<tr>
<td>DRIZO</td>
<td>99.99+</td>
<td>180–220</td>
</tr>
<tr>
<td>Stripping gas</td>
<td>99.2–99.98</td>
<td>100–150</td>
</tr>
</tbody>
</table>

Glycol Injection

Typical flow diagram for glycol injection dehydration system.
A well-designed glycol injection dehydration system should provide sufficient glycol delivery pressure from the glycol regeneration unit for an even spray pattern. There is a direct correlation between the glycol flow and the differential pressure across the nozzle. Therefore, both flow and differential pressure indication must be provided for monitoring spray performance. If there is a decrease in flow, this may be an indication that the spray nozzles are plugged. Filters provided upstream of the spray nozzles can avoid plugging of the spray heads.

Glycol injection heat exchanger design: The tube side inlet nozzle of the gas–gas exchanger is typically designed with a cone inlet channel designed with a specific angle and oriented in an axial position. A cone inlet channel provides better distribution of gas compared to a conventional radial nozzle. In addition, a cone inlet channel helps maintain uniform gas velocity profile so that the glycol spray pattern is not distorted. This allows for even glycol distribution.

The glycol spray systems are designed to have sufficient pressure drop across the nozzles for good atomization of the injected glycol and even coverage of the exchanger tubes. When glycol is sprayed onto the tube sheet and the glycol solution is cooled, the viscosity of the glycol solution increases. As the hydrocarbon starts condensing, the flow regime of the three-phase mixture changes, which affects heat transfer, pressure drop, and emulsion forming tendencies. These factors must be considered in the detailed design of these heat exchangers. Design techniques for these exchangers are based on empirical data.

Factors that must be considered in the design of these heat exchangers include:
- The effect of glycol on heat exchanger pressure drop
- The reduction in heat transfer due to the presence of glycol
- Continuous changes in flow regime across the heat exchanger tubes
- Maximum allowable tube length as a function of tube diameter
- Maximum allowable velocity to prevent emulsion formation
- Minimum allowable velocity to maintain the glycol in the annular flow regime

7.3.4 TEG unit design considerations

General equipment design guidelines of TEG dehydration unit can be found in the GPSA Engineering Data Book (2004). Following are some of the key design parameters that the designer should be aware of in the design of the TEG absorber and the regenerator.

Glycol circulate rate: The amount of water to be removed in a TEG dehydration system is calculated from the gas flow rate, the water content of incoming gas, and the desired water content of outgoing gas. The water removal rate, assuming the inlet gas is water saturated, can be determined as:

\[ W_r = \frac{Q_G(W_i - W_o)}{24} \]  

(7-2)
where $W_r$ is the water removed in lb/h, $W_i$ is water content of inlet gas in lb/MMscf, $W_o$ is the water content of outlet gas in lb/MMscf, and $Q_G$ is the gas flow rate in MMscfd.

The glycol circulation rate is determined on the basis of the amount of water to be removed, and is usually between 2 and 6 gallons of TEG per pound of water, which depends on the number of equilibrium stages in the absorber. For absorber with more than three equilibrium stages (a typical absorber design), 3 gallons TEG/lb water is sufficient.

Higher circulation rates provide little additional dehydration benefit while increasing reboiler heating duty and pumping requirements. The heat required by the reboiler is directly proportional to the glycol circulation rate. Therefore, an increase in circulation rate may decrease reboiler temperature, lowering lean glycol concentration, and actually decrease the amount of water that is removed by the glycol. On the other hand, problems can arise if the TEG circulation rate is insufficient; therefore, a certain amount of overdesign is required to accommodate changes in feed gas conditions. An optimal circulation rate for each dehydration unit typically ranges from 10% to 30% above the minimum glycol circulation rate (EPA430-B-03-013, 2003). The minimum glycol circulation rate can then be calculated as:

$$Q_{TEG,\,\text{min}} = G \times W_r$$  \hspace{1cm} (7-3)

where $Q_{TEG,\,\text{min}}$ is the minimum TEG circulation rate (gal TEG/h) and $G$ is the glycol to water ratio (gal TEG/lb water removed).

Figure 7-7 shows the effect of TEG circulation rate on water removal for an absorber with 2.5 equilibrium stages, at various glycol concentration levels. Note

![Figure 7-7](image-url)

**FIGURE 7-7**

Effect of TEG concentration and circulation rate on water removal ($N = 2.5$) (GPSA, 2004).
that the water removal curves become relatively flat at glycol circulation rate of 3–3.5 gal/lbm water.

Glycol purity: TEG purity can be controlled by the reboiler temperature and pressure in the regenerator. If necessary, the use of a stripping gas can remove the residual water in the lean glycol to produce a very lean glycol.

Figure 7-8 shows the relationship of regenerator reboiler temperature and glycol purity. Reboiler temperature for the TEG regenerator is typically limited to about 400 °F to minimize glycol degradation. Consequently, this temperature limits the lean glycol concentration to between 98.5% and 98.9%. Some operators limit the reboiler temperature to between 370 and 390 °F. If higher purity is required to meet stringent pipeline dew point specification, stripping gas may be necessary.

The effect of stripping gas on glycol purity is shown in Figure 7-9. As can been seen, when the number of equilibrium stages is greater than 3, increasing stripping gas can improve the glycol purity to as high as 99.95%. While stripping gas can be used to meet water specification of the product gas, it should be used only if the TEG dehydration unit fails to meet specifications. The use of stripping gas generates another gaseous waste stream. TEG overhead vapor is considered a source of emissions. The off-gas is typically sent to an incinerator to comply with VOC permits, or recycled to the front section of the unit.

FIGURE 7-8
Glycol purity versus reboiler temperature at different levels of vacuum (GPSA, 2004).
7.3.5 Operational problems

The operation problems of a TEG dehydration unit are typically related to the nonperformance of the absorber and regenerator, and summarized in the following sections.

7.3.5.1 Absorber
The main problems with the glycol absorber are described below.

*High feed gas temperature:* Feed gas temperature higher than design value has two negative impacts. First, the feed gas would contain more water at the higher temperature. Secondly, the absorber will operate at a higher temperature, which is unfavorable for the glycol–water equilibrium, reducing the dehydration capacity of glycol. To minimize the water content to the TEG dehydration unit,
particularly in hot climate areas, feed gas should be first cooled with a chilled water system. A chilled water system can easily cool the feed gas to about 70 °F, at which the bulk of its water content is condensed and can be removed.

Foaming problems: TEG foaming in contactor (absorber) will result in disruption of downstream operation. It is typically caused by dirty glycol and contaminants such as hydrocarbon condensate, well treating chemicals, salts, and pipe scales. To minimize this problem, the glycol must be maintained clean with filtration. This would require filtering 10% of the circulation flow through a mechanical (solid) filter and a carbon filter. Pressure drop across the filters should be monitored and the filters changed out as needed. The feed gas inlet separator should also be monitored to avoid contaminants from entering the glycol system.

BTEX content in feed gas: BTEX hydrocarbons are soluble in TEG, which is an environmental problem, as the absorbed gas is released to the atmosphere from the glycol regenerator. To avoid venting BTEX, glycol dehydration unit must send the overhead off-gas to an incinerator to destroy the BTEX components. BTEX absorption can be minimized by using more equilibrium stages in the absorber, which would lower the glycol circulation and reduce the BTEX absorption. Alternatively, the absorbed BTEX can be recovered as a liquid by-product with the DRIZO unit discussed in the previous section.

7.3.5.2 Regenerator
The main problem with the glycol regenerator is high glycol losses, which can be contributed by entrainment and high overhead temperature. The equilibrium glycol in the overhead gas increases significantly at higher temperature, particularly at higher than 250 °F. However, low overhead temperature can also be a problem. Low overhead temperature, especially during winter operation, may result in excessive water condensation, and may cause column flooding. The water balance of the glycol system must be monitored and excess water must be purged from the overhead.

The glycol regenerator design and heating duty must match the glycol flow. If the glycol circulation exceeds the regeneration duty, it will not help the dehydration process. Overcirculation without the required heating duty will lower the regenerator temperature, making it unable to produce the glycol purity required for drying.

7.3.5.3 Reboiler
The main problems with the glycol reboiler are:

Carryover of brine solutions: Contaminants from the field operation can lead to buildup of salts in the glycol reboiler system. Sodium salts (typically sodium chloride, NaCl) are a source of problems in the reboiler. The salt will precipitate from the solution at reboiler temperatures of 350–400 °F, and can deposit on the reboiler tube, reducing the heat exchanger performance and may cause corrosion. Salt content in the glycol reboiler system should be maintained at no more than 1% by removal of some of the glycol content.
Glycol degradation: Glycol degradation is caused primarily by oxidation or thermal degradation. Glycol readily oxidizes to form corrosive acids. Oxygen can enter the system with incoming gas, from unblanketed storage tanks or sumps, and through packing glands. Oxidation problem can be eliminated by blanketing with fuel gas or inert gas. Thermal degradation can be reduced by proper filtration, and maintenance of equipment.

Acid gas: H₂S and CO₂ are absorbed by glycol to some degrees and may cause corrosion in the reboiler and the regenerator overhead system. Typically, the reboiler and overhead system should be constructed of stainless steel or other suitable materials to resist acid gas corrosion.

7.3.6 Future technology developments
Glycol dehydration will continue to be a workhorse in the gas processing industry due to proven record and relatively low cost. Over the past years, most of the research works are on glycol–BTEX equilibrium and design that produce high purity glycol to meet deep water dew pointing requirement. Future developments most likely will be on standardization and modular design that can be prefabricated to meet the accelerating shale gas field development.

7.4 Solid-bed dehydration
Solid-bed dehydration is the process where a solid desiccant (adsorbent) is used for the removal of water vapor from a gas stream to meet water dew points less than −40 °F. The desiccant material becomes saturated as moisture is adsorbed onto its surface. A good desiccant should therefore have the greatest surface area available for adsorption.

The mechanisms of adsorption on a desiccant surface are of two types: physical and chemical. In physical adsorption (or physisorption), the bonding between the adsorbed species and the solid-phase holds liquids (condensed water vapors) and solids together and gives them their structure. In chemical adsorption, involving a chemical reaction that is termed “chemisorption,” a much stronger chemical bonding occurs between the surface and the adsorbed molecules. Chemical adsorption processes find very limited application in gas processing. This section considers only physical adsorption, and all references to adsorption mean physical adsorption.

Physical adsorption is an equilibrium process, where for a given vapor-phase concentration (partial pressure) and temperature, an equilibrium concentration exists on the adsorbent surface that is the maximum concentration of the adsorbate on the surface. The measurement of the amount of gaseous compound adsorbed over a range of partial pressures at a single (fixed) temperature results in a graph known as an adsorption isotherm, which can have very different shapes depending on the type of adsorbent, the type of adsorbate, and intermolecular interactions between the adsorbate and the adsorbent surface. In addition to
concentration (i.e., partial pressure for gases), two properties of the adsorbate (polarity and size) dictate its concentration on the adsorbent surface (Kidnay and Parrish, 2006).

7.4.1 Adsorption capacity

Adsorption capacity (or loading) is the amount of adsorbate taken up by the adsorbent per unit mass (or volume) of the adsorbent. The adsorption capacity of a solid desiccant for water is expressed as the mass of water adsorbed per mass of desiccant. There are three water adsorption capacity terms used as below (Campbell, 2000):

- **Static equilibrium capacity**: The water capacity of new (fresh) solid desiccant (usually expressed in weight percent) as determined in an equilibrium cell at a fixed temperature and 100% relative humidity.
- **Dynamic equilibrium capacity**: The water capacity of solid desiccant as the fluid is flowing through the desiccant at the design flow rate, temperature, and pressure.
- **Useful capacity**: The design capacity that accounts for the loss of solid desiccant capacity with time due to the fact that the total desiccant bed cannot be fully utilized.

The static adsorption capacity is greater than the dynamic adsorption capacity. The dynamic equilibrium loading is generally 50–70% of the static equilibrium capacity. The static adsorption capacity is the maximum theoretical capacity of the desiccant and can be used for comparison of different desiccants while the dynamic adsorption capacity is used to calculate the required filling amount of adsorbents.

The dynamic moisture adsorption capacity of a desiccant depends on a number of factors, such as the relative humidity of the inlet gas, the gas flow rate, the temperature of the adsorption zone, the mesh size of the adsorbent, and the length of service and degree of desiccant degradation, and not the least on the type of desiccant. Moisture adsorption capacity is not materially affected by variations in pressure except where pressure may affect the other variables listed above.

7.4.2 Adsorbent selection

A variety of solid desiccants are available in the market for specific applications. Some are good only for dehydration of natural gas and NGLs while others are capable of performing simultaneous dehydration and desulfurization/hydrocarbon dew point controlling. The selection of proper desiccant depends on the given application and sometimes is a complex problem. For solid desiccants used in gas

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1 Adsorption capacity is of great importance to the capital cost because it dictates the amount of adsorbent required, which also fixes the volume of the adsorber vessels.
dehydration, the following properties are desirable (Campbell, 2000; Daiminger and Lind, 2004):

1. High adsorption capacity: This reduces the adsorbent volume, requiring smaller vessels, resulting in lower capital and operating costs.
2. High selectivity: This permits removal of only the undesirable components and reduces operating expenses.
3. Easy regeneration: Low regeneration temperature reduces the heating requirement for solid desiccant regeneration.
4. Low-pressure-drop design: This makes more pressure available for the turboexpander operation in an NGL recovery process.
5. Good mechanical integrity: High crush strength, low attrition, low dust formation, and high stability against aging reduces the frequency of adsorbent change-out and downtime-related losses in production.
6. Environment-friendly properties: The materials should be noncorrosive, nontoxic, and chemically inert to permit safe handling.
7. Reasonable adsorbent price

The common commercial adsorbents used in solid-bed dehydrators (that possess the above-mentioned characteristics in a satisfactory manner) are molecular sieves, silica gels, and activated alumina, which are described below.

### 7.4.2.1 Molecular sieves

Molecular sieves or zeolites are crystalline alkali metal aluminosilicates with a typical structure of the form $\text{M}_x/\text{n}[(\text{AlO}_2)_x(\text{SiO}_2)_y]z\text{H}_2\text{O}$, where $n$ is the valence of the cation, and $\text{M}$ is the metal ion inside each aluminosilicate cage. The zeolite is a three-dimensional tetrahedral structure of silica and alumina. The alumina tetrahedra carry a net negative charge, and need to be balanced by a cation such as $\text{Na}^+$. The tetrahedra build up to form a truncated octahedra, which are either stacked in a cubic structure to make the A-type zeolite, or in a tetrahedral structure to make the X-type zeolite (Figure 7-10).

![Molecular sieve types A and X structures](Secker and Zafirakis, 2011).
In the zeolite structure, the cations determine the pore opening size, and depending on the type of cation, molecular sieves with a specific pore opening can be created (Table 7-2). The benefit of having an adsorbent with a specific pore opening can be seen when observing nominal diameters of typical molecules within a hydrocarbon feed stream. To be efficiently adsorbed, the polar molecules must be small enough to enter the cavities via their pores. Hence, water molecule (with the nominal size of 2.6 Å) can be adsorbed on 3A, 4A, 5A, and 13X molecular sieves. Molecular sieve is the most versatile adsorbent because it can be manufactured for a specific pore size, depending on the application. Molecular sieves are the only choice for natural gas dehydration to cryogenic processing standards (less than 0.1 ppmv water or −150 °F dew point). Molecular sieves can also provide a solution for removal of carbon dioxide and sulfur compounds such as hydrogen sulfide, mercaptans, carbonyl sulfide, and other sulfides (with the exception of carbon disulfide) from natural gas and NGLs to very low outlet specifications, either as a stand-alone unit or as a polishing unit within a combination of gas treating processes. The applicability of a molecular sieve unit for sweetening purposes will be based on the required outlet specification for the product stream, and the level of contaminants and operating conditions of the feed (see Tables 7-3).

### Tables 7-2 Most Common Types of Molecular Sieves (Angstrom = 10⁻¹⁰ m)

<table>
<thead>
<tr>
<th>Zeolite Type</th>
<th>Cation</th>
<th>Nominal Pore Size (Å)</th>
<th>Measured Pore Size (Å)</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>K</td>
<td>3</td>
<td>3.3</td>
</tr>
<tr>
<td></td>
<td>Na</td>
<td>4</td>
<td>3.9</td>
</tr>
<tr>
<td></td>
<td>Ca</td>
<td>5</td>
<td>4.3</td>
</tr>
<tr>
<td>X</td>
<td>Na</td>
<td>10</td>
<td>7.4–12.5</td>
</tr>
</tbody>
</table>

2Although carbonyl sulfide can be adsorbed on zeolites, its adsorption kinetics are very poor and it will slip very fast through the molecular sieve vessel.

3As carbon and sulfur have the same electronegativity, carbon disulfide cannot be polarized and therefore is not adsorbed by molecular sieves.
The choice of molecular sieve type depends on the type and concentration of contaminants in the feed gas. Depending on the type of feed and species to be removed, the different grades of adsorbents can be selected based on the general guideline shown in Figure 7-11.

The adsorption characteristics of the different molecular sieve sizes can be summarized as following:

- In general, type 4A is the most common sieve for dehydration of gases and liquids, but the smaller pore 3A is sometimes preferred to standard 4A for minimizing coadsorption of CO2 and H2S. If both oxygen and H2S are present, 3A reduces the production of elemental sulfur, which can block the adsorbent pores. The 3A form is normally used to treat reactive streams such as olefins since it excludes most molecules except water. It is also widely used for alcohol dehydration since it excludes the highly polar alcohol molecule.

- Molecular sieves 5A and 13X are commonly used for desulfurization (5A for light sulfur, while 13X for heavy and branched sulfur species). Type 13X has larger pore openings and therefore has better kinetics than type 5A; however, coadsorption of benzene, toluene, and xylene components, which can block pores and deactivate the molecular sieve, may hinder removal of heavy mercaptans (Mokhatab and Meyer, 2009).

It is sometimes better to use a compound bed with successive layers of molecular sieves for adsorbing the different impurities. This combination increases the useful capacity of the bed, which is typically the case when water, H2S, mercaptans and other sulfur species need to be removed at the same time.
When CO$_2$ is present in the feed gas, the molecular sieve may promote the hydrolysis reaction of H$_2$S and CO$_2$ to form carbonyl sulfide (COS). COS is a main concern as it can be later converted back to H$_2$S in the presence of water, causing corrosion and emission problems. Molecular sieve manufacturers have developed specific products, which are 3A based, for minimizing the COS reaction.

### 7.4.2.2 Silica gels

Silica gel is a generic name for a gel manufactured by adding aqueous sodium silicate to sulfuric acid. It is an amorphous product similar to alumina. There are different grades of silica gel commercially available for natural gas drying. Silica gels can be used for heavy hydrocarbon and water removal to meet a water dew point of $-40$ °F or lower with specialty grades silica gel.

Silica gel can be used to process sour gases, but not alkaline materials such as caustic or ammonia. Although there is no reaction with H$_2$S, sulfur can deposit and block their surface pores. H$_2$S content in the feed gas should stay below 5%.

Generic grades of silica gel will break up when exposed to liquid water; however, there are specialty grades commercially available that are more liquid water stable. Engelhard (now acquired by BASF) manufactures an improved silica gel called Sorbead$^{\text{TM}}$, which is supposed to have a higher performance in heavy hydrocarbons and water removal than generic silica gel.

Silica gel is typically used in feed gas pretreatment units for removal of trace components of heavy hydrocarbons and water, such as feed gas to a membrane separation unit, or conditioning flash gas to the fuel gas system. It is not used in drying feed gas to NGL and LNG production plants because of its lower adsorption capacity, and inability to meet low water dew point.

The adsorption capacity of silica gel is typically exhausted in a short time, in the range of dozens of minutes up to 2–3 h,$^4$ which increases the number of adsorption cycles and reduces the lifetime of the silica gel (Kane et al., 2004).

### 7.4.2.3 Activated alumina

A hydrated form of aluminum oxide (Al$_2$O$_3$.3H$_2$O), alumina, is the least expensive adsorbent for gas dehydration. Alumina, which is alkaline, should not be used for drying gas with high concentrations of acid gases (CO$_2$ and H$_2$S). Alumina does not have a precise pore opening the way molecular sieves do. Consequently, it is not as selective since more molecules have access to the active sites.

There are several types of alumina available for use as a solid desiccant. Activated alumina is a manufactured or natural occurring form of aluminum oxide that is activated by heating. The structure of the product is amorphous rather than crystalline. Activated alumina holds onto water less strongly than molecular sieves and therefore requires lower regeneration temperature and less

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$^4$The adsorption cycle times of the silica gels are very short due to lower useful capacity of the desiccant for hydrocarbons relative to water. Hence, the silica gel units are often called “short cycle” or “quick cycle” units.
regeneration heat duty. Similar to molecular sieves, feed temperature should be kept at below 120 °F to avoid overloading the adsorbents.

Activated alumina is typically used to dry a wet lean natural gas to a dew point of −60 °F. Activated alumina has a higher equilibrium water capacity than molecular sieves. It can adsorb water of 35–40% of its own weight. Activated alumina is usually applied as a top layer on the molecular sieves to reduce cost, and to protect the molecular sieves.

7.4.2.4 Comparison of different adsorbents

Table 7-4 summarizes some of the key properties of three above-mentioned solid desiccants. These properties are for comparison purposes and may differ among solid desiccant manufacturers. Note, the selection of these solids desiccants depends on economics. The aluminas have the lowest cost per unit of dehydration capacity. The silica gels are next. Molecular sieves are the most expensive and must be justified by their special characteristics.

7.4.3 Adsorption technology

7.4.3.1 Adsorption principle

There is an equilibrium relationship between the fluid and the solid adsorbents in the mass transfer applications, which can be expressed in the form of isotherms as shown in Figure 7-12. As the isotherm shows, the concentration of the adsorbate, which in this case would be water, is a function of the concentration of water in the fluid phase and the adsorbing temperature.

The level to which gas is dried is a function of the regeneration conditions. A thoroughly regenerated bed will be in equilibrium with the gas used for regeneration. The concentration of water left on the bed at the end of the regeneration cycle determines the performance of the bed because the outlet gas will be in equilibrium with the last particle of adsorbent. The adsorption process is dynamic as the adsorbent removes the water content from a continuously flowing stream. This process is intrinsically an unsteady state operation.

<table>
<thead>
<tr>
<th>Table 7-4 Properties of Solid Desiccants (Campbell, 2000)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Desiccant</td>
</tr>
<tr>
<td>-----------</td>
</tr>
<tr>
<td>Pore Diameter, Å</td>
</tr>
<tr>
<td>Water Dew Point, °F</td>
</tr>
<tr>
<td>Bulk Density, lb/ft³</td>
</tr>
<tr>
<td>Heat Capacity, Btu/lb°F</td>
</tr>
<tr>
<td>Design Capacity, wt%</td>
</tr>
<tr>
<td>Regeneration Stream Temperature, °F</td>
</tr>
<tr>
<td>Heat of Adsorption, Btu/lb</td>
</tr>
</tbody>
</table>
In commercial practice, adsorption is carried out in a vertical, fixed bed of adsorbent, with the feed gas flowing down through the bed (Figure 7-13). The feed gas entering the bed from the top and the upper zone becomes saturated first, where equilibrium between the water partial pressure in the gas and the water adsorbed on the desiccant is established and no additional adsorption occurs. This zone is called the equilibrium zone (EZ).

With the adsorption going on, the EZ will grow and more and more water will be adsorbed. The length of the adsorber bed across which the concentration of the adsorbate is reduced from inlet to outlet conditions is known as the mass transfer zone (MTZ). This is simply a zone or section of the bed where a component is transferring its mass from the gas stream to the surface of the solid desiccant.

The nonutilized zone is the part of the bed that has not encountered water yet because all water molecules were adsorbed in the EZ and MTZ. This zone will only serve for drying when the EZ and MTZ have lost their adsorption capacity.

As the flow of gas continues, the MTZ moves downward through the bed and water displaces the previously adsorbed gases until finally the entire bed is saturated with water vapor. When the leading edge of the MTZ reaches the end of the bed, breakthrough occurs (see Figure 7-14). At this time, the adsorber vessel should be switched to the regeneration mode.

### 7.4.3.2 Solid-bed design considerations

Optimization of solid desiccant beds size is a key factor in reducing the overall cost of a solid-bed dehydration unit. For a given performance, shorter beds mean smaller vessels and lower utilities requirements. However, adsorbent requirements must be sufficient to accommodate the equilibrium section as well as the mass transfer zone’s length. The MTZ is usually assumed to form quickly in the
FIGURE 7-13
Schematic of the adsorption zones in an adsorber vessel (Smit et al., 2009).

FIGURE 7-14
Variation of adsorption zone front with time (Savary, 2004).
adsorption bed and to have a constant length as it moves through the bed, unless particle size or shape is changed. The length of the MTZ is mainly a function of the feed flow rate, but is also influenced by the feed composition and temperature, and the adsorbent type, bulk density, particle size, and shape. The length of the MTZ is usually 0.5–6 ft, and the gas is in the zone for 0.5–2 s (Trent, 2004). To maximize bed useful capacity, the MTZ needs to be as small as possible because the water loading in this zone is low when compared to the equilibrium (saturation) zone. Normally, the most efficient adsorbent bed will be a tall, thin bed using the smallest particle size that has a larger surface area and thus improves the adsorption kinetics, resulting in a shorter MTZ. This higher bed efficiency has to be paid for by a higher pressure drop. The total pressure drop across an adsorbent bed should not exceed 8–10 psi. The desired bed length to bed diameter ratio should also fall between about 2.5 and 6.

Note should be made that compound desiccant beds, which use more than one desiccant size or desiccant type, increase the useful capacity of the bed by increasing the equilibrium capacity or shortening the MTZ, or both. The most common example of a compound bed is the use of large-sized desiccant at the top of the bed (which minimizes pressure drop and forces on the bed supports) and small-sized one at the bottom, resulting in longer cycle times and/or shorter bed lengths. Another compound desiccant bed application involves the use of activated alumina at the top of the molecular sieve beds. This application has been used in commercial natural gas dryers for decades.

There are a number of benefits to use an activated alumina/molecular sieve compound bed, including lower overall adsorbent cost, and higher resistance to liquid carryover into the bed from upstream separation facilities. Also, activated alumina has a higher static equilibrium capacity for water than molecular sieve when the feed gas is near saturation (as shown by the typical isotherms in Figure 7-15) and a lower heat of adsorption. This results in a higher useful capacity and lower

![Typical isotherms of activated alumina, silica gel, and molecular sieve.](image)
regeneration heating requirements. The higher water capacity of activated alumina also ensures that the mercaptans’ removal specification will be met by reducing the chance of water breakthrough from a compound bed of activated alumina/different types of molecular sieves (Northrop and Sundaram, 2009).

7.4.4 Operation of solid-bed dehydrator

Optimizing the performance of the solid-bed dehydration unit requires a detailed understanding of its operation. The process flow schematic of a typical molecular sieve dehydration unit is shown in Figure 7-16. The picture of the corresponding dehydration unit is shown in Figure 7-17. The illustrated dehydration system is for a large gas processing plant, which typically consists of four solid desiccant beds, packed with molecular sieve and activated alumina. The number of drying beds varies depending on the feed gas flow rate and the mode of operation, as described below.

7.4.4.1 3 + 1 mode of operation

The dehydrators can operate in a 3 + 1 mode, i.e., three parallel beds drying and one bed on regeneration. In fact, three vessels will be on-line removing water vapor from the feed gas, while one bed is being regenerated with a slipstream of the dry (product) gas. For example, in processing 600-MMscfd feed gas, three adsorber vessels can be in parallel, handling 200 MMscfd of feed gas each. The

![Diagram of molecular sieve dehydration unit]

FIGURE 7-16
Process flow schematic of a typical molecular sieve dehydration unit. HC, Hydrocarbon.
advantages of three-bed parallel operation are lower pressure drop and an extended molecular sieve lifetime, due to less frequent regeneration.

The direction of flow for the solid-bed gas dehydration system is designed so that the adsorption, depressuring, and repressuring steps are downward flow through the bed, and the heating and cooling steps are upward flow. The dehydrator beds are configured to operate under a timed cycle. The control of each step is accomplished using a dedicated programmable logic controller.

**Adsorption step**: For three-bed parallel operation, the approximate cycle time for adsorption is 18 h followed by a 6-h regeneration cycle. The regeneration is composed of a 30-min depressuring step, a 3-h heating step, a 2-h cooling step, and a 30-min repressuring step. The cycle times are generally set by the molecular sieve vendors and can be adjusted during unit start-up. See Table 7-5 for a typical operating sequence of the four-bed dehydration system.

The dry gas leaves the drying beds and enters the outlet filters. The function of these filters is to remove molecular sieve or any other solid particles.

**Depressurization step**: At the completion of the 18-h adsorption cycle, the dehydrator must be depressurized to prepare the environment for the regeneration heating step, which uses dry residue gas at low pressure. The normal depressurization lines for the four dehydrators should have flow restriction orifices to effectively limit the rate of depressurization. In accordance with recommendations from molecular sieve vendors, the depressurization rate should not exceed 75 psi/min to prevent bed movement and fluidization. The gas will flow downward via the restriction orifices in order for the vessel to depressure within the allotted time of 20 min.

During the depressurization step, the regeneration gas is compressed in the regeneration compressor, and routed through the bypass valve to the regeneration
gas cooler where it is cooled. The cooled gas flows to the regeneration gas separator, and returns to the residue gas stream.

Heating step: Once the dehydrator has been depressurized, the heating step begins. The residue gas is continuously compressed by the regeneration gas compressors and flows through the bypass valve to the regeneration gas heater. The gas outlet temperature is controlled to 570–600 °F. The hot gas flows upward through the dehydrator being regenerated and heats up the bed to a temperature at which the water content of the molecular sieve is reduced to the required low level.

As a result, the water leaves the surface of the molecular sieve and is removed by the regeneration gas flowing upward through the bed. The heat required to desorb the water from the sieve can be as high as 1800 Btu/lb. The wet regeneration gas is then sent to the regeneration gas cooler. The gas is cooled to ambient temperature to form a two-phase stream; the liquid is the water desorbed from the dehydrator plus liquid hydrocarbons coadsorbed during the drying step. The two-phase stream is then separated in the regeneration gas separator, and the vapor is returned to the residue gas line. The condensed water and coadsorbed liquid hydrocarbons are removed from the separator on level control.

Cooling step: Following the heating step, the dehydrator is cooled with cold compressed regeneration gas to cool the molecular sieve to the normal adsorption temperature prior to repressurizing and returning it to normal service. The cool gas flows upward through the dehydrator and then to the regeneration gas heater,
so that the fired heater can be operated continuously. It is important for the fired heater to operate continuously to minimize thermal stress on the heating system, which would be the case if the heater is completely shut down during the cooling cycle.

At the beginning of the cooling cycle, the temperature of the gas from the dehydrator outlet to the fired heater is about 540 °F. To maintain the minimum fired heater duty required for continuous burner operation, the cold regeneration gas is split, sending about 75% into the dehydrator and 25% to bypass the dehydrator. The two streams combine downstream of the dehydrator before entering the regeneration gas heater. It leaves the heater at about 570–600 °F. With these conditions, the burner main fuel valve will remain almost closed most of the time and the burners will be firing near their lower operating point.

Repressurization step: After the cooling step, the dehydrator must be pressurized to prepare the vessel to be brought back online. The normal repressurization lines for the four dehydrators have flow restriction orifices to effectively limit the rate of repressurization to no more than 75 psi/min downward.

### 7.4.4.2 2 + 2 mode of operation

The dehydration process can also be arranged in the 2 + 2 mode of operation. In this operating mode, two dehydrators are in the adsorption cycle, and two dehydrators are in the regeneration cycle at any given time. In the regeneration cycle, one dehydrator is in the heating cycle and one dehydrator is in the cooling cycle, simultaneously. This allows circulation of the regeneration gas in series, which flows first to the vessel that is being cooled, then to the heater, and finally to the other vessel that is being heated. The flow sequence is shown in Table 7-6.

The regeneration in series reduces the regeneration gas consumption and uses the energy of the dehydrator that is being cooled down. The only drawback is the higher capital cost compared to the 3 + 1 mode, but it provides a more robust operation.

### 7.4.4.3 Other mode operations

There are other operating modes, depending on the number of beds and the dehydration unit operating conditions. It can be operated on 2 + 1 mode where two dehydrators are on the adsorption cycle, and one dehydrator is on the

<table>
<thead>
<tr>
<th>Time (hour)</th>
<th>Bed A</th>
<th>Bed B</th>
<th>Bed C</th>
<th>Bed D</th>
</tr>
</thead>
<tbody>
<tr>
<td>6</td>
<td>A</td>
<td>A</td>
<td>C</td>
<td>R</td>
</tr>
<tr>
<td>12</td>
<td>D</td>
<td>H</td>
<td>A</td>
<td>A</td>
</tr>
<tr>
<td>18</td>
<td>C</td>
<td>R</td>
<td>D</td>
<td>H</td>
</tr>
<tr>
<td>24</td>
<td>A</td>
<td>C</td>
<td>R</td>
<td>D</td>
</tr>
</tbody>
</table>

Table 7-6 Typical Operating Sequence of 2 + 2 Bed Dehydration System
regeneration and heating cycle. For smaller gas dehydration units, it can be arranged on $1 + 1$ mode, with one dehydrator on adsorption cycle and one on regeneration and heating cycle. For smaller solid-bed dehydration units, it may be more economical to use electric heater for sieve regeneration, which would also minimize emissions from the conventional fired heater.

### 7.4.5 Dehydration unit design considerations

The capital and operating costs of a solid-bed dehydration unit are high, considering the high-pressure adsorption beds, the regeneration heater, recycle compressor, and separation system. Some of the design considerations and options are summarized in the following discussion.

**Feed gas preconditioning:** It is crucial in the operation of a solid-bed dehydration unit to provide an efficient knockout drum to remove any liquid entrainment. Free water can damage the molecular sieves. On the other hand, entrained water droplets can often increase the water load by 200–300%. The knockout system must also have inlet filter separation equipment to remove solids and pipe scales to the parts per million levels.

The water content in the feed gas can be reduced by compression, chilling, and separation. The dehydration unit location in the gas processing plant should be placed at the highest pressure possible, which would reduce the saturated gas water content significantly. If the gas has to be compressed for NGL recovery, the solid-bed dehydrators should be located downstream of the compressors.

**Feed gas chilling:** Feed gas to the dehydrators typically comes from the acid gas removal unit and is saturated with water. The treated gas temperature is cooled by lean amine which is, in turn, cooled by ambient air. In hot climate operation, air temperature in the summer can be very high, and consequently, impacting the feed gas temperature.

To reduce the water content in the feed gas, the gas to the dehydrators can be cooled with cooling water or a chilled water system. As shown in Figure 7-1, reducing the feed gas temperature, say by 20 °F, can reduce the water content in the gas by 50%. Cooling the feed gas to the lowest possible temperature to about 70 °F or before the hydrate point may be the best way to reduce dehydration unit cost. The cooling always helps to remove heavy hydrocarbons and other contaminants that might come in with the feed gas.

The refrigeration duty in the feed gas chiller can be by propane refrigeration or can be from the cold residue gas stream from the NGL recovery unit. Potential energy savings by integrating with the other processing units can significantly reduce the cost of the dehydration unit.

**Pressure drop:** Pressure drop across an adsorber can be a significant cost penalty. It is desirable in an NGL recovery turboexpander unit to provide a high inlet pressure to the NGL distillation column. A high pressure drop in the dehydrator would lower the expansion ratio across the turboexpander and would reduce cooling by the expander, consequently lowering the NGL recovery.
efficiency. This is the reason for the popularity of the $3 + 1$ mode operation. With three beds operating in parallel rather than two, there is a lower operating pressure drop. The flow rate to each bed is one-third of the total rather than one-half. The pressure drop can be reduced to as low as 3 to 5 psi. The lower flow rate and pressure drop in each bed can also extend the useful life of the adsorbents.

The shape of the molecular sieve can have an impact on pressure drop. The beaded sieve typically has a lower drop than the pellet type, and may reduce up to 20% of the bed pressure drop. However, its performance may not be as predictable as the regular pellet type. The standard pellet-shaped molecular sieves are more common due to their proven performance in mass transfer, higher surface area, a shorter diffusion path and MTZ, and consistent physical strength over the other types.

*Cycle time:* Another obvious area for cost savings is to operate the adsorbers to their water breakthrough points. Molecular sieve deterioration increases with increase in the frequency of molecular sieve regeneration cycle. Each regeneration cycle deposits some carbon due to thermal decomposition of the heavy hydrocarbons from the feed gas. Reducing the number of regeneration cycles by operating the adsorbers to close their water breakthrough points will prolong the life of the molecular sieve. Bed temperatures and moisture analyzers should be used to vary the operation to operate at close to its water breakthrough point.

*Regeneration methods:* There are two basic ways to regenerate a solid desiccant bed. They are temperature swing adsorption (TSA) and pressure swing adsorption (PSA) in which the change in the adsorption equilibrium is obtained by increasing the temperature and decreasing the pressure, respectively. TSA is generally used for natural gas drying or removal of $\text{CO}_2$ from natural gas streams to meet stringent product specifications. PSA is typically used for drying of air and industrial gases for general use, and not used for natural gas plants where stringent specifications are required.

*Regeneration gas heater:* Depending on the mode of operation, the fired heater can be subjected to cyclical operation. A heater is considered cyclical when it is shutoff and then brought back on by reigniting the burners. The three-bed parallel arrangement operates in this manner, since the heater shuts down while the regenerating bed is in the cooling cycle. A heater endures significantly less thermal stress as long as the burners remain on at all times. This is the case for the $2 + 2$ mode operation where the heater is always kept at its minimum turndown capability. Minimum turndown is defined as the minimum heat release, which can maintain a stable burner flame. The typical heater turndown ratio is 10:1.

A cyclical heater has several issues that must be addressed to avoid the thermal stress due to cycling operation. By properly addressing the heater for cyclical operation, the heater can be designed with reliability and durability close to a heater in continuous service.

*Regeneration gas flow direction:* The regeneration gas flow direction for regeneration heating and cooling affects the efficiency of water removal from the molecular sieve bed. For three-bed parallel operation ($3 + 1$ mode operation), the
heating and cooling are in the upflow direction because it results in better overall regeneration than the alternative, heating upflow and cooling downflow. Specifically, upflow cooling ensures good dew point control and prevents the potential leakage of wet feed gas into the cooling gas across a defective switching valve.

For two-bed series operation (2 + 2 mode operation), upflow heating and downflow cooling are typical because the potential leakage of wet feed gas into the cooling gas does not exist for series operation. However, different manufacturers can recommend different flow directions for regeneration heating and cooling to maintain the performance of their molecular sieves.

*Insulation:* Solid desiccant vessels are insulated externally or internally. Internal refractory requires careful installation and curing, usually before the desiccant is installed. It saves energy and can dramatically reduce the required heating and cooling times. This is an important benefit for systems where regeneration times are limited. However, internal insulation is prone to cracking due to thermal stress, which may retain moisture in the cracks during the regeneration cycle, resulting in off-spec product during the absorption cycle.

*Regeneration gas compressor:* The pressure ratio across the compressor depends on the required regeneration pressure. Lower pressure improves regeneration but it would also increase the regeneration compressor horsepower. Conversely, increasing the regeneration pressure makes regeneration more difficult and will take longer and higher temperature, but the compression horsepower would be reduced.

The regeneration gas compressor can also be integrated with the upstream facility where there is a recompressor. For example, the stabilizer overhead or stripper overhead in the condensate stabilization unit typically includes an overhead compressor that can be used to recompress the regeneration gas back to the inlet of the of the acid gas removal unit.

### 7.4.6 Operational problems

The molecular sieve design details and operation have been described in the previous sections. Problems associated with molecular sieve operation generally relate to misoperation of the upstream units, and improper installation and operation of the molecular sieve beds. The following sections summarize these problems and provide guidelines to mitigate these problems.

*Inlet conditions:* Often a small change in inlet temperature can significantly increase the water content in the feed gas, which can impact the solid-bed dehydration system performance (see Figure 7-1). Also, the adsorbents capacity drops as the temperature increases (see Figure 7-12).

A drop in inlet gas pressure will evaporate more water, increasing its water content. The lower feed pressure will also increase the gas velocity through the bed and will extend MTZ, which may result in an earlier breakthrough.

Changes in gas composition should be monitored. If the inlet gas source is changed from lean gas fields to rich gas fields, more liquids can enter the
dehydration unit. The slug catcher and inlet separator may not be adequate, allowing more liquid hydrocarbons and water to enter the dehydration unit.

High-pressure gas dehydration may reduce the water content in the feed gas, which is advantageous, but may create a different problem. Depending on the feed gas composition, there is a point where hydrocarbon liquid may drop out as the pressure is reduced when transversing the retrograde condensation region. This can result in wetting the adsorbents due to the normal pressure drop gradient throughout the bed. Retrograde condensation is very difficult to predict. The condensate may be detected by a steadily increasing pressure drop during the adsorption cycle due to liquid accumulation on the adsorbent beds. The only solution is to avoid the retrograde region as much as possible by increasing the feed temperature.

Contamination: The most frequent cause of solid desiccant bed failure is the incomplete removal of contaminants from the inlet gas. Below is a list of the most common contaminants.

Liquid carryover

Liquid carryover can be due to design deficiency of the feed separator or malfunctioning of the upstream amine treating unit. The cause can also be due to upset of the regeneration gas separator. During system start-up or shut-down, the regeneration compressor can surge and can entrain liquid from the regeneration gas separator. In the worst scenario, the liquid (hydrocarbon liquids and water) can flood the dehydrators and pulverize the adsorbents. The sieves can agglomerate and form a pastelike substance, as can be seen in Figure 7-18.

To avoid the liquid carryover, the regenerator gas separator should be designed with emergency shutdown system with high liquid level alarms that should shut down the compressor and stop the flow to the dehydrator in case of high liquid level.

Figure 7-18

Damaged sieve at bottom of dehydrator.
As for the adsorbent materials, some manufacturers can provide more robust and liquid-tolerant adsorbents that can handle the transient problems and minimize the adsorbent damage.

**Salt entainment**

Salt contamination comes from treating the units from salt-water-bearing formations, or from salt dome storage caverns. Salt will accumulate and block the opening and pores in the sieves. In extreme cases, all the voids between the molecular sieve beads will plug up, resulting in an increase in pressure drop. Eventually, the adsorbent must be replaced.

To minimize the salt content in the feed gas, the design of inlet scrubbers and separator coalescer is extremely important. In some gas plants, water washing of the feed gas coupled with inlet cooler is used to protect the solid-bed dehydrators.

**Corrosion products**

Corrosion materials can come from upstream equipment and piping. H₂S in the feed gas can react with carbon steel equipment during regeneration to produce pyrite flakes. These can deposit on the molecular sieves surface and can block the bottom outlet screen, which can result in maldistribution among the dehydrators.

The corrosion materials can also deposit on the fire tube on the regeneration heater, which may result in hot spots damaging the heater.

**Oxygen**

If there is any oxygen in the feed gas stream, it may react with H₂S and other sulfur compounds on the molecular sieve surface to produce elemental sulfur. In extreme cases, this will not only block macro- and micropores but also the spaces between the sieves, resulting in one large “lump” blockage, which is difficult to remove.

Oxygen can also react with heavy hydrocarbons in the feed gas to form coke during the regeneration cycle, causing fouling of the molecular sieves and the regeneration heater.

Oxygen can be present in the feed or enter through the vapor recovery system, or other means. It is good practice to monitor the oxygen content in the feed gas during daily stream analysis. If trace oxygen is detected early, the unit can be repaired, and the source can be isolated.

*Incomplete regeneration:* Insufficient or incomplete regeneration of the adsorbent beds will lead to a sudden loss in adsorption capacity and premature breakthrough. Insufficient adsorbent bed regeneration is a result of one or all of the following factors:

- Low regeneration gas flow rate/temperature
- Insufficient regeneration time
- Change of regeneration gas composition
To fully regenerate the adsorbents, the inlet and outlet temperatures of the adsorber in the regeneration step should be monitored. In order to make sure that the adsorbents are properly regenerated, three points have to be checked:

1. The inlet temperature should reach the temperature required to adequately regenerate the bed (depending on the desiccant type).
2. The outlet temperature should show an almost constant value (during the heating step) for 30–120 min, depending on the vessel size and cycle time (see Figure 7-19). This is necessary to make sure that the adsorbents near the vessel walls are fully regenerated.
3. The temperature difference between inlet and outlet streams at the end of the heating cycle should not be more than 50–60 °F depending on the vessel insulation (see Figure 7-19).

**Bed refluxing:** Heating up the adsorber without using a heating ramp or an intermediate heating step may result in a large temperature difference across the bed. During regeneration, as the regeneration gas flows from the bottom to the top, water is desorbed and readSORBED many times. Often it is carried off at such a rapid rate that it condenses on the cool top head of the vessel and dripped back down onto the bed. This phenomenon is called water refluxing, which increases with high pressures and low regeneration flow rates. At some point, the ascending heat zone mixes with the descending water resulting in boiling, which may cause hydrothermal damaging of the molecular sieves. Hydrothermal damaging may alter the binder material and destroy the crystal structure of the zeolite.

In order to prevent hydrothermal damaging of molecular sieves, it is important to choose not only the right formulation of the molecular sieve (binder and zeolite) but also the proper regeneration conditions. The adsorbent supplier should be consulted in troubleshooting the problems. Most suppliers have seen

![Figure 7-19](image-url)

**FIGURE 7-19**

Typical inlet and outlet temperatures during solid-bed regeneration cycle.
many different misoperations and molecular sieves damages and their experience can save time and money.

**Improper bed loading:** The same quantity of each adsorbent grade in each vessel must be used to prevent unequal flow distribution among the different beds to avoid premature breakthrough. For optimum bed loading, the bed may require a mixed bed of large and small particles, typically with the small particles being at the bottom of the bed. If the adsorbents are loaded differently among the beds, the MTZ may be different among beds, which may prolong the adsorption time to meet dew point specification.

**Unequal flow distribution:** Sometimes premature breakthrough occurs on only one vessel and this is the case when working with several vessels in adsorption in parallel. If there is a problem with one vessel only it is in general an issue of bed configuration and not of the molecular sieves. Unequal flow distribution due to bed configuration or internal misalignment can be detected by a high pressure drop difference in the vessels during adsorption.

**Bottom support:** Experience shows that more and more problems of solid-bed failure are caused by failure of the support grid. Nevertheless, very often the reason is not the support grid itself but the way how it was installed. The mechanical design of the support bed must be strong, and quantity and size of ceramic balls must be installed according to manufacturer guidelines.

**Bed bumping:** A rapid depressurization possibly from a sticking valve can lift the bed and even fluidize it. Normally a solid desiccant bed should not be depressurized at a rate greater than 50 psi/min, and under normal operation the operating pressure must be within 30 psi before the process valves are opened. Bed pumping will crush the molecular sieves making them in-operable. Severe damage may require dumping and replacing the sieves.

**Switching valve leakage:** The switching valves on the molecular sieve beds ensure extreme temperature and pressure fluctuation on a daily basis, varying from ambient temperature during the adsorption cycle to 500 °F during the regeneration cycle. These cyclic operations impose undue stress on these valves, which may affect the valve performance. If leakage occurs, this may result in an off-spec product. The orbit valves are more reliable but they are also expensive. High-performance butterfly valves are less costly but more prone to leakage. The switching valves must be monitored and maintained routinely.

**Moisture analyzer:** To optimize the life of the molecular sieves, reduce energy, and prevent wet gas from contaminating downstream equipment, a highly reliable and rapid-response moisture analyzer is crucial. Some of the recent advances in noncontact laser-type detector are now available. The moisture analyzer must be selective so that interferences from other compounds (such as methanol) are eliminated. The response rate of the analyzer must be fast, detecting rapid breakthrough in the molecular sieve beds.

**Internal insulation failure:** The internal insulation in the molecular sieve beds also goes through severe thermal stress during the adsorption and regeneration cycles. If the insulation is damaged, moisture may be retained within the
insulation, which may be the cause of off-spec products. The insulation should be repaired and replaced as necessary.

*Molecular sieve handling safety:* Many vintage gas processing plants have been operating for many years without any incidence. Typically, the H2S content of the inlet feed gas has increased over the years. Contractors and staff in these plants must be constantly reminded of the danger and hazards associated with operating a molecular sieve unit.

Typically, during shutdown, molecular sieve units are purged with nitrogen to remove H2S content from the vessel before entering. However, there is still H2S that is tightly adsorbed by the molecular sieve. When molecular sieves are changed out, they are usually dumped into water to avoid the pyrophoric reaction. When molecular sieves are in the water, H2S will be displaced from them, as sieves have higher affinity toward water than H2S. There were accidents in the past relating to fatality from H2S poisoning during the change-out operation. The molecule sieve operation program must insist personnel to carry H2S monitors and escape masks even when the unit feed gas contains low levels of H2S.

### 7.5 Other gas dehydration processes

Other less frequently used dehydration methods can be applicable to some operations listed below.

**Calcium chloride:** Calcium chloride can be used as a consumable desiccant to dry natural gas. Anhydrous calcium chloride combines with water to form various calcium chloride hydrates. As water absorption continues, calcium chloride is converted to successively higher states of hydration, eventually forming a calcium chloride brine solution. Outlet water contents of 1 lb/MMscf can be achieved with calcium chloride dehydrators.

Calcium chloride dehydrators may offer a viable alternative to glycol units on small-feed-rate, remote dry gas wells. The calcium chloride must be changed out periodically. Brine disposal is an environmental issue.

**Methanol refrigeration:** Dehydration of natural gas to \(-150 \, ^\circ F\) using methanol as the hydrate inhibitor has been used for ethane recovery in the Empress plant. The condensed water and methanol streams can be decanted in the cryogenic unit and can be separated by distillation. This can be found in a patented process called IFPEX-1®.

The difficulty in the methanol refrigeration process is the high methanol losses associated with the high vapor pressure. The process must work with cryogenic temperature refrigeration to minimize losses. The methanol process is more complex in terms of operation than the molecular sieves process, and is seldom used in NGL recovery processes today.

**Membrane and Twister:** Membranes and Twister technology can be used to remove water and hydrocarbons to meet pipeline water and hydrocarbon dew point. Please see Chapter 8 for details of the discussions on these subjects.
7.6 Gas dehydration process selection

The natural gas dehydration process is selected based on the water dew point requirements, as shown in **Figure 7-20**.

The process selection can be quite straightforward. If dehydration is required only to meet the pipeline specification of 4–7 lb/MMscf, any of the above-mentioned processes can be applicable. Typical glycol dehydration process is suitable for meeting pipeline gas specification; as low as $-40^\circ F$ and is more economical than molecular sieves technology. Solid desiccant dehydrators are chosen for deep dehydration to meet a low water dew point for NGL recovery or LNG production. Membrane and Twister processes are suitable for small gas plants and offshore installation where space is limited.

If hydrocarbon dew point control is required, cooling with inhibitor, such as the use of ethylene glycol injection with propane chilling or the use of silica gel, can be suitable. It can also be accomplished with TEG dehydration unit followed by propane chilling.

As for the use of silica gel dehydration process, because of its limited short cycle time, it is more suitable for polishing services, such as membrane separator pre-treatment, and fuel gas conditioning rather than feed gas to a gas processing plant.

For water dew pointing lower than $-40^\circ F$ and down to $-80^\circ F$, the enhanced TEG dehydration process, such as DRIZO, is suitable. TEG dehydration unit is lower in Capital Expenditure (CAPEX) and Operating Expenditure (OPEX) than the molecular sieve unit, especially for offshore installation where plot space is limited. There are other considerations that may favor the use of molecular sieves. Molecular sieves are selective and do not appreciably adsorb BTEX components hence avoiding the emissions problem in conventional TEG dehydration unit. However,
advances in enhanced glycol dehydration processes, such as the DRIZO system, can solve the problem by production of a BTEX hydrocarbon liquid. Molecular sieves can be designed to remove mercaptans to meet the sulfur specifications. The mercaptans are concentrated in the regeneration gas, and can be selectively removed using a physical solvent process, such as the DEPG process. The mercaptan components can then be converted to sulfur in a Claus sulfur recovery unit to eliminate sulfur emissions.

7.7 Mercury removal

Almost every NGL production plant will have a mercury removal unit installed. This is because the consequence for the NGL recovery unit from mercury attack is severe and because it is difficult to predict the mercury contents from production reservoirs. Low levels of mercury can result in amalgamate corrosion of the brazed aluminum heat exchangers in cryogenic systems, and potentially can pose environmental and safety hazards. The presence of mercury in the feedstocks to petrochemical plants will also cause poisoning of precious metal catalysts. For this reason, the NGL plant is conservatively designed for mercury removal to levels below 0.01 \( \mu \text{g/Nm}^3 \).

Most of the current mercury methods use fixed-beds method. The gas flows through the mercury removal beds, operating in a lead–lag manner to produce a mercury free gas.

There are two types of mercury removal materials: nonregenerative sorbents and regenerative adsorbents.

7.7.1 Nonregenerative mercury sorbents

In nonregenerative mercury removal process, the mercury reacts with the sulfur to form a stable compound on the sorbent surface. A number of different mercury removal sorbents are available with various tolerances to operating temperature, liquid hydrocarbons, and water.

The use of sulfur-impregnated activated carbon is a common process for mercury removal, in which mercury is securely attached to the sulfur compound in the microporous structure of the carbon bed. However, there are drawbacks to this method, as the sulfur-impregnated carbon can only be used with dry gas due to its small pore size. Any water will be preferentially adsorbed, which would restrict the access of mercury to the sulfur sites. Sulfur can also be lost by sublimation and dissolution in hydrocarbon liquids. This again reduces the life of the mercury beds and its mercury removal capacity. The mercury-laden carbon bed is hazardous and difficult to dispose.

Recognition of these problems has led to the development of a range of nonregenerable absorbents utilizing transition metal oxides and sulfides that can be safely handled. In these systems, the reactive metal is incorporated in an inorganic support and the absorbent is supplied with reactive sulfide component.
by either ex situ or in situ sulfiding. Johnson Matthey has taken this concept further and supplied an established range of absorbents marketed under the PURASPEC™ brand. The PURASPEC™ materials are a mixture of copper sulfide/copper carbonate, zinc sulfide/zinc carbonate, and aluminum oxide, which can operate in a wet gas environment.

Note that the nonregenerative methods appear to be simple since no regeneration equipment is required. However, disposal of the used sorbent must follow a safety procedure since the sorbent will contain mercury and other hazardous material such as benzene and cyanides or other contaminants.

7.7.2 Regenerative mercury adsorbents

The regenerative mercury removal process utilizes silver impregnated on molecular sieve (such as UOP’s HgSIV sieve) to chemisorb elemental mercury. The mercury-saturated bed is regenerated by hot regeneration gas typically at 550 °F. The mercury content now shows up in the condensed water in the regeneration gas separator. This method avoids accumulation of mercury on the adsorbent, but has not eliminated the mercury disposal problems. Trace of mercury may still show up in the regeneration gas.

7.7.3 Process selection considerations

There are four possible options to handle to the mercury problem in natural gas plant feed and product streams:

Option 1: Installing nonregenerative mercury removal sorbents at the gas plant inlet, before the amine unit. This option removes all the mercury and ensures no mercury contamination for the rest of the plant. The large volume of feed gas requires large vessels which is very costly, but this is a very safe and conservative approach in handling mercury in the feed.

Option 2: Installing a nonregenerative mercury removal sorbent downstream of the Acid Gas Removal Unit (AGRU), just before the molecular sieve unit. This option reduces the size of the beds to some extent, but it poses the risks of mercury contamination in the AGRU solvent system.

Option 3: Add a silver-impregnated mercury sieve section to the molecular sieve beds. While this option can remove water, mercaptans, and mercury at the same time, and avoid the need of a separate mercury bed, it presents problem with a high mercury content in the regeneration water that would pose operating hazards unless treated by another mercury removal step.

Option 4: Installing a nonregenerative mercury removal bed or a silver-impregnated molecular sieve bed after the molecular sieve unit. This option allows the use of low-cost carbon beds, as the feed gas is dried and clean. It may only be applicable if the feed gas has negligible amount of mercury. If there are significant amounts, there can be mercury contamination in upstream units, including the AGRU, which may pose health and safety hazards.
Life cycle costs, adsorbent disposal methods, mercury levels, environmental limits, operating hazards, and plant operator procedure must be evaluated in the selection of a suitable mercury removal system. The optimum mercury removal method can also be a combination of nonregenerative and regenerative mercury removal systems.

References


Carroll, J.J., March 11–13, 2002. The water content of acid gases and sour gas from 100 to 220°F and pressures to 10,000 Psia. Paper Presented at the 81st GPA Annual Convention, Dallas, TX, USA.


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8.1 Introduction

Most natural gas sources contain hydrocarbon liquids that must be removed to meet the hydrocarbon dew point and heating value specifications of the pipeline gas before they can be used by the consumers. The removal of heavy hydrocarbons is necessary to ensure pipeline transportation safety. The hydrocarbon liquids consist of two components, the natural gas condensate (C_{5+}) and the natural gas liquids (NGLs) (C_{2–C_{4}}). These components can be sold at a premium over natural gas for the equivalent heating value. The C_{5+} condensate is separated in the condensate stabilization unit that has been discussed in Chapter 5. The C_{3–C_{4}} liquids are valued as a liquid fuel. The C_{2} component can be sold as petrochemical plant feedstocks. Separation of the NGL components are discussed in this chapter.

The richness of the NGL components can be identified by the term “GPM,” that is, gallons per mole of hydrocarbon (C_{2+}) liquid. The value of “GPM” and the higher heating value of the hydrocarbon component are shown in Table 8-1.

The C_{2+} NGL recovered from an NGL recovery unit is termed “Y-grade” NGL. The acceptable Y-Grade is a mixture of NGLs composed principally of ethane, propane, butane, pentanes, and natural gasoline, which typically meets the specifications given in Table 8-2.

The Y-grade liquids must be free from sand, dust, gums, gum-producing substances, oil, glycol, inhibitor, amine, caustics, chlorides, oxygenates, heavy

| Table 8-1 Recoverable Hydrocarbons and Heating Values of NGL |
|-------------------------------------------------------------|----------------|
| Recoverable Hydrocarbons, Gal/lb mole | HHV, Btu/scf |
| C_{1} | 1010.0 |
| C_{2} | 1769.6 |
| C_{3} | 2516.1 |
| iC_{4} | 3251.9 |
| nC_{4} | 3262.3 |
| iC_{5} | 4000.9 |
| nC_{5} | 4008.9 |
| nC_{6} | 4755.9 |

*HHV, High Heating Value.*
metals, and any other contaminants or additive to the product added to enhance the ability to meet specifications.

This chapter covers the production of NGL from hydrocarbon dew pointing to propane and ethane recovery, including their history, and various technologies and design options as well as NGL fractionation.

8.2 Refrigeration processes

Refrigeration units are designed to meet the process cooling temperature requirements. In NGL recovery or natural gas liquefaction plant, a three-stage propane compressor is the most common, typically designed to meet the process temperature of about $-20^\circ F$. If lower temperatures are required, ethane refrigeration or mixed refrigeration system can be used. Alternatively, gas expander can also produce deep refrigeration using feed gas as the working fluid.

8.2.1 Propane refrigeration

Propane is a common refrigerant and has a very low environmental impact compared to other alternatives. The main difference is that propane is a flammable refrigerant, while the fluorocarbon refrigerant is nonflammable. Operating a propane refrigeration system requires proper training on safety requirements.

Refrigeration system operates on a vapor compression refrigeration cycle principle. Figure 8-1 plots the horsepower required per MMBtu/h of cooling duty against the propane evaporating temperature for various propane condensing temperatures. In a gas processing plant, propane refrigerant is typically

<table>
<thead>
<tr>
<th>Table 8-2 Y-Grade NGL Specifications</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Characteristics</strong></td>
</tr>
<tr>
<td>Composition</td>
</tr>
<tr>
<td>Methane, maximum</td>
</tr>
<tr>
<td>Aromatics, maximum</td>
</tr>
<tr>
<td>Olefins, maximum</td>
</tr>
<tr>
<td>Carbon dioxide</td>
</tr>
<tr>
<td>Corrosiveness</td>
</tr>
<tr>
<td>Total sulfur</td>
</tr>
<tr>
<td>Distillation: end point at 14.7 psia</td>
</tr>
<tr>
<td>Free water</td>
</tr>
<tr>
<td>Product temperature</td>
</tr>
</tbody>
</table>

*LV, Liquid Volume.*
condensed by cooling water or ambient air. For a fixed condensation temperature, the compression horsepower increases with lower refrigerant evaporating temperatures. This is mainly because a lower evaporating temperature also requires a lower evaporating pressure, which requires a higher compression ratio. Conversely, refrigeration horsepower is lower with higher evaporating...
temperature. Refrigeration unit, which is typically designed for the summer operation, would have more cooling capacity during winter.

Propane refrigeration can provide coolant at $-40\,^\circ F$ at atmospheric pressure. Lower chilling temperatures could be achieved with a fourth stage compression operating under vacuum pressure. However, most gas plants would avoid the use of vacuum pressure because of the potential explosion hazards from air ingress into the system.

Most gas plants use a three-stage propane compression system as shown in Figure 8-2. Centrifugal compressors are used since they typically have higher efficiency than screw or rotary compressors.

The feed gas cooling system typically consists of two chillers operating in series. The first chiller uses the refrigerant from the Medium Pressure (MP) drum, while the second chiller uses the atmospheric drum liquid. Cooling in two stages improves the thermodynamic efficiency, lowering the power consumption.

The chillers are typically kettle-type exchangers with propane evaporating on the shell side. The evaporated propane from the first-stage chiller is sent to the second-stage compressor suction, and the vapor from the second-stage chiller is sent to the first-stage compressor suction. The propane compressor also has a third-stage that receives vapor from the High Pressure (HP) suction drum.

The aerodynamics of a centrifugal compressor typically requires the compressor flow to stay above 60% of the design flow. Below this point, the compressor will be in a “surge” condition and become unstable. Compressor recycle vapor (RV) is required for compressor protection, when the gas plant is operating under turndown conditions, during plant start-up or shutdown. A desuperheating station (DS) must be installed, which cools the compressor recycle gas by injecting as desuperheating liquid (DL) to maintain the design suction temperatures. The propane compressor control and the propane desuperheating schematic are shown in Figure 8-2.

### 8.2.2 Cascade refrigeration

Propane refrigeration can produce refrigeration at $-40\,^\circ F$, which is adequate for the hydrocarbon dew pointing operation. However, when high propane recovery and ethane recovery is required, lower refrigerant temperature is required. Cascade refrigeration can be an option to generate a low-temperature refrigerant. A cascade system consists of two refrigeration circuits, with one higher boiling refrigerant (propane) being used to condense a lower boiling refrigerant (ethane).

A typical cascade refrigeration schematic is shown in Figure 8-3. The feed gas is first cooled in the first-stage chiller and second-stage chiller in the propane refrigeration circuit, and then in the first-stage chiller and second-stage chiller in the ethane refrigeration circuit. The propane condenser is cooled by ambient air or cooling water and the ethane condenser is cooled by evaporating propane at $-40\,^\circ F$. With such arrangement, ethane refrigerant at $-120\,^\circ F$ can be produced for cooling in the demethanizer in a gas plant.
FIGURE 8-2
Process flow schematic of a three-stage propane refrigeration system. RL, Recycle Liquid.
Because of the high equipment counts, capital and operating costs, and complexity in operation, cascade refrigeration system is seldom used in gas plants for NGL recovery.

8.2.3 Mixed refrigerants

An alternative to the multistage cascade refrigeration is to use a single mixed refrigerant. Multiple component refrigeration is common in natural gas liquefaction plants. To maximize the efficiency of the mixed refrigerant process, refrigerant composition must be adjusted such that the evaporating heat curve of the mixed refrigerant matches closely with the natural gas condensing heat curve. Compact heat exchangers such as brazed aluminum heat exchangers (BAHXs), spiral wound exchangers, or other proprietary exchangers are required. Although equipment count is lower than cascade refrigeration, mixed refrigeration system is also seldom used in NGL recovery, because of the operation complexity. Details on the design and operation of a mixed refrigerant
process can be found in the *Handbook of Liquefied Natural Gas* by Mokhatab et al. (2014).

### 8.3 Liquid recovery processes

There are different levels of NGL recovery. It can vary from the simple hydrocarbon dew pointing unit to the more complex cryogenic turboexpander units. Within each category, there are variation in the designs that can be applied to meet specifications and operation requirement. These processing units are discussed in the following sections.

#### 8.3.1 Hydrocarbon dew pointing with Joule–Thomson cooling

If feed gas is available at high pressure or at supercritical pressure, the gas pressure can be used to generate cooling by isenthalpic expansion. The gas cooling effect will cause heavy hydrocarbons to condense. Glycol injection is typically required to avoid hydrate formation due to the presence of water in the feed gas. As discussed in Chapter 7, glycol injection with feed gas cooling can be used to control the water and hydrocarbon dew points at the same time. Alternatively, the feed gas can be dried using triethylene glycol (TEG) dehydration upstream as shown in Figure 8-4.

As shown in the flow diagram, dried gas from the TEG dehydration unit is cooled with the cold separator vapor in a gas/gas shell and tube exchanger. The chilled gas is reduced in pressure using a Joule–Thomson (J–T) valve. The J–T letdown operation cools the gas further, producing a liquid condensate in the cold separator.

The pressure of the cold separator pressure is typically maintained between 600 and 700 psig, depending on the feed gas composition. High separator

![Flow diagram](image-url)  
**FIGURE 8-4**  
Hydrocarbon dew pointing using J–T process.
pressure would reduce the recompression horsepower and operating cost. However, high pressure also makes phase separation more difficult, requiring a larger separator. For practical purposes, the separator pressure should stay at 10–20% below the critical pressure of the gas mixture.

The liquid from the cold separator is further processed in a stabilizer, which separates the methane component from the NGL, and recompressed to the sales gas pipeline. The stabilizer is heated with steam or heat medium to produce a Y-grade NGL, which must meet the specifications in Table 8-2. The stabilizer typically operates between 100 and 200 psig. The stabilizer pressure is selected based on NGL liquid composition. It is desirable to operate the stabilizer pressure as high as possible to minimize gas compression cost. However, high pressure also increases the bubble point temperature of the stabilizer bottom, which requires a larger reboiler, or may also result in thermal degradation. Therefore, the stabilizer pressure should be optimized to minimize compression while maintaining a reasonably sized reboiler exchanger.

In many midstream dew pointing units, J–T units are the most common. They can be acquired as low cost standard modularized units. With adequate pressure drop, relatively low temperatures can be produced for hydrocarbon dew point control.

The main difference between a J–T unit and a turboexpander unit is that the gas expansion is adiabatic across the J–T valve, while the expander operates on an isentropic path. Thus, the J-T design tends to be less efficient and generates less cooling than turboexpander. However, the J–T process offers several advantages over turboexpander and refrigeration processes, such as low cost, simple operation, no rotating equipment and low maintenance, high turndown, as well as capability of operating a wide range of flow and gas compositions. For small gas processing plant operation, J–T is the technology of choice to produce on-spec hydrocarbon dew point gas for sales.

8.3.2 Hydrocarbon dew pointing with propane refrigeration

If the feed gas is supplied at low pressures, there will not be sufficient pressure to operate the J–T process. In this situation, the feed gas can be chilled at pressure using propane refrigeration. The basic process schematic is shown in Figure 8-5. The process configuration is similar to the J–T process, with the exception that the J–T valve is replaced by a propane chiller. The propane chiller is a kettle-type exchanger with propane evaporating on the shell side. The liquid level on the kettle can be varied to control the cold separator temperature.

The use of the propane refrigeration to recover liquids is a more efficient process than the J–T process. The feed gas pressure can be maintained, which would minimize recompression requirement. It is a trade-off of propane compression to feed gas compression, which can be evaluated in an optimization study.
8.3.3 Deep hydrocarbon dew pointing

J-T units or refrigeration units can be used to meet hydrocarbon dew point specification of the sales gas. Typically, with a low ethane content gas, the heating value specification of 1100 Btu/scf can be met. However, with a higher ethane content feed, the heating value of the residue gas may exceed specification. Particularly, in western United States and Europe, a very lean gas with low British thermal unit content is required. California pipeline typically operates with heating values as low as 970 Btu/scf (Mak et al., 2004b), mainly to minimize emissions.

To meet the heating value specifications for these markets, most of the propane and butane components must be removed. The conventional J–T process or the refrigeration process may not be enough. Consequently, a deeper hydrocarbon dew pointing unit is required, as shown in Figure 8-6. The process requires both

![Figure 8-5](Image)

**Figure 8-5**
Hydrocarbon dew pointing using propane refrigeration.

![Figure 8-6](Image)

**Figure 8-6**
Deep hydrocarbon dew pointing process.
propane refrigeration and J–T expansion. This deep hydrocarbon dew pointing unit requires two columns: an absorber and a demethanizer operating at below −50 °F temperature. The feed gas must be dried to below −60 °F to avoid hydrate formation in the absorber. Therefore, molecular sieve dehydration or Drizo dehydration unit must be used instead of conventional TEG dehydration unit. Please refer to earlier chapter for discussion of these two designs.

The innovation of this process is the reflux exchanger, which uses the absorber overhead vapor to cool and condense the stabilizer overhead vapor, forming a cold reflux stream to the absorber. The reflux system serves two functions. First, it recycles the C₂⁺ components to the absorber. Second, the reflux stream condenses the C₃⁺ component from the feed gas in the absorber. Consequently, with sufficient J–T cooling, the process can remove most of the propane and heavier components and meet the heating value specification of the sales gas. Propane recovery of over 90% can be achieved with this deep hydrocarbon dew pointing process.

8.3.4 Turboexpander NGL recovery processes

The term “turboexpander” refers to an expander/compressor machine as a single unit. It consists of two primary components: the radial inflow expansion turbine and a centrifugal compressor integrated as a single assembly (see Figure 8-7). The expansion turbine is the power unit and the compressor is the driven unit.

In cryogenic NGL recovery processes, the turboexpander achieves two different but complementary functions. The main function is to generate refrigeration to cool the gas stream. This is done by the expansion turbine end that
expands the gas isentropically by extracting the enthalpy from the gas stream, causing it to cool. The other function is the use of the extracted energy to rotate the shaft to drive the compressor end of the turboexpander, which recompresses the residue gas stream.

The first turboexpander unit was built in 1964 for NGL recovery in the city of San Antonio, Texas. The gas is supplied at 700 psig pressure and is letdown in pressure to about 300 psig to the demethanizer. Methanol injection was used for hydrate inhibition. Until this time, Liquefied Petroleum Gas (LPG) recovery was mainly achieved with refrigerated lean oil, which is described in the later section.

The first turboexpander process patent was issued to Bucklin (Fluor) in 1966. The flow schematic is shown in Figure 8-8. The concept was to use turboexpander, which is a more efficient method of cooling, instead of the J–T valve. The feed gas is cooled by the cold demethanizer overhead, and separated in the HP cold separator. The separator vapor is letdown in pressure using the turboexpander and fed to the top of the demethanizer as a reflux. The HP cold separator liquid is letdown in pressure to the Low Pressure (LP) cold separator. The separator liquid is further letdown and used to cool the feed gas before it is fed to the lower part of the demethanizer. The demethanizer bottom product is heated with steam that controls the methane content (see Table 8-2).

Earlier NGL recovery units did not have the benefits of the compact BAHX and would require multiple shell and tube heat exchangers to achieve the chilling requirements. Even with extensive heat integrated scheme as shown in Figure 8-9,
FIGURE 8-9
Process flow schematic of earlier turboexpander NGL recovery unit.
high NGL recovery could not be achieved. The other limitation of earlier expander
designs is the low expansion ratio, which required two expanders operating in
series to achieve a high expansion ratio. A higher expansion ratio would generate
more cooling that is necessary for high NGL recovery.

The process flow schematic of a typical turboexpander NGL recovery
unit built in the 1970s is shown in Figure 8-9. The process uses two
expander–compressor sets operating in series and several separators to produce
various liquid streams that are fed to different locations in the demethanizer.
The process also requires propane refrigeration for feed gas cooling. The NGL
recovery process could achieve about 70% ethane recovery and was typically
based on a relatively rich feed at about 950 psig pressure. Most of these NGL
recovery plants are still operating today, but they are processing much leaner
gas as the reservoirs are depleted. The leaner gas has an impact on the NGL
recovery level. The heat exchanger surface areas are no longer adequate and the
expanders are operating at lower efficiency. Consequently, ethane recovery
drops to about 55%. A picture of the earlier NGL recovery unit is shown in
Figure 8-10.

The main contributors to the success of today’s NGL recovery plants are the
turboexpanders and the brazed aluminum exchangers. The application of tur-
boexpanders to the natural gas industry began in the early 1960, which was
followed by the development of BAHXs.
8.3.4.1 Turboexpander
Today’s turboexpander designs can yield very high adiabatic efficiencies (over 85%). However, there are aerodynamic limitations for both the expander and compressor. Machinery efficiencies will drop if the gas composition or flow rate were different than the design points. In most instances, when the changes are temporarily, they can be managed by the expander bypass J–T valves. During the expander bypass operation, the NGL unit temperature profile will increase and operation of the turboexpander will be inefficient; consequently, NGL recovery and plant capacity will be reduced. If the flow conditions are expected to be permanent, rewheeling of the expander and compressor can be economically justified.

8.3.4.2 Brazed aluminum heat exchanger
The application of BAHX to the gas processing industry is another important advance. The BAHX can significantly reduce the size of heat exchangers while achieving a close temperature approach on the heat exchanger. The exchanger is compact, requiring about 20% the size of a shell and tube exchanger of comparable performance. Furthermore, the alternating plate fin construction offers multiple stream capability and can replace multiple shell and tube heat exchangers, significantly reducing the space requirement and simplifying installation. A typical BAHX is shown in Figure 8-11.
BAHXs are constructed of aluminum material and are light in weight. The relatively high conductance of aluminum and the fin configuration can produce very close temperature approaches among different passes, which contributes to the lower power consumption and higher NGL recovery. The disadvantages are that they are prone to fouling from hydrates or other foreign matters and they are susceptible to aluminum corrosion by mercury contaminant. They require feed gas filtration and pretreatment with mercury removal beds. Aluminum is also a weaker material than stainless steel and is prone to thermal stress. The maximum operating temperature is limited to 150 °C/14 °F and the temperature differentials among exchanger passes must be kept below 50 °F.

When all these design limits are circumvented, BAHX can be effectively used to produce a highly efficient NGL recovery design. The exchanger designs are well known and are available from several manufacturers in different countries.

8.3.5 Lean oil absorption

The lean oil absorption process was developed in the early 1910s and was used exclusively until the 1970s. The absorption unit uses a lean oil to absorb the C₃⁺ components followed by a deethanizer and a rich oil still to regenerate the rich oil. Propane and butane products can be produced. A typical refrigerated lean oil absorption process is shown in Figure 8-12.

To allow the unit to operate at low temperatures, the feed gas must be injected with ethylene glycol solution to avoid hydrate formation in the heat exchangers. The feed gas is cooled by propane refrigeration and separated in a cold separator, typically at about 0 °F. The separator liquid is sent to the deethanizer while the separator vapor is routed to the absorber operating at 400 psig. Refrigerated lean oil is used to absorb the C₃⁺ content from the feed gas, producing a lean gas and a propane-rich bottom, which is sent to the deethanizer. The deethanizer operates at a lower pressure, typically at 200 psig, producing an ethane-rich gas and a rich oil bottom containing the C₃⁺ components.

The deethanizer overhead is compressed to the sales gas pipeline or used as fuel gas. The bottoms is further processed in a rich oil still, which regenerates a lean oil to be recycled back to the absorber, and an overhead distillate containing the C₃⁺ components. The C₃⁺ stream can be fractionated in a depropanizer, which produces the propane and butane product. Because of the high boiling material of the lean oil, a fired heater is used in the rich oil still. If necessary, the lean oil composition can be controlled using a lean oil still (not shown) to remove the heavy tails of the lean oil from the process.

A typical refrigerated lean oil process can achieve 50–60% propane recovery, depending on the feed gas composition. Because of the high equipment counts and the process complexity, lean oil absorption processes are not cost competitive to expander plants and are seldom used today.
FIGURE 8-12
Typical lean oil absorption process.
8.3.6 Modern NGL recovery processes

Modern NGL recovery processes are turboexpanders based on various reflux configurations. There are many patented processes that can be used to improve NGL recovery, either for propane recovery or ethane recovery. In addition, some ethane recovery units can operate on ethane rejection mode with minimum loss in propane. Similarly, some of the propane recovery units can also be operated on ethane recovery. These processes are discussed in detail in the following sections.

8.3.6.1 Dual column reflux process

The dual column process was quite common among NGL recovery units. Typically, the first column acts as an absorber recovering the bulk of the NGL components and the second column serves as a deethanizer during propane recovery and as a demethanizer during ethane recovery. Addition of refluxes to the dual column design process was originally configured for high propane recovery. The process is very efficient and can achieve over 99% propane recovery. The process flow schematic of the dual column reflux process is shown in Figure 8-13.

Feed gas, typically supplied at about 1000 psig, is first dried using molecular sieve dryers, and then cooled in a feed exchanger (BAHX). The feed gas is chilled with the absorber overhead vapor, cold separator liquid, and the absorber bottoms. Refrigeration is done using turboexpander by expanding the separator gas to the absorber typically operating at 450 psig. When processing a rich feed gas, propane refrigeration is often used to supplement the cooling requirement.

The absorber bottoms liquid is pumped by the bottoms pump to the deethanizer, which operates at a slightly higher pressure than the absorber. The

![Figure 8-13](attachment:FIGURE_8-13.png)

Dual column reflux process (Mak, 2005).
deethanizer overhead is refluxed using propane refrigeration, producing a C₂-rich liquid that is used as reflux to the deethanizer and the absorber. The absorber overhead is used to subcool the absorber reflux stream, which further enhances the separation process. With the dual reflux process, a very high propane recovery can be achieved with very low energy consumption.

Figure 8-14 shows a picture of a dual column process. The deethanizer is the taller column behind the absorber.

**8.3.6.2 Gas subcooled process**

The gas subcooled process (GSP) is a common method of ethane recovery in the gas processing industry. The process was invented by Ortloff in the late 1970s. The schematic of a typical GSP process is shown in Figure 8-15. The unique feature of the GSP process is the split flow arrangement using a portion of cold separator vapor (about 30%) to form a lean reflux stream to the demethanizer. The refrigeration content of the residue gas from the demethanizer is used in subcooling the reflux stream in the gas subcooler. The remaining portion is expanded in a turboexpander, generating the low-temperature refrigeration for the demethanizer. One- or two-side reboiler on the demethanizer is typically used to reduce the refrigeration requirement.

The demethanizer reboiler is used to produce an ethane-rich liquid that must meet methane and CO₂ specifications (see Table 8-2). The column typically
operates between 350 and 450 psig, depending on the feed gas inlet pressure, the feed gas composition, and the ethane recovery level.

### 8.3.6.3 Ortloff single-column overhead recycle process

The single-column overhead recycle process (SCORE) licensed by Ortloff Engineers, Ltd, is based on the same reflux principle as the dual column reflux process. The SCORE process is designed to recover over 99% propane from the feed gas in a single-column configuration.

This process recovers the C$_3^+$ hydrocarbons from the feed gas and produces a lean residual gas for sales. Alternatively, the residue gas can be sent to the natural gas liquefaction plant. It has been used in natural gas liquefaction unit to accomplish the function of the scrub column for the removal of C$_5^+$ hydrocarbons. The recovered NGL can be exported or used for blending with LNG if an LNG of higher heating value is required. Figure 8-16 shows a simplified process schematic of a typical SCORE process (Thompson et al., 2004).

### 8.3.6.4 Residue gas recycle

When high ethane recovery is required, additional cooling is required by recycling a portion of the residue gas as reflux to the absorber. The process flow schematic of a residual gas recycle process is shown in Figure 8-17. With this configuration, C$_2$ recovery, as high as 95%, can be achieved.
FIGURE 8-16
Ortloff SCORE process (Thompson et al., 2004).

FIGURE 8-17
Residue gas recycle for high ethane recovery.
Chilling can be produced by lowering the column pressure or increasing the recycle gas rate. However, if the feed gas contains CO₂, lower demethanizer pressure would lower the column temperature, which is more prone to CO₂ freezing. Lowering the demethanizer pressure also requires more residue gas compression. Typically, the optimum pressure of the demethanizer column falls between 350 and 450 psig.

Optimum C₂ recovery levels should be evaluated based on the project economics. Typically, C₂ recovery level greater than 95% would require a significant increase in gas recycle and equipment cost, which may not be economically justified.

### 8.3.6.5 Fluor twin column high absorption Process

When the sales gas must be compressed to the pipeline pressure, it is desirable to operate the demethanizer at as high a pressure as possible. However, this may not be feasible in conventional demethanizer designs, such as that shown in Figure 8-15.

Conventional demethanizer is a single-column design, which operates between 350 and 450 psig. The operating pressure should stay way below the critical pressure. More importantly, the critical region of the demethanizer column is the column bottom, which has the highest temperature. Increasing the column pressure will increase the bottom temperature, further moving the operating point closer to the critical region, which makes fractionation of methane from ethane difficult. For this reason, most demethanizers typically operate between 350 psig and 450 psig, with 500 psig as the upper limit.

To overcome the fractionation difficulty, Fluor has developed the twin column high absorption process (TCHAP) using a dual column approach. The first column operates as an absorber at 600 psig or higher pressure and is designed for bulk absorption. The second column operates at a lower pressure at about 450 psig, as a demethanizer or deethanizer. To improve NGL recovery, the overhead vapor from the second column is recycled using a small overhead compressor. The recycle gas is chilled using the absorber overhead vapor and used as a cold reflux to the absorber.

The process schematic of the TCHAP process is shown in Figure 8-18. This NGL process may eliminate residue gas compression, which can significantly reduce the overall power consumption of the facility. The TCHAP process can achieve over 80% ethane and 99% propane recovery.

### 8.3.6.6 Fluor twin-reflux absorption process

NGL recovery units are frequently required to operate on ethane rejection mode when the profit margin of ethane production is low. During these periods, the NGL recovery units are required to reject their ethane content to the sales gas pipeline. Ethane recovery process, such as the GSP, can typically recover about 70–80% of the ethane content and about 98% of propane content. When operated
on ethane rejection mode, some propane is lost with the rejected ethane, resulting in a loss in liquid revenue.

To circumvent this problem, Fluor has developed the twin reflux absorption process (TRAP). The process can be operated on ethane recovery and can also be operated on ethane rejection while maintaining high propane recovery (Mak et al., 2004a). The process flow schematic is shown in Figure 8-19.

The process is a dual column design, with the second column acting as a demethanizer during ethane recovery and as a deethanizer during ethane rejection.
rejection. The front section of the process is similar to the GSP process. The feed gas is chilled using the cold residue gas and propane refrigeration and separated. About 30% of the expander drum vapor is subcooled in the subcooler forming a subcooled reflux stream to the absorber. The remaining is expanded across a turboexpander producing the refrigeration.

The absorber produces a bottom liquid enriched in the C$_2$+ liquids, which are routed to the second column. During ethane recovery, the second column operates as a demethanizer. The demethanizer overhead vapor is routed to the bottom of the absorber where the ethane content is reabsorbed. During this operation, the second column reflux system can be bypassed.

During ethane rejection, the second column operates as a deethanizer. The overhead vapor is rerouted from the absorber bottom to a deethanizer reflux system. The propane content is recovered and refluxed back to the deethanizer. The deethanizer reflux drum vapor is cooled in the absorber subcooler and recycled as a cold reflux to the absorber. During ethane rejection, the TRAP process can achieve over 99% propane recovery.

### 8.3.7 Other hydrocarbons removal processes

For removal of small quantities of heavy hydrocarbons, such as in fuel gas conditioning, there are other simpler processes that can be used, instead of the traditional NGL recovery processes. Some of these processes are briefly described in the following sections.

#### 8.3.7.1 Solid-bed adsorption

Solid-bed adsorption process can be designed to selectively remove specific hydrocarbons. The adsorbent can be silica gel (i.e., Sorbead) that can be designed to remove most of the C$_6$+ hydrocarbons. A typical two-bed adsorption process is shown in Figure 8-20. Regeneration is accomplished by passing heated recycle gas through the bed. The heavy hydrocarbon is recovered from the regeneration gas by cooling, condensation, and separation.

The solid-bed adsorption process can be used to adsorb hydrocarbons at high pressure. This process has an advantage over the refrigeration process, which depends on low pressure for phase separation (Parsons and Templeman, 1990). The disadvantage is the high-pressure vessels, which can be expensive. Details of the design and operation of a solid-bed adsorption system can be found in Chapter 7.

#### 8.3.7.2 Membrane separation

The membrane-based process is a compact design, particularly suitable for offshore installation. The process is simple and suitable for unmanned operation. Figure 8-21 shows the flow scheme of a membrane system used to produce a fuel gas for the engine of the pipeline compressor.

In this process, a slip stream of the pipeline gas is processed in the membrane module, which removes the heavy hydrocarbons, producing a lean gas as
gas engine fuel. The hydrocarbons can be recycled back to the compressor suction and recompressed back to the pipeline system. Alternatively, the hydrocarbon contents can be recovered by condensation and chilling to produce a liquid by-product.

The hydrocarbon permeation membranes are typically made with vitreous polymers that exhibit high permeability with respect to heavy hydrocarbons.
New membrane materials are continuously being developed. Membrane Technology and Research Inc. has recently developed a new membrane that claims to be applicable for separating C$_3^+$ hydrocarbons. However, the commercial application has yet to be proven.

**8.3.7.3 Twister® supersonic separation**

Twister separation technology is based on a supersonic mechanism (a combination of aerodynamics, thermodynamics, and fluid dynamics) to condense and remove water and heavy hydrocarbons from natural gas. It is based on the concept that condensation and separation at supersonic velocity reduces the residence time to milliseconds, allowing no time to form hydrates.

Twister supersonic separation technology can potentially offer significant costs and environmental benefits for offshore operation. More detail discussion of the Twister technology can be found in Chapter 4.

**8.4 Selection of NGL recovery process**

NGL recovery unit is a significant investment in a gas processing plant. The NGL recovery levels must be determined by life cycle cost analysis, considering the value of the liquid products over sales gas for the life of the project. Cost of the gas plant also includes supporting facilities, such as inlet gas compression, sales gas compression, liquid handling, liquid treating, liquid storage, and transportation systems. The cost is site dependent, such as ambient temperatures, soil conditions, and proximity of plant site to the nearest liquid pipeline. Selection of the NGL recovery process must consider all these facilities and not just the NGL unit itself.

If the gas plant is located in a remote area and is unmanned, simplicity in operation and low maintenance are important. The use of J–T units may be ideal because of the low equipment counts. J–T units do not need an expander, which requires operator attention and maintenance. Shell and tube exchangers are preferred because of the robust nature of the mechanical design when compared to BAHX. Propane refrigeration is required if the feed gas pressure is low and gas recompression is restricted.

Lean oil absorption is seldom used due to its low efficiency, process complexity, and high maintenance. Cascade refrigeration or mixed refrigeration systems are rarely used for the same reasons. For traditional high-NGL recovery plants, turboexpander in combination with propane refrigeration is the best solution.

The capital cost of a propane recovery unit is less than that of an ethane recovery unit. If the ethane product is not expected to be profitable, the unit should be designed for propane recovery to minimize capital cost. If necessary, additional piping can be installed to allow the propane recovery unit to operate on ethane recovery, such as the TRAP process. If ethane recovery is required, the
unit should be designed using the GSP process, which typically can achieve 75% ethane recovery. If very high ethane recovery is required, residue gas recycle can be added, which can achieve over 90% ethane recovery.

8.5 NGL recovery technology development

The NGL recovery technology is continuously evolving. Today’s NGL processes are using the high-efficiency expanders and compressors and compact heat exchanger to reduce capital and operating costs.

There are many proprietary NGL technologies available. They typically use multiple refluxes, column design, and heat exchanger configuration closely integrated to reduce cost. While these innovations may reduce capital cost, they may prove to be difficult and complex to operate. Most gas plants prefer operation simplicity and minimum maintenance design. The development efforts must emphasize on safety, simplicity, and ease of operation. Very high efficiency may not be necessary due to the projected low margin of the NGL products.

Another development is on modularization. Most of the midstream developers expect a high return of their investments. They also desire that the NGL recovery units can be standardized and mobilized quickly. Modular units minimize site work and reduce installation costs. They can be designed to be truckable such that they can be deployed in other sites when the current resource is depleted.

8.6 NGL recovery unit design considerations

The process performance of an NGL recovery unit is very sensitive to feed gas compositions. As the unit matures, the feed gas will contain less of the heavy hydrocarbons. When nitrogen injection is used for enhanced oil recovery, the nitrogen content in the feed gas will increase. These changes will negatively affect the NGL recovery unit performance.

Another design consideration, as mentioned in the earlier sections, is the requirement to operate on lower levels of ethane recovery to meet ethane demands. The unit must be designed to operate on feed gas with varying “GPM.” The ability to operate on ethane rejection without a loss in propane recovery, such as the TRAP process, is desirable.

8.7 NGL recovery unit operating problems

The operating problems in an NGL recovery unit are unstable operation, and failure to meet NGL recovery targets and product specifications. The following checklists can be used to identify the root cause of the problems.
If a high pressure drop occurs in the unit, check the CO₂ content in the feed gas. This may be due to CO₂ freezing in the demethanizer. Check the performance of the amine unit.

Check the moisture content in the dried gas from the molecular sieve unit. The water content should register below 1 ppmv to avoid hydrate formation. Formation of hydrate will plug the heat exchange equipment, resulting in high pressure drop. Methanol injection can be started to temporarily relieve the problems.

If the unit fails to achieve the design capacity, check the feed gas compositions. If the feed gas is leaner and contains more nitrogen, vapor loading on the demethanizer will increase, reducing the capacity of the unit.

Check the temperature and pressure profile of the NGL unit. If the deethanizer operates at high temperature, propane recovery will be reduced. If the column pressure is too high, separation may become difficult, resulting in lower recovery.

To debottleneck the demethanizer, check the tray design and conduct a column scan. Follow some of the procedures discussed in the fractionation handbook by Kister (2006).

If the exchanger fails to achieve design performance, check the temperature profile of the NGL unit to determine if there is a temperature pinch on heat exchangers.

Verify the expander compressor performance. Is the bypass valve around the expander opened?

Check the refrigeration system performance. Check the contaminant levels in the propane refrigeration system. Sometimes, leakage of lube oil into the propane system can be a problem.

8.8 NGL fractionation

NGL produced from the NGL recovery unit is typically sent to a fractionation center for further processing into individual products. A simplified NGL fractionation flow schematic is shown in Figure 8-22. A picture of a fractionation center is shown in Figure 8-23.

NGL is fractionated by heating the NGL stream and processing through a series of distillation columns. Fractionation takes advantage of the differing boiling points of the various NGL components. As the NGL stream is heated, the lightest (lowest boiling point) NGL component boils off first and is separated. The overhead vapor is condensed and a portion is used as reflux and the remaining portion is routed to product storage. The heavier liquid mixture at the bottom of the first column is routed to the second column where the process is repeated and a different NGL component is separated as product. This process is repeated until the NGL is separated into its individual components.
In some cases, where isobutane is desired, the mixed butane stream can be further separated with a deisobutanizer or butanes splitter to produce the required product for refinery consumption. There are many options in configuring the NGL fractionation train and the optimum arrangement may vary depending on the NGL compositions, and the product specifications.

8.8.1 Fractionation column design and operation

Fluctuation in feed conditions and compositions, which may happen on a daily basis, has significant impacts on the NGL fractionation unit operation. Their
impacts on the design can be evaluated on the process simulators and the optimum operating variables can be assessed. The simulation software can now be integrated to the distributed control system allowing real-time optimization.

Three distillation books published by Kister (1989, 1992, 2006) can be used as references and guidelines in the design and operation of distillation columns. These books provide design parameters for equipment, discuss limitations of the design methods, and suggest solutions in troubleshooting columns. They also contain process calculations on column hydraulic and tower performance, tray and packing design details, and the methods to maximize column operating efficiency.

8.9 Liquid product processing

The liquid products from the NGL fractionation unit seldom meet customer’s specifications without further treatment. The following sections discuss some of the processing methods to produce marketable liquid products.

8.9.1 NGL contaminant treating

If acidic and sulfur compounds are present in the feed gas and have not been removed prior to NGL recovery, they will end up in the NGL products. The distribution of the contaminants for the various NGL products is summarized in Table 8-3. H₂S will show up in the deethanizer overheads, while most COS is concentrated in the propane fraction. The mercaptans tend to split between the various NGL fractions, depending on their molecular weight, with most of the heavy molecular weight mercaptans ending up in the condensate.

These contaminants not only can lead to odor problems but also can form sulfur oxides on combustion. They can cause corrosion of equipment unless they

| Table 8-3 NGL Contaminant Distribution (Modified after John M. Campbell & Co, 2003) |
|--------------------------------|-------|-------|-------|-------|
|                               | Ethane | Propane | Butane | Condensate |
| H₂O                           | x      | x       | x      | x         |
| CO₂                           | x      | x       |        |           |
| H₂S                           | x      | x       |        |           |
| COS                           | x      | x       |        |           |
| CH₃SH                        | x      | x       | x      |           |
| C₂H₅SH                        | x      | x       | x      | x         |
| C₃H₇SH⁺                      |        | x       | x      | x         |
| CS₂                           | x      | x       |        |           |
| DMS                           | x      | x       |        |           |
| Others                        |        |         |        | x         |
are dehydrated. The presence of significant quantities of CO₂ can increase the vapor pressure and lower the heating value of the hydrocarbon liquids. Carbonyl sulfide (COS) and carbon disulfide (CS₂), although not corrosive in LPG, will hydrolyze slowly to H₂S, resulting in off-spec products.

8.9.1.1 Caustic processes

The hydrocarbon processing industry has historically used caustic solutions to extract or treat acidic impurities in liquid hydrocarbon streams. A number of caustic processes, both regenerative and nonregenerative, can be used to remove sulfur compounds from hydrocarbon liquids. The simplest process is the use of a nonregenerative solid KOH bed, which is effective for removal of H₂S but not for other sulfur compounds. One of the common processes for treating hydrocarbon liquids is the use of regenerative caustic wash with sodium hydroxide. Typically it is located downstream of an amine unit that is used for bulk acid gas removal. Such arrangement would minimize the treating duty required by caustic wash. When acid gas quantities are small, a simple caustic wash is effective and economical. However, as the quantity of contaminants increases, caustic wash process can be expensive due to high chemical and disposal costs.

Several organic materials may be added to caustic solutions (as promoters) to increase their solubility for mercaptans. Details on various promoter processes for caustic treating can be found in a paper by Maddox (1982). Mercaptans can be converted to disulfides by several methods. These disulfides will remain in the sweetened hydrocarbon product. The overall sulfur content, therefore, remains unchanged. However, the sulfur leaves as disulfide (no odor) rather than as mercaptans. The method or combination of methods that can be used depends on the mercaptan content of the product to be treated and the specification that must be met (Fischer et al., 1993). Among the processes, which convert (oxidize) mercaptans to disulfides, the “doctor sweetening” process is the earliest. Doctor treatment ordinarily will leave 0.0004% mercaptan at which level there is negligible effect on the tetraethyl lead susceptibility of the gasoline. In this process, an alkaline solution of lead oxide (usually sodium plumbite) contacts the hydrocarbon stream forming lead mercaptides (soluble in the oil) with the mercaptans. The mixture is then treated with powdered sulfur (which has a high affinity for lead) and the conversion of the mercaptide into the so-called disulfide (which remains in solution in the gasoline stream) occurs. The reactions of the “doctor treating” process are considered to be:

\[ 2RSH + \text{Na}_2\text{PbO}_2 \rightarrow (RS)_2\text{Pb} + 2\text{NaOH} \quad (8-1) \]

\[ (RS)_2\text{Pb} + S \rightarrow R_2\text{S}_2 + \text{PbS} \quad (8-2) \]

Note that sulfur should be added in stoichiometric excess in order to get maximum conversion of the mercaptides to disulfides, where too much excess sulfur will cause formation of polysulfides (Maddox, 1982). Sometimes with no sulfur added, but in the presence of atmospheric oxygen and sodium hydroxide.
solution, the same conversion (Eqn (8-2)) occurs, but only slowly, and not completely (McBryde, 1991).

The Merox™ process, a UOP trademark, developed and commercialized over 50 years ago by UOP, is used to treat end product streams by rendering any mercaptan sulfur compounds inactive. This process can be used for treating LPG, gasoline, and heavier fractions. The method of treatment is the extraction reaction of the sour feedstock containing mercaptans (RSH) with caustic soda (NaOH) in a single, multistage extraction column using high-efficiency trays (UOP’s Extractor Plus). The extraction reaction is shown by the following equation:

$$RSH + NaOH \leftrightarrow NaSR + H_2O$$ (8-3)

After extraction, the extracted mercaptans in the form of sodium mercaptides (NaSR) are catalytically oxidized (by Merox WS™ catalyst, which is dispersed in the aqueous caustic solution) to water-insoluble disulfide oils (RSSR), as shown in the equation given below:

$$4NaSR + O_2 + 2H_2O \rightarrow 2RSSR + 4NaOH$$ (8-4)

The disulfide oil is decanted and sent to fuel or to further processing in a hydrotreater. The regenerated (lean) caustic is then recirculated to the extraction column. Typical product mercaptan levels can be controlled to less than 10 ppmw (UOP, 2003). A typical flow schematic of the UOP’s Merox process is shown in Figure 8-24.

The Sulfrex™ process (developed by Axens) is another technology similar to the Merox process, which has demonstrated its industrial performance, robustness, simple design, and ease in operation since 1975.

![Diagram](image-url)

**FIGURE 8-24**

Liquid Merox treating process (UOP, 2003).
8.9.1.2 Molecular sieve technology
Molecular sieve technology is commonly used for treating NGLs. Molecular sieves can be used for removal of sulfur compounds (H₂S, COS, and mercaptans) either in the gas or liquid phase. There are advantages and disadvantages for both options. The adsorber efficiency is higher in the gas phase since the mass transfer rate of sulfur compounds is much faster in the gas phase. However, the size and number of adsorber is smaller for the liquid phase as the liquid rate tends to be significantly lower than the gas rate.

Molecular sieves can dry the feed stream in addition to contaminant removal. Their drawbacks are high capital and operating costs. The formation of COS is another concern if both H₂S and CO₂ are present. COS formation is to be avoided or minimized as much as possible, as it converts back to H₂S in the presence of water, resulting in off-spec product and corrosion problems in downstream units. Different grades of molecular sieves have varying levels of conversion of H₂S to COS. Molecular sieve manufacturers can supply customized sieves to minimize COS formation. Usually these products are 3A-based, but they also have less adsorption capacity for CO₂ and H₂S.

For liquid-phase adsorbers, adsorption flow tends to be upward, and regeneration downward. Molecular sieve regeneration is similar to that for gas-phase application. In addition, liquid draining and filling time must be taken into account when determining the cycle times. If regeneration gas is supplied at a different pressure, pressurization and depressurization must be programmed in the setup. Care must be taken to ensure that the maximum velocities are not exceeded, which may cause bed lifting (or fluidizing) and attrition of molecular sieves. More details on this process can be found in Chapter 7.

8.9.1.3 Amine processes
Amine treating is an attractive alternative, especially when an amine gas treating unit is already onsite. In such cases, the liquid treating unit can often be operated using a slipstream of lean amine from the main amine regeneration unit.

Amine treating is often used upstream of caustic treaters to minimize caustic consumption caused by irreversible reactions with CO₂. In this process, H₂S and CO₂ from the sour liquid feed are absorbed by liquid–liquid contacting the sour liquid with the lean amine solvent. The amine unit design should consider the suitable amine solution and the type of contacting device. In most cases, generic amines (including DGA and DIPA, and the MDEA-based specialty solvents) will perform satisfactorily. The liquid–liquid contacting devices include packed towers, traysed towers, jet eductor–mixers, and static mixers. Most installation uses randomly packed columns as they are less expensive and avoid the potential backmixing problems with tray column.

A typical LPG treating unit utilizing amine solvent is shown in Figure 8-25. As can be seen, raw LPG enters the bottom of a packed absorber and lean amine
enters the top of the absorber. Treated LPG leaves the absorber from the top and rich amine leaves the absorber from the bottom. The treated LPG is washed using a recirculating water-wash system to recover the entrained amine and protect downstream caustic treaters. The treated LPG and the wash water are mixed in the water-wash static mixer, which is then coalesced into two liquid phases and separated in the water-wash separator. Makeup water is added to the circulating water-wash circuit to maintain the concentration of the amine system.

Carbonyl sulfide (COS) is a stable, unreactive compound that is very difficult to reduce to concentration levels below 1 ppmv using conventional amine and molecular sieve processes. The ADIP process is a regenerative amine process to selectively reduce COS to very low levels (5 ppm wt as S) in liquid hydrocarbons such as LPG and NGL. Numerous nonregenerative metallic oxide processes are also available to remove COS from liquid products. Some of these processes remove the COS directly, and others require water to hydrolyze the COS to H2S before it is reacted (Maddox, 1982).

8.9.2 Dehydration

NGLs must be dehydrated to meet requirements of a handling chain to a direct consumer. The acceptable water content in light hydrocarbon liquid streams varies from no free water present to very low levels of moisture in liquid products. For example, most liquid sales specifications require the LPG to yield a negative
result to the cobalt bromide test, which is equivalent to a water content of 15–30 ppm.

Alumina is a satisfactory desiccant for liquid drying. If simultaneous treating and drying is required, molecular sieve type 13X is also a good candidate. There are other advanced processes to be used for contaminant removal and dehydration, which have been discussed in Chapter 7.

References


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9.1 Introduction

Roughly 25% of the natural gas brought into production from new sources requires some degree of treatment to remove hydrogen sulfide (H₂S) and recover elemental sulfur from acid gas streams. H₂S is a flammable gas with a wide limit of flammability, and unlike elemental sulfur, it is extremely toxic. In fact, 5-min exposure to only 800 ppm results in death and a single breath of 1000–2000 ppm may cause coma. For environmental and safety reasons, venting or flaring H₂S to the surroundings is unacceptable; therefore, its conversion to the harmless elemental sulfur is necessary. Elemental sulfur can be stored, handled, and transported in bulk. Ease of storage is an important advantage as it enables sulfur to be stockpiled in periods of reduced demand. The use of sulfur is not very large compared to the large sulfur contents in the sour gas. The primary use of sulfur is for the production of sulfuric acid, and other minor uses include manufacture of fertilizer, fungicide, pesticide, bactericide in food preservation, and other fine chemicals.

Chemical conversion of H₂S for disposal as solid waste (such as calcium sulfate) is technically feasible but uneconomical and impractical. In remote locations, or where there is no market for sulfur, acid gas injection is another option instead of sulfur recovery.

This chapter discusses the properties of elemental sulfur and then describes some of the common methods for processing H₂S.

9.2 Sulfur properties

Elemental sulfur, which usually occurs in the form of eight-member rings, is a relatively nontoxic and chemically inert substance, and insoluble in water and most other liquids. It is a solid at ambient temperatures and exists in two crystalline and amorphous forms. Sulfur in crystalline form has different shapes. Rhombic and monoclinic are the most common forms of crystalline sulfur. These two forms differ in the way in which the rings are stacked. Rhombic sulfur is the stable form of the element at room temperature. If rhombic sulfur is heated to about 203 °F, it changes into monoclinic crystals. Above 203 °F and up to 239 °F, monoclinic sulfur becomes the more dominant structural form. Amorphous sulfur is formed when liquid sulfur, which has been heated to elevated temperatures, is allowed to cool rapidly. Amorphous sulfur slowly changes to the rhombic crystalline form at ambient temperatures.
When melted, sulfur becomes a brownish-yellow transparent liquid whose molecular structure also consists of eight-member rings. As liquid sulfur is heated to about 320 °C14, the structure of the liquid undergoes a change (forming long-chain sulfur polymers in conjunction with the eight-member sulfur rings); its color becomes a dark reddish brown and its viscosity increases significantly. The effect of temperature on the liquid sulfur viscosity is shown in Figure 9-1.

Sulfur vapor exists as $S_x$ (where $x$ can have values from 1 through 8), but the principal sulfur species are $S_2$, $S_6$, and $S_8$. The distribution of sulfur vapor species varies significantly with changes in temperature, where at lower temperatures, $S_8$ dominates, but as the temperature rises $S_8$ is converted to $S_6$, and finally to $S_2$.

The physical and chemical properties of sulfur are unusual. Therefore, any attempt to understand sulfur handling processes requires an understanding of the behavior of the various sulfur species. The properties of elemental sulfur are well described in the literature (Meyer, 1976; Chao, 1980; Shuai and Meisen, 1995).

**9.3 Sulfur recovery**

Sulfur recovery refers to the conversion of $H_2S$ to elemental sulfur. The original sulfur recovery process was developed in 1883 by Carl Friedrich Claus, a London chemist, and described in a British patent application as a means to recover sulfur as a by-product in the manufacture of soda ($Na_2CO_3$).
In the original Claus process, elemental sulfur was produced by the partial oxidation of \( \text{H}_2\text{S} \) in a single step over a preheated catalyst bed, according to the following exothermic reaction:

\[
\text{H}_2\text{S} + \frac{1}{2} \text{O}_2 \rightarrow \frac{1}{8} \text{S}_8 + \text{H}_2\text{O} \quad (9-1)
\]

However, since the reaction was extremely exothermic, the original process was severely limited by inability to control the reaction temperature, and consequently sulfur recovery was limited.

In 1938, I.G. Farbenindustrie A.G. in Germany described a modification to the Claus process in which the oxidation of one-third of the hydrogen sulfide to sulfur dioxide was carried out in a boiler, and the remaining two-thirds of hydrogen sulfide reacted with the sulfur dioxide over a catalyst. The modified Claus process has become the standard sulfur recovery technology of today.

The following sections cover the conventional Claus process and its various modifications and extensions, as well as some other less-utilized sulfur recovery processes.

### 9.3.1 Modified Claus process

In the modified Claus sulfur recovery process, the overall reaction is carried out in two stages. In the first stage (thermal section), enough air is added to oxidize only one-third of the incoming \( \text{H}_2\text{S} \) to \( \text{SO}_2 \). This reaction is highly exothermic and is not limited by equilibrium.

\[
\text{H}_2\text{S} + \frac{3}{2} \text{O}_2 \rightarrow \text{SO}_2 + \text{H}_2\text{O} \quad (9-2)
\]

In the reaction furnace, the unburned \( \text{H}_2\text{S} \) in the acid gas reacts with the produced \( \text{SO}_2 \) to yield elemental sulfur vapor. This reaction is referred to as the Claus reaction and is shown by Reaction (9-3). This reaction is endothermic and is limited by equilibrium.

\[
\text{SO}_2 + 2\text{H}_2\text{S} \leftrightarrow 3/2 \text{S}_2 + 2\text{H}_2\text{O} \quad (9-3)
\]

Usually, 60–70% of the total conversion of \( \text{H}_2\text{S} \) to elemental sulfur is achieved in the thermal stage (Gamson and Elkins, 1953; Opekar and Goar, 1966). The theoretical equilibrium percentage of \( \text{H}_2\text{S} \) conversion to sulfur is shown in Figure 9-2.

#### 9.3.1.1 Process description

A typical modified Claus process flow schematic is shown in Figure 9-3. The picture of a typical sulfur recovery unit (SRU) is shown in Figure 9-4. In this process, the overall reaction is carried out in two stages, thermal and catalytic.

**Thermal section**

The hot combustion products from the reaction furnace flow through the tubes of waste heat boiler (WHB) and are partially cooled to 600–800 °F by generating
As the gases from the reaction furnace are cooled in the WHB, the S\textsubscript{2} sulfur vapor species shift through a number of exothermic reactions to other sulfur species, primarily S\textsubscript{6} and S\textsubscript{8} (Goar and Fenderson, 1996).

The cooled gases from the WHB are further cooled (to 340–375 °F typically) in the first sulfur condenser by generating low-pressure or medium-pressure steam, where most of the sulfur vapor formed in the reaction furnace is...
condensed. The condensed sulfur is separated from the gas and drained from the condenser through a hydraulic seal and rundown line to the sulfur collection pit.

Catalytic section
The catalytic section has three processing steps. The first is the reheater step, which raises the temperature of the gas from the sulfur condenser to avoid condensation of sulfur vapor, as sulfur continues to form by the Claus reaction. Typically, the first reheater is required to increase the temperature to about 600 °C, which is necessary to promote the destruction of COS and CS2.

In the second step, the hydrogen sulfide and sulfur dioxide are reacted over an activated alumina catalyst in a series of catalytic reactors (1–3). Other catalysts may be used to promote COS and CS2 destruction but they are typically very expensive.

The Claus reactions in the catalytic section are slightly different from the thermal section reaction. These reactions are both exothermic and are represented as:

\[
2\text{H}_2\text{S} + \text{SO}_2 \leftrightarrow \frac{3}{6}\text{S}_6 + 2\text{H}_2\text{O} \quad (9-4)
\]

\[
2\text{H}_2\text{S} + \text{SO}_2 \leftrightarrow \frac{3}{8}\text{S}_8 + 2\text{H}_2\text{O} \quad (9-5)
\]

In the third step, the sulfur condenser is used to remove liquid sulfur, the product of reaction. This shifts the equilibrium so that further reaction may take place in the next catalytic converter (reactor).

The final sulfur condenser is typically at 270 °F to minimize sulfur vapor losses in the vent gas (Goar and Fenderson, 1996). The effluent (tail) gas, which still contains appreciable amounts of sulfur compounds and a small amount of sulfur as vapor and entrained liquid, is routed either to a tail gas treating unit (TGTU) for further sulfur recovery and minimizing impacts on the environment or to a thermal oxidizer to incinerate all the sulfur compounds in the tail gas to sulfur dioxide before dispersing the effluent to the atmosphere.
The sulfur recovery depends on such things as feed composition, age of the catalyst, and number of reactor stages. Typical sulfur recovery efficiencies for standard Claus units are 90–96% for a two-stage catalytic conversion and 95–98% for a three-stage catalytic conversion. Because of equilibrium limitations and other sulfur losses, overall sulfur recovery efficiency in a standard Claus unit is typically limited to 98% (Lagas et al., 1989).

Claus burner performance
The reaction furnace is the critical piece of equipment in the Claus process. For Claus units processing ammonia-laden sour water stripper off-gas, or feeds containing heavy hydrocarbons, benzene, toluene, ethylbenzene, and xylenes (BTEX), and cyanide, the minimum furnace operating temperature required must be increased above 925 °C (1700 °F) to ensure destruction of these components. Ammonia, in particular, requires a high combustion temperature, and SRUs with inadequate temperature are prone to shutdowns from plugging due to deposition of ammonia salts. Coke formation from cracking and incomplete reaction of heavy hydrocarbons can cause deactivation of the downstream Claus catalyst.

Figure 9-5 provides an operating guideline for the theoretical, adiabatic flame temperature as a function of the H₂S content in the acid gas feed. The two horizontal dashed lines provide operator with the temperature targets, depending on the nature of the hydrocarbons contained in the acid gas feed.

COS and CS₂ destruction
Traditional activated alumina Claus catalyst gives high conversion of H₂S, but only about 65% hydrolysis of COS and about 30% hydrolysis of CS₂ in the first Claus converter. Conversions are reduced in subsequent Claus stages. For the second stage, activated alumina catalyst gives about 20% hydrolysis of COS and only about 5% hydrolysis of CS₂.

For the catalytic stages, some technology licensors such as Fluor employ titanium oxide (TiO₂) catalyst in one or more of the Claus converters, which

![FIGURE 9-5](image)

Flame temperature versus H₂S content. HC, Hydrocarbon (Flood et al., 2011).
greatly improves the ability to hydrolyze COS and CS$_2$ (Flood et al., 2011). On the other hand, when applied properly, the TiO$_2$ catalyst is capable of achieving over 90% hydrolysis of COS and CS$_2$ in the first Claus converter. This would increase the overall sulfur recovery efficiency, which may be necessary to meet the sulfur emissions target.

9.3.1.2 Lean acid gas operations

The amount of heat generated in the reaction depends on the amount of H$_2$S available to the burners. With rich acid gas (60–100% H$_2$S), the reaction heat keeps the flame temperature above 2200 °F. When the gas is leaner, the flame temperature is reduced, where at 1800–1900 °F the flame starts to become unstable. This point is usually reached when the H$_2$S content is lower than 50%. However, when the H$_2$S in the acid gas falls below 40%, the reaction furnace may become inoperable, and may cause operating problems in the downstream Claus reactors (Goar, 1974). To overcome this problem in processing lean acid gases, there are several design options that can be incorporated to the SRU design, which are discussed in the following sections.

Feed preheating

The low furnace temperature problem can be overcome, within limits, by preheating the acid gas and/or air before it enters the burner.

Both the combustion air and the acid gas can be preheated in order to raise the flame temperature. Usually, combustion air is the first choice since it is more benign than the acid gas. Furthermore, there are pressure drop considerations. The upstream amine unit often limits the available acid gas pressure. The extent of combustion air preheater is an economic choice. Steam at a suitable pressure level is preferred over a fired heater due to ease of operation and lower cost. It is also desirable to use carbon steel material rather than the more exotic and expensive piping and equipment material.

Hot gas bypassing

The use of hot gas bypass is another solution to raise the furnace temperature, particularly where there is limited access to external heat sources, such as high-pressure steam and power. It is the lowest cost option to process medium acid gas composition, since no additional equipment and operating cost is required.

It uses a hot stream from the first pass outlet of the WHB (1000–1200 °F) to mix with process gas streams from the sulfur condensers. By bypassing part of the feed gas, the furnace will operate nearer to complete combustion. Consequently, the flame temperature is increased. However, the bypass is limited to two-thirds of the total feed because it is undesirable to operate the furnace under oxidizing conditions to a point beyond complete combustion.

However, the downside of this scheme is the contaminants in the feed gas. If present, they will not be destroyed in the furnace zone. The detrimental effects are reflected on the backside of the Claus unit, causing catalyst deactivation, high corrosion rate as well as equipment and line plugging, which will jeopardize the operation of the SRU.
Acid gas/natural gas fuel burner

Acid gas fired burners have also been extensively used in many SRUs. Acid gas burners are usually operated far away from stoichiometry to minimize oxygen slippage. When the acid gas is lean in H₂S, the burner is usually designed for firing nearer to stoichiometry to reduce the amount of acid gas. Consequently, only an acid gas with stable composition should be used to avoid oxygen slippage.

Their primary advantage is the ability to achieve the required catalytic reactor inlet temperature. To operate the inline burner, the air/fuel ratio must be closely monitored to avoid oxygen breakthrough and soot formation.

Both the hot gas bypass and the inline burner option have lower sulfur recovery efficiency than the straight-through Claus process. If hydrocarbons and other contaminants are present in the feed gas, they may also cause deactivation of the catalysts.

Acid gas enrichment

The lean acid gas (typically 10–20%) can be processed in an acid gas enrichment unit (AGEU) to increase its H₂S content to above 50% in order to operate a straight-through Claus unit.

By using a selective amine, such as proprietary MDEA or the Flexsorb processes, H₂S can be selectively absorbed from the acid gas, producing two acid gas streams. The treated gas from the amine absorber is primary CO₂, with H₂S content of less than 10 ppmv, which can be vented, incinerated, or used for enhancing oil recovery. The gas stream leaving the amine regenerator is the acid gas recycled to the SRU, which is now enriched in H₂S to over 50 mol%.

9.3.1.3 Oxygen enrichment

Oxygen enrichment can be used to sustain a higher reaction furnace temperature. It can be used to process a lean acid gas, destroy contaminants, and increase the SRU throughput.

With lower nitrogen content in the combustion oxygen, a higher flame temperature can be reached without additional fuel gas firing. However, if the H₂S content drops below 20 mol%, there will not be sufficient heating value in the feed gas and additional fuel firing may still be necessary (Goar and Fenderson, 1996).

The main advantage of oxygen enrichment is the smaller sizes of the SRU equipment. Oxygen enrichment is economically attractive for debottlenecking an existing SRU, and it may not require any changes to the existing equipment. For grassroots installation, the lower SRU cost can be offset by the cost of an air separation unit. In this case, oxygen enrichment can be economical if there is a low-cost oxygen supply.

There are three levels of oxygen enrichment: low (less than 28 vol% oxygen), medium (28–45 vol% oxygen), and high (greater than 45 vol% oxygen). Figure 9-6 shows the increases in sulfur capacity and furnace flame temperature with various levels of oxygen enrichment.
For an increase of up to 20–25%, low-level oxygen enrichment technology is generally adequate. The capital cost for the SRU is mainly in the installation of an oxygen supply system to the reaction furnace burner and no change to the burner is necessary. For a 50% increase, medium-level oxygen enrichment is required. The capital cost is in the installation of an oxygen supply system and a new oxygen combustion burner (with more exotic material). Typically, the cost associated with an oxygen enrichment revamp is 10–15% of a new air-based SRU (excluding the cost of the oxygen supply).

A challenge for oxygen-blown Claus process is controlling the high flame temperatures in the front-end furnace (FEF) to avoid overheating and damage of the reaction furnace refractory. There are several design and control options. A common design is to recycle the cooled gas to the FEF (Fluor COPE process); a unit picture is shown in Figure 9-7. Special oxygen burners (developed by Lurgi) or temperature control using staged combustion process (SURE process developed by British Oxygen Corp. and Parsons Corp.) can also be used.

### 9.3.2 Direct oxidation processes

For lean acid gases that contain 5–30 mol% H₂S, the straight-through Claus process is not workable without additional equipment or the AGEU. The alternative is to replace the thermal section of the modified Claus process by a direct oxygen process. There are several commercially proven processes including UOP’s Selectox process and Linde’s Clinsulf process, which are described in the following sections.
9.3.2.1 Selectox process

For lean acid gases that contain 5–30 mol% H₂S, the straight-through Claus process is not workable without additional equipment or the AGEU. The alternative is to replace the thermal section of the modified Claus process by a direct oxygen process. UOP offers the Selectox process using their proprietary catalysts.

Over the catalysts, air oxidizes H₂S to SO₂, which reacts with additional H₂S to produce elemental sulfur, as shown in the following reactions. Overall sulfur recoveries of 95–97% are possible with one Selectox process followed by a two-stage Claus process (Goar and Fenderson, 1996).

\[
H₂S + 3/2 O₂ \rightarrow SO₂ + H₂O \quad (9-6)
\]
\[
2H₂S + SO₂ \rightarrow 3/x S_x + 2H₂O \quad (9-7)
\]
Overall : \[
3H₂S + 3/2 O₂ \rightarrow 3/x S_x + 3H₂O \quad (9-8)
\]

Two versions of the process are offered by UOP. If the H₂S content in the acid gas feed is 5 mol% or less, the “once-through” mode can be used for the catalysts. The gas stream from the Selectox converter is cooled, then reheated, and treated in one or two additional conventional Claus catalytic stages. If the H₂S content is above 5 mol%, the “recycle mode” Selectox process can be used, where the
cooled gas is recycled from the first condenser outlet to maintain the Selectox converter temperature at about 700 ºF (Gowdy and Bertram, 1998).

The drawback of the catalytic process is the inability to handle contaminants. Contaminants, such as hydrocarbons and aromatics, will react on the Selectox and Claus catalysts, causing hot spots and coke formation (Jones and Bertram, 2001). Thus, the process is not as robust and forgiving as the thermal reactor in the Claus process. The Selectox units are generally smaller in capacities compared to the Claus units with a maximum of 45 long tons/day (LT/D) of sulfur production.

9.3.2.2 Clinsulf process

Linde A.G. has developed a nonthermal, direct oxidation process called the Clinsulf process, which is typically applied to syngas operation. In the Clinsulf process, H₂S is directly oxidized to sulfur and H₂O in a reactor filled with a suitable catalyst. In this process, the heat of reaction is removed directly from the Claus reactor with a coil-type tubular heat exchanger arranged within the catalyst bed. The internal cooling keeps the reaction temperature close to the sulfur dew point, in order to maximize sulfur yield.

A single-stage Clinsulf unit claims to be capable of attaining up to 92% sulfur recovery for very lean acid gases (<5% H₂S) and up to 95% for gases with 10–20% H₂S content. With the cooling provided by the Linde reactor, it is theoretically possible to control the reaction temperature to below the sulfur dew point, which can further improve the recovery to 99% (Gemmingen and Lahne, 1994; Heisel et al., 1999).

9.3.3 Small- and medium-scale processes

While the Claus process is still the main process for recovering sulfur from H₂S, several new processes are vying for acceptance for the recovery of sulfur from low-quality acid gas streams (less than 40 mol% H₂S), or if recovery is required for very low mass flow rates of sulfur production.

9.3.3.1 H₂S scavenger

Fixed bed scavenger

For small amounts of sulfur (less than 0.2 LT/D), nonregenerable H₂S scavengers such as activated carbon, iron oxide, other metal oxides, and caustic solution are frequently used with good success (Fisher et al., 1999). The details of several nonregenerable fixed bed processes are discussed in Chapter 6.

Liquid injection scavenger

Although fixed bed scavengers are proven commercially, the direct injection method is a lower cost alternative when a suitable pipeline length is available for contact. The direct scavenger injection process consists of a chemical injection pump and injection nozzles for injecting the liquid scavenger in a sour gas pipeline. The pipeline must be of sufficient length for good gas/liquid contact.
A separator is located downstream to separate the scavenger agent from the treated gas. The amount of gas/liquid contact time depends on the type of contacting device, the gas velocity, and the residence time. Because of this, the degree of mixing and efficiency is sensitive to changes in gas flow. Good separation of the scavenger and treated gas is required since some scavengers can cause foaming problems in downstream equipment.

Triazine is a common liquid chemical that is used for instantaneous removal of H2S from natural gas. Triazine is an H2S-scavenging chemical that will not absorb CO2. The injection method provides a single stage of contact and is not as efficient as an absorber. Sulfur scavenger injection can produce a treated gas with H2S content at around 20–40 ppm. Lower content may be achievable if sufficient time, pipe length, and injection agent are provided. The cost of sulfur scavenger is relative low for treating parts per million levels of H2S; however, the disposal cost of the spent scavenger can be very high.

9.3.3.2 Redox process

For the midsize SRUs (producing between 0.2 and 25 LT/D of elemental sulfur) that are required to process a lean acid gas (with 10% H2S), the operating cost of the H2S scavengers is very high and the straight-through Claus unit by itself is not operable. The niche application can be the “liquid redox” technology. Liquid redox sulfur recovery processes are based on reduction–oxidation (redox) chemistry to oxidize the H2S to elemental sulfur in an alkaline solution containing a redox catalyst.

The liquid redox technology was introduced in the beginning of the eighteenth century and was developed to process coal gas and synthesis gas for the removal of H2S. There are many versions of the redox processes, but most have limited commercial success due to various operational and environmental problems. Currently, only a few liquid redox technologies are being practiced. This includes the LO-CAT® process from Merichem and the SulFerox® process from Shell. Both the LO-CAT and the SulFerox processes are essentially the same in principle.

The process flow schematic of a typical redox process is shown in Figure 9-8. The sour gas stream is fed to the redox absorber, which employs a dilute water solution of ferric iron, held in solution by organic chelating agents, which are proprietary compounds that maintain iron ions in solution and prevent the precipitation of Fe(OH)3 and FeS.

The redox technology represents a single step that can be used to replace the traditional H2S removal unit, Claus unit, and the TGTU. In principal, the redox process is a very versatile process that is suitable to treat different feed gases, not limited to acid gases, to meet very low H2S specifications. Unlike traditional acid gas removal solvents, coabsorption of CO2 and hydrocarbons by the chelated solution is relatively low and hence, the energy consumption is low. Liquid redox can achieve 99% sulfur recovery efficiency. The disadvantages of the redox process are: complex operation, high consumption of catalysts and chemicals, and low sulfur quality.
9.3.3.3 CrystaSulf

The CrystaSulf process is a licensed process from CrystaTec Inc., which is based on nonaqueous liquid-phase Claus reaction that combines H₂S with SO₂ to form dissolved elemental sulfur in a single step. Using SO₂ as an oxidant, CrystaSulf converts inlet H₂S to elemental sulfur through a modified liquid-phase Claus reaction. Elemental sulfur is soluble in CrystaSulf solution, which avoids circulating solids in high-pressure equipment. The sulfur is crystallized and separated from the process using solid handling equipment, maintaining the rest of the process in a solids-free state. It is suitable for the medium-scale sulfur removal market (0.2–25 tons/day). The process can meet H₂S specifications of less than 4 ppmv at pressure above 150 psig, and CO₂ has no negative effect on the CrystaSulf process. The basic flow schematic is shown in Figure 9-9.

FIGURE 9-8
Typical redox unit.

FIGURE 9-9
CrystaSulf process flow schematic (Courtesy of CrystaTec Inc).
The elemental sulfur crystals are not high quality and are suitable to be used as a soil amendment for agricultural use, blended with other sulfur streams, or disposed as a nonhazardous waste.

9.3.4 Microbiological treatment processes

The Bio-SR process was developed in Japan in 1984 to recover sulfur from sour gas (Satoh et al., 1988). In this process, the sour gas is contacted with a solution containing ferric sulfate \([\text{Fe}_2\text{(SO}_4)_3]\) in an absorber, where a chemical reaction occurs between the ferric sulfate and the \(\text{H}_2\text{S}\), yielding sulfur and ferrous sulfate \((\text{FeSO}_4)\). The elemental sulfur is conglomerated to a large particle size in the slurry tank. Sulfur is separated from solution through filter and settler. The ferrous sulfate solution is then introduced into the bioreactor where \(\text{Thiobacillus ferrooxidans}\) bacteria are employed for the regeneration of the solution. The bacteria oxidize ferrous sulfate into ferric sulfate. The regenerated solution is then recycled to the absorber to repeat the cycle. In this closed-loop system, there is no degradation of the solution, no waste is generated, and essentially neither catalyst nor any chemical needs to be added.

The THIOPAQ process (codeveloped by Shell Global Solutions and Paques B.V) is one of the common technologies for biological treatment of sour gas, which can be used to remove \(\text{H}_2\text{S}\) from sour gas or from acid gas (see Section 6.11 in Chapter 6 for description of the THIOPAQ process).

9.4 Tail gas cleanup

Tail gas from a Claus unit invariably contains small quantities of sulfur compounds whether it is from the conventional Claus process or its extended version. In the earlier SRUs, removal of these trace quantities was not considered since there were no strict regulations that limit sulfur emissions. However, as air became more polluted, environmental regulations were imposed and the unconverted sulfur emissions from the Claus unit tail gas became the target of regulation. This led to the development of Claus tail gas cleanup processes in order to remove the residual sulfur species to meet emissions regulations. The tail gas cleanup processes can be categorized into three general groups: reduction, \(\text{SO}_2\) scrubbing, and catalytic oxidation.

9.4.1 Reduction processes

Reduction processes, which are commonly called the tail gas treating processes, convert all sulfur compounds in the tail gas to \(\text{H}_2\text{S}\) by hydrogenation followed by \(\text{H}_2\text{S}\) scrubbing by one of the selective amine-type processes so that \(\text{H}_2\text{S}\)-rich gas can be recycled to the inlet of the Claus unit for destruction. Therefore, the only emission is from the \(\text{CO}_2\) vent gas. Processes of this type include Shell Claus Off-gas Treating (SCOT) and the Parsons/Unocal’s Beavon
Sulfur Removal series of processes. The SCOT process was the first to be developed and is the most widespread, with more than 200 units constructed all over the world since 1973.

Two processes involved in the TGTU are described in the following sections.

9.4.1.1 Hydrogenation section

The hydrogenation section (Figure 9-10) is composed of three steps: Claus tail gas preheat, hydrogenation of SO$_2$ and other sulfur species to H$_2$S, and quenching and removal of water from the effluent gas.

The fundamental process would require an in-line fired heater to raise the tail gas to 550–650 °C by substoichiometric combustion of natural gas in a reducing gas generator (RGG) for subsequent catalytic reduction of virtually all non-H$_2$S sulfur components to H$_2$S. Conversion of SO$_2$ and elemental sulfur is by hydrogenation over CoMo catalyst.

Alternatively, a low-temperature CoMo catalyst can be used. The reactor can be operated at 410 °F using steam heating. This option avoids the use of an RGG, saving capital and operating costs while also minimizing the dilution effect from the combustion by-products. However, the use of a steam reheater will limit the ability to compensate for normal catalyst activity loss with age, potentially limiting its useful life. A bottom layer of COS hydrolysis catalyst in the first Claus converter may be required for COS/CS$_2$ hydrolysis.

![Figure 9-10: Typical hydrogenation section of TGTU. CW, Cooling Water; DCC, Direct Contact Condenser.](image-url)
The CoMo catalyst in the reactor hydrogenates/hydrolyzes essentially all the sulfur species in the Claus tail gas to \( \text{H}_2\text{S} \) according to the following reactions:

**Hydrogenation reaction:**

\[
\begin{align*}
\text{SO}_2 + 3\text{H}_2 & \rightarrow 2\text{H}_2\text{O} + \text{H}_2\text{S} \\
\text{S}_\text{X} + x\text{H}_2 & \rightarrow x\text{H}_2\text{S}
\end{align*}
\]

**Hydrolysis reaction:**

\[
\begin{align*}
\text{COS} + \text{H}_2\text{O} & \leftrightarrow \text{CO}_2 + \text{H}_2\text{S} \\
\text{CS}_2 + 2\text{H}_2\text{O} & \leftrightarrow \text{CO}_2 + 2\text{H}_2\text{S} \\
\text{CO} + \text{H}_2\text{O} & \leftrightarrow \text{H}_2 + \text{CO}_2
\end{align*}
\]

CO and \( \text{H}_2 \) naturally present in the Claus tail gas are sufficient to meet most of the hydrogenation demand, with the balance produced in the RGG.

The hydrogenated tail gas is cooled by generating steam in the effluent cooler. The effluent gas is further cooled in a quench tower to condense the reaction water. The hot gas is directly contacted with circulation water in the quench tower where heat and condensate are removed. Caustic solution can be injected to the circulation water to ensure that any \( \text{SO}_2 \) breakthrough can be eliminated under any operations. This design feature is essential, as it would avoid contamination of the expensive amine solution downstream.

### 9.4.1.2 Selective \( \text{H}_2\text{S} \) removal section

After quench cooling, the hydrogenated tail gas is treated by a selective amine treating unit to remove \( \text{H}_2\text{S} \) (Figure 9-11). The common selective amines include generic MDEA (40–45 wt%), Flexsorb\textsuperscript{®}/SE and SE Plus, UCARSOL\textsuperscript{®} HS-101 and 103, GAS/SPEC\textsuperscript{®} SS and TG-10, and Huntsman MS-300.

The amine is typically designed for a maximum rich loading of 0.1 mol acid gas (\( \text{H}_2\text{S} + \text{CO}_2 \)) per mole amine. Cooling of the lean amine to at least 100 \(^\circ\text{F}\) is critical for favorable \( \text{H}_2\text{S} \)–amine equilibrium that is necessary to reduce the residual \( \text{H}_2\text{S} \) in the \( \text{CO}_2 \) vent. If necessary, a chilled water system can be used. Specialty TGTU amines are essentially pH-modified MDEA to facilitate stripping that is necessary to treat the \( \text{CO}_2 \) gas to below 10 ppm \( \text{H}_2\text{S} \), allowing it to be vented to atmosphere, rather than incinerated. Venting would save the fuel cost required in incineration or thermal oxidizers.

Proprietary solvents, such as ExxonMobil’s Flexsorb, which is a sterically hindered amine with a very high selectivity toward \( \text{H}_2\text{S} \), can meet very low emission requirements. The other advantage is equipment cost savings from a 20–30% reduction in circulation rate. Flexsorb SE Plus is available for treatment to below 10 ppmv \( \text{H}_2\text{S} \). Both solvents require a license agreement with...
ExxonMobil. The Flexsorb solvent is more stable than MDEA, but the main drawback is that the solvent is very expensive.

In tail gas applications, FLEXSORB® SE solvents can use as little as one-half of the circulation rate and regeneration energy typically required by MDEA-based solvents. CO₂ rejection in TGTU applications is very high, typically greater than 90% rejection, which reduces the size of the amine system as well as the size of the SRU.

9.4.2 SO₂ scrubbing processes

The basic principle behind these processes lies in the incineration or catalytic oxidation of Claus tail gas streams to convert all sulfur species to SO₂. If the resulting SO₂ concentration is below emission limit, the gas can be discharged to atmosphere. If the SO₂ concentration following oxidation is higher than permissible, the gas must be treated to remove SO₂ and then recycle the SO₂ stream to the Claus unit for conversion to elemental sulfur.

The most common SO₂ removal process is caustic wash. Spent caustic is nonregenerable and is seldom used because of the waste disposal problem. Regenerative processes such as the CanSolv or Clintox process that use a proprietary solvent for SO₂ removal, can achieve high sulfur recovery up to 99.9%. However, the residual SO₂ content in the vent gas is about 80 ppmv, which may require further treatment to meet the 10 ppmv requirement for venting.
9.4.3 Catalytic oxidation

Sulfur recovery of the conventional Claus unit can be further increased to over 99% by the addition of one or more catalytic stages, such as the SuperClaus process or a sub-dew-point reaction process. These two processes are described in the following sections.

9.4.3.1 SuperClaus process

The SuperClaus process was developed by Comprimo B.V. (now Jacobs Comprimo Sulfur Solutions) to increase the overall sulfur recovery of the Claus process by reducing its inherent thermodynamic limitations (Lagas et al., 1994). The SuperClaus process consists of a conventional Claus unit thermal stage, followed by two reactors filled with standard Claus catalyst, and a final reactor filled with the SuperClaus catalyst. In the thermal stage, the acid gas is burned with a substoichiometric amount of controlled combustion air such that the tail gas leaving the second reactor contains typically 0.8–1.0 vol% of H₂S. In cases where substoichiometric combustion in the reaction furnace is not possible due to minimum flame temperature requirements, the 0.8–1.0 vol% H₂S concentration is maintained by routing a portion of the acid gas feed to downstream of the first condenser. A total sulfur recovery efficiency of up to 99% can be obtained.

9.4.3.2 Sub-dew-point processes

The conventional Claus sulfur recovery process is limited by reaction equilibrium to sulfur recoveries in the range of 94–97%. When a sub-dew-point reaction stage is added to the Claus unit, sulfur recovery can be increased to 97.5–99.5%, depending on the H₂S concentration in the acid gas and the number of catalytic stages.

The sub-dew-point reactor operates below the sulfur dew point (260–320 °F), resulting in a favorable shift in the reaction equilibrium and higher sulfur conversion. The liquid sulfur deposited in the sub-dew-point catalyst bed is removed cyclically by passing hot gas through the bed to vaporize the adsorbed liquid sulfur, which is then condensed and removed in the sulfur condensers. The catalyst bed is then cooled before returning for adsorption.

Cyclical operation requires two beds: one operating on conversion and the other on regeneration. Due to the cyclic nature of the sub-dew-point process, the reactor switching valves are subjected to thermal stress and corrosion from the sulfur vapor service. The sub-dew-point units have encountered significant operation and maintenance problems in some installations.

Processes of this type include cold bed absorption (Goddin et al., 1974), Maximum Claus Recovery Concept (MCRC) (Heigold and Berkeley, 1983), and Sulfreen technology (Willing and Linder, 1994).

9.4.4 Other tail gas treating configurations

The TGTU can be integrated with other upstream acid gas removal unit (AGRU) or AGEU to eliminate or reduce emissions, simplify operations, and reduces capital cost.
9.4.4.1 Integration with AGRU for zero emissions
The tail gas from hydrogenation section can be compressed and recycled back to inlet of the AGRU for selective H$_2$S removal, as shown in Figure 9-12. This would eliminate the tail gas amine treating unit and the incinerator. The main advantage is the elimination of an emission source from the typical TGTU, which makes permitting easier. In this configuration, the treated gas from the selective AGRU contains all the CO$_2$ and the residual H$_2$S (4 ppm), while most of the sulfur species are destroyed in the Claus SRU, achieving a “zero emissions” status. However, compressing the tail gas from atmospheric pressure to the feed gas pressure requires several compression stages. There is an associated capital and operating cost to operate the recycle gas process.

9.4.4.2 Integration with AGEU
The TGTU and AGEU can be designed as an integrated unit, as shown in Figure 9-13. As can be seen, the AGEU absorber can be used to selectively absorb H$_2$S from the lean acid gas feed, producing an H$_2$S-rich solvent and a CO$_2$ overhead gas. Similarly, the H$_2$S in the tail gas can be removed in the TGTU absorber. A common solvent regenerator can be used to regenerate the rich solvents from both absorbers. The CO$_2$ gases from the AGEU and TGTU absorbers containing trace amounts of H$_2$S (10 ppmv) can be directly vented to the atmosphere.

It should be noted that the acid gas enrichment process can also be integrated to advanced tail gas absorption processes, including the Fluor-patented Double Absorption Process℠ (DAP). Details of the DAP process are described in Chapter 6.

![Diagram of integration with AGRU for zero emissions](image-url)

**FIGURE 9-12**
Zero emission with tail gas recycle (Flood et al., 2011).
9.5 Sulfur degassing

Sulfur pits are used to collect raw liquid sulfur from the sulfur condensers under gravity. Raw liquid sulfur produced from the Claus sulfur condensers typically contains about 300 ppmw of soluble H₂S as hydrogen polysulfide (H₂Sₓ) and about 20–30 ppmw SO₂. H₂Sₓ is a weakly bound polymeric sulfur compound formed by the equilibrium reaction between sulfur and H₂S following the below reaction:

$$\text{H}_2\text{S} + (X - 1)\text{S} \leftrightarrow \text{H}_2\text{S}_X$$  \hspace{1cm} (9-14)

This reaction proceeds to the right with increasing temperature conditions. During storage or transport, H₂Sₓ compounds will decompose. This results in formation of dissolved H₂S in the liquid sulfur, which will vaporize by physical desorption. H₂S and SO₂ can accumulate to dangerous levels in the storage pits. If liquid sulfur is not degassed, H₂S will be released during storage, handling, loading, and transport. Undegassed sulfur can also create an explosive mixture of H₂S in air, and poses a toxicity hazard and a noxious odor when H₂S is released from the liquid sulfur.

The desirable liquid sulfur product should contain H₂S concentration of 10 ppmw or less, suitable for transport. To meet the 10 ppmw total H₂S requirement for sulfur transport, sulfur degasification processes employ a combination of residence time, agitation, and sometimes catalysts. All degassing processes involve agitation of the liquid sulfur and removal of the evolved H₂S with a sweeping gas. Generally, air is used as the sweep gas since oxygen helps to release the H₂S from the polysulfide molecule. The contaminated degassing off-gas is typically vented to the thermal oxidizer for oxidation to SO₂, or directed to the front end of the SRU. A properly sized wire mesh mist eliminator is utilized to minimize potential liquid sulfur entrainment in the overhead vapor stream.
The most common degassing processes capable of meeting a 10 ppmw total H₂S specification are the D’GAASS process (licensed from Fluor) and the Aquisulf process (licensed from Lurgi).

9.5.1 D’GAASS process
The D’GAASS process (developed and commercialized by Fluor in 1996) can achieve 10 ppmw residual H₂S/H₂Sₓ in liquid sulfur without catalyst addition. There are over 80 D’GAASS units worldwide with a total capacity of over 40,000 LT/D sulfur. The picture of a typical D’GAASS module is shown in Figure 9-14.

In this process, degasification is carried out in a pressurized vertical vessel, using instrument air or clean plant air. In fact, the key to the D’GAASS process is the elevated operating pressure typically over 60 psig, using the proprietary vessel internals for degassing. Operation at elevated pressure facilitates degassing operation, and results in low capital and operating costs. Due to the elevated operating pressure, the D’GAASS contactor is also small. The contactor vessel is typically located outside the pit, which makes it easy to install in a grass root unit or for retrofitting an existing facility.

FIGURE 9-14
D’GAASS module (Courtesy of Fluor).
For SRUs with TGTUs that are required to meet more than 99.9% sulfur recovery, conventional degassing systems that vent the vapor stream to the incinerator can be a significant emission source. In these cases, emission from the degassing systems is the only emission source from the SRU/TGTU. For the D’GAASS process, more than 90% of the H2S/H2SX feed and the overhead stream can be recycled back to the SRU, resulting in a zero-emission design.

9.5.2 Aquisulf process

Among these degassing processes, the earliest process is the Aquisulf process (Nougayrede and Voirin, 1989), with more than 80 units in operation ranging from 15 to 1200 LT/D sulfur (single train). This process is based on the two principles: mechanical degassing by agitation and pulverization to promote gas–liquid contact and chemical degassing by catalyst injection to speed up the decomposition of H2Sx. The Aquisulf process is available in batch and continuous versions. In the continuous version (see Figure 9-15), the liquid sulfur pit consists of two or more compartments. The sulfur in the first compartment is pumped and mixed with the catalyst and sprayed back into the compartment. The liquid sulfur overflows from a weir to the second compartment. The liquid sulfur is again pumped and sprayed in this tank to provide more agitation. The original catalyst was ammonia. However, problems associated with ammonium salts have resulted in the development of an improved version of the proprietary Aquisulf liquid catalyst.

9.6 Sulfur storage and handling

Liquid sulfur, degassed to below 10 ppmw H2S, has a strong equilibrium driving force to react with air (oxygen) to form SO2 in air-purged storage tanks.
The degassed liquid sulfur, containing traces of SO$_2$, is typically directed to an insulated and heated carbon steel storage tank to provide a buffer between the SRU and the downstream system. Some H$_2$S will evolve inside the tank, due to the air/sulfur/H$_2$S equilibrium.

Despite the fact both H$_2$S and sulfur are flammable in air, the current industry practice is to use an air sweep to the vapor space in the sulfur tank to maintain the H$_2$S concentration to below the lower explosive limit, which is 3.4% volume at the storage temperature of 330 °F (Iyengar et al., 1995).

The stored liquid sulfur can be shipped to the market by tank trucks or by railcars or pipeline. In most cases, it is converted into a solid form for ease of handling and transportation. The liquid sulfur can also be poured to a block for short-term storage during emergency. The sulfur fines reclaimed from the sulfur block can be melted to provide feed for shipments or to sulfur forming facilities.

**9.6.1 Molten sulfur handling system**

Molten sulfur transmission pipelines must be maintained within the upper and lower temperature limits that are necessary to prevent solidification. In fact, to pump sulfur, a minimum temperature (255 °F) is needed, but if heated to an excessive temperature (320 °F), sulfur degrades and thickens. Therefore, all lines and equipment must be adequately heated to maintain the sulfur at the optimum temperature. A number of heating systems have been successfully employed, including steam, heating medium, glycol, hot water, and skin effect electrical heat tracing (SEEHT).

It is widely accepted by pipeline operators that SEEHT system, monitored and controlled by fiber optic technology, is cost effective and reliable for long sulfur transmission pipelines (Beres et al., 2004). An SEEHT system is typically designed to maintain the sulfur at an average fluid temperature of 275–284 °F. The fiber optics temperature measurement should be accurate to ±1 °F, for identification of cold spots throughout the pipeline network (Johnson, 2008).

When the molten sulfur is supplied to a forming facility, it is cooled with air or water to just above its solidification temperature (250 °F), and is filtered to avoid plugging up of the sulfur spray nozzles in the forming facility (Johnson, 2009).

**9.6.2 Sulfur forming**

Bulk solid sulfur is produced in many different forms, which vary in their friability and mechanical strength, using different technologies and processes.

Slating was the sulfur industry’s first attempt at developing a formed product, which generates less dust than the traditional crushed bulk sulfur method. Slate form of sulfur is produced by pouring and spreading liquid sulfur onto a slow-moving belt while simultaneously cooling the belt by air from above and water from below. There is a tendency for slates to break into flake and dust, and therefore, slated sulfur is a less desirable form.
Prilled sulfur is created when sulfur droplets are cooled by a surrounding fluid, which may be air or water. Air-prilled sulfur is created when spraying liquid sulfur from the top of a tower against an upward flow of air. However, the facility is costly and generates sulfur emissions. This method is not common these days.

Wet-prilled sulfur is created when sulfur droplets fall into an agitated water bath, which solidifies the sulfur into uniformly sized prills. The prills are then separated from the water in high-frequency vibrating screens.

Granulated sulfur is created through the gradual buildup of successive layers of sulfur around a central core. The ultimate product size can be controlled. In the granulation process, small particles of sulfur are introduced to a long horizontal rotating drum where the particles are spray coated with molten sulfur.

9.6.3 Conveying formed sulfur
From the forming facility, the bulk sulfur is transported to the storage area. Typically, rubber belt conveyors are used in transferring the sulfur from production to storage. Front-end loaders are used to transfer bulk sulfur from a stockpile to a hopper feeding a conveyor.

During the normal course of handling solid sulfur, some sulfur dust will be generated. To keep this fugitive dust from accumulating within the conveyors, the conveyor enclosures are swept with air. This dust-laden air system, if vented directly, would be a major source of particulate emissions. Therefore, a dust suppression system shall be installed at conveyor transfer points and at load out stations. This system typically uses a combination of water, compressed air, and a surfactant mixed in a proprietary foam generator where the foam is applied to the formed sulfur at the above-mentioned points and helps suppress dust (Riegel and Kent, 2003).

9.6.4 Storage of formed sulfur
Sulfur storage has been incorporated into the design of many SRUs to allow for accumulation of sulfur during interruptions in the sulfur shipping system, or during periods of low sulfur demands. The sulfur is often stored in the open in huge stockpiles where it is exposed to wind, rain, dust, etc. In some cases, sulfur is stored indoors where some protection from the elements is available. Although this refinement may be expensive, it is the preferred modern method of storing formed sulfur product awaiting delivery to end users. However, if indoor storage cannot be provided, a location shielded from prevailing winds should be selected to reduce the migration of fugitive dust. Also, the capability to collect and dispose of water effluent off the pad will be required.

9.7 SRU design considerations
Overall SRU reliability can be increased through proper design methods and practices. Unit reliability can be significantly improved by attention to details in designing the unit, and typically requiring on a modest increase in the initial
capital cost. However, many of the features that can be incorporated into a design are difficult and more expensive to implement later on into an existing unit. Therefore, it is extremely important for the designer to design the SRU for optimum reliability at the earliest stage of design. The following discussion identifies some of the key design parameters that need to be considered.

9.7.1 Piping

Piping design, equipment layout, and plot layout are the most important aspects of the SRU design. The rules for piping design are relatively simple such as:

- Steam jacketing liquid sulfur lines
- Piping to be free draining
- Piping runs to be as short as practical
- Utilize crosses at all direction changes in all liquid sulfur lines
- Slope all liquid sulfur lines to promote gravity flow.

Sulfur will solidify at temperatures between 245 and 250°F. It is therefore imperative to maintain the sulfur piping above 250°F using steam-jacketed piping. The steam supply and condensate piping, and particularly the steam trap, should be properly installed. Steam traps should be provided at frequent intervals to maintain the sulfur temperature.

It is important to avoid liquid sulfur from being cooled in piping, as it will result in additional pressure drop through the SRU, causing equipment damage, corrosion, and damage to personnel. The piping design should ensure that any liquid is free drained and pumped back to the process equipment.

The pipe runs should be kept short and the process should be kept hot. It is easier to keep the process hot if the pipe runs are short. Heat loss from uninsulated flanges and valves would promote condensation, corrosion, and plugging in the piping system. All lines containing ammonia and sulfur vapor should use only steam-jacketed control and block valves.

It should be noted that most lines would plug at some point during the SRU life. To avoid creating a permanent plug, any change in flow direction in liquid sulfur service should be provided with a cross, which allows the piping to be rodded out to break the plug. In addition, the upstream valves should be steam jacketed to allow rodding.

Liquid sulfur is a viscous fluid, and can be become very viscous at some temperatures. Gravity flow liquid sulfur piping should be designed to run only partially full. All liquid sulfur piping should be sloped at minimum 1/8” per foot to promote free draining.

9.7.2 Acid gas feed drums

The acid gas feed drum will remove the entrained liquids from the amine regenerator, during normal operation and from gas processing plant upsets. Entrained liquids are sour water, hydrocarbons, and amines. If not removed,
entrained liquids can cause problems with the feed metering system, causing plugging in the burners, refractory damage, and eventual failure of the burners and reaction furnace.

The drums should be equipped with pumps to remove the collected liquids to the sour water system or waste tanks. There should be spare pumps, and these should be operated on a start stop basis, being activated on high liquid level.

9.7.3 Combustion air blowers

For reliability, there should also be spare combustion air blowers, either two 100% capacity blowers or three 50% blowers. Multistage centrifugal motor driven blowers are common.

9.7.4 Main burner and reaction furnace

The burner must be of high-intensity design with efficient mixing, which is critical to ensure complete reaction. It must perform the function of burning one-third of the feed H₂S to SO₂ to satisfy the modified Claus process. The temperature must be high enough to destroy impurities and contaminants in the acid gas (see previous section). The burners must be capable of performing efficiently at normal operation and during turndown.

The feed impurities and contaminants must be destroyed in the reaction furnace or they will cause fouling problem in the downstream equipment. There must be adequate time for the reaction to reach equilibrium. The reaction furnace is typically designed for 0.6–1.5 s residence time.

The burner and furnace are typically constructed of carbon steel. The extremely high temperature, reducing atmosphere, and acid gas environment are very corrosive. The furnace refractory lining must be designed with proper material for the corrosive environment. The furnace is typically equipped with proprietary internal components supplied by licensor to promote mixing and the designs depend on feed gas compositions, temperature, and the amounts of contaminants.

The burner and furnace system must be equipped with necessary instrumentation for control and monitoring. The other critical burner instruments are the flame scanners. If the flame scanners are defective, it can result in nuisance shutdowns.

It is necessary to purge each instrument nozzle when the SRU is shutdown. The sulfur content in the process gas will condense and solidify if allowed to enter the instrument leads.

9.7.5 Waste heat boiler

SRU WHBs are typically of the fire tube design. They serve the dual purposes of heat recovery by generating steam and cooling the process gas from 2000 °F to
9.7.6 Sulfur condensers
Sulfur condensers serve the primary function of cooling and condensing sulfur from the upstream reaction. Sulfur condensers are normally kettle-type shell and tube exchangers. The sulfur condensers are designed to condense and separate product sulfur from the process gases. The sulfur condenser design must operate within the range of the design mass velocities, typically ranging from 2 to 6 lb/s-ft².

9.7.7 Sulfur pit
Product sulfur is normally collected in a below-grade, concrete pit equipped with steam coils to keep the sulfur molten. The pit does not directly affect the SRU process operation until the SRU must be shutdown because of problems with the pit.

Some common sulfur pit problems are steam coil leakage, sulfur pump failure, internal sulfur fires, and even internal explosions. There are design features that will improve the reliability of the sulfur pit including:

- Construct the pit using sulfate-resistant concrete with limestone-free aggregate.
- Use alloy piping for the steam supply piping and internal components.
- Install dual-steam-jacketed sulfur transfer pumps.
- Use a fully steam-jacketed steam eductor to continuously provide sweeping air to the pit to prevent accumulation of H₂S.
- Steam snuffing connections for extinguishing internal sulfur fires.

9.8 SRU operation problems
Problems associated with SRU operation generally relate to misoperation of the furnace and the reactor systems, reduced catalyst activity, and throughput limitation due to high pressure drop. The end result is failure to meet the SRU throughput, the sulfur recovery target, and sulfur emissions. Emissions from the SRUs are continuously monitored. If they do not meet the emission permits requirements for a certain amount of time, and remedies fail to resolve the problems, the complete facility has to be shutdown. The following sections summarize these problems and provide guidelines to mitigate the problems.

9.8.1 Proper air ratio
For optimum conversion, the ratio of H₂S/SO₂ must be 2:1. This ratio is measured in the tail gas from the final condenser. SRUs are best operated on closed-loop
control based on tail gas analyzer. In practice, other methods can also be used to adjust air to the reaction furnace. Excessive airflow to the reaction furnace will result in formation of SO$_3$.

Incinerator temperature can be used to adjust the airflow to the reaction furnace. A high incinerator temperature, coupled with low fuel consumption, is a sign of insufficient air to the reaction furnace. Alternately, if a large amount of fuel gas is used to maintain the incinerator temperature, this may indicate too much air used in the reaction furnace.

### 9.8.2 Reactor activity

To verify the reactor activity, the temperature rise across each reactor should be checked. If there is a temperature drift from the design, it may indicate that the effluent from the first stage is not reaching equilibrium. Sulfur formation in the first reactor may have decreased. The overall catalyst effectiveness may have declined. The problem may be due to catalyst deactivation caused by sulfur precipitation. This can be a result of low reactor feed temperatures. Check the operation of the reheat exchanger upstream of the reactor. If catalyst activity has been irreversibly lost, the catalyst may need to be changed.

The catalyst performance can be monitored by reading the vertical temperature profile through the first catalyst bed. A typical temperature profile of the fresh catalyst and the damaged catalyst is shown in Figure 9-16. If the catalyst is in good condition, 90% of the heat of reaction is released in the top 6 inch of the bed. If catalyst activity is impaired, the reaction is shifted down the bed. The loss of activity can be the result of many factors such as carbon deposits, leaking

![FIGURE 9-16](image)

**FIGURE 9-16**

Vertical temperature profile shows the condition of catalyst.
condenser tubes, damaged reactor support screens or internals, sulfuric acid formation, and operation below the sulfur dew point.

9.8.3 Excessive COS and CS₂
The presence of hydrocarbons and CO₂ in the acid gas promotes the formation of COS and CS₂ in the reaction furnace. These compounds can significantly contribute to SO₂ formation in the incinerator. The COS and CS₂ can be eliminated by operating the first reactor at an outlet temperature of 650 °F. At this temperature these compounds are converted to H₂S and CO₂.

9.8.4 Leakage of reheat exchanger
Some multistage SRUs reheat the third-stage reactor feed with first-stage reactor effluent. The reheat exchangers are prone to leakage due to the corrosive environment, which may contribute to lower conversion. If there is a large increase in the third-stage inlet temperature, there is a possibility that the reheat exchanger is leaking.

9.8.5 Reactor pressure drop
For properly operating an SRU, it is important that the pressure drop in the unit must be monitored. A sudden increase in pressure in the SRU is a warning of the equipment being fouled. Pressure drop in an SRU is proportional to the square of the gas flow. If the high pressure drop is not due to the higher flow, it may suggest that the catalysts are plugged with carbons or some equipment component fails.

9.8.6 Carbon deposits
Heavy hydrocarbons and BTEX may be contained in the feed gas from the production wells. Some of them are absorbed by amines in the amine absorber, which will end up in acid gas to the SRU.

Ten times more air is needed to oxidize propane than H₂S. When the light hydrocarbon vapors are reacted in the converters, carbon black is formed. A gradual increase in pressure drop is an early warning sign of carbon deposit. When this happens, check for low steam production rates from the high-pressure boiler. If both steam production and reactor outlet temperatures are low and the pressure drop is high, it may indicate tube leakage. The SRU may need to be shutdown for repair.

If sulfur precipitates on the catalyst, it would stop gas flow through the unit. The worst thing that can happen to an SRU is a crash shutdown. When the unit suddenly shuts down, precipitating sulfur will solidify in the catalyst beds. SRUs should be cleared of sulfur by burning natural gas instead of H₂S before a shutdown.
9.8.7 Catalyst support screens

SRU catalyst is supported by thin flexible screens. These screens are lapped and folded over to keep the catalyst from leaking through the support grating. Improper installation of screens occurs frequently during catalyst change-out. The catalyst may wash down into a seal leg and result in plugging.

9.8.8 Water vapor and carbon dioxide

The capacity of the SRU is negatively impacted by an increase in water and CO₂ vapor in the acid gas. Reducing the water content can be done by more cooling in the regenerator overhead condenser. For example, lowering the overhead temperature from 135 to 110 °F can reduce the water content by about 5%, which can be translated to a 2% increase in SRU throughput.

By lowering the feed location of the lean amine to the amine absorber, more CO₂ can be slipped and rejected to the absorber overhead, which would correspondingly reduce the CO₂ content in the acid gas, opening up more capacity to the SRU. The use of selective amine such as formulated MDEA or Flexsorb would further reduce the CO₂ content to the SRU.

9.8.9 Steam heater

High-pressure steam is the most common source of heat for the SRU reheators. Steam reheaters are very reliable and easy to operate. However, the maximum reheater outlet temperature is limited to about 460 °F, when using high-pressure steam. This low temperature can be a problem for the first-stage reactor, particularly if significant quantities of COS and CS₂ are formed in the burner/reaction furnace.

9.8.10 Combustion air control

Control of combustion air feed is the most critical part of a sulfur recovery process to ensure efficient and reliable operation. The combustion air control should be split into two sections, a main airflow loop based on acid gas flow rate and a trim airflow loop based on the tail gas analyzer.

Air demand is calculated from the acid gas flows and used as a feed forward ratio set point for the main air control loop. The main air loop supplies about 90% of the total air to the burner.

The trim air loop operates on feedback control from the tail gas analyzer. The analyzer measures the relative amount of H₂S and SO₂ in the tail gas. The analyzer controller provides a remote set point signal to the trim air loop based on the relationship 2SO₂ = H₂S = 0.

When this relationship is satisfied, the optimum amount of combustion air is being supplied to the SRU. If the result is positive, too much air is being fed and the rate should be reduced. Likewise, if the result is negative, too little air is being
fed and the rate should be increased. It is apparent that the tail gas analyzer must work properly to achieve optimum air control. If the air controls are off, inadequate impurity destruction with equipment plugging and/or equipment corrosion may result.

### 9.9 Selecting the sulfur recovery process

The large variations in concentrations and flows require different methods for H₂S removal and sulfur recovery. For relatively small quantities of H₂S/sulfur, scavenger processes are often used. For sulfur quantities up to 10–20 T/D of sulfur, liquid reduction–oxidation (redox) processes are common but the sulfur is produced as aqueous slurry. Direct oxidation can sometimes be utilized for low H₂S concentrations to produce high-quality liquid sulfur. The selection criteria based on H₂S concentration in the acid gas is shown in Figure 9-17.

Typical acid gas in a gas processing plant contains hydrocarbons, ammonia, and BTEX component. The thermal section of the Claus process is the most robust process stage that can be used to destroy the hydrocarbons and contaminants. When the Claus process is followed by a tail gas selective treating unit, 99.99% sulfur recovery and 10 ppmv H₂S in the vent gas can be achieved. If the gas is lean, AGEU can be installed prior to the Claus SRU.

The use of direct oxidation most likely is applicable to gasification of syngas due to the absence of hydrocarbons, but is rarely used in natural gas processing plants.

![Figure 9-17: Selection of sulfur recovery process.](image-url)
For new SRUs, most countries today require the best available technologies to minimize sulfur emissions. Some of the key parameters affecting the selection of the sulfur recovery process are:

- Acid gas composition, including H₂S content and hydrocarbons and other contaminants
- Required recovery efficiency
- Concentration of sulfur species in the stack gas
- Turndown requirements
- Ease of operation
- Remote location
- Sulfur product quality
- Capital and operating costs
- Is this a revamp unit?

When required recovery efficiency and concentration of sulfur species in the stack gas is known, the tail gas cleanup process licensor can be evaluated. When concentration of impurities in the acid gas such as COS and CS₂, H₂S content, and the treated gas specifications are established, the type of amine used for a particular application could be selected.

Finally, the selection of similar process will be based on the ease of operation, robustness, experience, capital and operating cost, and remote location. For revamp units, minimum equipment modifications and process configuration should be considered as a main key factor.

### 9.10 Sulfur disposal by acid gas injection

Acid gas injection into depleted oil and gas reservoirs and into deep saline aquifers is a viable process for gas processing plants located in remote location where there is no demand for the sulfur product. This is also the only option if the gas plant is located in an environment-sensitive region, which mandates the zero emissions policy. This is the situation in Norway where Statoil has to inject about 55 MMSCFD acid gas (mainly CO₂) to the reservoir. Generally, acid gas injection unit is lower in capital cost than an SRU, if the injection destinations are nearby. In addition to providing a cost-effective alternative to sulfur recovery, the deep injection of acid gas also helps to maintain the oil field pressure, which enhances oil recovery and increases revenue.

In the acid gas injection process, the acid gas mixture is typically compressed through several stage of compression (at least three or four stages) to the reservoir pressure. Dehydration of the acid gas may not be required, depending on the acid gas composition, the injection interstage pressure, and temperature, but each case needs to be evaluated on a case-by-case basis (Carroll, 2009).

While high-pressure subsurface gas injection has been used extensively around the world, it is commonly used for injection of sweet gas or low flow rates of acid gas (5–10 MMSCFD). Only in recent years the compression technology
has been developed to permit high-flow-rate/high-pressure sour gas injection using large centrifugal compressors. As an example, the Sour Gas Injection project at Tengiz in the Republic of Kazakhstan that was started in 2006 is a demonstration of the ability to expand the operating envelope for sour gas injection: high capacity (275 MMSCFD), high H2S content (25 mol%), and high compressor discharge pressure (620 bar) (Block et al., 2005).

The success of sour gas or acid gas injection process depends on selecting an injection location that can maintain field pressure and avoid the potential gas breakthrough. This would require detailed subsurface modeling and analysis to manage the reservoir response to the injected gas. It is also very important to select a compressor supplier who has experience with high-pressure acid gas particularly in the dense-phase operation.

References


Willing, W., Linder, T. Lurgi’s TGT processes and new operational results from Sulfreen plants. Paper Presented at the Sulfur 94 International Conference & Exhibition, Tampa, FL, USA, November 6–9, 1994.
Nitrogen Rejection

10.1 Introduction

Nitrogen can be naturally occurring in high concentration in production fields, such as in the U.S. Midwest, North Sea, Eastern Europe, and South East Asia. Approximately 15% of the nonassociated gas reserves in the world are high in nitrogen content and would not meet the specification for normal pipeline quality gas (typically 3–4 mol%). Nitrogen, being an inert, will not support combustion, and the gas when burned will lead to unstable combustion. In addition, nitrogen will strip off the NGL components in the NGL recovery plants, and would increase the size of the compression equipment and transmission pipeline. In LNG plant, high nitrogen is not desirable as it will create liquefaction problem and increase boiloff gas from the storage tanks.

In production areas, where nitrogen injection is used for enhanced oil recovery, nitrogen content in the production gas would gradually increase over time as nitrogen breaks through from the injection wells to the producing wells. In earlier production, the production gas can be blended with low-nitrogen sources to meet pipeline specification. However, if the nitrogen content continues to increase, and in the absence of low-nitrogen natural gas sources, blending is no longer an option, and the nitrogen content must be removed in a nitrogen rejection unit (NRU).

This chapter reviews the current state-of-the-art nitrogen rejection technologies, their applicability, and their limitations. The technologies of cryogenic separation process are introduced and discussed on the design for high and low levels of nitrogen content gases.

10.2 Nitrogen rejection options

Nitrogen rejection methods can be classified into cryogenic and noncryogenic processes.

10.2.1 Cryogenic processes

Cryogenic nitrogen rejection technology, which is the most common method of removing nitrogen from natural gas, uses the difference of boiling points between nitrogen and methane for separation. Cryogenic processes use Joule–Thomson (JT) cooling of the high-pressure gas, and separation in a distillation column to produce a nitrogen overhead product and a methane-rich bottom. The nitrogen product contains less than 1 mol% methane, which can be safely discharge to...
atmosphere. The methane-rich bottom contains less than 1 mol% nitrogen that can be sent to the pipeline or natural gas liquefaction plant.

Cryogenic processes are typically used to treat natural gas containing more than 10 mol% nitrogen and are proven to be economical in processing gas flow rate ranging from 30 MMscfd up to 900 MMscfd.

10.2.2 Noncryogenic processes

Noncryogenic nitrogen separation processes include mainly pressure swing adsorption (PSA) and membrane separation.

PSA uses a zeolite adsorbent to selectively separate nitrogen from methane in a cyclic process. The process includes adsorption of methane at high pressure, during which nitrogen is separated at pressure. Natural gas with low-nitrogen content is produced at low pressure. The adsorbent can be regenerated with a combination of pressure and thermal changes. PSA technology is limited in capacity and can typically handle 2–15 MMscfd gas flow rate. The drawback of this process is that natural gas is produced at low pressure requiring recompression to pipeline pressure. Also, the hydrocarbon content in the rejected nitrogen is significant which represents a production loss. The losses can be somewhat reduced by recompression to the feed section, which is seldom justified because of the high compression cost. For coal bed methane applications where there are few impurities, the Molecular Gate™ adsorption-based technology (developed by Engelhard) can be used to handle small gas flow of 0.5 MMscfd. The Molecular Gate technology traps nitrogen in a unique adsorbent material while letting methane flow through. The Molecular Gate™ adsorbent is a titanium silicate molecular sieve designed with a 3.7-Å pore size. Because nitrogen has a smaller molecular diameter (3.6 Å) than methane (3.8 Å), nitrogen can enter the pore and be adsorbed while methane passes through the fixed bed of adsorbent with minimal pressure losses.

In membrane process, the feed gas is compressed and passed across the surface of the membrane elements. It separates the hydrocarbon permeate which is compressed back to the pipeline pressure while the nonpermeate, nitrogen-rich waste gas can be used as fuel. Membrane can handle small flow rates varying from 0.5 to 25 MMscfd. The membrane units are compact and can be supplied in modules, suitable for offshore installation. However, membrane separation systems would require feed gas pretreatment such as aromatics removal and dew point depression to remove impurities that may cause damage to the membrane. Membrane has the same drawback as PSA process, that is, the hydrocarbon losses in the nitrogen waste stream is relatively high.

PSA and membrane separation methods are limited in commercial applications, and have not found acceptance, especially for larger NRUs.

10.3 Nitrogen rejection unit integration

Since NRU operates at cryogenic temperatures, the feed gas pretreatments as shown in Figure 10-1 are required. Pretreatment generally consists of amine
scrubbing to remove carbon dioxide, followed by TEG dehydration, the molecular sieve and mercury removal units. Depending on the process design, total removal of CO$_2$ may not be necessary as the use of prefractionator can avoid CO$_2$ freezing in the low-pressure cryogenic column. TEG unit can remove some of the BTEX components which may minimize freezing problem in the cryogenic unit. Molecular sieves must meet $\text{CO}_2/\text{C}_14$ water dew point, the lowest operating temperature of the cryogenic unit. In processing rich gases, a hydrocarbon dew pointing unit may also be required before the nitrogen removal unit to remove the heavier hydrocarbons (BTEX).

The gas leaving the pretreatment section is cooled and liquefied in a series of operations involving expansion across JT valves to generate cooling. Depending on the feed gas compositions and pressure and the product specifications, the cryogenic separation system can be different.

The NRU design can be integrated with an NGL recovery process, to improve energy efficiency and reduce equipment counts. However, the integrated complex may prove to be difficult in operation, particularly when the NGL recovery unit needs to operate at different NGL recovery modes (ethane or propane recovery mode). To reduce process complexity, most of the NRUs are designed to operate independently from the NGL recovery units.

If the feed gas contains helium above 0.5 vol%, it can be recovered as a product in a helium recovery unit located downstream of the nitrogen rejection unit. The design of the helium recovery unit is dependent on the required helium product purity. If the helium product purity is in the 50% range, 98% recovery of the helium in the feed gas is economically feasible. For higher purity, some of the helium is lost in the nitrogen vent, reducing the recovery level.

When processing a feed gas with high CO$_2$ and significant nitrogen and helium contents, the overall process flow diagram can be configured as shown in Figure 10-2. Where CO$_2$ is used for enhanced oil recovery, it is more energy efficient to fractionate CO$_2$ in a fractionation column. The column bottom liquid can then be pumped to high pressure for reinjection. The column overhead vapor, which contains residual acid gas, is further scrubbed by a methanol wash.
system to produce a clean gas. The acid gas from the methanol wash unit is recycled back to the inlet feed gas. The clean gas is fractionated in the NRU-distillation column, producing a bottom stream containing the methane and hydrocarbons, which can be sent to the sales gas pipeline. The NRU-column overhead, containing nitrogen and helium, is fractionated in a helium recovery unit into the helium overhead product and a nitrogen bottom stream.

10.4 Cryogenic nitrogen rejection

Several cryogenic distillation schemes are known to reject nitrogen from natural gas stream. The state of the art of cryogenic nitrogen rejection processes are as follows: single-column process, double-column process, and two-column process. These processes vary in complexity and efficiency and are discussed individually in the following sections.

10.4.1 Classical single-column design

The classical single-column nitrogen rejection process uses a heat pump to provide the reflux duty and reboiler duty of the fractionation column, as shown in Figure 10-3. The single column typically operates between 300 and 400 psig. Methane can be used as the working fluid that is compressed to supply the reboiler heating duty, condensed and then evaporated at low pressure providing the condensation duty. Brazed aluminum heat exchangers (BAHXs) are used to achieve a close temperature approach required by the process.

Feed gas is typically supplied at 800–1000 psig. The gas is chilled by several effluent streams, and then letdown in pressure to the HP column in a JT valve, JT1. The column produces a nitrogen overhead vapor, and a bottom methane-rich liquid. The column bottom is letdown in pressure using valve JT2, providing further chilling to heat exchanger E-1. The low-pressure gas is then compressed to the sales gas pressure.
Due to the large boiling point difference between nitrogen and methane, sharp separation of nitrogen from methane is feasible. Typically, the single-column design can produce a nitrogen overhead product with less than 1% methane, and a methane bottom product with less than 1 mol% nitrogen. The single column process can be used to process variable nitrogen content gas ranging from 3 mol% to as high as 30 mol%. The disadvantage of this process is the high power consumption required by the heat pump and the LP gas compressor. In addition, internal reflux exchanger works best when feed gas composition does not vary. It lacks the controllability to react to varying feed gas changes, particularly its nitrogen content which is common in pipeline operation.

10.4.2 Modified single-column design

To circumvent some of the difficulties in the classical single-column nitrogen rejection process, and reduce the power consumption, the single-column design can be modified for natural gas pipeline operation, as shown in Figure 10-4.
In the modified version, the high-pressure feed gas is chilled to an intermediate temperature, typically $-60^\circ C$, and separated in cold separator V-1. Refrigeration is provided by the nitrogen vent, HP gas and LP gas, and the reboiler. By cooling to $-60^\circ C$, most of the heavy hydrocarbons (BTEX) and CO$_2$ are removed, thus avoiding solidification by the cryogenic temperature in the NRU column. Removal of the CO$_2$ content also makes the fractionation column resistant to CO$_2$ freezing, avoiding potential CO$_2$ problems. The chilled feed gas is letdown in pressure to 300 psig and 400 psig to the NRU column using valve, JT1. The overflash vapor is condensed by the reflux condenser E-2. The refrigeration duty is provided by letdown of a portion of the HP column bottom liquid in a second valve, JT2.

The modified single-column design uses the pipeline compressor to generate cooling, instead of a dedicated methane compressor. This is thermodynamically more efficient as the heat composite curves can be closely matched, minimizing thermodynamic losses. The heat composite curves of the feed gas and the multi-stream (LP gas, MP gas, nitrogen, and reboiler) are shown in Figure 10-5. Because of the close approaches of the heat composite curves, power consumption by the pipeline compressor is significantly less than the earlier single-column design.

### 10.4.3 Double-column design

The double-column nitrogen rejection process was developed in the late 1970. The process is used for air separation plant and can be modified for fractionation
of nitrogen from hydrocarbon streams. In this process, nitrogen is separated at low pressure at cryogenic temperature, typically $-310\, ^\circ F$, which requires the feed gas to contain no more than 40 ppmv CO$_2$, in order to avoid CO$_2$ freezing in the cryogenic column. This scheme takes advantage of the high nitrogen content in the feed gas to produce refrigeration by the JT effect for refluxing the cryogenic column. It has the advantage of low power consumption and a simple configuration that is inherited from the air separation plant design. When adapting the air separation plant design concepts to natural gas plants, the design must be modified to ensure that a low-hydrocarbon level in the nitrogen vent to minimize greenhouse gas emissions.

The double-column process, as shown in Figure 10-6, uses two columns stacked together operating at different pressures, where the reboiler of the LP column is thermally linked to the condenser of the HP column. The hydrocarbon liquid from the LP column is pumped by pump P-1 to an intermediate pressure, and vaporized in the exchangers E-2 and E-1, supplying cooling to the inlet gas and the reflux streams. The pump discharge pressure must be controlled to avoid the temperature pinch on the heat exchangers. Higher pump discharge pressure can reduce compression horsepower, but may result in a temperature cross on the heat curves. In most designs, the optimum discharge pressure is somewhere between 60 and 100 psig.

Nitrogen vent stream from the LP column overhead is used to provide cooling the feed gas and reflux streams in exchangers. The feed gas is cooled in these exchangers, condensed and letdown in pressure in valve JT1 to the HP column, typically operating at 300–400 psi. The HP column is designed to produce an

![Composite heating curve](image) ![Composite cooling curve](image)

**FIGURE 10-5**
Heat composite curves for exchanger E-1 in modified single-column nitrogen rejection.
overhead stream, typically about 50–60 mol% nitrogen, which must contain sufficient refrigeration for fractionation in the LP column. The HP column produces a bottom product containing methane and nitrogen, is subcooled in heat exchanger E-2 and then fed to the LP column at close to atmospheric pressure. The LP column fractionates the gas into methane and nitrogen products.

The overhead product from HP column, relatively pure nitrogen, is condensed in E-4, which is internally built-in to the column. This condenser also serves as a reboiler for the LP column. A portion of this liquid nitrogen is used as reflux to the HP column while the remainder is subcooled in heat exchanger E-3, letdown in pressure in valve JT2 and used as reflux to the LP column. The HP column bottom is subcooled in heat exchanger E-2 and letdown in pressure in valve JT3 and fed to the lower section of the LP column.

In principle, when the nitrogen content in the feed gas is high, all the refrigeration requirement can be provided by the JT valves via pressure letdown of the feed gas and products. No external refrigeration should be required. However, when nitrogen content in the feed gas is low, JT cooling is limited, which may not be enough for fractionation purposes, resulting in lower recoveries.

**FIGURE 10-6**

Typical double-column nitrogen rejection process. NRU, nitrogen rejection unit.
10.4.4 Two-column design

The two-column process is an adaptation of the double-column process where the two columns are not thermally linked. This configuration, as shown in Figure 10-7, uses two separate columns, which provides flexibility for heat recovery and integration among different streams, improving the efficiency of the process. The HP column can be considered as a preseparation column as this column removes most of the heavy hydrocarbons and CO₂.

The HP column is a nonrefluxed stripper, operating between 300 and 400 psig. It produces a bottom liquid with most of C₂⁺ hydrocarbons, and an overhead nitrogen-rich vapor, typically consisting of 60 mol% nitrogen and 40 mol% methane. The refrigerant content in the HP bottom liquid is recovered by cooling the feed gas before it is fed to the HP stage of the pipeline gas compressor. The overhead vapor is chilled in exchanger E-2, letdown in pressure to about 100 psig, and separated. The separator produces a reflux to the LP column, and a vapor stream that is further cooled and letdown in pressure to the LP column using valve JT2. The overhead gas is fractionated into an overhead nitrogen stream and a methane bottom stream. The separation can be very sharp, minimizing the methane losses in the nitrogen vent gas.

Although the two-column design is more efficient than the classical double-column process, the process design still has the same limitation. When processing low-nitrogen gas, there is not sufficient nitrogen in the HP overhead vapor, and as a result, there is not enough chilling from the JT operation; and

![FIGURE 10-7](image_url)  
Two-column nitrogen rejection process. NRU, nitrogen rejection unit.
consequently, the methane content in the nitrogen stream is higher than desirable. This process is more suitable to process feed gas with nitrogen content higher than 25 mol%.

10.4.5 Process selection

Process selection for the NRU should be based on the life-cycle costs on CAPEX and OPEX. The more important criterion today is the methane content in the nitrogen vent, which is responsible for the greenhouse gas emissions.

The two process configurations that can be considered are as follows: the single-column design and the two-column design. The methane content in the nitrogen vent gas for these two designs can be compared in Figure 10-8. The single column design can achieve less than 1% methane content in the nitrogen vent, for a feed gas with a variable nitrogen content up to 60 mol%. However, for the two-column design, only when the nitrogen content is higher than 25 mol%, the methane content can be kept at below 1 mol%. When operating at lower than 25 mol% nitrogen, the methane content in the nitrogen vent increases significantly.

The power consumption for these two designs is shown in Figure 10-9. The power consumption is based on processing 70 MMscfd feed gas with variable nitrogen content at 800 psig, and the residue gas is recompressed back to the pipeline. As shown in Figure 10-9, the two-column design requires about 50% of the power required by the single-column design. For this reason, unless the feed gas contains lower than 25 mol%, the two-column design is the right choice.

**FIGURE 10-8**

Methane content in nitrogen vent gas.
However, when the nitrogen content is lower than 25 mol%, the single column is most likely the right choice that is governed by the requirement to meet hydrocarbon emissions of less than 1 mol%.

10.5 Design considerations
The following describes some of the cryogenic NRU design issues.

10.5.1 Feed gas characteristics
The NRU operation is very sensitive to changes in feed gas compositions, nitrogen content, and inlet temperature and pressure. If the feed gas is expected to process a variable nitrogen content gas, the NRU design must include sufficient design margins and capacity to accommodate the expected changes.

Higher feed gas pressure would generate more cooling when feed pressure is letdown to the NRU column. Lower feed gas temperature would reduce the refrigeration requirement and may require more reboiler duty. The NRU heat and material balance must consider the varying feed gas compositions in the equipment design. Sufficient chilling must be provided to ensure the column reflux and reboiler duty be met for various operation including plant start-up. If necessary, a heater should be provided on the NRU column to ensure sufficient heat is available during start-up.

10.5.2 Reflux and reboiler duties
Methane content in the nitrogen vent stream must meet hydrocarbon emissions requirements, and the nitrogen content in the hydrocarbon stream must meet the
sales gas specification. The reflux system is typically controlled by the high nitrogen case, while the reboiler design is controlled by the low nitrogen case. The design of BAHX must include the extreme range of operation, as it will influence the NRU-column design, and the refrigeration requirements.

10.5.3 Temperature and pressure control
The NRU-column shall be equipped with temperature indicators for start-up and normal operation. The BAHX should be provided with temperature indicators that are used to avoid temperature excursion during start-up or upset conditions. The NRU-column pressure must be designed with advanced control on reflux and reboiler, as the fractionator temperature is very sensitive to a slight change in column pressure.

Pressure differential across the BAHX, column pressure and feed strainers shall be continuously monitored to detect hydrate formation or blockage by foreign materials.

10.5.4 Insulation
Cold conservation is critical in an autorefrigerated system. While a design margin can be applied to a mechanically refrigerated system to compensate for external heat leak, an autorefrigerated system relies on efficient insulation to limit heat gain from the environment.

The equipment can be mechanically insulated with advanced efficient insulation materials, or they can be enclosed in a cold box filled with perlite. For a cold box design, an internal, dry nitrogen atmosphere must be maintained to ensure insulation integrity and safety. Leakage of hydrocarbons must be monitored to avoid hydrocarbon accumulation within the cold box.

10.5.5 Reboiler hydraulics
Sufficient heat must be provided to the NRU bottom to meet product specifications. Because of the expansion loops in the reboiler piping, excessive pressure drop can slow down the reboiler flow. A hydraulic study of the reboiler system must be carried out for the various operations, particularly during start-up, to ensure adequate heating is provided to the NRU column.

10.6 Operating problems
The following describes some of the cryogenic NRU operating issues.

10.6.1 Feed contaminants
Feed gas to NRU can contain impurities such as water, BTEX, pipe scales, and other contaminants, resulted from poor-performance of the upstream units that
will cause blockage of the BAHX. Feed strainers must be provided to avoid fouling of the BAHX. They should be removed for cleaning online when high-pressure differentials are detected.

10.6.2 Foaming
Foaming is an unlikely event. However, if the NRU-column pressure drop is higher than design, the column should be shut down for inspection.

10.6.3 High methane content in nitrogen vent
If the methane content in the nitrogen vent gas is high, it is most likely caused by insufficient reflux, which may be due to the limitation of the compression cooling equipment. The other cause could be due to reflux pump capacity limitation, lower fractionation efficiency and maldistribution in the column that may be caused by hydrates.

10.6.4 Hydrate formation
During NRU start-up, the system may not be completely dried. Therefore, methanol injection should be used to remove any hydrate.

10.6.5 Nitrogen safety
During plant upset, the nitrogen vent gas can drop to very low temperature, which may cause a safety hazard. Nitrogen vent must be routed to a safe location, away from the operating plant. Nitrogen must not be allowed to accumulate in confined spaces, which may cause accidental asphyxiation.

10.6.6 BAHX failure
BAHX is known to fail from severe temperature excursion during plant start-up. As such, monitoring of these exchangers is critical, particularly the rate of temperature change. Avoid the “the cold death spiral” on the exchanger. The manufacturer’s guideline on cooling down rate and derimming must be strictly followed.
11 Natural Gas Compression

11.1 Introduction

“Compression” is used in all aspects of the natural gas industry including gas lift, reinjection of gas for pressure maintenance, gas gathering, gas processing operations (gas loading and discharge), transmission and distribution systems, boil-off system (gas storage and tanker for vapor control and to avoid releasing gas to atmosphere). In recent years, there has been a trend toward increasing pipeline-operating pressures. The benefits of operating at higher pressures include the ability to transmit larger volumes of gas (referred at base conditions) through a given size of pipeline, lower transmission losses due to friction, and the capability to transmit gas over long distances without requiring or even reducing additional compressor stations. In gas transmission, two basic types of compressors are used: reciprocating and centrifugal compressors. Reciprocating compressors are usually driven by either electric motors or gas engines, while centrifugal compressors use gas turbines or electric motors as drivers. The key variables for equipment selections are life-cycle cost, capital cost, maintenance costs including overhaul and spare parts, fuel or energy costs. The unit’s level of utilization, as well as demand fluctuations, plays an important role. Both gas engines and gas turbines can use pipeline gas as a fuel, but electric motor has to rely on the availability of electric power. Due to the number of variables involved, the task of choosing the optimum driver can be quite involved and a comparison between the different types of drivers should be done before a final selection is made (Lubomirsky et al., 2010). An economic feasibility study is of fundamental importance to determine the best selection for the economic life of a project. Furthermore, it must be decided whether the compression task should be divided into multiple compressor trains, operating in series or in parallel (Santos, 1997, 2004).

This chapter presents a brief overview of the two major types of compressors, a procedure for calculation of the required compression power, as well as additional and useful considerations for the design of compressor stations. All performance calculations should be based on compressor suction and discharge flanges conditions. For reciprocating compressors, pressure losses at the cylinder valves as well as the pressure losses in pulsation dampeners have to be included in the calculation. Additional losses for process equipment such as suction scrubbers, intercoolers, and aftercoolers, have to be accounted to define compressor design conditions.
11.2 Reciprocating compressors

A reciprocating compressor is a positive displacement machine in which the compressing and displacing element is a piston moving linearly within a cylinder. The reciprocating compressor uses automatic spring-loaded valves that open when the proper differential pressure exists across the valve. Figure 11-1 describes the action of a reciprocating compressor using a theoretical pressure–volume (PV) diagram. In position A, the suction valve is open and gas will flow into the cylinder (from point 1 to point 2 on the PV diagram) until the end of the reverse stroke at point 2, which is the start of compression. At position B, the piston has traveled the full stroke within the cylinder and the cylinder is

![Figure 11-1](https://example.com/figure11-1.png)

**FIGURE 11-1**
Reciprocating compressor compression cycle.
full of gas at suction pressure. Valves remain closed. The piston begins to move to the left, closing the suction valve. In moving from position B to position C, the piston moves toward the cylinder head, reducing the volume of gas with an accompanying rise in pressure. The PV diagram shows compression from point 2 to point 3. The piston continues to move to the end of the stroke (near the cylinder head) until the cylinder pressure is equal to the discharge pressure and the discharge valve opens (just beyond point 3). After the piston reaches point 4, the discharge valve will close, leaving the clearance space filled with gas at discharge pressure (moving from position C to position D). As the piston reverses its travel, the gas remaining within the cylinder expands (from point 4 to point 1) until it equals suction pressure and the piston is again in position A.

The flow to and from reciprocating compressors is subject to significant pressure fluctuations due to the reciprocating compression process. Therefore, pulsation dampeners have to be installed upstream and downstream of the compressor to avoid damages to other equipment. The pressure losses (several percent of the static flow pressure) in these dampeners have to be accounted for in the station design (Botros, 2013).

Reciprocating compressors are widely utilized in the gas processing industries because they are flexible in throughput and discharge pressure range. Reciprocating compressors are classified as either “high speed” or “slow speed.” Typically, high-speed compressors operate at speeds of 900–1200 rpm and slow-speed units at speeds of 200–600 rpm. High-speed units are normally “separable.” That is the compressor frame and driver are separated by a coupling or gearbox. For an “integral” unit, power cylinders are mounted on the same frame as the compressor cylinders and the power pistons are attached to the same drive shaft as the compressor cylinders. Low-speed units are typically integral in design.

### 11.3 Centrifugal compressors

A centrifugal compressor achieves the compression task by converting the mechanical energy from the driver to kinetic energy in the gas, using the forces that moving and stationary airfoils exert on this gas. We want to introduce now the essential components of a centrifugal compressor that accomplish the tasks specified above (Figure 11-2). The gas entering the inlet nozzle of the compressor is guided to the inlet of the impeller. An impeller consists of a number of rotating vanes that impart the mechanical energy to the gas. The gas will leave the impeller with an increased velocity and increased static pressure. In the diffuser, part of the velocity is converted into static pressure. Diffusers can be vaned, vaneless, or volute type. If the compressor has more than one impeller, the gas will be again brought in front of the next impeller through the return channel and the return vanes. If the compressor has only one impeller, or after the diffuser of the last impeller in a multistage compressor, the gas enters the discharge system.
The discharge system can either make use of a volute, which can further convert velocity into static pressure, or a simple cavity that collects the gas before it exits the compressor through the discharge flange.

The rotating part of the compressor consists of all the impellers. It runs on two radial bearings (on all modern compressors, these are hydrodynamic tilt pad bearings), while the axial thrust generated by the impellers is balanced by a balance piston, and the residual force is balanced by a hydrodynamic tilt pad thrust bearing. To keep the gas from escaping at the shaft ends, dry gas seals are used. The entire assembly is contained in a casing (usually barrel type).

A compressor stage is defined as one impeller, with the subsequent diffuser and (if applicable) return channel. A compressor body may hold one or several (up to 8 or 10) stages. A compressor train may consist of one or multiple compressor bodies. It sometimes also includes a gearbox. Pipeline compressors are typically single body trains with one or two stages.

The different working principles cause differences in the operating characteristics of the centrifugal compressors compared to those of the reciprocating unit. Centrifugal compressors are used in a wide variety of applications in
chemical plants, refineries, onshore and offshore gas lift and gas injection applications, gas gathering, and in the transmission and storage of natural gas. Centrifugal compressors can be used for outlet pressures as high as 10,000 psia, thus overlapping with reciprocating compressors over a portion of the flow-rate/pressure domain. Centrifugal compressors are usually either turbine or electric motor driven. Typical operating speeds for centrifugal compressors in gas transmission applications are about 14,000 rpm for 5000 hp units and 8000 rpm for 20,000 horsepower (hp) units.

11.4 Comparison between compressors

Figure 11-3 illustrates the approximate application ranges of reciprocating and centrifugal compressors in terms of actual inlet flow and compressor discharge pressure. Although there is a significant overlap, some of the secondary considerations, such as reliability, availability of maintenance, reputation of vendor,
price, etc., will allow one to choose one of the acceptable compressors. Note should be made that in Figure 11-3, the actual gas flow rate\(^1\) at suction conditions can be calculated using Eqn (11-1)

\[
Q_G = 0.0283 \frac{Z_1 T_1}{p_1} Q_{G,SC}
\]

where \(Q_G\) is an actual cubic feet per minute flow rate of gas, \(T_1\) is the suction temperature in °R, \(p_1\) is suction pressure in psia, \(Z_1\) is the gas compressibility factor at inlet temperature and pressure, dimensionless; and \(Q_{G,SC}\) is the standard volumetric flow rate of gas in MMSCFD.

The advantages of a reciprocating compressor over a centrifugal machine include the following:

- Ideal for low volume flow and high pressure ratios
- High efficiency at high pressure ratios
- Relatively low capital cost in small units (less than 3000 hp)
- Less sensitive to changes in composition and density

The advantages of a centrifugal compressor over a reciprocating machine include the following:

- Ideal for high volume flow and low head
- Simple construction with only one moving part
- High efficiency over normal operating range
- Low maintenance cost and high availability
- Greater volume capacity per unit of plot area
- No vibrations and pulsations generated

11.5 Compressor selection

The design philosophy for choosing a compressor should include the following considerations:

- Good efficiency over a wide range of operating conditions
- Maximum flexibility of configuration
- Low maintenance cost
- Low life-cycle cost
- Acceptable capital cost
- High availability and reliability

\(^1\) While the sizing of the compressor is driven by the actual volumetric flow rate \((Q_G)\), the flow in many applications is often defined as standard flow. Standard flow is volumetric flow at certain, defined conditions of temperature and pressure \((60 °F\) or \(519.7 °R\) and \(14.696 \text{ psia})\) that are usually not the pressures and temperatures of the gas as it enters the compressor.
However, additional requirements and features will depend on each project and
on specific experiences of the pipeline operator. In many cases, the decision
whether to use a reciprocating compressor or a centrifugal compressor, as well as
the type of driver, will be made based on operator strategy, emissions restrictions,
general life-cycle cost assumptions, etc. However, a hydraulic analysis should be
made for each compressor selection to insure the best choice. In fact, compressor
selection can be made for an operating point that will be the most likely or most
frequent operating point of the machine. Selections based on a single operating
point have to be carefully evaluated to provide sufficient speed margin (typically 3–10%) and surge margin to cover other potentially important operating condi-
tions. For reciprocating compressors, operational restrictions, for example, due to
rod load limits, or pulsations, have to be considered. In general, centrifugal
machines have continuous head–flow map coverage, whereas reciprocating
machines often have maps with areas where the machine cannot operate.
A compressor performance map (head vs flow map) can be generated based on
the selection, and is used to evaluate the compressor for other operating condi-
tions, by determining the head and flow required for those other operating condi-
tions. In many applications, multiple operating points are available based on
hydraulic pipeline studies or reservoir studies. Some of these points may be
frequent operating points while some may occur during upset conditions. With
this knowledge, the selection can be optimized for a desired target, such as lowest
fuel consumption.

Selections can also be made based on a “rated” point, which happens to be the
most onerous operating conditions (highest volumetric flow rate, lowest molec-
ular weight, highest head or pressure ratio, highest inlet temperature). In this
situation however, the result may be an oversized machine that does not perform
well at the usual operating conditions.

Once a selection is made, the manufacturer is able to provide parameters such
as efficiency, speed, power requirements, and based on these and the knowledge
of the ambient conditions (prevailing summer and winter temperatures and site
elevation) can size the drivers. At this point, the casing arrangement, the number
of required units based on desirable flexibility, growth scenarios, and sparing
considerations will play an important role in this decision.

11.6 Thermodynamics of gas compression

The task of gas compression is to bring gas from a certain suction pressure to a
higher discharge pressure by means of mechanical work. The actual compression
process is often compared to one of two ideal processes: isothermal and isen-
tropic compression.

Isothermal compression occurs when the temperature is kept constant during
the compression process. It is not adiabatic, because the heat generated in the
compression process has to be removed from the system.
The compression process is isentropic (or adiabatic reversible) if no heat is added to or removed from the gas during compression and the process is frictionless. With these assumptions, the entropy of the gas does not change during the compression process.

The actual or polytropic compression is similar to isentropic compression, but it is not adiabatic. It can be described as an infinite number of isentropic steps; each interrupted by isobaric heat transfer. This heat addition allows the process to yield the same discharge temperature as the actual compression process.

Figure 11-4 shows the PV diagram for three types of compression. As can be seen, when the gas compression process is assumed to be isentropic, the calculated work (horsepower (HP)) is the maximum while under the assumption of isothermal condition; the required work is less than for other types of compression. Therefore, the actual work required to compress a given gas to obtain the same final pressure is between these maximum and minimum values.

### 11.6.1 Basic relations

The design of a compressor begins with the first law of thermodynamics, the conservation of energy. It is written for the steady-state gas flow through the compressor as follows:

\[
\left( h_2 + \frac{u_2^2}{2} + gz_2 \right) - \left( h_1 + \frac{u_1^2}{2} + gz_1 \right) = q_{12} + W_{t,12}
\]

where \( h \) is enthalpy, \( u \) is velocity, \( g \) is gravitational acceleration, \( z \) is elevation coordinate, \( q \) is heat, and \( W_t \) is work done by the compressor on the gas. For gas compressors, the subscripts 1 and 2 refer to the inlet into and the discharge from the compressor.
Neglecting the changes in potential energy (because the contribution due to changes in elevation is not significant for gas compressors) the energy balance equation for adiabatic processes \((q_{12} = 0)\) can be written as follows:

\[
\left( h_2 + \frac{u_2^2}{2} \right) - \left( h_1 + \frac{u_1^2}{2} \right) = W_{t,12}
\] (11-3)

\(W_{t,12}\) is the amount of work\(^2\) that we have to apply to change the enthalpy in the gas. The work \(W_{t,12}\) is related to the required power, \(P\), by multiplying it with the mass flow.

\[
P = \dot{m} W_{t,12}
\] (11-4)

Combining enthalpy and velocity into a total enthalpy \(h_t = h + \frac{u^2}{2}\), power and total enthalpy difference are thus related by

\[
P = \dot{m} (h_{t,2} - h_{t,1})
\] (11-5)

The total enthalpy difference (head) can be expressed as follows:

\[
\Delta h = h(p_2, T_2) - h(p_1, T_1)
\] (11-6)

Use of Eqn (11-6) requires that the inlet pressure and temperature and the outlet pressure and temperature be known to evaluate the change in enthalpy. However, in design work, only the first three of these items are known the outlet temperature cannot be arbitrarily specified. Hence, a compression model is needed.

If we can find a relationship that combines enthalpy with the pressure and temperature of a gas, we have found the necessary tools to describe the gas compression process. For an ideal gas, with constant heat capacity, the relationship between enthalpy, pressures, and temperatures is

\[
\Delta h = C_p (T_2 - T_1)
\] (11-7)

where \(T_1\) is suction temperature, \(T_2\) is discharge temperature, and \(C_p\) is heat capacity at constant pressure.

### 11.6.2 Isentropic model

For an isentropic compression, the discharge temperature is determined by the pressure ratio as

\[
\frac{T_2}{T_1} = \left( \frac{p_2}{p_1} \right)^{\frac{k-1}{k}}
\] (11-8)

\(^2\)Physically, there is no difference between work, head, and change in enthalpy. In systems with consistent units (such as the SI system), work, head and enthalpy difference, have the same unit (e.g., kJ/kg in SI units). Only in inconsistent systems (such as the U.S. customary units), we need to consider that the enthalpy difference (e.g., in BTU/lbm) is related to head and work (e.g., in ft lb/lbm) by the mechanical equivalent of heat (e.g., in ft lb/BTU).
where $k$ (isentropic exponent) is ratio of the heat capacities of gas at constant pressure and temperature ($k = \frac{C_p}{C_v}$), $p_1$ is suction pressure, and $p_2$ is discharge pressure.

Note, the heat capacities of real gases are a function of the pressure and temperature; however, they are functions of temperature only for ideal gases and can be related together with $C_p - C_v = R$, where $R$ is universal gas constant. The isentropic exponent ($k$) for ideal gas mixtures can therefore be determined as

$$k = \frac{\sum y_i C_{Pi}}{\sum y_i C_{Pi} - R} \quad (11-9)$$

where, $C_{Pi}$ is the molar heat capacity of the individual component at constant pressure, and $y_i$ is the molar concentration of the component. Since the temperature of the gas increases as it passes from suction to discharge in the compressor, common practice is to determine the heat capacities of gas at the average of suction and discharge temperatures.

If the gas composition is not known and the gas is made up of pure hydrocarbons (methane, ethane, propane, and butane), the following simple correlation can be used for determining the isentropic exponent of ideal gases (Moshfeghian, 2013):

$$k = \left(1.6 - 0.44\gamma + 0.097\gamma^2\right)\left(1 + 0.0385\gamma - 0.000159T\right) \quad (11-10)$$

where $\gamma$ is the gas specific gravity, and $T$ is gas temperature in °R. This correlation covers gas specific gravities in the range of 0.55–2.0 and temperatures up to 302 °F.

Combining Eqns (11-7) and (11-8), the isentropic head ($\Delta h_s$) for the isentropic compression of an ideal gas can thus be determined as

$$\Delta h_s = C_p T_1 \left[\left(\frac{p_2}{p_1}\right)^\frac{k}{k-1} - 1\right] \quad (11-11)$$

For real gases (where $k$, and $C_p$ in the above equation become functions of temperature and pressure), the enthalpy of a gas, $h$, is calculated in a more complicated way using equations of state.

The isentropic head can also be determined by

$$\Delta h_s = h(p_2, s_1) - h(p_1, T_1) \quad (11-12)$$

where entropy $^3$ of the gas at suction condition ($s_1$) is

$$s_1 = s(p_1, T_1) \quad (11-13)$$

The relationships described above can be easily seen in a Mollier diagram (Figure 11-5).

$^3$ Entropy is a quantitative measure of the disorder in a system.
The performance of a compressor can be assessed by comparing the actual head (which directly relates to the amount of required compression power) with the calculated head for an ideal, isentropic compression. This defines the isentropic efficiency ($\eta_s$) as follows:

$$\eta_s = \frac{\Delta h_s}{\Delta h} \quad (11-14)$$

For ideal gases, the actual head can be calculated from

$$\Delta h = \frac{1}{\eta_s} C_p T_1 \left[ \left( \frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] \quad (11-15)$$

and further, the actual discharge temperature ($T_2$) becomes

$$T_2 = \frac{T_1}{\eta_s} \left[ \left( \frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] + T_1 \quad (11-16)$$

The second law of thermodynamics tells us

$$\dot{m}(s_2 - s_1) = \int_{1}^{2} \frac{dq}{T} + s_{irr} \quad (11-17)$$
where $q$ is heat added to the process, $T$ is temperature of the process and $s_{irr}$ is the entropy of the irreversible process\(^4\).

For adiabatic flows, where no heat enters or leaves ($dq/T = 0$), the change in entropy simply describes the losses generated in the compression process. These losses come from the friction of gas with internal surfaces and the mixing of gas of different energy levels. The isentropic (adiabatic reversible) compression process therefore does not change the entropy of the system.

Equation (11-6) for calculating the actual head implicitly includes the entropy rise $\Delta s$, because

$$
\Delta h = h(p_2, T_2) - h(p_1, T_1) = h(p_2, s_1 + \Delta s) - h(p_1, s_1)
$$

(11-18)

If cooling is applied during the compression process (for example, with intercoolers between two compressors in series), then the increase in entropy is smaller than an uncooled process. A gas cooler therefore allows removing some of the entropy generated by the compression from the system. Therefore, the power requirement will be reduced.

### 11.6.3 Polytropic model

Since polytropic compression is similar to adiabatic compression, we can easily calculate the discharge gas temperature in polytropic compression by substituting polytropic exponent ($n$) for “$k$” in Eqn (11-8).

$$
\frac{T_2}{T_1} = \left(\frac{p_2}{p_1}\right)^{\frac{n-1}{n}}
$$

(11-19)

However, compressor manufacturers do not supply values for $n$; instead they provide a polytropic efficiency. The polytropic efficiency ($\eta_p$) is constant for any infinitesimally small compression step, which then allows us to write

$$
\eta_p = \frac{dT_s}{dT}
$$

(11-20)

where $dT_s$ is elemental temperature rise for isentropic compression, and $dT$ is elemental temperature rise for actual compression.

Integration of Eqn (11-20) and use of Eqn (11-8) yields

$$
\frac{T_2}{T_1} = \left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k\eta_p}}
$$

(11-21)

Comparison of Eqns (11-19) and (11-21) shows

$$
\frac{n - 1}{n} = \frac{k - 1}{k\eta_p}
$$

(11-22)

\(^4\) The second law of thermodynamics states that all natural processes are irreversible.
The polytropic head ($\Delta h_p$) can be calculated from

$$\Delta h_p = Z_{ave}RT_1 \left( \frac{p_2}{p_1} \right) \left[ \left( \frac{p_2}{p_1} \right)^{\frac{n}{n-1}} - 1 \right] \quad (11-23)$$

In the above equation, $Z_{ave}$ is the average compressibility factor for the gas.

Then the actual head of compression is

$$\Delta h = \frac{\Delta h_p}{\eta_p} \quad (11-24)$$

Finally, comparing Eqns (11-14) and (11-24), the isentropic and polytropic efficiencies are related by

$$\eta_s = \eta_p \left[ \left( \frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] \left/ \left[ \left( \frac{p_2}{p_1} \right)^{\frac{n-1}{n}} - 1 \right] \right. \quad (11-25)$$

For compressor designers, the polytropic efficiency has an important advantage. If a compressor has five stages, and each stage has the same isentropic efficiency $\eta_s$, then the overall isentropic compressor efficiency will be lower than $\eta_s$. If, for the same example, we assume that each stage has the same polytropic efficiency $\eta_p$, then the polytropic efficiency of the entire machine is also $\eta_p$. As far as power calculations are concerned, either the approach using a polytropic head and efficiency, or using isentropic head and efficiency will lead to the same result:

$$P = \dot{m} \cdot \Delta h = \dot{m} \left( \frac{\Delta h_s}{\eta_s} \right) = \dot{m} \left( \frac{\Delta h_p}{\eta_p} \right) \quad (11-26)$$

### 11.6.4 Real gas behavior

Understanding gas compression requires an understanding of the relationship between pressure, temperature, and density of a gas. An ideal gas exhibits the following behavior:

$$\frac{P}{\rho} = RT \quad (11-27)$$

where $\rho$ is density of gas, and $R$ is gas constant (as long as the gas composition remains unchanged). Any gas at very low pressures ($p \to 0$) and high temperatures (ambient and higher) can be described by this equation.

For the elevated pressures we see in natural gas compression, Eqn (11-27) becomes inaccurate, and an additional variable, the gas compressibility factor ($Z$), has to be added:

$$\frac{P}{\rho} = ZRT \quad (11-28)$$
Consequently, the gas compressibility factor itself is a function of pressure, temperature, and gas composition.

A similar situation arises when the enthalpy has to be calculated. For an ideal gas, we find

\[ \Delta h = \int_{T_1}^{T_2} C_p(T) \, dT \]  \hspace{1cm} (11-29)

where \( C_p \) is only a function of temperature. This is a better approximation of the reality than the assumption of an ideal gas used in Eqn (11-7).

In a real gas, we get additional terms for the deviation between real gas behavior and ideal gas behavior (Poling et al., 2001)

\[ \Delta h = \frac{T_1}{T_2} \left( h^0 - h(p_1) \right)_{T_1} + \int_{T_1}^{T_2} C_p dT - \frac{T_1}{T_2} \left( h^0 - h(p_2) \right)_{T_2} \]  \hspace{1cm} (11-30)

The terms \( (h^0 - h(p_1))_{T_1} \) and \( (h^0 - h(p_2))_{T_2} \) are called departure functions, because they describe the deviation of the real gas behavior from the ideal gas behavior. They relate the enthalpy at some pressure and temperature to a reference state at low pressure, but at the same temperature. The departure functions can be calculated solely from an equation of state, while the term \( \int C_p dT \) is evaluated in the ideal gas state. For gas compression applications, the most frequently used equations of state are Redlich–Kwong, Soave–Redlich–Kwong, Benedict–Webb–Rubin, Benedict–Webb–Rubin–Starling and Lee–Kesler–Ploecker. Kumar et al. (1999) have compared these equations of state regarding their accuracy for compression applications. In general, all of these equations provide accurate results for typical applications in pipelines, i.e., for gases with a high methane content, and at pressures below 3500 psia.

Modern compressors often operate at pressures higher than the critical pressure of the gas. This is often referred to as “dense phase.” All procedures discussed above are not affected by this.

11.7 Compression ratio

Compression ratio (CR) is the ratio of absolute discharge pressure to the absolute suction pressure. Mathematically

\[ CR = \frac{P_2}{P_1} \]  \hspace{1cm} (11-31)

By definition, the CR is always greater than one. If there are “n” stages of compression and the CR is equal on each stage, then the CR per stage is given by

\[ CR_{\text{stage}} = \left( \frac{P_2}{P_1} \right)^{1/n} \]  \hspace{1cm} (11-32)
If the CR is not equal on each stage, then Eqn (11-31) should be applied to each stage.

The term CR can be applied to a single stage of compression and multistage compression. When applied to a single compressor or a single stage of compression, it is defined as the stage or unit CR; when applied to a multistage compressor it is defined as the overall CR. The CR for typical gas pipeline compressors is rather low (usually below 2). Low pressure ratios can be covered in a single compression stage for reciprocating compressor, and in a single body (with one or two impellers) in a centrifugal compressor.

While the pressure ratio is a valuable indicator for reciprocating compressors, the pressure ratio that a given centrifugal compressor can achieve depends primarily on gas composition and gas temperature. The centrifugal compressor is better characterized by its capability to achieve a certain amount of head (and a certain amount of head per stage). For natural gas (with specific gravity in the range of 0.58–0.70), a single centrifugal stage can provide a pressure ratio of 1.4. The same stage would yield a pressure ratio of about 1.6 if it would compress air. The pressure ratio per stage is usually lower than the values above for multistage machines. A multistage centrifugal compressor (i.e., a machine with multiple stages) can achieve, with natural gas, a pressure ratio of about 4–5, which equates to about 70,000–90,000 ft lb/lb of head. The temperature rise resulting from this amount of head is usually a limiting factor.

For reciprocating compressors, the pressure ratio per compressor is usually limited by mechanical considerations (rod load), and temperature limitations. Reciprocating compressors can achieve cylinder pressure ratios of 3–6. The actual flange-to-flange ratio will be lower (due to the losses in valves and bottles). For lighter gases (i.e., natural gas), the temperature limit will often limit the pressure ratio before the mechanical limits do. Centrifugal compressors are also limited by mechanical considerations (rotor dynamics, maximum speed) and temperature limits. Whenever any limitation is involved, it becomes necessary to use multiple compression stages in series, and intercooling. Furthermore, multistage compression may be required from a purely optimization standpoint. For example, with increasing CR, compression efficiency decreases and mechanical stress and temperature problems become more severe. For reference, if we assume natural gas at 100 °F suction temperature, a pressure ratio of 3 will—depending on the compressor efficiency—lead to about 275 °F discharge temperature.

For pressure ratios higher than 3, it may be advantageous to install intercoolers between the compressors. Intercoolers are generally used between the stages to reduce the power requirements as well as to lower the gas temperature that may become undesirably high. Theoretically minimum power requirement is obtained with ideal intercooling and no pressure loss between stages by making the ratio of compression the same in all stages. However,

---

5 After the cooling, liquids may form. These liquids are removed in interstage scrubbers or knockout drums.
intercoolers invariably cause pressure losses (typically between 5 and 15 psi), which is a function of the cooler design. For preliminary design considerations, a value of 10 psi can be used (coolers—especially gas-to-air coolers for lower pressure drop tend to become expensive).

Note that an actual compressor with an infinite number of compression stages and intercoolers would approach isothermal conditions (where the power requirement of compression cycle is the absolutely minimum power necessary to compress the gas) if the gas were cooled to the initial temperature in the intercoolers.

Interstage cooling is usually achieved using gas-to-air coolers. The gas outlet temperature depends on the ambient air temperature. The intercooler exit temperature is determined by the cooling media. If ambient air is used, the cooler exit temperature, and thus the suction temperature to the second stage will be about 20–30 °F above ambient dry bulb temperature. Water coolers can achieve exit temperatures about 20 °F above the water supply temperature, but require a constant supply of cooling water. Cooling towers can provide water supply temperatures of about wet bulb temperature plus 25 °F.

For applications where the compressor discharge temperature is above some temperature limit of downstream equipment (a typical example are pipe coatings that limit gas temperatures to about 125–140 °F), or has to be limited for other reasons (for example, to not disrupt or damage the permafrost), an aftercooler has to be installed.

If we trace the compression process in a Mollier diagram (e.g., Figure 11-5), the compression process will always move away from the two-phase dome of the gas. However, if the gas is cooled after compression, the gas can enter the two-phase region, and liquids can form. It is important that the liquids are removed (for example, by a separator or coalescer) before the gas is fed into the next compressor downstream.

11.8 Compressor design

Compressor design involves several steps including selection of the correct type of compressor as well as the number of compression stages required. Depending on the capacity, there is also a need to determine the HP required to compress the gas.

11.8.1 Determining number of compression stages

For reciprocating compressors, the number of compression stages is determined from the overall CR as follows (Arnold and Stewart, 1999):

\[ \text{Number of Stages} = \left\lfloor \frac{\text{Overall CR}}{\text{Stage CR}} \right\rfloor \]

A stage for a reciprocating compressor is usually a cylinder. For centrifugal compressor, a stage is one impeller with its inlet and diffusor. A compressor body can contain multiple stages (up to 10).
1. Calculate the overall CR. If the CR is under 4, consider using one stage. If it is not, select an initial number of stages so that \( CR < 4 \). For initial calculations it can be assumed that the CR per stage is equal for each stage. CRs of 6 can be achieved for low pressure applications; however, at the cost of higher mechanical stress levels and lower volumetric efficiency.

2. Calculate the discharge gas temperature for the first stage. If the discharge temperature is too high (more than 300 °F), we will either have to increase the number of stages, or reduce the suction temperature through precooling. It is recommended that the compressors be sized so that the discharge temperatures for all stages of compression be below 300 °F. It is also suggested that the aerial gas coolers be designed to have a maximum of 20 °F approach to ambient, provided the design reduces the suction temperature for the second stage, conserving HP and reducing power demand. If the suction gas temperature to each stage cannot be decreased, increase the number of stages by one and recalculate the discharge temperature.

For centrifugal compressors, the number of compressor bodies is determined in similar fashion:

1. Calculate the overall isentropic head using Eqn (11-11) or Eqn (11-12). Typically a centrifugal compressor casing can provide about 70,000 ft lb/lb\(^7\) of head, due to limitations in speed, number of impellers, and discharge temperature limits (although 100,000 ft lb/lb might be possible in some cases).

2. If the overall head exceeds 70,000 ft lb/lb, take the square root of the overall pressure ratio (using Eqn (11-32)), and calculate the isentropic head for the new pressure ratio. The compression train will consist of two compressors.

3. If the required isentropic head still exceeds 70,000 ft lb/lb, take the cube root of the overall pressure ratio and calculate the isentropic head for the new pressure ratio. The compression train will consist of three compressors. More than three compressors per train are usually not practical, so if the required head per compressor still exceeds 70,000 ft lb/lb, a second train may have to be considered.

4. For cases 2 and 3, i.e., multiple compressors per train, the head requirement for the second or third body must be recalculated considering about 20 °F approach to ambient for gas-to-air coolers, and about 10-psi pressure drop. Unless the gas has a very high inlet temperature, or the compressor is low, the 70,000 ft lb/lb limit will keep discharge temperatures at an acceptable level below 350 °F. In any case, the discharge temperature should be checked, using Eqn (11-16).

\(^7\) 70,000 ft lb/lb is equivalent to a pressure ratio of about 4 for pipeline quality natural gas.
The 300 °F temperature limit is used for reciprocating compressors because the packing life gets shortened above about 250 °F and the lube oil, being directly involved in the compression process will degrade faster at higher temperatures. The 350 °F temperature limit pertains to centrifugals, and is really a limit for the seals (although special seals can go to 400–450 °F) or the pressure rating of casings and flanges. Since the lube oil in a centrifugal compressor does not come into direct contact with the process gas, lube oil degradation is not a factor.

If oxygen is present in the process gas in the amount that it can support combustion (i.e., the gas-to-oxygen ratio is above the lower explosive limit), much lower gas temperatures than mentioned above are required. In reciprocating machines, oil-free compression may be required (no lube oil can come into contact with the process gas). This requires special piston designs that can run dry. Also, special precaution has to be taken to avoid hot spots generated by local friction.

11.8.2 Compression power calculation

Once we have an idea about the type of compressor we will select, we also need to know the power requirements so that an appropriate prime mover can be designed for the job. After the gas horsepower (GHP) has been determined by either method, HP losses due to friction in bearings, seals, and speed increasing gears must be added. Bearings and seal losses can be estimated from Scheel’s equation (GPSA, 2004). For reciprocating compressors, the mechanical and internal friction losses can range from about 3% to 8% of the design GHP. For centrifugal compressors a good estimate is to use 1–2% of the design GHP as mechanical loss.

To calculate brake horsepower (BHP) the following equation can be used:

\[
BHP = GHP + \text{Mechanical losses}
\]  \hspace{1cm} (11-33)

The detailed calculation of BHP depends upon the choice of type of compressor and number of stages. The BHP per stage can be determined from Eqn (11-34) (GPSA, 2004):

\[
BHP = 0.0854 \cdot Z_{\text{ave}} \left[ \frac{(Q_{G,SC})(T_1)}{E \cdot \eta} \right] \left[ \frac{k}{k - 1} \right] \left[ \left( \frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] \]  \hspace{1cm} (11-34)

where BHP is brake horsepower per stage; \( Z_{\text{ave}} \) is average compressibility factor; \( Q_{G,SC} \) is standard volumetric flow rate of gas, MMSCFD; \( T_1 \) is suction temperature, °R; \( p_1 \) and \( p_2 \) are pressure at suction and discharge flanges, respectively, psia; \( E \) is parasitic efficiency (for high-speed reciprocating units—use 0.72–0.82; for low-speed reciprocating units—use 0.72–0.85; and for centrifugal units—use 0.99); and \( \eta \) is compression efficiency (1.0 for reciprocating and 0.80–0.87 for centrifugal units).
In Eqn (11-34), the parasitic efficiency (E) accounts for mechanical losses, and the pressure losses incurred in the valves and pulsation dampeners of reciprocating compressors (the lower efficiencies are usually associated with low pressure ratio applications typical for pipeline compression). Hence, suction and discharge pressures may have to be adjusted for the pressure losses incurred in the pulsation dampeners for reciprocation compressors. The compression efficiency accounts for the actual compression process. For centrifugal compressors, the lower efficiency is usually associated with pressure ratios of 3 and higher. Very low flow compressors (below 1000 acfm) may have lower efficiencies.

The total HP for the compressor is the sum of the HP required for each stage. Reciprocating compressors require an allowance for interstage pressure losses. It can be assumed that there is a 3% loss of pressure in going through the cooler, scrubbers, piping, etc., between the actual discharge of the cylinder and the actual suction of the next cylinder. For a centrifugal compressor, any losses incurred between the stages are already included in the stage efficiency. However, the exit temperature from the previous stage becomes the inlet temperature in the next stage. If multiple bodies are used, the losses for coolers and piping have to be included as described above.

11.9 Compressor control

To a large extent, the compressor operating point will be the result of the pressure conditions imposed by the system. However, the pressures imposed by the system may in turn be dependent on the flow. Only if the conditions fall outside the operating limits of the compressor (e.g., frame loads, discharge temperature, available driver power, surge, choke, speed), control mechanisms have to be in place. On the other hand, the compressor output may have to be controlled to match the system demand. The type of application often determines the system behavior. In a pipeline application, suction and discharge pressure are connected with the flow by the fact that the more flow is pushed through a pipeline; the more pressure ratio is required at the compressor station to compensate for the pipeline pressure losses. In process-related applications, the suction pressure may be fixed by a back pressure controlled production separator. In boost applications, the discharge pressure is determined by the pressure level of the pipeline the compressor feeds into, while the suction pressure is fixed by the process. In oil and gas field applications, the suction pressure may be depending on the flow, because the more gas is moved out of the gas reservoir, the lower the suction

---

8 Many calculation procedures for reciprocating compressors use numbers for E that are higher than the ones referenced here. These calculations require however, that the flange-to-flange pressure ratio (which is used in Eqn (11-33)) is increased by the pressure losses in the compressor suction and discharge valves, and pulsation dampeners. These pressure losses are significant, especially for low head high flow applications.
pressure has to be. The operation may require constant flow despite changes in suction or discharge pressure. Compressor flow, pressure, or speed may have to be controlled. The type of control also depends on the compressor driver. Both reciprocating compressors and centrifugal compressors can be controlled by suction throttling or recirculating of gas. However, either method is very inefficient for process control (but may be used to protect the compressor), because the reduction in flow or head is not accompanied by a significant reduction in the power requirement.

11.9.1 Reciprocating compressors

The following control mechanisms may be used to control the capacity of reciprocating compressors: suction pressure, variation of clearance, speed, valve unloading, and recycle. Reciprocating compressors tend to have a rather steep head versus flow characteristic. This means that changes in pressure ratio have a very small effect on the actual flow through the machine.

Reciprocating compressors generate flow pulsations in the suction and discharge lines that have to be controlled to prevent over and underloading of the compressors, avoid vibration problems in the piping or other machinery at the station, and to provide a smooth flow of gas. The flow pulsations can be greatly reduced by properly sized pulsation bottles or pulsation dampeners in the suction and discharge lines.

Controlling the flow through the compressor can be accomplished by varying the operating speed of the compressor. This method can be used if the compressor is driven by an internal combustion engine, or a variable speed electric motor. Especially internal combustion engines, but also variable speed electric motors, produce less power if they operate at a speed different from their optimum speed. Internal combustion engines allow for speed control in the range of 70–100% of maximum speed. Since reciprocating compressors generate pulsations in the piping system, pulsation dampeners have to be installed. Since these have to be optimized based on the frequencies of the pulsations, which are speed dependent, the allowable speed range may be limited.

If the driver is a constant speed electric motor, the capacity control consists of either inlet valve unloaders or clearance unloaders. Inlet valve unloaders can hold the inlet valve into the compressor open and thereby prevent compression. Clearance unloaders consist of pockets which are opened when unloading is desired. The gas is compressed into them at the compression stroke and expands back into the cylinder on the return stroke, thus reducing the intake of additional gas, and subsequently, the compressor capacity. Additional flexibility is achieved by using several steps of clearance control and combinations of clearance control and inlet valve control. Figure 11-6 shows the control characteristic of such a compressor.
11.9.2 Centrifugal compressors

As with reciprocating compressors, the compressor output must be controlled to match the system demand. The operation may require constant flow despite changes in suction or discharge pressure. Compressor flow, pressure or speed may have to be controlled. The type of control also depends on the compressor driver. Centrifugal compressors tend to have a rather flat head versus flow characteristic. This means that changes in pressure ratio have significant effect on the actual flow through the machine.

Compressor control is usually accomplished by speed control, variable guide vanes, suction throttling, and recycling of gas. Only in rare cases, adjustable diffuser vanes are used. To protect the compressor from surge, recycling is used. Controlling the flow through the compressor can be accomplished by varying the operating speed of the compressor. This is the preferred method of controlling centrifugal compressors. Two-shaft gas turbines and variable speed electric motors allow for speed variations over a wide range (usually from 50% to 100% of maximum speed or more). Virtually any centrifugal compressor installed in the past 15 years in pipeline service is driven by variable speed drivers, usually a two-shaft gas turbine or a variable speed electric motor. Older installations and installations in other than pipeline services sometimes use single-shaft gas turbines (which allow a speed variation from about 90% to 100% speed) and constant speed electric motors. In these installations, suction throttling or variable inlet guide vanes are used to provide means of control.

**FIGURE 11-6**
Control characteristic of a reciprocating compressor with constant speed driver and pockets (GPSA, 2004).

### 11.9.2 Centrifugal compressors

As with reciprocating compressors, the compressor output must be controlled to match the system demand. The operation may require constant flow despite changes in suction or discharge pressure. Compressor flow, pressure or speed may have to be controlled. The type of control also depends on the compressor driver. Centrifugal compressors tend to have a rather flat head versus flow characteristic. This means that changes in pressure ratio have significant effect on the actual flow through the machine.

Compressor control is usually accomplished by speed control, variable guide vanes, suction throttling, and recycling of gas. Only in rare cases, adjustable diffuser vanes are used. To protect the compressor from surge, recycling is used. Controlling the flow through the compressor can be accomplished by varying the operating speed of the compressor. This is the preferred method of controlling centrifugal compressors. Two-shaft gas turbines and variable speed electric motors allow for speed variations over a wide range (usually from 50% to 100% of maximum speed or more). Virtually any centrifugal compressor installed in the past 15 years in pipeline service is driven by variable speed drivers, usually a two-shaft gas turbine or a variable speed electric motor. Older installations and installations in other than pipeline services sometimes use single-shaft gas turbines (which allow a speed variation from about 90% to 100% speed) and constant speed electric motors. In these installations, suction throttling or variable inlet guide vanes are used to provide means of control.
The operating envelope of a centrifugal compressor (typically shown in Figure 11-7) is limited by the maximum allowable speed, the minimum allowable speed, the minimum flow (surge flow), and the maximum flow (choke or stonewall). Another limiting factor may be the available driver power. Only the minimum flow requires special attention, because it is defined by an aerodynamic stability limit of the compressor. Operating the compressor at lower flows than defined by the surge line will lead to surge. Surge, which is the flow reversal within the compressor, accompanied by high-fluctuating load on the compressor bearings, has to be avoided to protect the compressor. Modern control systems can detect this situation and shut the machine down, or prevent it entirely by automatically opening a recycle valve. For this reason, virtually all modern compressor installations use a recycle line (Figure 11-8) with a control valve that allows the flow to increase through the compressor if it comes near the stability limit. Modern control systems constantly monitor the operating point of the compressor in relation to its surge line, and automatically open or close the recycle valve if necessary. The control system is designed to compare the measured operating point of the compressor with the position of the surge line (Figure 11-7). To that end, flow, suction pressure, discharge pressure and suction temperature, as well as compressor speed have to be measured (Kurz et al., 2013).
Typical control systems use suction and discharge pressures and temperatures, together with the flow through the compressor to calculate the relative distance (“turndown”) of the present operating point to the predicted or measured surge line of the compressor. The turndown is defined by Eqn (11-35).

\[
\text{Turndown} = \left. \frac{Q_{\text{op}} - Q_{\text{surge}}}{Q_{\text{op}}} \right|_{H_s = \text{const}}
\]

Equation (11-35)

If the turndown reaches a preset value (often 10% turndown), the antisurge valve starts to open, thereby reducing the pressure ratio of the compressor and increasing the flow through the compressor. The situation is complicated by the fact that the surge valve also has to be capable of precisely controlling flow. Additionally, some manufacturers place limits on how far into choke (or overload) they allow their compressors to operate.

As part of the control system, the antisurge system has to fulfill three tasks:

- Task 1: Allow the starting and stopping of the compressor
- Task 2: Keep the compressor out of surge during changes in process
- Task 3: Keep the compressor out of surge during emergency shutdowns (ESD)

Putting it differently, the surge control system has to protect the compressor and the process. Therefore, it has to do its tasks without causing fluctuations or

FIGURE 11-8
Antisurge and recycle system.
disruptions of the process. Task 2 requires a valve that can be precisely controlled, so that neither opening the valve nor closing the valve upsets the process. In a well-designed system, the compressor can operate from maximum flow all the way down to no station flow at all. This also requires a well-designed control algorithm.

The design of the antisurge and recycle system also impacts the start-up of the station (Task 1). Particular attention has to be given to the capability to start-up the station without having to abort the start due to conditions where allowable operating conditions are exceeded. Problems may arise from the fact that the compressor may spend a certain amount of time recycling gas until sufficient discharge pressure is produced to open the discharge check valve (Figure 11-8), and gas is flowing into the pipeline.

Virtually all of the mechanical energy absorbed by the compressor is converted into heat in the discharged gas. In an uncooled recycle system (Figure 11-9), this heat is recycled into the compressor suction and then more energy added to it. Low pressure ratio compressors often do not require aftercoolers. There are three primary strategies that can be employed to avoid overheating the uncooled compressor during start-up: accelerate quickly, delay hot gas reentering the compressor, and throttled recycle.

**FIGURE 11-9**
Recycle system layout for centrifugal compressors: cooled (top) and hot (bottom) recycle lines (Courtesy of Solar Turbines Inc.).
If aftercoolers are used (Figure 11-9), the heat from recycling can be removed effectively. However, a cooler in the recycle loop increases the volume of the system, and, as we will see, impacts the dynamic behavior of the system, for example, during ESD.

A very critical situation arises upon ESD, Task 3. Here, the fuel supply to the gas turbine driver, or the electric current to an electric motor drive, is cut off instantly, thus eliminating the power to the driven compressor. The inertia of compressor, coupling and power turbine or motor balance the compressor absorbed power, causing a rapid deceleration. Because the head-making capability of the compressor is reduced by the square of its running speed, while the pressure ratio across the machine is imposed by the upstream and downstream piping system, the compressor will surge if the surge valve cannot provide fast relief of the pressure, and the discharge check valve closes fast enough. The deceleration of the compressor as a result of inertia and dissipation are decisive factors. The speed at which the pressure can be relieved not only depends on the reaction time of the valve, but also on the time constants imposed by the piping system. The transient behavior of the piping system depends largely on the volumes of gas enclosed by the various components of the piping system, which may include, besides the piping itself, various scrubbers, knockout drums, and coolers. With the initiation of a shutdown, the compressor can be expected to decelerate by approximately 30% in the first second. With a 30% loss in speed, the head the compressor can develop will drop by approximately 50%. The recycle control valve must, therefore, reduce the pressure ratio across the compressor by one-half in that first second to avoid surge.

Surge control valves must be sized to meet two diverse objectives. During steady-state recycling, the required capacity of the recycle valve can be directly derived from the compressor map (Figure 11-10): the smaller the valve, the smoother the control. During transient conditions, the required capacity increases due to the volumes on either side of the compressor. Therefore, to avoid surge during a shutdown, the bigger the valve, the better. To facilitate both smooth throttling at partial recycle and the need to reduce the pressure differential across the compressor quickly during a shutdown, control valves with an equal percentage characteristic are recommended. With an equal percentage characteristic, the more the valve is opened, the greater the increase in flow for the same travel. We recommend two types of valves for surge control: globe valves and noise-attenuating ball valves.

Surge control valves are primarily sized to fit the compressor (Task 2). In other words, the valve characteristic, that is the flow through the valve for a given pressure ratio over the valve, and a given valve position, has to be matched with

\[ \text{flow} = \text{characteristic} \times \text{pressure ratio} \]

9 Some installations maintain fuel flow to the turbine for 1–2 s while the recycle valve opens. However, this can generate a safety hazard.
the compressor map (Figure 11-10). To handle transient conditions, the required capacity must be greater to allow for the volumes on either side of the compressor.

In some cases, a solution using a single valve will not allow to combine the objectives of a smooth process control and the protection of the compressor during ESD. In this case, it may be necessary to separate the tasks by installing a valve to handle ESD (“hot bypass valve”) and a separate valve for process control (“recycle valve”). The recycle valve ideally would include the aftercooler (where available) into the recycle loop, therefore allowing extended recycle without the risk of exceeding allowable gas temperatures. The hot bypass line would seek the shortest connection between compressor discharge and compressor suction. This valve is only activated for ESD. Other options involve two valves in parallel. Above recommendations pertain to single compressor bodies. Compressor trains with multiple compressor bodies, or compressor sections, involving intercooling or multiple services, require a more detailed analysis.

While the valve sizing for Task 2 essentially requires to match the valve with the steady-state compressor map, the evaluation of the system behavior in an ESD (Task 3) requires an assessment of the dynamic behavior of the system. While it is possible to perform a computational fluid dynamics simulation for the compressor with all associated pipes, valves, and coolers, simpler methods are available.

**FIGURE 11-10**

Matching of valve and compressor. The valve characteristic for a number of opening positions (60%, 70%, and 100%) is superimposed to the compressor performance map (Kurz et al., 2013).
11.10 Compressor performance maps
11.10.1 Reciprocating compressors

Figure 11-11 shows some typical performance maps for reciprocating compressors. Since the operating limitations of a reciprocating compressor are often defined by mechanical limits (especially maximum rod load), and the pressure ratio of the machine is very insensitive to changes in suction conditions and gas composition, we usually find maps depicting suction and discharge pressures and actual flow. Maps account for the effect of opening or closing pockets and for variations in speed.

11.10.2 Centrifugal compressors

For a centrifugal compressor, the isentropic or polytropic head (rather than the pressure ratio) is relatively invariant with the change in suction conditions and gas composition. As with the reciprocating compressor, the flow that determines the operating point is the actual flow as opposed to mass flow or standard flow. Head versus actual flow maps are therefore the usual way to describe the operating range of a centrifugal compressor. These maps change very little even if the inlet conditions or the gas composition changes. They depict the effect of changing the operating speed, and define the operating limits of the compressor such as surge limit, maximum and minimum speed, and maximum flow at choke conditions. Every set of operating conditions, given as suction pressure, discharge pressure, suction temperature, flow, and gas composition, can be converted into isentropic head and actual flow using the relationships described before. Once the operating point is located on a head–flow map, the efficiency of the compressor, the required operating speed as well as the surge margin can be determined.

FIGURE 11-11
Typical maps for a speed-controlled reciprocating compressor for two different load steps (Courtesy of Solar Turbines Inc.).
**11.11 Example for operating a compressor in a pipeline system**

To illustrate how the considerations regarding operating characteristics, system behavior, and control work together, we can look into the operation of a typical compressor station in a pipeline (Kurz et al., 2010).

For a situation where a compressor operates in a system with pipe of the length $L_u$ upstream and a pipe of the length $L_d$ downstream, and further where the pressure at the beginning of the upstream pipe $p_u$ and the end of the downstream pipe $p_e$ are known and constant, we have a simple model of a compressor station operating in a pipeline system (see Figure 11-12).

For a given constant flow capacity $Q_{std}$, the pipeline will then impose a pressure $p_s$ at the suction and $p_d$ at the discharge side of the compressor. For a given pipeline, the head ($H_s$)–flow ($Q$) relationship at the compressor station can be approximated by

$$H_s = C_p T_s \left[ \left( \frac{1}{1 - \frac{C_3 + C_4 Q}{p_d}} \right)^{\frac{k-1}{k}} - 1 \right]$$  \hspace{1cm} (11-36)

where $C_3$ and $C_4$ are constants (for a given pipeline geometry) describing the pressure at either ends of the pipeline, and the friction losses, respectively.

Among other issues, this means that for a compressor station within a pipeline system, the head for a required flow is prescribed by the pipeline system (Figure 11-13). In particular, this characteristic requires the capability for the compressors to allow a reduction in head with reduced flow, and vice versa, in a prescribed fashion. The pipeline will therefore not require a change in flow at constant head (or pressure ratio).

![Figure 11-12](https://via.placeholder.com/150)

*Figure 11-12*

Conceptual model of a pipeline segment.
In transient situations (for example, during line packing), the operating conditions follow initially a constant power distribution, i.e., the head–flow relationship follows Eqn (11-37) and will asymptotically approach the steady-state relationship. This is universally valid for speed-controlled centrifugal compressors. For reciprocating compressors, there might be additional constraints due to rod load and rod load reversal concerns, limitations in clearance control, and others.

\[
P = \frac{\dot{m} \cdot H_s}{\eta_s} = \text{const}
\]

\[
H_s = \frac{\eta_s \cdot \text{const}}{\rho} \frac{1}{Q}
\]  

(11-37)

Based on the requirements above, the compressor output must be controlled to match the system demand. This system demand is characterized by a strong relationship between system flow and system head or pressure ratio (Figure 11-13). Given the large variations in operating conditions experienced by pipeline compressors, an important question is how to adjust the compressor to the varying conditions, and, in particular, how does this influence the efficiency.

Centrifugal compressors tend to have rather flat head versus flow characteristic. This means that changes in pressure ratio have a significant effect on the actual flow through the machine. For a centrifugal compressor operating at a
constant speed, the head or pressure ratio is reduced with increasing flow. Controlling the flow through the compressor can be accomplished by varying the operating speed of the compressor. This is the preferred method of controlling centrifugal compressors. Two-shaft gas turbines and variable speed electric motors allow for speed variations over a wide range (usually from 40% or 50% to 100% of maximum speed or more). It should be noted that the controlled value is usually not speed, but the speed is indirectly the result of balancing the power generated by the power turbine (which is controlled by the fuel flow into the gas turbine) and the absorbed power of the compressor.

Assuming the pipeline characteristic derived in Eqn (11-36), the compressor impellers will be selected to operate at or near their best efficiency for the entire range of head and flow conditions imposed by the pipeline. This is possible with a speed-controlled compressor, because the best efficiency points of a compressor are connected by a relationship that requires approximation (fan law).

\[
\frac{H_s}{N^2} = C_5 \quad \frac{Q}{N} = C_6
\]

\[
H_s = \frac{Q^2}{N} \cdot \frac{C_5}{C_6}
\]

(11-38)

For operating points that meet the above relationship, the absorbed gas power \( P_g \) is (due to the fact that the efficiency stays approximately constant):

\[
P_g = C_7 \cdot H_s \cdot Q = C_7 \cdot \frac{C_5}{C_6} \cdot Q^3 = C_5 \cdot C_6 \cdot C_7 \cdot N^3
\]

(11-39)

As it is, this power–speed relationship allows the power turbine to operate at, or very close to its optimum speed for the entire range. The typical operating scenarios in pipelines therefore allow the compressor and the power turbine to operate at their best efficiency for most of the time. The gas producer of the gas turbine will, however, lose some thermal efficiency when operated in part load.

Figure 11-14 shows a typical example based on an existing pipeline, where pipeline-operating points for different flow requirements are plotted into the performance map of the speed-controlled centrifugal compressor used in the compressor station.

Reciprocating compressors will automatically comply with the system pressure ratio demands, as long as no mechanical limits (rod load, power) are exceeded. Changes in system suction or discharge pressure will simply cause the valves to open earlier or later. The head is lowered automatically because the valves see lower pipeline pressures on the discharge side and/or higher pipeline pressures on the suction side. Therefore, without additional measures, the flow would stay roughly the same—except for the impact of changed volumetric efficiency which would increase, thus increasing the flow with reduced pressure ratio. The control challenge lies in the adjustment of the flow to the system demands. Without additional adjustments, the flow throughput of the compressor
changes very little with changed pressure ratio. Historically, pipelines installed many small compressors and adjusted flow rate by changing the number of machines activated. This capacity and load could be fine-tuned by speed or by a number of small adjustments (load steps) made in the cylinder clearance of a single unit. As compressors have grown, the burden for capacity control has shifted to the individual compressors.

Load control is a critical component to compressor operation. From a pipeline operation perspective, variation in station flow is required to meet pipeline delivery commitments, as well as implement company strategies for optimal
operation (i.e., line packing, load anticipation). From a unit perspective, load control involves reducing unit flow (through unloaders or speed) to operate as close as possible to the design torque limit without overloading the compressor or driver. Critical limits on any load map curve are rod load limits and HP/torque limits for any given station suction and discharge pressure. Gas control generally will establish the units within a station that must be operated to achieve pipeline flow targets. Local unit control will establish load step or speed requirements to limit rod loads or achieve torque control.

The common methods of changing flow rate are to change speed, change clearance, or deactivate a cylinder-end (hold the suction valve open). Another method is an infinite-step unloader, which delays suction valve closure to reduce volumetric efficiency. Further, part of the flow can be recycled or the suction pressure can be throttled, thus reducing the mass flow while keeping the volumetric flow into the compressor approximately constant.

Control strategies for compressors should allow automation, and be adjusted easily during the operation of the compressor. In particular, strategies that require design modifications to the compressor (i.e., rewheeling of a centrifugal compressor, changing cylinder bore, or adding fixed clearances for a reciprocating compressor) are not considered here. It should be noted that with reciprocating compressors, a key control requirement is to not overload the driver or to exceed mechanical limits.

As mentioned above, we are describing the situation for a pipeline operating at or near steady-state conditions. For conditions where the compressor operates at low load in steady state, and the power supply is increased instantly, the compressor operating point will move to an increased flow, without initially seeing a change in pressure ratio (because the gas volume in the pipeline has to be increased first). Then the pressure ratio dictated by the pipeline will cause the compressor operating point to drift back to the steady-state line, following a line of more or less constant absorbed power (see Eqn (11-37)).

References
Botros, K., October 6–9, 2013. Predicting the power loss in reciprocating compressor manifolds. Paper Presented at the GMRC Gas Machinery Conference, Albuquerque, NM, USA.
Kurz, R., White, R.C., Mokhatab, S., June 2013. Surge Prevention in centrifugal compressor systems. Pipeline Gas J. 240 (6), 77, 78, 86.
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12.1 Introduction
Natural gas continues to play a great role as a worldwide energy supply. In fact, major projects are being planned to move massive amounts of high-pressure sales gas from processing plants to distribution systems and large industrial users through large diameter buried pipelines. These pipelines utilize a series of compressor stations along the pipeline to move the gas over long distances. In addition, gas coolers are used downstream of the compressor stations to maintain a specified temperature of the compressed gas for pipeline pressure drop reduction and to protect gas pipeline internal and external coating against deterioration due to high temperatures. This chapter covers all the important concepts of sales gas transmission from a fundamental perspective.

12.2 Gas flow fundamentals
Optimum design of a gas transmission pipeline requires accurate methods for predicting pressure drop for a given flow rate or predicting flow rate for a specified pressure drop in conjunction with installed compression power and energy requirements, e.g., fuel gas, as part of a technical and economic evaluation. In other words, there is a need for using analytical methods to relate the flow of gas through a pipeline to the properties of both the pipeline and gas, and to the operating conditions such as pressure and temperature. Isothermal steady-state pressure drop or flow rate calculation methods for single-phase dry gas pipelines are the most widely used and the most basic relationships in the engineering of gas-delivery systems (Beggs, 1984; Smith, 1990; Aziz and Ouyang, 1995). But close attention should be given while designing a new gas pipeline project since with the technology now available is nonisothermal and transient analysis may be used providing better results and lowering economic risks to the gas pipeline project (Santos, 1997). When enough information is not available for modeling a gas pipeline for transient analysis designers can make appropriate assumption of future expected operation conditions or make use of load (or swing) factor that will increase design capacity for steady-state calculation of a pipeline diameter.
12.2.1 General flow equation

Based on the assumptions that there is no elevation change in the pipeline and that the condition of flow is isothermal, the integrated Bernoulli’s equation is expressed by Eqn (12-1) (Uhl, 1965; Schroeder, 2001):

\[
Q_{sc} = C \left( \frac{T_b}{P_b} \right) D^{2.5} \left( \frac{P_1^2 - P_2^2}{f \gamma_G T_a Z_a L} \right)^{0.5} E
\]

where \( Q_{sc} \) is standard gas flow rate measured at base temperature and pressure, \( \text{ft}^3/\text{day} \); \( T_b \) is gas temperature at base condition, \( 519.6 \, ^\circ\text{R} \); \( P_b \) is gas pressure at base condition, \( 14.7 \, \text{psia} \); \( P_1 \) is inlet-gas pressure, \( \text{psia} \); \( P_2 \) is outlet-gas pressure, \( \text{psia} \); \( D \) is inside diameter of pipe, inches; \( f \) is Moody friction factor; \( E \) is flow efficiency factor; \( \gamma_G \) is gas specific gravity; \( T_a \) is average absolute temperature of pipeline, \( ^\circ\text{R} \); \( Z_a \) is average compressibility factor; \( L \) is pipe length, miles; and \( C \) is constant for the specific units used (77.54).

Although the assumptions used to develop Eqn (12-1) are usually satisfactory for a long pipeline, the equation contains an efficiency factor, \( E \), to correct for these assumptions. Most experts recommend using efficiency factor values close to unity when dry gas flows through a new pipeline. However, as the pipeline ages and is subjected to varying degrees of corrosion, this factor will decrease (Campbell et al., 1992). In practice, and even for single-phase gas flow, some water or condensate may be present if the necessary drying procedure for gas pipeline commissioning is not adopted or scrubbers are not installed. This puts compression equipment at risk of damage and also allowing localized corrosion due to water spots (wetting of the pipe surface). The presence of liquid products in the gas transmission lines can also cause drastic reduction in the flow efficiency factor. Typically, efficiency factors may vary between 0.6 and 0.92 depending on the pipelines liquid contents (Ikoku, 1984). As the amount of liquid content in the gas phase increases, the pipeline efficiency factor can no longer account for the two-phase flow behavior and two-phase flow equations must be used (Brill and Beggs, 1991; Asante, 2002).

Pipelines are usually not horizontal; however, as long as the slope is not too great, a correction for the static head of fluid (\( H_c \)) may be incorporated into Eqn (12-1) as follows (Schroeder, 2001):

\[
Q_{sc} = C \left( \frac{T_b}{P_b} \right) D^{2.5} \left( \frac{P_1^2 - P_2^2 - H_c}{f \gamma_G T_a Z_a L} \right)^{0.5} E
\]

where

\[
H_c = \frac{0.0375 \cdot g(H_2 - H_1)P_2^2}{Z_a T_a}
\]

And \( H_1 \) is inlet elevation, ft; \( H_2 \) is outlet elevation, ft; and \( g \) is gravitational constant, \( \text{ft/s}^2 \).
The average compressibility factor, $Z_a$, is determined from the average pressure ($P_a$) and average temperature ($T_a$), where $P_a$ is calculated from Eqn (12-4) (Campbell et al., 1992):

$$P_a = \frac{2}{3} \left( \left( \frac{P_1 + P_2}{P_1 + P_2} \right) - \left( \frac{P_1 P_2}{P_1 + P_2} \right) \right)$$

(12-4)

where $P_1$ and $P_2$ are the upstream and downstream absolute pressures, respectively. The average temperature is determined by Eqn (12-5):

$$T_a = \left[ \frac{T_1 - T_2}{\ln \left( \frac{T_1 - T_S}{T_2 - T_S} \right)} \right] + T_S$$

(12-5)

In the above equation, parameter $T_S$ is the soil temperature, and $T_1$ and $T_2$ are the upstream and downstream temperatures, respectively.

Having obtained $P_a$ and $T_a$ for the gas, the average compressibility factor can be obtained using Kay’s rule and gas compressibility factor charts (Campbell et al., 1992).

### 12.2.2 Friction factor correlations

The fundamental flow equation for calculating pressure drop requires a numerical value for the friction factor. However, because the friction factor, $f$, is a function of flow rate, the whole flow equation becomes implicit. To determine the friction factor, the fluid flow is characterized by a dimensionless value known as the Reynolds number, Eqn (12-6):

$$N_{Re} = \frac{\rho V D}{\mu}$$

(12-6)

where $N_{Re}$ is Reynolds number, dimensionless; $D$ is pipe diameter, ft; $V$ is fluid velocity, ft/s; $\rho$ is fluid density, lbm/ft$^3$; and $\mu$ is fluid viscosity, lbm/ft s.

For Reynolds numbers less than 2000 the flow is considered laminar. When the Reynolds number exceeds 2000, the flow is characterized as turbulent. Note, in high-pressure gas transmission pipelines with moderate to high flow rates, only two types of flow regimes are observed: partially turbulent flow (smooth pipe flow) and fully turbulent flow (rough pipe flow). For gases, the Reynolds number is given by Eqn (12-7) (Kennedy, 1993):

$$N_{Re} = \frac{0.7105 \ P_b \ \gamma_G \ Q_{sc}}{T_b \ \mu_G \ D}$$

(12-7)

where $D$ is pipe diameter, inches; $Q_{sc}$ is gas flow rate, standard ft$^3$/day; $\mu_G$ is gas viscosity, cp; $P_b$ is base pressure, psia; $T_b$ is base temperature, °R; and $\gamma_G$ is gas specific gravity, dimensionless.
For the gas industry, Eqn (12-7) is a more convenient way to express the Reynolds number since it displays the value proportionally in terms of the gas flow rate.

The other parameter in the friction factor correlation is pipe roughness ($\varepsilon$), which is often correlated as a function of the Reynolds number and the pipe relative roughness (absolute roughness divided by inside diameter). Pipe roughness varies considerably from pipe to pipe and Table 12-1 shows the roughness for various types of new (clean) pipes. These values should be increased by a factor ranging between 2 and 4 to account for age and use.

The Moody (1944) friction factor, $f$, in Eqn (12-1) is determined from the Moody diagram. The Moody correlation is shown in Figure 12-1. The Moody diagram consists of four zones: laminar, transition, partially turbulent, and fully turbulent zones.

The laminar zone, the left side, is the zone of extremely low flow rate in which the fluid flows strictly in one direction and the friction factor shows a sharp dependency on flow rate. The friction factor in the laminar regime is defined by the Hagen–Poiseuille equation (Streeter and Wylie, 1979):

$$f = \frac{64}{N_{Re}}$$  \hspace{1cm} (12-8)

The fully turbulent zone, the right side, describes fluid flow that is completely turbulent (back mixing) laterally as well as in the primary direction. The turbulent friction factor shows no dependency on flow rate and is only a function of pipe roughness, as an ideally smooth pipe never really exists in this zone. The friction factor to use is given by the rough pipe law of Nikuradse (1933), Eqn (12-9):

$$\frac{1}{\sqrt{f}} = 2 \log \frac{D}{\varepsilon} + 1.14$$ \hspace{1cm} (12-9)

The above equation shows that if the roughness of the pipeline is increased, the friction factor increases and results in higher-pressure drops. Conversely, by decreasing the pipe roughness, lower friction factor or lower pressure drops are obtained. Note, most pipes cannot be considered ideally smooth at high Reynolds numbers (Schlichting, 1979), therefore, the investigations of Nikuradse (1933) on flow through rough pipes have been of significant interest to engineers.

<table>
<thead>
<tr>
<th>Table 12-1</th>
<th>Pipe Roughness Value (Norsok Standard, 1996)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Type of Pipe (New, Clean Condition)</strong></td>
<td><strong>$\varepsilon$ (inches)</strong></td>
</tr>
<tr>
<td>Carbon steel corroded</td>
<td>0.019685</td>
</tr>
<tr>
<td>Carbon steel noncorroded</td>
<td>0.001968</td>
</tr>
<tr>
<td>Glass fiber reinforced pipe (GRP)</td>
<td>0.0007874</td>
</tr>
<tr>
<td>Steel internally coated with epoxy</td>
<td>0.00018—0.00035</td>
</tr>
</tbody>
</table>
The partially turbulent (transition) zone is the zone of moderately high flow rate in which the fluid flows laterally within the pipe as well as in the primary direction, although some laminar boundary layer outside the zone of roughness still exists. Partially turbulent flow is governed by the smooth pipe law of Karman and Prandtl (Uhl, 1965), Eqn (12-10):

$$\frac{1}{\sqrt{f}} = 2 \log\left(\frac{\nu}{\sqrt{f}} \right) - 0.8$$

(12-10)

The above correlation has received wide acceptance as a true representation of experimental results. However, a recent study by Zagarola (1996) on the flow at high Reynolds numbers in smooth pipes showed that the relevant correlation was not accurate for high Reynolds numbers, where the correlation was shown to predict too low values of the friction factor.

Consensus on how the friction factor varies across the transition region from an ideal smooth pipe to a rough pipe has not been reached. However, Colebrook (1939) presented additional experimental results and developed a correlation for the friction factor valid in the transition region between smooth and rough flow. The correlation is as follows:

$$\frac{1}{\sqrt{f}} = -2 \log\left(\frac{e/D}{3.7} + \frac{2.51}{N_{Re} \sqrt{f}}\right)$$

(12-11)
The above equation is universally accepted as standard for computing friction factor of rough pipes and incorporated in most thermohydraulic simulation softwares available at the market. Moody (1944) concluded that the Colebrook (1939) equation was adequate for friction factor calculations for any Reynolds number greater than 2000. Certainly the accuracy of the equation was well within the experimental error (about ±5% for smooth pipes and ±10% for rough pipes).

The friction factor is sometimes expressed in terms of the Fanning friction factor, which is one-fourth of the Moody friction factor. Care should be taken to avoid inadvertent use of the wrong friction factor.

### 12.2.3 Simplified flow equations

The Moody friction factor, $f$, is an integral part of the general gas flow equation. Since it is a highly nonlinear function, it must be either read from a chart or determined iteratively from a nonlinear equation. Approximations to the Moody friction factor have been widely used because they allow the gas flow equation to be solved directly instead of iteratively. The four most widely published friction factor approximations are: Weymouth, Panhandle A, Panhandle B, and Institute of Gas Technology (IGT)\(^1\) (Beggs, 1984; Ikoku, 1984). The Weymouth equation approximates the Moody friction factor by Eqn (12-12), and the remaining three equations approximate the friction factor by Eqn (12-13), where “m” and “n” are constants. These constants are given in Table 12-2.

$$f = m(D)^{-n} \quad (12-12)$$

$$f = m(N_{Re})^{-n} \quad (12-13)$$

The Reynolds number, $N_{Re}$, can be approximated by Eqn (12-7). In addition to the Reynolds number, the pipe roughness also affects the friction factor for turbulent flow in rough pipes. Hence, the efficiency factor is chosen to correctly account for pipe roughness (Ouyang and Aziz, 1996).

<table>
<thead>
<tr>
<th>Table 12-2</th>
<th>Constants in Eqns (12-12) and (12-13)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Equation</strong></td>
<td><strong>m</strong></td>
</tr>
<tr>
<td>Weymouth</td>
<td>0.032</td>
</tr>
<tr>
<td>Panhandle A</td>
<td>0.085</td>
</tr>
<tr>
<td>Panhandle B</td>
<td>0.015</td>
</tr>
<tr>
<td>IGT</td>
<td>0.187</td>
</tr>
</tbody>
</table>

\(^1\)The IGT equation proposed by the Institute of Gas Technology (IGT) is also known as the IGT distribution equation.
These approximations can then be substituted into the flow equation for \( f \), and the resulting equation is given by Eqn (12-14) (Kennedy, 1993; Towler and Pope, 1994):

\[
Q_{sc} = \frac{a_1 \left( \frac{T_a}{P_a} \right) E \left( P_1^2 - P_2^2 \right)^{0.5} D^{a_2}}{(\gamma_G)^{a_3} (T_a Z_a L)^{a_4} (\mu_G)^{a_5}}
\]

(12-14)

In Eqn (12-14) the values for \( a_1 \) through \( a_5 \) are constants that are functions of the friction factor approximations and the gas flow equation. These constants are given in Table 12-3.

Inspection of Table 12-3 shows that the gas flow rate is not a strong function of the gas viscosity at high Reynolds numbers because viscosity is of importance in laminar flow, and gas pipelines are normally operated in the partially/fully turbulent flow region. However, under normal conditions, the viscosity term has little effect because a 30% change in absolute value of the viscosity will result in only approximately a 2.7% change in the computed quantity of gas flowing. Thus, once the gas viscosity is determined for an operating pipeline, small variations from the conditions under which it was determined will have little effect on the flow predicted by Eqn (12-14) (Huntington, 1950).

Note that all of the equations noted above have been developed from the fundamental gas flow equation; however, each has a special approximation of the friction factor to allow for an analytical solution. For instance, the Weymouth (1912) equation uses a straight line for \( f \), and thus its approximation has been shown to be a poor estimation for the friction factor for most flow conditions (Kennedy, 1993). This equation tends to overpredict the pressure drop, and thus provides a poor estimate relative to the other gas flow equations. The Weymouth equation, however, is of use in designing gas distribution systems in that there is an inherent safety in over predicting pressure drop (Maddox and Erbar, 1982). In practice, the Panhandle equations are commonly used for large diameter, long pipelines where the Reynolds number is on the flat portion of the Moody diagram. The Panhandle “A” equation is most applicable for medium- to relatively large-diameter pipelines (12”–60” diameter) with moderate gas flow rate, operating under medium to high pressure (800–1500 psia). The Panhandle “B” equation is

<table>
<thead>
<tr>
<th>Equation</th>
<th>( a_1 )</th>
<th>( a_2 )</th>
<th>( a_3 )</th>
<th>( a_4 )</th>
<th>( a_5 )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Weymouth</td>
<td>433.46</td>
<td>2.667</td>
<td>0.5000</td>
<td>0.5000</td>
<td>0.0000</td>
</tr>
<tr>
<td>Panhandle A</td>
<td>403.09</td>
<td>2.619</td>
<td>0.4603</td>
<td>0.5397</td>
<td>0.0793</td>
</tr>
<tr>
<td>Panhandle B</td>
<td>715.35</td>
<td>2.530</td>
<td>0.4900</td>
<td>0.5100</td>
<td>0.0200</td>
</tr>
<tr>
<td>IGT</td>
<td>307.26</td>
<td>2.667</td>
<td>0.4444</td>
<td>0.5556</td>
<td>0.1111</td>
</tr>
</tbody>
</table>

*IGT, Institute of Gas Technology.*
normally appropriate for high flow rate, large diameter (>36") and high-pressure (>1000 psi) transmission pipelines (Maddox and Erbar, 1982; Kennedy, 1993). The IGT equation is particularly suitable for high pressure, high flow rates through steel or plastic/polyethylene pipes for most distribution design situations.

Since friction factors vary over a wide range of Reynolds number and pipe roughness, none of the gas flow equations is universally applicable. However, in most cases, pipeline operators customize the flow equations to their particular pipelines by taking measurements of flow, pressure, and temperature and back-calculating pipeline efficiency or an effective pipe roughness.

12.3 Predicting gas temperature profile

Predicting the pipeline temperature profile has become increasingly important both in the design and in the operation of pipelines and related facilities. Flowing gas temperature at any point in a pipeline may be calculated from known data in order to determine (1) location of line heaters for hydrate prevention (Towler and Mokhatab, 2004), (2) inlet-gas temperature at each compressor station, and (3) minimum gas flow rate required to maintain a specific gas temperature at a downstream point. To predict the temperature profile and to accurately calculate pressure drop, it is necessary to divide the pipeline into smaller segments. The temperature change calculations are iterative since the temperature (and pressure) at each point must be known to calculate the energy balance. Similarly, the pressure loss calculations are iterative since the pressure (and temperature) at each point along the pipeline must be known in order to determine the phase physical properties from which the pressure drop is calculated. Thus, generating a useable temperature profile requires a series of complex, interactive type calculations for which even the amount of data available in most cases is insufficient. Additionally the pipeline outer environment properties such as soil data and temperatures vary along the pipeline route and therefore play a very important role and require a consistent modeling to provide a reliable temperature profile evaluation. A simple and reasonable approach is to divide the pipeline into sections with defined soil characteristics and prevailing soil temperatures for summer and winter time and then to calculate the overall heat transfer that will be highly influenced by the outer conditions. The complexity of this method has led to the development of approximate analytical methods for prediction of temperature profile, which in most situations are satisfactory for engineering applications.

Basic relationships needed for these calculation methods are thermal, mechanical energy balances, and mass balance for the gas flow in pipelines. The general or thermal energy balance can be written as follows (Buthod et al., 1971):

\[
P_{C_P} \left( \frac{dT}{dx} \right) - \eta P_{C_P} \left( \frac{dP}{dx} \right) + \left( \frac{V dV}{g_c} \right) \left( \frac{dV}{dx} \right) + \left( \frac{g}{g_c} \right) \left( \frac{dH}{dx} \right) = - \frac{dq}{dx} \quad (12-15)
\]
where $T$ is gas temperature; $P$ is absolute pressure of gas; $V$ is gas linear velocity in pipeline; $q$ is heat loss per unit mass of flowing fluid; $C_p$ is constant pressure specific heat; $\eta$ is Joule–Thomson coefficient; $H$ is height above datum; $x$ is distance along pipeline; $g_c$ is conversion factor; and $g$ is gravitational acceleration.

The major assumption in the development of Eqn (12-15) is that the work term is zero between the compressor stations.

To calculate the heat transfer from the pipe to the ground (soil), per unit of pipeline length, the Kennelly equation, Eqn (12-16), is used (Neher, 1949):

$$\frac{dq}{dx} = \frac{2 \pi K (T - T_S)}{m_G} \ln \left( \frac{2H' + \sqrt{4H'^2 - D_0^2}}{D_0} \right)$$  \hspace{1cm} (12-16)

where $K$ is thermal conductivity of soil; $T_S$ is undisturbed soil temperature at pipe centerline depth; $m_G$ is mass flow rate of gas; $H'$ is depth of burial of pipe (to centerline); and $D_0$ is outside diameter of pipe.

A basic assumption in Eqn (12-16) is that the temperature of the gas is the same as the temperature of the pipe wall (the resistances to heat transfer in the fluid film and pipe wall are negligible).

The mechanical energy balance is given by Eqn (12-17) (Streeter and Wylie, 1979):

$$\frac{1}{\rho} \frac{dP}{dx} + \frac{VdV}{g_c dx} + 2fV^2 \frac{g_c}{D_i} = 0$$  \hspace{1cm} (12-17)

where $\rho$ is density of gas; $f$ is Fanning friction factor; and $D_i$ is inside diameter of pipe.

The continuity equation, Eqn (12-18), relates velocity to the pressure and temperature (Buthod et al., 1971):

$$\frac{dV}{dx} = \frac{-m_G}{\rho^2 A} \left[ \frac{\rho}{P} - \frac{\rho}{Z} \frac{\partial Z}{\partial P} \right] \frac{dP}{dx} - \left[ \frac{\rho}{T} + \frac{\rho}{Z} \frac{\partial Z}{\partial T} \right] \frac{dT}{dx}$$  \hspace{1cm} (12-18)

where $A$ is inside cross-sectional area of pipe; and $Z$ is gas compressibility factor.

Equations (12-15)–(12-18) are the basic equations that must be solved simultaneously for the calculation of the gas temperature and pressure profiles in pipelines. Details for solving this set of equations can be found in textbooks on numerical analysis, such as Constantinides and Mostoufi (1999).

The typical equations used to determine pipeline temperature loss are the integrated form of the general equations. However, assumptions and simplifications must be made to obtain the integrated equations, even though the effects of these assumptions or simplifications are not always known. A major advantage of numerical integration of the differential equation is that fewer assumptions are necessary. Considering this fact, Coulter and Bardon (1979) have presented an integrated equation as follows:

$$T_x = \left\{ T_1 - \left[ T_S + \left( \frac{\eta}{a} \right) \left( \frac{dP}{dx} \right) \right] e^{-ax} \right\} + \left[ T_S + \left( \frac{\eta}{a} \right) \frac{dP}{dx} \right]$$  \hspace{1cm} (12-19)
where, $T_1$ is the inlet-gas temperature and the term “a” is defined as below:

$$a = \frac{2\pi RU}{mG C_P}$$  \hspace{1cm} (12-20)

where $R$ is pipe radius; and $U$ is overall heat transfer coefficient.

Equation (12-19) can be used to determine the temperature distribution along the pipeline, neglecting kinetic and potential energy, and assuming that heat capacity at constant pressure, $C_P$, and Joule–Thomson coefficient remain constants along the pipeline. For most practical purposes, these assumptions are close to reality and generally do give quite good results. Moreover, for a long, gas transmission pipeline with a moderate to small pressure drop, the temperature drop due to expansion is small (Buthod et al., 1971) and Eqn (12-19) simplifies to Eqn (12-21):

$$T_x = T_S + (T_1 - T_S) \exp(-ax)$$  \hspace{1cm} (12-21)

Equation (12-21) does not account for the Joule–Thomson effect, which describes the cooling of an expanding gas in a transmission pipeline. Hence, it is expected that the fluid in the pipeline will reach soil temperature later than that predicted by Eqn (12-19).

While there has been extensive effort in the development of equations for predicting pipeline temperature profile, little attention has been paid to the fact that the Joule–Thomson coefficient and heat capacity at constant pressure are not constants. However, a new analytical technique for the prediction of temperature profile of buried gas pipelines has been developed by Edalat and Mansoori (1988) while considering the fact that $\eta$ and $C_P$ are functions of both temperature and pressure. Readers are referred to the original reference for a detailed treatment of this method.

12.4 Transient flow in gas transmission pipelines

Transients occur in natural gas pipelines during filling and line pressurization, emergency shutdown, gas blowdown, and line depressurization processes. Transient flow can also occur in the gas pipeline during normal operation due to variations in demand, inlet and outlet flow changes, compressor start and stop, pipe leak and/or rupture.

There are some reasons for applying transient analysis during design phase of gas transmission pipelines. The first reason is related to the large investment required by such projects which include pipeline and compression stations. The designed system must be able to operate under predicted different scenarios; otherwise the transmission company would face penalties for not delivering the contracted gas volumes and will probably have to make additional unpredicted capital investment on the system, dramatically affecting its cash flow. The second reason is related to the operation behavior of some end user such as gas-fired
power plant with different gas demand profile that may interrupt gas consumption up to zero for a certain period of time on a weekly basis. These scenarios must be taken into account since they directly affect pipeline capacity management and transmission costs and also affect the assembling schedule of compression stations and compressor units causing cyclic operation shutting units on and off. Transient analysis is used to help selecting turbo compressors that best fit system requirements arranged in series (few units and bigger machines) or parallel (more units and smaller machines). It is also necessary to run failure analysis for a single compressor unit or even for a complete station and predict system response in terms of remaining capacity versus time. Transient analysis is important to establish maintenance strategy and define whether standby unit will be installed to enhance system availability. Transient analysis will also be very useful along the negotiations process of ship-or-pay contracts that normally starts well before the design phase. Additional uses of transient analysis are related to the operation of the pipeline in control rooms, training operation personnel, and commercial planning (Santos, 1997; Santos and Mokhatab, 2008).

Accurate predictions of the gas flow rate, temperature, and pressure profiles along the pipelines under transient conditions, which are vital to the adequate operation of gas transmission pipelines, indicate the need for appropriate mathematical models to do a detailed analysis. When modeling pipelines, however, it is sometimes convenient to make the simplifying assumption that flow is isothermal and steady state as long as we incorporate a load or swing factor contingent to a latter transient design checking to prevent inadequate pipeline sizing with potential detrimental impact on the feasibility of pipeline projects (Santos, 1997). Steady-state models are widely used to design pipelines and to estimate flow and line pack. However, there are many situations where an assumption of steady-state flow and its attendant ramifications produce unacceptable engineering results. Note should be made that the flowing gas temperature in transmission pipeline does not remain constant over the length of the pipeline. Therefore, complete temperature profiles along pipeline length are required to adequately design the pipeline for operation under varying environmental conditions.

The unsteady nonisothermal flow of gas in transmission pipelines can be described by a one-dimensional approach and by using an equation of state and a set of partial differential equations expressing mass, momentum, and energy conservation laws (Osiadacz and Chaczykowski, 2010). In practice, the form of the mathematical relationships depends on the assumptions made based on the operating conditions of the pipelines. For the case of slow transient flows due to fluctuations in demand, it is assumed that the gas in the line has sufficient time to reach thermal equilibrium with its constant-temperature surroundings. Similarly, for the case of rapid transient flows, it is assumed that the pressure changes occur instantaneously, allowing no time for heat transfer between the gas in the pipeline and the surroundings. However, for this case, heat conduction effects cannot be neglected.
12.5 Compressor stations

A principal component of any gas transmission system is the compressor station. As natural gas flows through a pipeline, it loses pressure due to friction against the inside of the pipe. Therefore, compressor stations are installed along the pipeline to boost the gas pressure in the pipeline to maintain required delivery pressure and flow (Figure 12-2). A given gas transmission system may have anywhere from a few stations up to well over 50 depending on feasibility studies.

Compressor stations comprise one or more compressor units (designed with enough horsepower and throughput capacity to meet contractual requirements levied on the system), each of which will include a compressor and driver together with valves, control systems and exhaust ducts, and noise attenuation systems. Most compressors are powered by natural gas taken directly from the pipeline, but electric powered compressors are becoming more common. Figure 12-3 shows a typical compressor station with three gas turbine-driven centrifugal compressors.

A typical single-stage compressor station design, as shown in Figure 12-4, may consist of an inlet scrubber to collect entrained liquids (i.e., water, corrosion inhibitors, hydrocarbon liquids that may have formed in the gas transmission pipeline) followed by a coalescing filter to remove fine solids (i.e., pipe scale) and hydrocarbon mist from the gas that could otherwise contribute to compressor failure. A coalescing filter also cleans the gas in each fuel supply to the turbines and gas engines. From the scrubber, the gas is taken to the compressor unit(s) where it is compressed. At the compressor station discharge or between compressor units in case of series arrangement the gas is cooled down, typically with an air cooler, and then it passes through a scrubber allowing drainage of any formed liquid. In case of reciprocating compressor being used in the compressor station.
station, a coalescing filter shall be used after the scrubber to remove lube oil mist prior to the gas being introduced back into the pipeline.

12.5.1 Station facilities

Each compressor station will be built up from the same functional blocks of equipment. Each functional element plays a part in the work of the station and the design and sizing of each is essential to the efficient and safe operation of the plant.

The functional elements include gas scrubbing and liquid removal, compressor and driver units, aftercoolers, pipes and valves. Controls, including Supervisory Control and Data Acquisition (SCADA) system, monitoring and data recording, alarms and shutdown procedures, both routine and emergency, are an integral part of the station. Provision also has to be made for, venting, the compressor and driver housing and buildings, complete with ventilation and fire protection, and safety equipment.
12.5.1.1 Scrubbers

Individual scrubbers (with or without standby unit) may be installed for each stage of compression for each compressor unit. A common scrubber should be considered for multiple units on a common suction line. Scrubbers at compressors inlet and after gas coolers can take several forms, inertial with or without demister pads or the horizontal cyclonic type. The latter are commonly used on mainline transmission stations.

Efficient and safe handling of the liquids collected from the scrubbers in a compressor station is one of the keys to a good design. The liquids removed from the gas flow to a hydrocarbon storage tank to be separated by gravity and then be transported by truck as saleable hydrocarbon liquid or as disposed waste product. There are three main concerns that should be addressed in the liquid-handling design for any compressor station: safety, environmental impacts, and economics. An additional point of consideration should be operability, which includes issues like hydrate formation (when water is present in the gas composition), failure consequences, etc.

12.5.1.2 Compressors

In gas transmission, two basic types of compressors (reciprocating and centrifugal) are used. Reciprocating compressors are usually driven by either electric motors or gas engines, whereas centrifugal compressors use gas turbines or electric motors as drivers.

Compressor selection is based on detailed analysis of the operating conditions in terms of flow and pressure ratio needed by the pipeline hydraulics. Frequently these conditions will vary over time and the compressor selection will have to have flexibility, by restaging if necessary, to accommodate all the expected conditions. From transient analysis based on predicted ramp-up gas demand and flow profiles, a set of values are defined that will be used to preselect compressor units depending on the installation layout whether series or parallel (Santos, 1997, 2004).

Typically the conditions for most mainline stations will require compressors with one or two stages, and the compressor design may be of the overhung rotor or barrel design. The determination of operating conditions and hence the development of the compressor characteristics will have to take into account the gas characteristics, suction and discharge pressure, and suction temperatures, usually involving equations of state for the particular gas composition, and the process environment. (i.e., climate, altitude, location, etc.)

Pipeline and station designer will make sure that the equipment selection and arrangement including maintenance strategy and level of availability will be subject to a feasibility study (Santos, 2004).

12.5.1.3 Compressor drivers

Choices for drivers can be gas turbines or electric motors. The selection is usually based on considerations of cost, both capital and maintenance, fuel or energy
cost, availability of power supply, reliability, and availability. Gas turbines are high-speed machines and can be directly coupled to the compressor. Electric motors can be of several types with both fixed and variable speed options. In recent times, the high-speed variable speed electric motor operating at super synchronous conditions has become available at powers of more than 25 MW which makes this option competitive with larger gas turbines (Cleveland and Mokhatab, 2005). The variable speed planetary gear (VSPG from VOITH) that provides another economic alternative to be used with fixed speed electric motor is also available in the market.

Capital costs for gas turbines and electric motors are generally similar, electric drives generally have lower maintenance and operating costs than gas turbines. The decision to use gas turbine or electric motor drivers comes from the feasibility analysis for the alternatives available for the project and site logistics, capital expenditure (CAPEX) and operating expenditure (OPEX), availability and reliability for the compressor units and for the energy source are key items. Important items in favor of electric-driven compressor station that should be considered in the feasibility analysis is that the fuel gas used for gas turbine-driven compressor station will be transformed into capacity increase for the electric-driven compressor station and therefore will add revenue to this alternative and also the overhaul cost for the gas turbines that are expressive and will occur after completing around 40,000 running hours, accounted in the OPEX costs (Kurz and Ohanian, 2003). There will always be gas in the pipeline so the question of reliability and availability of the energy source for the gas turbine does not enter into the question. For the electric drive, there has to be a reliable electric grid within a reasonable distance from the compressor station, transmission lines costs need to be considered in the evaluation. When there is a reliable source of electric energy available, it would be worth to evaluate both alternatives (gas or electric drivers) supported by failure statistics for the electric grid such as Loss of Load Frequency, Loss of Load Duration, and Loss of Load Probability that will allow to define the reliability of the electric grid. Given that these conditions are satisfied, the decision then comes from the economic evaluation of all the alternative and selecting the one that will give the best economic result. The life-cycle cost must examine and test the results for sensitivity to cost escalation in power prices and gas price taking into account the correlation between these two commodities. A long-term power supply agreement would be required to mitigate risk. The question of using a diesel engine as power source has not been considered as it offers no advantages over a similar gas engine and it introduces another fuel, which invokes additional costs for transportation and storage.

Selection of the appropriate units has to consider all aspects of interest including operating cost, reliability and availability as well as capital cost. The process of selection should be formalized so as to remove as far as possible subjective elements and ensure the decisions are objectively based (Kurz et al., 2003).
12.5.1.4 Piping and valves
The compressor station will be connected to the gas pipeline via a set of block valves with a bypass. These will be typically ball valves with a block and bleed capability. The piping should be designed to minimize pressure loss as pressure loss represents inefficiency. The supports and anchors for the piping shall be designed to allow for expansion following compression and to keep the forces and moments on the compressor flanges within the limits set by the compressor manufacturer. For centrifugal compressors, there will be a surge valve and recycle line for surge protection, this recycle line will connect the discharge line to the suction line adjacent to the compressor and typically can go through the after-cooler. The sizing of this line and the valve is critical; and should be done in consultation with the compressor supplier and with knowledge of the compressor characteristics. Each compressor on the station will have its own set of isolation and block valves with a bypass.

12.5.1.5 Shutdown and venting systems
For most compressor stations there are two cases for shutdown and venting. The first can be stated as routine, in which a unit or a station has to be shutdown for repair or maintenance the second is in the case of a failure or emergency when the gas in the station needs to be evacuated safely and as quickly as possible. Most stations are now designed such that the compressors will remain pressurized when idle and routine venting will only be required when work has to be done on the units.

12.5.1.6 Emergency shutdown system
The compressor station shall be provided with an emergency shutdown system by means of which the station will be isolated from the pipeline and gas flow through the station will be stopped. A blowdown system will release the gas from station piping to the atmosphere. Operation of the emergency shutdown system also shall cause the shutdown of all gas compressing equipment and shall de-energize the electrical facilities located in the vicinity of gas headers and in the compressor room except those that provide emergency lighting for personnel protection and those that are necessary for protection of equipment. The emergency shutdown system shall be operable from any one of at least two locations outside the gas area of the station, preferably near exit gates in the station fence, but not more than 150 m (500 ft) from the limits of the station. Blowdown piping shall extend to a location where the discharge of gas is not likely to create a hazard to the compressor station.

12.5.1.7 Pressure relief systems
Pressure relief or other suitable protective devices of sufficient capacity and sensitivity shall be installed and maintained to assure that maximum allowable operating pressure of the station piping and equipment is not exceed by more than 5%.
12.5.1.8 Flare system
Flare lines shall be provided to exhaust the gas from the pressure relief and vent valves to atmosphere and shall be extended to a location where the gas may be discharged without hazard. Flare lines shall be sized to provide sufficient capacity so that they will not interfere with the performance of the relief and vent valves.

12.5.2 Compressors arrangements
Since gas pipeline projects demand high amounts of capital expenditure and therefore are involved with investments risks, project sponsors will try to maximize capacity usage and minimize investment so as to have a competitive transportation rate to offer to the market. At the same time, it will avoid the pipeline to operate with spare capacity.

The decision on the compressors arrangement whether series or parallel is mostly based on economics and on simulation of failure analysis (Santos, 2000). While series arrangement may present some advantages when standby compressor units are not required, parallel arrangement provides better results when standby units are required, and also provides more operating flexibility under failure scenario analysis. Pipeline designer should evaluate different configuration and compressor unit size and perform technical and economical feasibility study to identify which configuration presents a better overall economic result. The decision process has to take into account issues such as capacity ramp-up, further expansion, back-up strategies, operational strategy, and transient analysis.

For any given pipeline compressor station, two units in series will yield a higher specific speed than two units in parallel. Thus, once the driver size (and thus the power turbine speed) and the desired head and flow through the station are known, one can conceptually decide whether the series or the parallel approach would lead to better aerodynamic performance. With modern compressors and stages with a wide operating range, it is usually possible to have identical stages for both the low-pressure and the high-pressure compressor in a series application. Intercooling is usually not necessary nor does it typically yield significant savings in power demand (Kurz et al., 2003; Santos, 2004).

12.5.3 Station control
Control functions are typically based on personnel safety, the operating parameters of the station, and the types and number of compressor units installed at the station.

Compressor station controls can be divided into two sections, unit control and station control. Digital technology is now used throughout both systems, the unit control utilizes a microprocessor which will control the turbine compressor unit to run to set points under the direction of the operator or the station control system, the set points can be flow or pressure. Commonly, a flow or suction
pressure will be the control parameter with discharge pressure and/or suction pressure as overrides. The control protocol will include limits to ensure safe operation. These limits will include pressure, and temperatures on discharge and suction on the compressor as well as speed and flow and pressure ratio in relation to surge.

The unit control will monitor the compressor operation to ensure that it will not run into surge. If the operation of the compressor nears the surge line, the unit control will instruct the recycle valve to open and so maintain safe operation. Should the recycle condition continue for a time, and if coolers are not provided in the recycle line or compressor discharge, the unit will be shut down on high discharge temperature.

In addition to control and safety, the unit control will monitor key operating parameters and provide video output on demand and printout on a routine basis to provide a continuous record of operation. These readouts and records can be used for trouble shooting and maintenance. The station control system will oversee the unit operation and in addition provide the interface between the operators and the plant. It will also provide video and print data recording of all key station parameters. It has become practice to operate stations and units remotely from central dispatch stations and the station control systems will report to the central station via a SCADA link (Cleveland and Mokhatab, 2005).

The overall control of a major gas pipeline transportation system typically originates from a central “Gas Control” office that is remote from all of the compressor stations. Gas control monitors flow measurement for the total station throughput as well as each compressor’s throughput and fuel consumption. The programmable logic controller (PLC) in each compressor unit’s control panel communicates to gas control the operating parameters for that compressor and the positions of the valves controlling the gas flow through that compressor. This example station uses redundant communication using microwave, satellite, or conventional leased telephone systems. This station is designed for the option of completely unmanned operation by Gas Control. The PLC compressor on/off operation, performance set points, and all station-critical valves may be remotely controlled. Gas Control may monitor all station alarms and shutdowns.

Gas turbine-driven centrifugal compressor unit panels typically include unit protective functions, local and remote starting, automatic unit valve operation, and equipment and process monitoring instrumentation. If a multiunit station is involved, a separate station panel may be used to automatically control station operation (Mokhatab et al., 2008).

### 12.5.4 Acoustical treatment

Noise is a significant environmental pollutant and the reduction of noise is an essential part of the design of a compressor station. The technology of noise reduction has reached the level that for most practical purposes a compressor station can be designed to contribute less than 3 dB to the preexisting background
noise level. Local requirements should be taken into account for a proper design. The design of the unit enclosures, buildings, exhaust, and inlet silencers are subject to stringent specifications. Double-wall enclosures are frequently used to control unit noise emissions as afore mentioned.

Normally an acoustic simulation study is done. If an acoustic simulation study has not been made, the following requirements for the piping system should be made:

- The use of elbows in the piping should be minimized to the minimum number possible.
- Elbows eliminated between compressor cylinders and pulsation suppression devices.

Exhaust emission from the gas turbines now have to meet the environmental limits of the location and modern gas turbines are designed with low-emission combustion systems to meet these requirements. These systems may be dry or wet low NOx and are now becoming the standard equipment for all gas turbines.

Silencing equipment incorporated into the engine air intake ducting and filters, engine exhaust systems, vent and blowdown exits, equipment enclosures, and piping insulation accomplishes noise control for the surrounding community and station employees. Liners and baffles of noise-reduction material are used inside the engine air intake ducting and filters, engine exhaust, and vent and blowdown exits. Acoustical lining on the internal walls and silenced ventilation systems reduce noise from the compressor buildings.

As much station gas piping as possible is installed below grade to provide additional noise reduction. Above grade station, gas piping is acoustically insulated. Internal noise-reduction modifications are required in flow control valves. Fan tip speed limitations and low-noise fan drive designs are required for gas and oil coolers. Station noise control provides a day–night average sound level (Ldn) of less than 55 dB per the “A” rating (dBA) of human response to noise at the nearest noise sensitive area. This complies with the current US Federal Energy Regulatory Commission regulations. Noise levels within a 3-ft distance from equipment for the employee’s average time of exposure is less than 85 dBA. This complies with current US Occupational Safety and Health Administration requirements.

### 12.5.5 Reliability and availability

Compressor station reliability and availability are paramount to overall gas pipeline delivery dependability. Reliability considerations are incorporated into many areas of the facility.

- The total amount of installed compression (operating and standby units) must be more than the normal design requirement to allow for scheduled and unscheduled maintenance.
- Spacing between compressors or compressor groups aims at preventing fire damage to one compressor from harming others and to ease maintenance work.
Using redundant and parallel filter coalescers prevents unexpected large amounts of contaminants from impeding the gas flow and allows filtration cartridge replacement without interrupting compressor operation.

- Monitoring and trending of vibration, bearing temperatures, and other critical operational parameters by the compressor PLC identify service needs to prevent catastrophic equipment failures.
- Maintenance systems should be developed to manage all aspects of maintenance prior to station start-up.
- All below ground steel pipe, conduit and structures are coated with a corrosion protective and electrically insulating coating. Additionally, steel pipe installed below ground is protected from external corrosion using cathodic protection as part of the protection system.
- Sufficient operational and capital spare parts inventories should be available based on reliability-availability-maintenance analysis and life-cycle cost considerations.
- Equipment should be preferably standardized as feasible to minimize spare parts requirements.

Good human factors practices should be used in evaluating access to and viewing of operating data, manipulation of controls, installation of isolation devices, and removal and replacement of equipment. (e.g., equipment and personnel access and egress, lifting points, pull clearances, materials movement, etc.).

12.6 Reduction and metering stations

In addition to compressor stations, there may be gas injection and delivery points along the line where the pressure and flow will have to be monitored and controlled. Each of these locations will include pressure control facilities and flow measurement.

Each reduction and metering station branches off the pipeline and is used to reduce pressure and meter the gas to the various users. For the reduction and metering stations, the main equipments include filters, heaters, pressure reduction and regulators, and flow metering skids. In addition, each station is generally equipped with drains collection and disposal, instrument gas system, and storage tanks.

12.6.1 Filters

Natural gas filter units are installed at each station to remove any entrained liquids and solids from the gas stream. The filters may comprise cyclonic elements to centrifuge particles and liquids to the sides of the enclosing pressure vessel. These particles and liquids will then drop down for collection in a sump, which can be drained periodically.
12.6.2 Heaters
Natural gas heaters are installed to avoid the formation of hydrates, liquid hydrocarbons, and water as a result of pressure reduction. The gas heater is designed to raise the temperature of the gas such that after pressure reduction, the temperature of the gas will be above the dew-point temperature at operating conditions and maximum flow. The heater is a water bath natural circulation type maintained at a temperature between 158 and 176 °F. Where gas cost is high, an alternative is to use high-efficiency or condensing furnaces for the purpose of preheating the gas rather than the water bath heater.

12.6.3 Pressure reduction and regulation system
Pressure reduction system controls the supply pressure to the gas users at a regulated value. Each system consists of at least two trains of pressure reduction, one operating, and the other standby. Each train will normally comprise two valves in series, one being the “active valve,” the other the “monitor valve.” Each valve will be equipped with a controller to operate the valve to maintain the preset discharge pressure.

12.6.4 Metering system
The flow rate of the gas has to be measured at a number of locations for the purpose of monitoring the performance of the pipeline system and more particularly at places where “custody transfer” takes place that is where gas is received from the supply source and gas is sold to the customer for distribution. Depending on the purpose for metering, whether for performance monitoring or for sales, the measuring techniques used may vary according to the accuracy demanded. Typically, a custody transfer metering station will comprise one or two runs of pipe with a calibrated metering orifice in each run.

12.7 Design considerations of sales gas pipelines
The typical design of a gas transmission pipeline involves a compromise between the pipe diameter, compressor station spacing, fuel usage, and maximum operating pressure. Each of these variables influences the overall construction and operating cost to some degree, hence an optimized design improves the economics of the construction and operation of the system, and the competitiveness of the project.

12.7.1 Line sizing criteria
The pipe size generally is based on the acceptable pressure drop, compression ratio, and allowable gas velocities. Acceptable pressure drop in gas transmission pipelines must be one that minimizes the size of the required facilities and
operating expenses such as the pipe itself, the installed compression power, the size and number of compressors, and fuel consumption. In fact, a large pressure drop between stations will result in a large compression ratio and might introduce poor compressor station performance. Experience has shown that the most cost-effective pipeline should have a pressure drop in the range between 3.50 and 5.83 psi/mile (Hughes, 1993). However, for those pipelines (short ones) in which pressure drop is of secondary importance, the pipe could be sized based on fluid velocity only. The flow velocity must be kept below maximum allowable velocity in order to prevent pipe erosion, noise or vibration problems, especially for gases that may have a velocity exceeding 70 ft/s. In systems with CO₂ fractions of as low as 1–2%, field experiences indicate that the flow velocity should be limited to less than 50 ft/s, because it is difficult to inhibit CO₂ corrosion at higher gas velocities (Kumar, 1987; Arnold and Stewart, 1999).

The recommended value for the gas velocity in the transmission pipelines is normally 40–50% of the erosional velocity² (Mohitpour et al., 2002). As a rule of thumb, pipe erosion begins to occur when the velocity of flow exceeds the value given by Eqn (12-22) (Beggs, 1984):

\[
V_e = \frac{C}{\rho_g^{0.5}}
\]

(12-22)

where \( V_e \) is erosional flow velocity, ft/s; \( \rho_G \) is density of the gas, lb/ft³; and \( C \) is empirical constant. In most cases, \( C \) is taken to be 100. However, API RP 14E (1984) suggested a value of \( C = 100 \) for a continuous service and 125 for a noncontinuous service. In addition, it suggests that values of \( C \) from 150 to 200 may be used for continuous, noncorrosive, or corrosion-controlled services, if no solid particles are present.

After selecting the appropriate inside diameter for a pipe, it is necessary to determine the pipe outside diameter (wall thickness), which would result in the minimum possible fabrication cost while maintaining pipeline integrity.

### 12.7.2 Compressor station spacing

In long distance gas transmission systems with a number of operating compressors, there is a definite need to optimize the spacing between compressor stations. Compressor station spacing is fundamentally a matter of balancing capital and operating costs at conditions, which represent the planned operating conditions of the transmission system. The process can become somewhat involved and lengthy particularly as the selection of spacing needs to be designed in such a way to address a capacity ramp-up scenario that will cover not only the initial condition but also the future years associated with the economics of the

²It is recommended that a minimum gas velocity of 10–15 ft/s in the pipeline be maintained so as to minimize surging and to transport sand and other solids.
pipeline. In case of unexpected growth opportunities, we can also rely on loop lines that may be a better additional choice to increase capacity even more.

For a given pipe diameter, the distance between compressor stations may be computed from the gas flow equation, assuming a value of pipeline operating pressure (station discharge pressure) and a next compressor station suction pressure limited to the maximum compression ratio adopted for the project. Ideally, the pipeline should operate as close to maximum allowable operating pressure (MAOP) as possible, as high density in the line of the flowing gas gives best efficiency. This would point to the selection of close compressor station spacing but this approach would not be the best economical decision. A decision based on the pipeline economics is the recommended one. Based on the required gas flow, an initial diameter is assumed that results in a reasonable compression ratio (usually around 1.25–1.6 for transmission lines) and gas velocity, and the compressor station spacing is established by setting the maximum discharge pressure at the MAOP. Other diameters are tested and compressor station spacing calculations are performed again. The optimum diameter is determined based on minimizing capital cost and operating cost resulting in a chart (the so-called J-curves, because of their shape) that will plot present value of total project cost per transmission capacity, US$/CFD or transportation rate (tariff) in US$/MMBTU against transmission capacity, MMCFD, based on predefined economic assumptions and risks (Santos and Saliby, 2003). Such assumptions include the economic life of the facility, the required rate of return or discount rate on capital employed and on annual operating expenses incurred over time to a present value.

Figure 12-5 illustrates J-curves for a 1127-miles gas pipeline with MAOP of 1422.34 psig and compression ratio of 1.4. As can be seen, diameter 36” produces lower cost indexes, US$/CFD, from 1000 to 1200 MMSFD transmission capacity. Markers represent quantity of compressor stations starting with one (that produces the highest cost index) and adding one more as capacity increases dropping cost index up to minimum and then going up again.

After selecting the best configuration for the project (nominal diameter and compressor stations quantity) next step is to define the capacity buildup. Designer can start from the configuration with maximum capacity with compressor stations already located along the length of the pipeline and then remove gradually compressor stations (e.g., from 14 to 7 then to 4 then to 2 then to 1) keeping compressor spacing equals so as to avoid undesirable bottleneck. This approach will allow a better design and help defining equipments that would equal. This also allows defining the installation schedule for the compressor stations and compressor units as well.

Gas pipeline feasibility study with its resultant J-curves requires detailed technical, economical, and operational cost data to be accomplished. CAPEX include costs such as pipe, valves, fittings, compressors, turbines (or electric motors), control and construction, and assembly costs. OPEX include all maintenance and supervision, and fuel or energy cost. CAPEX can be derived from past experience and databases. OPEX has to be estimated based on the specific
The most significant part of operating costs is fuel or energy and equipment overhaul. Fuel cost is directly related to compressor horsepower. In order to illustrate how compressor station spacing influences the economics of pipeline operation, a simple model can be set up. This hypothetical pipeline model is based on a system 1000-miles long, operating at a maximum pressure of 1000 psig and flowing 1000 MMscfd. Assuming a uniform pressure loss per unit length of the pipe, and station spacing, the inlet pressure at the first station downstream can be calculated and the horsepower needed to bring the pressure up to the discharge pressure set point. The process is repeated and the total power needed is the sum of all the stations. Figure 12-6 shows the manner in which total power required increases with spacing of stations. Figure 12-7 based on some data shows horsepower in relation to number of stations. We need to keep in mind that even if we have an increase in required power for a pipeline with less compressor stations the overall cost tends to be lower than many compressor stations with lower horsepower requirements. The installed cost per horsepower will be lower for larger compressor units and thermodynamic performance will be much better pointing to the direction of having fewer compressor stations with larger compressor units. This explains why an economic evaluation has to be done for each project configuration taking into consideration all related information in terms of CAPEX and OPEX.

Since fuel use is related to horsepower, minimum-operating cost is associated with close compressor station spacing, which is logical since maximum
transmission efficiency is obtained at the highest mean line pressure although the pipeline will have larger diameter requiring higher CAPEX. However, the optimum is influenced by two factors, the first that small turbine compressor units are less efficient than large ones (have a higher specific fuel consumption), though this effect is small and almost negligible at unit powers above 20,000 horsepower. A much greater impact though lies in the cost of the stations and it is this capital cost, which declines as the number of stations is reduced (not linear since large stations are proportionately more costly than small ones). This tends to move the optimum spacing away from the minimum distance shown on Figure 12-6.
Every project has to be considered individually because of specific factors and the relationship between CAPEX and OPEX that will differ, but the general conclusion that close spacing results in greater transmission efficiency although may not be the best economic selection.

Experience shows that large units (compressor stations) are more efficient than smaller ones since larger centrifugal compressors and gas turbines have better efficiency. However, the impact of unit outage or failure must be simulated under a transient analysis (Santos, 1997) so that we can define the remaining capacity and therefore establish a maintenance criterion in terms of having standby units or spare equipment to allow quick replacing without affecting the contractual obligation in terms of transportation capacity.

When the preferred solution is found, it must be tested for robustness not only over the range of throughput anticipated but also for all credible upset conditions. Having developed the optimum solution, the result should then be applied to the practical case with elevation changes and other local factors, including the availability of sites, all of which may result in adjustments and minor changes.

12.7.3 Compression power

The next step in the design of a pipeline system is to calculate the maximum power required at the stations and set the design point(s). Typically, a new pipeline system will grow from a low flow condition to the maximum over a period of several years and the decisions on compressors and drivers have to take these changing conditions into account. Growth of flow can be accommodated in several ways. Initially, compressors may only be installed at alternate stations and the intermediate stations built as the growth of flow dictates. Another option is to install one unit at each station location and then add units at the stations as flow increases. On the design phase, the capacity ramp-up will determine the installation schedule for the compressor station and also the additional units that will be necessary at the stations. Hydraulic simulators both in steady and in transient states will help making an accurate design and will guarantee that the project will have a good perform along the operating years without any unexpected situation.

A preselection of the compressor units can also be performed while at the design phase (Santos, 1997). Another important job that can be checked in transient analysis is the pipeline operating points inside the performance maps of the compressors on an early basis so as to allow a proper selection of impellers and number of compressor units as the capacity increases yearly. Different compressor manufacturers can be modeled to check performance and fuel usage. Operations close to surge line or operations that would require recycling would also be identified during transient analysis underlying the importance of doing this kind of simulation on the design phase of a pipeline.

When the compressor design point is decided, the power required from the driver can be calculated using Eqn (11-34).
12.8 Pipeline operations

In the industry supply chain, pipeline operations is an integral part of the transportation between the “upstream-end” preceding it and distribution or the “downstream end” following it. Pipeline operations evolved from being prescriptive (i.e., defined by mandatory requirements) to its current stance of being performance-based driven by risk management principles.

These trends stemmed from competitive forces that decrease operating costs; they also have evolved because of the experience gained from several decades of pipeline operations along with the technologies and applications that developed along the way. These evolutions have given pipeline operators the tools they need to survive under these conditions. The pipeline facilities are mature to the point that many of them have exceeded their originally intended design life of approximately 25 years at the time of conception. Today, most of these facilities continue to operate, partly for economic reasons as they are too costly to replace and also partly because these facilities still remain worthy of continued use (i.e., they are still deemed to be safe). Recognizing this, operating companies continue to extract value from these facilities, but under tremendous scrutiny and heightened awareness of their existence and vulnerabilities.

Current pipeline operation activities have taken on a new dimension of performance. While the basic activities continue, such as, mechanical operations and maintenance of the facilities including line pipe, valves, valve actuators, etc., corrosion prevention and control, pipeline monitoring as well as the focus on safety, today optimization of resources is being considered while still achieving safety, reliability, and efficiency. These challenges become more daunting given the fact that these pipeline systems have expanded and merged, often acquiring systems built by others under different design, construction, and operating philosophies. Further, staff reorganization and attrition saw much of the corporate knowledge and information misfiled or discarded. Some companies remained as free-standing whereas many became a part of a larger corporate entity. To overcome these developments, pipeline operators now strive for standardization in their procedures and compare their performance to industry benchmarks to gauge their performance and identify areas for improvement.

Certain time-dependent defects such as corrosion and environmental concerns started to manifest themselves in unplanned incidents. Development of other infrastructure at or near pipeline right-of-way saw an increase in third-party incidents and close calls. Pipeline regulators too evolved over these times and increased their vigil over the industry but allowed them to formulate their own facilities management programs. Industry-sponsored research programs were founded to better understand the consequential effects of pipeline incidents for risk evaluation.

Consequently, pipeline operation has been transformed toward the following areas of focus:

1. making effective choices among risk-reduction measures;
2. supporting specific operating and maintenance practices for pipeline subject to integrity threats;
3. assigning priorities among inspection, monitoring, detection and maintenance activities; and
4. supporting decisions associated with modifications to pipelines, such as rehabilitation or changes in service.

These focus points require that pipeline operations activities include the following elements (Mokhtatab and Santos, 2005):

1. Baseline assessment and hazard identification
2. Integrity assessment by
   a. In-line Inspection
   b. Hydrostatic Testing
   c. Direct Assessment
   d. Defect Management and Fitness-for-Service
   e. Information Management and Data Integration
   f. Risk Management
3. Integrity Management Programs
4. Operator Qualification and Training
5. Operating Procedures including handling abnormal operating conditions
6. Change Management
7. Operating Excellence.

These elements constitute a broad makeup of pipeline operations. Not only must operators be aware of them but they must also be well versed in their application, continuously improve them and incorporate them into a comprehensive and systematic integrity management plan. Combined, these elements form the basis for directing a prevention, detection, and mitigation strategy for their system.

Pipeline operations will be the longest phase of the life cycle of a pipeline when cost management becomes a high priority. This priority will see pipeline operators performing many scenarios of life extension of the existing assets for enhancing value to its stakeholders.

References

API RP 14E, April 1984. Design and Installation of Offshore Production Platform Piping Systems, fourth ed. American Petroleum Institute (API), Dallas, TX, USA.
13.1 Introduction
Automation has become an increasingly important aspect of gas processing. The amount of specific types of automation is a major decision now in the construction of new plants. With many existing plants having debottlenecked and improved process efficiency through mechanical means, automation is now a focus for further improvements. Automation provides the means for fully utilizing the mechanical capabilities of the equipment at all times and to run the process at its most efficient points in a stable and reliable fashion. A good automation platform can be leveraged to provide the right information at the right time to the right personnel to make the right decisions in a timely manner. Historical data can be collected in virtually any time frame and analyzed statistically. With these historical data, upset situations can be reconstructed and production reports can be automatically generated just to name a couple of the numerous uses for these data. Many processors have upgraded their plants to higher-level control systems such as distributed control systems (DCS). Some of these processors ask, “Where are the benefits?” The correct questions to ask are, “How do I best leverage by automation equipment to maximize benefits?” and “What are the best methods for measuring these benefits?”

This chapter discusses the elements of automating today’s gas processing plants including considerations for instrumentation, controls, data collection, operator information, optimization, information management, etc. The advantages and disadvantages of various automation and control approaches are analyzed in this chapter. Also, strategies for identifying and quantifying the benefits of automation are discussed.

13.2 Early methods of gas plant automation
The earliest gas processing plants were typically controlled manually by opening, pinching, and closing valves to meet their operating requirements. Pneumatic control systems were quickly adopted. These pneumatic control systems allowed the use of proportional, integral, and derivative (PID) controllers to send analog
outputs to control valves to change their opening. As long as a sensor is available as a process variable for the controller, then a set point given by the operator can be targeted automatically. Discrete control could also be accomplished with pneumatic control systems by employing devices to fail with or without an air signal. In most cases, compressed air is used as the pneumatic conveyor, however natural gas is used in some remote operations and hydraulic oils are also employed.

A reliable source of clean, oil-free air is quite important in the operation of pneumatic air systems.

As electronic controls were developed, these systems became the standard, although some elements of the pneumatic control systems are still in use. Most control valves today are still pneumatically operated for new and existing plants even though electronic control valves have been on the market for decades. Electronic controllers were accepted due to their lower price and greater reliability. These controllers included fewer moving parts to maintain. For new installations, wire was now run instead of pneumatic tubing with electronic to pneumatic (I/P for current to pressure) transducers added at the control valves. Process sensors such as temperature, pressure, level and flow indicators were converted to electronic types as available and justified. Electronics also brought widespread use of safety systems such as vibration sensors, burner management, and emergency shutdown systems. Other special controllers such as dedicated surge controllers for rotating equipment and triple modular redundant shutdown devices were developed.

### 13.3 Microprocessor-based automation

#### 13.3.1 Programmable logic controllers

Programmable logic controllers (PLCs) employ programming of a microprocessor to mimic electronic relays. As PLCs gained wider acceptance, the functionality expanded to include PID control and other nondiscrete capabilities. PLCs are still quite effective for batch type operations in the gas processing industry such as solid bed gas dehydration and start-up as well as shutdown sequencing for rotating equipment. This automation platform is often used for smaller new facilities and smaller scope retrofits of existing facilities. A Human Machine Interface (HMI) is highly recommended to monitor the activities of the PLC.

#### 13.3.2 Distributed control systems

The definition of a DCS varies somewhat, but a rather simple definition is a control system method that is spread, of distributed, among several different unit processes on a common computer platform. These systems are typically hardwired and exist within finite boundaries, although wireless means are
becoming accepted, as data security issues are resolved. DCS offers the advantage of centralized control, while retaining the capability of local control. True DCS use localized control, which is in turn controlled by the operator located at a central location. A DCS consists of the following

- Remote control panel or device
- Communications medium
- Central control panel or facility
- Control, interface, and database software

A DCS may be as simple as one PLC remotely connected to a computer located in a field office. Larger systems may be PLC based, but will most likely consist of specially designed cabinets containing all of the equipment necessary to provide input–output (I/O) and communication. One point to consider during the design and specification of a control system is the level of autonomy each node will have in the event of a network or system failure. A true distributed system will allow most remote nodes to operate independently of the central control facility should the facility go off-line or lose communication capability. Each remote node should be able to store the minimum process data required to operate in the event of such a failure. In this manner, costly and potentially disastrous process upsets can be avoided. If the system is performing both monitoring and control of a process or facility, it is referred to as a Supervisory Control and Data Acquisition (SCADA) system. Most systems will transfer data and commands using communication protocols such as Ethernet, or some other open standard, depending upon the DCS vendor (Capano, 2002).

13.3.2.1 Remote control panel
The remote panel for a DCS or SCADA is typically referred to as a Remote Transmission Unit (RTU). A typical RTU contains terminal blocks, input/output modules (both analog and digital), a computer or proprietary processor, and a communications interface. An RTU, depending upon where it is located in the world, can perform both monitoring and control of a given process or processes.

13.3.2.2 Communications medium
The communications medium is a cable or wireless link that serves to connect the RTU to the central control facility. There are several methods of doing this; typically a cable, either a coaxial or twisted pair, is connected between the central control computer and the remote unit, or RTU. It is considered prudent to run two cables, on different routes, between the two in order to increase the reliability of the system. A network operates by taking data from the sending station, or node, and packaging and routing the information to the proper receiving station. The possibility of electrical noise, physical abuse, and software bugs should be considered when implementing a DCS.
13.3.2.3 Central control

The control room is the center of activity and provides the means for effectively monitoring and controlling the process or facility. The control room contains the HMI, a computer that runs specialized software designed for that purpose. There may be multiple consoles, with varying degrees of access to data. In most cases, each operator or manager is given specific rights to allow more or less access and control of the system. The plant superintendent, for instance, may have complete control over his facility, while a technician may only have access to specific data on a particular process. This is done to avoid accidents and process upsets. This scheme also affords a degree of security, ensuring that only properly trained and authorized personnel can operate the various parts of the facility. The HMI presents the operator with a graphical version of the remote process. Depending upon the skill of the operator and the level of sophistication of the interface, the process may be represented by anything from simple static graphics and displays to animation and voice alerts. Most packages afford the operator wide latitude on the design of the interface. The common thread to each system is the input–output (I/O) database.

The database contains all of the I/O defined for that DCS. This does not mean that all process data will be monitored and controlled; it means that only the data defined by the designers to be monitored and controlled will be available to the DCS. This database is a product of detailed evaluation of the process by the designer who typically has the responsibility, with operator input, to design the most effective control schemes for a particular facility. The database is the reference the control software uses to correctly address each remote I/O point. Each database entry corresponds to an entity on the system, whether it is a physical point or an internal, or “soft” point such as an alarm, timer, or screen entity. A disadvantage of the early generations of PLC and DCS was the proprietary communications protocols employed. Some protocols such as MODBUS were adopted as a de facto standard, but had limitations as communications complexity increased.

13.3.3 Standards and protocols

More recently control systems have standardized on protocols made popular through the wide use of personal computers. Ethernet and Object Linking and Embedding for Process Control (OPC) are two examples of highly accepted protocols. Ethernet is used mainly for device-to-device communications while OPC is used primarily for application-to-application communications. Various versions of field buses have been developed to further standardize device-to-device communications. Early DCS was also based on legacy computing platforms such as VMS and the many different UNIX versions. Although Solaris UNIX is still used widely, Microsoft’s Windows platform is gaining popularity (Poe and Harris, 2005).
13.4 Control of equipment and process systems
13.4.1 Gas gathering

Gas gathering systems are typically controlled through pressure control. Since, most gathering systems employ primarily reciprocating compressors (employing a piston or screw), the discussion will focus on control of this type of compression. Rod loadings, maximum discharge temperature, liquid entry as well as minimum and maximum speeds must be considered when operating these machines. It is quite important to prevent liquids from entering the compressor. An adequately sized upstream scrubber with mesh pad should be installed. A level controller to automatically dump liquids is required. The basic purpose of the gas gathering compressors is to keep the wellhead pressures down and the pressure to the downstream facilities, whether gas conditioning systems, liquids recovery facilities, or transmission pipelines, up to a minimum pressure. The speed of the reciprocating machine is the main manipulation to control the suction pressure while the discharge pressure is dictated by flow rate and downstream resistance. Speed range is rather limited for these compressors, but can be manipulated from about 90% to 100% of maximum speed. Other forms of capacity control include pockets and valve unloaders. These may be manual or automated. Rod loading, which is a function of the pressure differential across the compressor (discharge pressure less suction pressure), must not be exceeded or damage will result. Since there is no direct way to limit discharge pressure, a recycle line is installed to allow gas to recycle to the suction and raise the suction pressure. This line is often controlled by a minimum flow controller, but is more effective from an efficiency perspective when controlled off a differential pressure or even a rod load calculation where the control platform can be configured accordingly. The recycle line should be routed after the discharge cooler and discharge scrubber to increase its effectiveness. The temperature rise across the compressor is dictated by the ratio of discharge pressure to suction pressure. High temperatures can warp piping and destroy packings. Discharge temperature should be monitored and the speed reduced or recycle rate increased to keep the temperature below the maximum allowable. The primary medium for provision of cooling the gas discharged from gas gathering systems is air. The following typically controls these coolers:

- Louvers and
- On–off fans
- Multispeed fans
- Variable pitch fans
- Variable speed fans, or gas recirculation.

In most cases, lower temperatures of gas exiting the cooler are preferred for downstream processes. However, temperatures below the hydrate or freezing point are not desired. In these cases, a temperature controller manipulating any of
the above can maintain the desired temperature. Logic may be employed to
determine when fans are turned on, off, or selected to run a discrete speed.

13.4.2 Gas treating
The primary method of gas treating is with chemical absorbents. This process
is analogous to gas dehydration with absorbents. Some differences

1. The objective is acid gas removal and therefore, the contactor outlet gas
   should be analyzed for H2S and/or CO2.

2. Reflux temperature is set lower to maximize the retention of water.

3. In some cases, selective treating of H2S versus CO2 is desired and the
   contactor outlet analyzer can drive a stripper overhead temperature set point,
   which in turn drives the ratio of heat medium used for the reboiler to
   absorbent flow.

Physical absorption is sometimes employed that utilizes a series of flash tanks,
which yield better absorbent regeneration at lower temperatures. Vacuum is often
pulled on the final stage of flash. In this case, the vacuum driver may need to be on
speed control or eductor on flow control to prevent implosion of the vessel.
Hydraulic turbines that use the energy of depressuring to drive recirculation
pumps are quite effective. A “helper” pump is always required to make up the
horsepower deficiency.

13.4.3 Sulfur recovery
The most common sulfur recovery process is the Claus process where one-
third of the H2S must be converted to SO2 for proper stoichiometry. The
theoretical sulfur recovery efficiency drops sharply when the stoichiometry
differs from the desired 2 mol of H2S to 1 mol of SO2. A proper amount of
oxygen, typically in the form of air, must be introduced to the reaction furnace.
A tail gas analyzer is installed after the final reaction stage and before incin-
eration. Two parallel valves, a main valve, and a trim valve, are usually avail-
able on the discharge of the combustion air blower. Ideally, the main
combustion air is manipulated on a feed forward basis as acid gas feed rate and
H2S percentage varies. Feedback from the tail gas analyzer will control the trim
valve. Since the time delay between the air introduction and tail gas analysis
may be significant, a model predictive scheme may improve the ability to
maintain proper stoichiometry.

Reaction furnace temperature must be maintained between a minimum and
maximum temperature. Infrared measurements are quite effective in this service.
Air preheat can be increased or decreased to raise or lower the reaction furnace
temperature. Otherwise, air input requires adjustment. Conversion in the reactors
is a trade-off between equilibrium favored at lower temperatures and kinetics
favored at higher temperatures. Equilibrium usually dictates when catalyst is
fresh while kinetics typically dictates when catalyst is near the end of life. Another factor when the temperatures are low is the converter outlet temperature must be maintained above the sulfur dew point. Converter outlet temperature is controlled primarily by the reheat, which directly controls the converter inlet temperature. Reheat can be classified as direct or indirect. Direct reheat is when hot gas bypasses the waste heat boiler at the outlet of the reaction furnace and is injected at the inlet of each converter. Lowering the converted inlet temperature, when kinetics allow, causes less gas to bypass and gives more shots at additional stages of conversion. Indirect reheat is with a heat medium such as steam or hot oil. Control of the reheat is straightforward with the indirect methods. Condensers are air cooled or water cooled and operated at their lowest temperature to achieve minimum dew point.

13.4.4 Gas dehydration

This section will cover two types of gas dehydration typically employed in gas processing operations. The first is absorption, typically with a glycol, and the second is fixed bed absorption, typically with molecular sieve.

13.4.4.1 Absorption

Dehydration by absorption has several aspects, which require control. These include the following:

- Lean absorbent flow rate and temperature
- Contactor pressure
- Flash tank pressure (where applicable)
- Stripper pressure
- Stripper reboiling
- Stripper reflux

Since the objective of gas dehydration is removal of water from the gas stream, the outlet gas stream should be monitored continuously with a moisture indicator. This indication should be monitored to adjust the flow of lean absorbent and the heat input to the stripper reboiler. The flow of absorbent should be based on a ratio of the gas flow that is corrected based on the moisture indicator reading. This adjustment should take precedence when the moisture is lower than required, since sensible heat will be saved in the stripper in addition to latent heat. The flow of heating medium to the stripper should be on ratio control and corrected based on the moisture indicator reading. This adjustment should take precedence when the moisture is higher than required, since the ability to dry is driven largely by the water content of the lean absorbent. If the lean absorbent is increased without increasing the ratio of heat to flow, then the lean absorbent’s moisture content may not be reduced. Bottom temperature is not a good control basis as it will only indicate the boiling temperature of water at the pressure encountered at the bottom of the stripper. A better indication is a pressure compensated top tray
(or above the packing height) temperature. This has direct correlation with the water content of the lean absorbent. A low absorbent temperature improves its ability to hold water. However, too low a temperature may lead to condensation of hydrocarbons into the absorbent causing foaming. An ideal strategy is to control the temperature of the absorbent about 5–10 °F above the temperature of the inlet gas.

Feed forward strategies that take into account the water content, flow rate, and temperature of the inlet gas to the absorber can also be employed. Model predictive strategies can account for the relative effects of lean absorbent temperature, flow, and stripper heat. A high pressure on the contactor is desirable to increase the contactor capacity and enhance the absorption of water. A back-pressure controller should be employed to maintain a high pressure without causing the relief valve to function. The pressure should be maintained as low as possible when a flash tank exists between the contactor and stripper. A minimum pressure is required to “push” liquids into the contactor while decreasing the demand for stripper reboiler duty. One strategy will raise or lower the flash tank pressure when the liquid level exceeds a desired deadband around the set point. If the liquid rises above the deadband, the pressure is increased. Once the liquid falls within the deadband, the pressure can be slowly decreased. This strategy works best with nonlinear level control, which is always recommended for flash or feed tanks.

The stripper pressure should also be maintained as low as possible to lower the boiling point of the water stripped and allow the flash tank, if installed, to run at a lower pressure. Differential pressure measurements should be installed to indicate the onset of column flooding. When the differential pressure approaches set point the stripper pressure should be raised to alleviate flooding. The reflux condenser should be set at a temperature to attain maximum recovery of entrained glycol without condensing excessive water unless required by environmental considerations.

### 13.4.4.2 Adsorbents

Since drying with fixed bed adsorbents is a multibed process, the main control is cycle and bed switching. The dryer modes are: drying, regenerating, cooling, and standby. Typically, each cycle is set for a fixed time and the beds are cycled through the use of switching valves per this timer.

### 13.4.5 Liquids recovery

#### 13.4.5.1 Condensate stabilization

Condensate is stabilized by stripping light hydrocarbon components in a fractionation tower. Nonlinear level control is recommended for the upstream flash or feed tank to provide a steady feed rate. Due to the nature of the condensate, online analysis is very difficult. Typically a bottom temperature, preferably pressure compensated, is used to control the input of heat to the reboiler. A laboratory
analysis is required to verify the adequacy of the bottom temperature set point. An inferential property predictor can be added to drive the temperature set point in between laboratory updates. The reflux temperature, when employed, should be controlled by a sensitive tray above the feed tray. Tower pressure should be driven as low as possible to enhance separation subject to constraints on an overhead compressor.

13.4.5.2 Refrigeration
Refrigeration is used to achieve the bulk condensation of natural gas liquids. Propane refrigerant is the primary medium used in gas processing. The main control aspects are: compression, compression driver, refrigerant condenser, economizers, and chillers. Both centrifugal and reciprocating compressors are commonly employed in this service with turbine, electric motor, or gas engine drivers.

Lower temperatures are achieved at lower compressor suction pressures subject to surge conditions on centrifugal compressors and rod loading of reciprocating compressors. The suction pressure directly affects the pressure on the chillers.

Level control of the chillers is cascaded to the flow of refrigerant to the chiller. This control is critical to assure that the chiller tubes are covered without carryover of liquids. A scrubber or economizer before each compression stage is necessary to dump liquids whenever encountered. Economizers should be employed on multistage systems. The economizer pressure should be set to accommodate the compressor load and minimize kickback of high stage vapor to lower stages. Refer to the sections in this chapter for compressor and driver control considerations.

13.4.5.3 Cryogenic recovery (turboexpander processes)
Expansion with turboexpanders is now the main process employed for recovering natural gas liquids. Turboexpanders can be controlled for various objectives: inlet pressure, demethanizer pressure, or residue pressure are the most common. Guide vanes are manipulated to control the speed of the expander. A Joules–Thompson (JT) valve is always included to allow rapid unloading of the expander. One split range controller typically operates the guide vanes and JT valve so that the JT valve will open when the manipulation of the guide vanes has been exhausted. The compressor driven by the turboexpander in either inlet compression or residue compression mode requires a recycle valve to maintain a minimum flow for surge protection. Depending on the exact cryogenic recovery processing scheme, additional controls may be required for heat exchanger circuit flow splits, chillers, separator levels, and pressure profiles. Heat exchanger flow splits are typically configured as flow ratios. This ratio may be overridden to prevent “cold spins” or prevent temperatures below the critical temperature in the cold separator upstream of the turboexpander.
13.4.5.4 Demethanizer

The demethanizer is integral to the turboexpansion process. The various feeds to the column are created at multiple points in the process. Side reboil heat sources and sometimes the bottom reboiler heat are integral to the heat exchanger circuit. Seldom are the side reboiler temperatures controlled. Manipulating the heat to the bottom reboiler controls a bottom temperature, preferably pressure compensated. An online chromatograph monitors the methane and/or carbon dioxide content. Ideally, this output would reset the bottom temperature. Demethanizers are good candidates for model predictive control due to the disturbances caused by the side reboilers, inlet flow rates, and inlet compositions. Minimizing the pressure of the demethanizer based on turboexpander constraints and residue compression constraints is a major opportunity for increasing liquids recoveries.

13.4.6 NGL fractionation

NGL fractionation consists of deethanization, depropanization, debutanization, and butane splitting (or deisobutanization). The control schemes for each are analogous. The major control points for these fractionators are reboiling heat, reflux, and pressure. Again, nonlinear level control is recommended for feed tanks and bottom surge levels. Reboiling heat is manipulated to control the bottom composition. The composition is cascaded to a sensitive temperature below the feed tray. Preferably the temperature is pressure compensated.

Reflux flow is manipulated to control the bottom composition. The composition is cascaded to a sensitive temperature above the feed tray. Preferably the temperature is pressure compensated. Minimum reflux schemes are employed to assure that reboiling load is not increased due to excessive reflux and conversely, excessive reboiling leading to greater reflux rates for a given separation. Internal reflux calculations and multivariable control schemes can achieve minimum energy consumption for a given separation.

Pressure should be minimized on these towers subject to constraints such as flooding, condenser temperature, and bottom hydraulics. Flooding is indicated by delta pressure measurements across the tower. Reflux is more difficult at lower temperatures as the available duty of the condenser may be limited. There must be sufficient head on the bottom of the tower to allow liquids to feed downstream towers or satisfy minimum head requirements for pumps. Again, multivariable control schemes handle the pressure minimization issue elegantly.

13.4.7 Centrifugal compressors

Centrifugal compressors (utilizing an impeller to increase the kinetic energy of a vapor) are gaining wider acceptance in a variety of gas processing services including feed, residue, and refrigeration compression. These compressors are typically driven by gas turbines or electric motors, but can sometimes be steam turbines. There are several control considerations for this unit operation.
Upstream liquids removal, surge prevention, suction pressure, discharge pressure, and driver speed control are the primary issues. Upstream liquid removal is accomplished with vessels upstream of the compressor that remove any entrained liquids and automatically dump liquids based on a liquid level. A simple on/off level control scheme is adequate for this purpose.

Surge is caused by excessive head requirements for a given suction pressure. The horsepower delivered by the compressor driver can be reduced to prevent surge or gas can be recirculated from a higher stage of compression. Slowing the compressor driver is usually the most energy efficient means. However, some drivers are limited in their speed range. There are many schemes that use anti-surge or kickback valves to quickly increase the suction pressure by recirculating gas. The most sophisticated will take gas density and head curve characteristics into account for a wide variety of operating conditions. The simplest schemes assure a minimum flow or minimum suction pressure for the compressor. These simpler schemes yield horsepower inefficiencies due to a more conservative approach. Kickback schemes that use cooler gases and minimize the number of stages that gas is recycled are also the most energy efficient.

Suction and discharge pressures can be controlled by adjusting the driver speed, recycling gas, or with throttling valves. Driver speed adjustments are the most energy efficient. Gas turbines will typically have a wide speed range. Steam turbines have a moderate speed range. Electric motors may be constant speed; however variable speed and variable frequency drives are becoming more popular.

### 13.4.8 Centrifugal pumps

Centrifugal pumps are analogous to centrifugal compressors, but are seldom driven by a gas turbine. The same control considerations exist except, of course, the requirement to remove liquids upstream is replaced by a need to remove entrained vapors upstream. The net positive suction head dictates the surge point. Therefore, the level and density of the liquid at the suction of the pump is important.

### 13.4.9 Reciprocating pumps

Reciprocating pumps are typically driven by electric motors, but some applications employ gas engines or steam turbines. These pumps are rather forgiving in their operation and require minimal control. The most common control required is capacity control, since these are positive displacement machines. Variable speed and liquid recycle based on minimum flow or upstream level considerations are the main forms of capacity control.

### 13.4.10 Utilities

The most common utilities found in gas processing plants for process purposes are refrigeration systems, heating (hot oil or steam) systems and cooling water systems.
systems. Refrigeration systems have been covered previously. Hot oil systems employ heaters, mixing tanks, and headers. It is not uncommon for a plant to employ at least two levels of hot oil temperatures. The various temperatures are distributed through separate header systems. Some processes, such as amines and glycols, will degrade when high skin temperatures are encountered. Lower temperature heat medium minimizes reboiler skin temperatures. A common hot oil temperature scheme is to accumulate all heat medium returning from the process in a surge tank. A portion of the hot oil is routed through the heater to the temperature target for the high-temperature header. Enough of the liquid in the surge tank bypasses the heater and is mixed with the right amount of heated oil to achieve the low-temperature header target. Steam systems also typically employ several levels of temperature (or pressure). Boilers produce sufficient steam to satisfy a high-pressure header. Steam users, heat exchangers, and steam turbines, discharge the exhausted steam into a lower pressure header to be reused or into the condensing system for collection and reuse. Steam turbines that exhaust steam into a lower pressure header for reuse at a lower temperature are called topping turbines. Otherwise, the steam turbines are total condensing turbines.

In order to satisfy the balance for the entire steam system for all the temperature (or pressure) levels required, let down valves are employed to route higher-pressure steam into lower-pressure headers based on pressure control. Exhausted system is collected in the lowest pressure header, typically at atmospheric pressure, deaerated, replenished with make-up boiler feedwater to replace losses and boiled again. Process temperature control is usually achieved by simply regulating the flow of heat medium to the heat exchanger.

13.5 Automation applications

A central control room using electronic means to transmit data prompted the advent of applications to reside on the automation platform. These applications are focused on collecting information and using it to operate more profitably (Poe and Harris, 2005).

13.5.1 Data historians

The ability to collect and store a large amount of data on a disk is a key advantage of the microprocessor-based automation platforms. Even so, several vendors have specialized in developing historians to more efficiently and effectively store and analyze the data. Data compression techniques are used to store a maximum amount of information in the minimum space while maintaining resolution of the data. These historian packages come with tools to assist in mining the data, graphing, tabulating, and analyzing statistically. The historical database can be manipulated to automatically generate reports as well.
13.5.2 Asset and performance management

Asset and performance management software has been developed to also tap into the wealth of information that is now available with microprocessor-based automation systems. These solutions include

- Computerized maintenance management
- Work order generation
- Predictive maintenance
- Control loop performance and tuning
- Online equipment health monitoring
- Process performance monitoring

Asset management focuses on maintaining the plant equipment. Inventory management, work order generation, predictive maintenance programs, and turnaround planning can be accomplished with these tools. Many of these packages include hooks into enterprise planning systems. Control loop performance monitoring and tuning packages are available. These applications can determine whether a control loop is experiencing problems with a valve, positioner, or controller tuning for example. Other applications trend the vibration, temperatures, and other key parameters for rotating equipment to determine when a failure is expected to occur.

Process models can be run online to determine how well a plant is performing compared to an expected performance. Heat exchanger fouling, expander and compressor efficiencies as well as tower efficiencies can be calculated and monitored. These packages include data reconciliation features to overcome the problem of how to adjust process models for inconsistent, missing, or bad data.

The opportunities for application of the process performance audit initiative for increased plant profitability in the gas processing and NGL fractionation industries can be defined by analysis of the individual facilities in order to pinpoint the control loops that are the economic drivers of each facility. The following is a list of applications that require minimum control variability and tight adherence to hard spec limits in order to maximize economic performance (Kean, 2000):

**Distillation Towers**

- Feed and reflux flow control loops
- Reflux temperature control loop
- Reboiler temperature control loop
- Reboiler level control loop
- Pressure control loop

**Gas Compressors**

- Flow and pressure control loops
- Surge control loops
- Station recycle control loops
- Gas temperature control loops
Acid Gas Treating Systems

- Stripper reboiler temperature control loop
- Stripper overhead temperature control loop
- Stripper reflux flow control loop
- Contactor and flash drum level control loops
- Hot oil heater fuel and air flow control loops

Steam Boiler Systems

- Steam drum level/feedwater flow control loops
- Steam pressure and fuel flow control loops
- Feedwater heater train control loops
- Combustion air/O\textsubscript{2} control loops

Plant Utility Systems

- Cooling water flow control loops
- Fired heater fuel and air flow control loops
- Refrigeration chiller level control loops

Control loop optimization through the employment of a formal process performance audit by skilled process consultants and control engineers can be an effective route to increased plant profitability. This economic improvement initiative could make the difference between a plant being economically viable, or one that is considered for temporary shutdown or asset disposition (Kean, 2000).

As discussed above, the basic control loop greatly affects plant performance due to the following facts:

- Process optimization requires optimization of the entire process, both hardware and software (Rinehart, 1997).
- The final control loop plays a significant role in process optimization (Rinehart, 1997).
- Control loop optimization reduces process variability and also increases process reliability.
- Optimization of the control loop is an essential step for successful application of advanced control.
- Large economic returns result from proper sizing, selection, and maintenance of the process control equipment.
- Continuous, online monitoring of both loop equipment and loop performance is a key element for achieving the lowest cost of production, while minimizing the life-cycle cost of the processing facility.

13.5.3 Statistical process control

Data from the automation system can be interfaced to statistical process control packages. This software is used to generate run charts, process capability
analyses, process characterization, experimental design, and cause and effect diagrams. This type of information is quite valuable to determine the causes of plant instability and off-specification products. It is also an excellent tool when baselining the plant performance and determining the benefits of improved control. Statistical process control concepts form a foundation for many of the Six Sigma and other quality initiatives when applied to continuous processes.

### 13.5.4 Advanced regulatory control

Advanced regulatory control was made much easier with the advent of microprocessor-based controllers. This control methodology basically turns single input, single output control into multiple input, single output control through the use of cascading controllers, selectors, feed forward, ratios, etc. Shinskey (1996) discusses the variety of control strategies that can be employed in this way. Although not impossible with single loop pneumatic and electronic controllers, the software configuration approach with microprocessor-based controllers superseded the tubing runs, wiring, and other devices necessary to accomplish these strategies with earlier controller forms.

### 13.5.5 Multivariable predictive control

A more elegant and robust form of control is multivariable predictive control. This form of control has been used in the petroleum refining industry since the 1970s and provides true multiple input, multiple output control. Multivariable predictive process control provides a structured approach to managing process constraints, such as limits on valves and rates of change of temperature and pressures. A model for long-range prediction is used to ensure that the constraints upon these variables are not violated. This enables the maintenance of an operating envelope within which the process is constrained. Recently introduced technology that enhances capability in this area includes constrained quadratic programming. In order to determine the optimal set points and constraint values for the controller, an outer optimization can be performed. This optimization can be described as a linear programming (LP) technique which is combined with a steady-state model and a cost function, determining the optimum operating point to be derived from a strategy based on minimum energy usage, maximum throughput, or a balance between these or other objectives.

Figure 13-1 shows a multivariable control strategy for a typical cryogenic demethanizer. A multivariable controller takes advantage of the interactive nature of the process. Key controlled variables are modeled dynamically as a function of key manipulated and disturbance variables. Flows depend on the pressure profile, the compression horsepower available, and the efficiency of the turboexpander. When more horsepower is available such as at night and in cooler weather, then the flow can be increased or the pressure on the demethanizer can be lowered to increase NGL recovery for a given flow demand. Control of the NGL quality, typically for methane or carbon dioxide in ethane, becomes more difficult as tower
FIGURE 13-1

Multivariable control on a typical cryogenic demethanizer.
pressure is adjusted. The multivariable controller can determine the correct heat input as tower pressure is adjusted to maintain maximum recovery and product quality. Pressure compensated temperature is a key element of this strategy.

Other controlled variables can be minimum and maximum flows; pressures, temperatures, levels, speeds, etc. Demethanizer pressure, flow splits, reboiler flow, plant inlet pressure are some manipulated variables while disturbances such as residue pipeline pressure, inlet flow, and inlet composition are considered.

In the above manner, a “team” of key controllers are pushing the plant to its optimum operating point at all times. Controllers are effectively decoupled thereby making adjustments similar to an automobile on cruise control. In many regards, the objectives of statistical process control are achieved.

Figure 13-2 shows the beneficial effects of advanced controls. First the variability of the process is reduced by the ability to control closer to set points with the model predictive capability. Once variability is reduced, then the process can be pushed closer to the operating constraints where maximum profitability is attained. This is an area where operators hesitate to operate due to the possibility of exceeding a constraint. With advanced process control, set points can be put closer to these constraints without fear of exceeding the constraints.

Examples of the benefits from the application of model predictive have been widely documented (Alexander et al., 2007).

13.5.6 Optimization

The next level of automation is optimization. For simpler optimization problems, LP techniques can be employed. Nonlinear techniques may be warranted for
more complex optimization opportunities where linear methods would miss significant benefits. With the advances in computing power and optimization mathematics, online, rigorous optimization has become a reality. Again, petroleum refining is leading the way, but several gas processing applications have been developed. Optimization systems must rely on a multivariable predictive control system. As the robustness of the multivariable predictive control system improves, so does the effectiveness of the optimization system. With the business focus that optimization brings, multivariable predictive control developments should address the requirements of operations management. Table 13-1 details recent technology developments that have improved operations management and the benefits of multivariable predictive control applications. These advancements have enhanced basic process operations. This enables staff to be reployed and provide more valuable operations functions that provide increased efficiency, reliability, reduced cost, improved quality, ideal staffing, and responsiveness to changing business requirements.

Networking and Web capabilities continue to impact advanced applications. Some companies have central support centers or have contracted application service providers. Automated testing can be easily monitored remotely. Performance-monitoring technologies and tools provide clear benchmarks of expected performance. Updates and improvements can be applied remotely. These capabilities reduce project and support costs and further improvement the economics of multivariable predictive control initiatives. Mixed integer programming and enhanced state-space methods are now employed to improve models and decrease the impacts of unmeasured disturbances. New developments in sensor technology supply new valuable, low-cost process information. Methods for property estimation and predicting the behavior of multiphase and complex reaction systems are also improving. Multivariate statistical methods continue to progress. Visualization technology may be a key to making the models transparent to the end user.

<table>
<thead>
<tr>
<th>Development</th>
<th>Technology</th>
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</thead>
<tbody>
<tr>
<td>Nonlinear controllers</td>
<td>Neural nets and other empirical modeling techniques</td>
</tr>
<tr>
<td>Automated process testing</td>
<td>Multistep, random frequency testing</td>
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<td>Performance monitoring</td>
<td>Metrics for model adequacy</td>
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<td>Remote implementation and support</td>
<td>Web-enabled applications</td>
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<tr>
<td>Adaptivity</td>
<td>Background testing and model identification</td>
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<tr>
<td>Control system embedded</td>
<td>Powerful processors coupled with programming efficiencies</td>
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13.5.7 Leveraging automation

In order to get the most benefit, the automation platform should be leveraged to its maximum capability. The platform itself is about 80% of the cost for about 20% of the potential benefits. A multitude of additional benefits are there for a fraction of the platform cost by adding applications as shown in Figure 13-3. Include adequate instrumentation up front for grassroots plants and during retrofits of existing plants to support the advanced applications. It is much more cost-effective to add instrumentation during construction rather than adding it later. Use control system consultants rather than relying solely on equipment manufacturers. Many equipment manufacturers are penalized for including more instrumentation than their competitors. For control system upgrades, do not just replace in kind; replace and enhance. A DCS gives limited additional value when used simply as a panel board replacement (Poe and Harris, 2005).

13.5.7.1 Automation upgrade master plans

Many gas processors have developed automation upgrade master plans for existing facilities where the relative benefits of each unit and the value of enhancements are estimated. In this way, the area with the most benefits can bear the majority of the initial cost of the platform. Additions to the platform, once the infrastructure is in place, typically cost less per I/O than the initial installation. Master plans are quite effective when the gas processor teams with an automation consultant. The consultant can share their past experience and advise of future developments that should be considered.
13.5.7.2 Determining the benefits

The benefits of a fully leveraged automation system are numerous and include the following:

- Control room consolidation (reduced manpower)
- Plant reliability (uptime)
- Plant stability (better efficiency due to fewer process upsets)
- Maintenance management (reduced inventories)
- Product quality (fewer off-specification penalties or give-away of over processed product)
- Continuous constraint pushing (increased throughput and recoveries as well as energy savings)
- Optimization (energy savings and increased recoveries)

When approached appropriately, these benefits are easily quantified. The key is measuring through baselining and monitoring the results through proper metrics.

Baselining

A good baseline is a crucial element of determining benefits. Historical data are necessary to derive a good baseline. The historical data should include the primary measurement as well as any factors that should be used to normalize the result. For instance, ethane recovery should be normalized for inlet gas flow and composition as well as ambient temperature. Data during upset periods and outliers should be discarded. True process variability is measured best with frequent data capture on the order of every minute. High-quality baselining requires an excellent understanding of the operation in addition to statistical analysis. In addition to measuring mean performance of a process, baselining will reveal the amount of variability in the process and the source of the variability. Reducing variability is one of the main benefits of automation. With reduced variability, the process has freedom to shift to a more beneficial operating point.

Statistical analysis

Statistical analysis is crucial in determining the benefits of automation. A good statistical analysis will give the most accurate assessment of process performance under changing conditions. Mean or average performance is always the final measure between before-and-after performance. Again, the mean should be evaluated on a normalized basis so that the performance is assessed on a fair or equivalent basis. Total liquids or even total ethane recovered is not a normalized evaluation. Barrels per million cubic feet (MMCF) of inlet is a better metric, but does not take into account changes in inlet composition. Recovery levels take into account inlet composition, but not process capability under variable inlet rates or ambient conditions that provide more mechanical horsepower availability. Standard deviation gives a measure of the amount of variability for a normal (bell-shaped) response curve. Depending on the process and the sources of variability anywhere from one-half to two standard deviation improvement is reasonably expected. One way to check this is to compare best performance or process
capability (entitlement) to the mean performance. This difference can be compared to the standard deviation to give an idea of the maximum standard deviation improvement is possible. Since standard deviation gives an absolute number and is difficult to compare to the mean, coefficient of variation, which is the standard deviation divided by the mean gives a relative measure of standard deviation. In other words, the coefficient of variation gives a percentage improvement potential. All responses are not normal or bell-shaped. Kurtosis is a measure of the skewness of the response curve and should be considered when evaluating standard deviation. A common example of an abnormal response is when a process runs close to its physical limits to the upside or downside and far from its physical limits to the opposite side. Elimination of upset conditions can sometimes shift a response curve close to normality. Multivariable control packages also include the capability to further identify the dynamic performance of a process with tools such as autocorrelation, power spectrum, and Fourier transforms. Determining financial benefits in the face of uncertainty is always a challenge. Even though we may have a good gauge on process improvement potential from the standpoint of increased product, the value of these improvements is dependent on economic conditions like pricing, feedstock availability, and product demand. Risk management tools, such as Monte Carlo simulations, can be used to establish the range and certainty within these ranges of financial benefits over the variety of conditions anticipated.

Performance improvement initiatives
Many companies are adopting Six Sigma and other statistically founded performance improvement initiatives. Automation can greatly enhance the effectiveness of Six Sigma and other performance improvement initiatives. Conversely, the rigor of these performance improvement methodologies is useful in documenting the benefits of automation. Take Six Sigma as an example of how performance improvement methodologies and automation complement each other. The four phases of Six Sigma include the following: measure, analyze, improvement, and control (Breyfogle, 1999).

Measurement can sometimes be difficult and painful without automation. With automation and data historians, the measurement and analysis task becomes much easier. After the collection of data, the process capability assessment, run charts, cause and effect matrices, etc., can be developed to assess the performance improvement opportunity and where to focus the effort. The many elements of automation can be the enabling technology for the implementation and control phases of Six Sigma. Stabilizing the process is always a key element of performance improvement and is enabled by automation, especially advanced controls. The control phase pretty well speaks for itself.

13.6 Condensate stabilizer case study
The following case study reviews an application of some of the advanced control features and the benefits. This case study is for an actual condensate stabilization process. The main reason for advanced process control on this unit was for quality
purposes. Condensate product must meet a Reid Vapor Pressure (RVP) specification as defined by the customer. The previous method of production relied on laboratory sampling to verify the RVP, which was infrequent. Online analyzers are available but can be expensive. Instead of an online analyzer, the important quality measurement was derived through inferential means. The APC solution provided a stable plant and reduced the RVP variation.

BG Tunisia implemented multivariable predictive control on their gas condensate production system at the Hannibal plant in Sfax, Tunisia. The aims of the project were to maximize condensate yields, improve the stability of the condensate stabilization process, and ensure quality limits for the product are adhered to at all times. BG Tunisia’s Hannibal terminal processes gas at 5.4 million standard cubic meters per day (MMSCM/D) with the condensate column operating at a typical rate of 550 l/m controlling RVP to a maximum limit of 12 psia. Condensate can be sold as crude oil and is therefore more valuable than the alternative natural gas liquids products (Hotblack, 2004).

Figure 13-4 shows a detailed schematic for the condensate process. The liquid hydrocarbon condensate is brought into the plant from the slug catcher and preheated in the stabilizer feed/bottoms exchanger E-201. The preheated feed is flashed in the stabilizer feed drum D-201. Liquid from the feed drum is further preheated through the feed/bottoms exchanger and fed to the stabilizer T-201 on local flow control. The feed drum level controller maintains level by manipulating the feed to the drum. The flashed vapor is sent to the high-pressure fuel gas system. A boot is provided on the feed drum to separate any entrained water, which is sent to the warm flare drum under level control. Condensed hydrocarbon liquid from the warm separator in the nitrogen rejection unit (NRU) is fed to the NRU liquids flash drum D-202 under level control—via the warm gas/liquid exchanger in the NRU. The flashed vapor is sent to the high-pressure fuel gas system and liquid from the drum is fed to the stabilizer.

The condensate stabilizer T-201 reduces the vapor pressure of the condensate by removing the lighter components. It is a stripper column with 24 trays. Liquid from the stabilizer feed drum is fed at the midpoint on tray 9 and liquid from the NRU liquids flash drum is fed to the top tray. Overhead vapor from the stabilizer is sent to the low-pressure fuel gas system through a back-pressure control valve that maintains the tower pressure to set point under the action of pressure controller.

The bottom part of the tower is divided into two sections by a baffle—the baffle does not extend to the very bottom of the tower, so there is some mixing between the two sections. Liquid from the bottom tray flows into the section that is preferentially pumped through the fired stabilizer reboiler H-201. The two-phase stream from the reboiler is returned to the other compartment where the liquid is separated as stabilized product and the vapor flows up the tower to provide stripping action. The glycol/condensate exchanger provides additional heat duty if required.

Stabilized condensate leaving the stabilizer is cooled in the condensate cooler E-202 and by exchange with cold inlet feed condensate in the exchanger E-201, and is sent to the condensate storage tanks TK-201 A/B.
The main objectives of the condensate stabilizer connoisseur controller are as follows:

- Control RVP to an operator specified target value
- Enforce any specified unit operating constraints at all times
- Stabilize the unit operation

The controller uses a real-time online estimate (inferential) of RVP. By controlling more tightly to the target RVP value, flashing of stabilized condensate in the storage tanks and during tanker transfer can be minimized thus reducing losses. The product specification for RVP is 10–12 psia. Figure 13-5 shows the before-and-after trends of key process variable. The baseline assessment before APC was a mean RVP of 12.25 psia with one standard deviation of 0.43. After APC was installed, the standard deviation dropped to 0.26, enabling control to 11.5 psia with 95% confidence on the 12-psi limit (Hotblack, 2004).

Constrained optimization drives the process to an optimum without violating process constraints such as the RVP limit imposed by the condensate purchaser.

A hybrid RVP sensor using both first principle and neural network technology provides a continuous measurement for control. Laboratory samples taken on a periodic basis are integrated into the control scheme improving accuracy and ensuring product quality is maintained.
FIGURE 13-5
Reid Vapor Pressure (RVP) at BG Tunisia’s plant, before-and-after trends of key process variable (Hotblack, 2004).

The inferential estimate is implemented directly within the controller as a hybrid of a first principles model based on the Antoine vapor pressure equation and a radial basis function neural network. Both the first principles model and the neural network require the stabilization column base temperature and top vapor pressure. Compensation for process drift and process measurement error is via feedback from the laboratory analysis. A weekly lab sample is inserted as a minimum for this purpose.

References
Alexander, B., Harris, S., Poe, W., Toillion, M., Whatley, L., March 11–14, 2007. Maximizing helium and NGL recovery at DCP midstream’s Panhandle region gas plants with model predictive control. Paper Presented at the 86th GPA Annual Convention, San Antonio, TX, USA.
Canney, W., November 2004. Advanced process control powers developments in operations management. Oil Gas J. 102 (42), 50.
Hotblack, C., September 2004. BG Tunisia’s advanced process control improves condensate product stability. World Oil 225 (9).
14.1 Introduction

A well-designed gas processing plant is not successful until it is operating safely and profitably. This requires a smooth start-up as well as a productive and safe environment for the operations. In order to sustain the operation, good maintenance practices are required. Troubleshooting is invariably required to detect and fix issues that occur when the performance of engineered equipment degrades.

An exhaustive coverage of gas plant operations topics would require an entire book. Therefore, the objective of this chapter is to provide an introduction into commissioning and start-up, control room management, maintenance, and troubleshooting techniques applicable to gas processing plants. Commissioning and start-up activities covered include mechanical completion and pre-commissioning, control systems testing, initial start-up procedures, process commissioning, and performance testing. Control room management topics focus on areas addressed by the U.S. Department of Labor Occupational Safety & Health Administration (OSHA) and Department of Transportation. These specific areas include roles and responsibilities, process safety management (PSM) focusing on hazards and operability reviews, fatigue mitigation, alarm management, and training for operational excellence.

14.2 Commissioning and start-up

14.2.1 Mechanical completion and precommissioning

Mechanical completion is the stage of a project between equipment installation and process commissioning. Components of the plant are proven to be mechanically ready for their process duty (Horsley, 1997). This stage of plant commissioning may include some specific performance tests on individual components of the plant. Precommissioning is the stage where process materials are introduced to the plant and any preliminary issues are resolved.

The object of mechanical completion is to prove that each installed component is ready for commissioning. This stage includes:

- checking that equipment is installed correctly,
- proving that equipment operates acceptably for commissioning, and
- demonstrating that instruments, controls and safety system work.
Much of the work to start the checkout of equipment can be completed at the factory before delivery to the site, and the completion of this work should be encouraged as it has the potential to save cost and time if vendor equipment does not perform initially. Activities performed at the factory acceptance test can include leak tests, initial operational runs, control system checkout, cleanliness checks, and test fitting of internals.

Inspection of the equipment is carried out before testing and may require specialists depending on the type of equipment. This inspection assures that the equipment meets the specifications as installed and is complete. Also, assurance is made that the drawings pertaining to this equipment is in as-built status.

Safety and operability checks are keys at this stage including adequacy of access, handrails, drains, vents, and so on. Orientation of valves, safety devices, and fire protection equipment are critically inspected. Tests of the component parts are witnessed as acceptable in meeting specifications.

All electrical grounding is checked. Wiring is checked for continuity and insulation resistance. Motors are inspected for correct direction of rotation.

Before pipes and equipment are placed in service, it is important that good cleaning procedures be conducted to help ensure a successful and trouble-free start-up. Often initial plant commissioning is marred by foreign materials that have been left inside pipe work and then find their way to pumps and other equipment, causing significant damage and schedule delay as a result of rectification work. The following methods of cleaning are common to fulfill this task: cleaning by blowing (air or nitrogen), steam blowing, cleaning via flushing (water), chemical cleaning, use of a pipeline integrity gage (pig), mechanical cleaning, and visual inspection.

Hydrostatic tests are conducted to prove pipe integrity; however, many potential leak path points will still exist in an untested state. The object of a full system leak test is to test all leak potential suitably and practically before the introduction of process and hazardous chemicals.

The necessary laboratory facilities should be verified and laboratory personnel trained to carry out the planned tests. Laboratory testing is critical for plant start-up to monitor raw materials and assure that products meet specifications before delivery to customers.

Sampling points and analyzer equipment must be ready once the plant is in operation.

All required raw materials and chemicals need to be on-site in adequate quantities to satisfy the commissioning and initial start-up time frame.

Finally plant operators should have completed all training to assure that they understand the plant systems, start-up procedures, and response to abnormal situations.

14.2.2 Control systems testing

Control systems testing, as with all commissioning activities, requires careful planning and coordination. Electrical, instrumentation, process control, process
engineering, and computer specialists are required and must work closely together.

Precommissioning of field instruments and factory acceptance testing of the distributed control system (DCS), programmable logic controllers (PLCs), and other computer-based equipment are independent activities. This fact makes it even more critical for a comprehensive site acceptance test of the control systems.

Operator screens, databases, configuration, applications, and communications between computer components can be checked during the factory acceptance test. DCS simulators are available that can generate a wide range of representations of the process for testing, validation, and training purposes. Simple models can be built through the use of transfer functions or physical flow networks. The control system configuration can be tested in an off-line simulation environment before or after the system has been installed in the field. Special attention should be paid to emergency shutdown and equipment protection logic.

All issues found during factory acceptance testing must be corrected and retested before final acceptance.

Motor control systems, switchgear and other electrical equipment are often installed in a different location from the DCS.

Inspecting the installation of instruments is critical. Power and air supplies, mounting as well as associated piping, tubing, and wiring should be checked thoroughly. The flow lines should be pressure tested for leakage especially at fittings. Primary sensors must be consistent with the type and range of transmitters. Control and solenoid valves must fail safe and positioners must operate correctly.

Instrumentation is often calibrated by the manufacturer, but needs to be checked and recalibrated as necessary on-site. When recalibrating, it is important to:

- calibrate over the entire range and set any bias correctly according to specifications,
- use the correct process medium when calibrating analyzers, and
- check all configurable settings.

Full loop tests of communication with field devices to the final actuator will be checked during site acceptance testing.

Documentation of the instrumentation and control systems must reflect as built conditions. Typical documents include:

- process and instrumentation diagrams,
- loop diagrams,
- interconnection schematics,
- database tables,
- sequential flow diagrams,
- wiring diagrams,
- termination rack layouts,
- tag number reference lists, and
- system and operation manuals.
14.2.3 Initial start-up procedures

When possible, “safe” chemicals are introduced into water, steam, and air to simulate closely the unit in actual operation and to provide an indication of how the plant will perform when the process chemicals are introduced before the main commissioning and start-up event takes place. During commissioning execution, all safety-related systems are checked rigorously, including confirmation of alarm activation points by means of manipulation of the actual process variables; confirmation of the operation of all control system software trip logic by various means, including variability of the process conditions both manually and via the control system; confirmation of all hardwired emergency shutdown systems by various documented operational means; and confirmation of the operation and control of all DCS sequences, including full testing of all failure monitoring. All operating modes are tested, including starting up, scheduled and emergency shutdown, and the actions required after a loss of site services, including power and instrument air. The emergency power generation system must be fully tested.

14.2.4 Process commissioning

Process commissioning can begin after precommissioning tests are complete and all identified issues are corrected. Plant equipment is placed into normal operation during this phase.

Process commissioning should not commence until the final checklist is reviewed and verified. This list will include:

- all safety equipment are in place and functional,
- emergency procedures are known,
- medical and first aid considerations have been arranged,
- all inspection equipment has been removed,
- test orifices and blanks removed,
- vents and overflows are not obstructed,
- strainers cleaned and replaced,
- methanol injection system for hydrate control,
- liquid seals are at correct levels,
- emergency and normal lighting systems are functional,
- rotating equipment guards are in place,
- source of ignition have been removed,
- access for emergency vehicles is clear, and
- all operations documentation is in place.

The typical sequence of commissioning will include:

1. Utilities including:
   a. electrical service,
   b. water and water treatment,
   c. compressed air,
d. nitrogen,
e. flare system,
f. backup fuel gas,
g. fire water system, and
h. steam system.

2. Storage tanks for raw materials
3. Catalysts and chemicals
4. Process equipment
5. Product storage
6. Emergency power generator
7. Any ancillary equipment.

A system for recording times and stages of the process where samples are taken along with their results will accelerate start-up. Many more analyses than during normal operation are required and the lab should be staffed accordingly or third party labs contracted. The lab results will provide information for initial troubleshooting of equipment and instrumentation.

It is very important that scrubbers, separators, and filters are working satisfactorily before introducing material into the main plant equipment.

Raw materials may be consuming faster than normal during initial commissioning due to charging the system and any process issues encountered. The purchasing department should be advised and ready to schedule delivery for additional raw materials.

Often during start-up, material form the process may not meet product specifications. Considerations should be in place for recycling, disposal, or arrangements with buyers for off-specification material. Also, any environmental impacts should be discussed with the regulatory agencies prior to encountering these issues.

The main objective of process commissioning is to achieve stable and reliable operating conditions while meeting product specifications. Most efficient or higher throughput operating conditions will be explored later.

Often a decision must be made to shutdown and correct a problem or whether to continue running and work through or around an issue. This decision must be made carefully taking into account safety and hazards. Most process accidents occur during shutdown and start-up than during normal, continuous operation. However, a specific issue’s consequences on safety and hazards may outweigh start-up and shutdown safety concerns.

A list of items to resolve before commissioning is complete will be compiled and handled by the construction contractor. Commissioning is complete once the plant can achieve its intended purpose and can be taken over by operations. The intended purpose and conditions of commissioning completion will be stated in the construction contract. Other items that would improve the operability of the plant that were not included in the construction scope should be documented as future enhancements to the plant.
Following commissioning, the start-up and shutdown procedures should be reviewed and modified based on learnings from the commissioning process. For NGL production plants, the cooldown procedure must be strictly followed to avoid thermal stress in the processing equipment, such as the brazed aluminum exchangers.

14.2.5 Performance testing

Performance tests are carried out with the plant as close to design conditions as possible. If the plant or process has been licensed, the licensor usually guarantees the process. Guarantee tests are witnessed by the licensor or engineering contractor as well as plant operators and engineering personnel. Raw material and other conditions are generally outlined in the contract and efforts are made to replicate these conditions as closely as possible. Duration of the tests, performance criteria, conditions causing postponements and interruptions, methods of measurement and analysis, and allowable tolerances will usually be stipulated.

Performance criteria may include:

- plant capacity,
- quality of products,
- yields,
- utilities consumption, and
- quantity of emissions.

It is important that a mass and energy balance can be calculated or reconciled within a reasonable tolerance before the performance tests are conducted. All data required to calculate the metrics dictated by the performance criteria should be accurately captured by the plant historian or otherwise recorded at an acceptable frequency. After collection and processing of the data, some corrections may need to be made for off-design conditions such as feed gas composition, feed temperature and pressure, ambient temperature, and so on.

There are often contractual consequences for not meeting the performance criteria including resolution, monetary penalties, and liquidated damages. In cases where performance criteria are not part of the construction contract such as the plant is designed by the operating company, performance tests are usually conducted to document the plant capabilities.

14.3 Control room management

The activities conducted in control rooms must assure compliance with all federal, state, and local regulations. The most important regulations apply to personnel and process safety as this affects the well-being of plant employees and the public. Two federal regulations that have the great applicability to the

The U.S. Department of Transportation Pipeline and Hazardous Materials Safety Administration (PHMSA) sponsored 49 CFR Parts 192 and 195. This regulation has been adopted by natural gas processing facilities in the United States as well as other parts of world. This regulation addresses human factors and other aspects of control room management for pipelines where operators use supervisory control and data acquisition (SCADA) systems. Under the final rule, affected pipeline operators must define the roles and responsibilities of operators and provide operators with the necessary information, training, and processes to fulfill these responsibilities. Operators must also implement methods to prevent operator fatigue. The final rule further requires operators to manage SCADA alarms, assure control room considerations are taken into account when changing pipeline equipment or configurations, and review reportable incidents or accidents to determine whether control room actions contributed to the event (US DOT, 2009).

The OSHA first issued 29 CFR Part 1910.119 in 1992. Even prior to the official release major oil and gas companies began compliance with the intent of the regulation. The methods used to comply with this regulation are now well established.

14.3.1 Roles and responsibilities

According to 49 CFR, each operating company must define the roles and responsibilities of an operator during normal, abnormal, and emergency conditions. To provide for an operator’s prompt and appropriate response to operating conditions, an operating company must define each of the following:

- An operator’s authority and responsibility to make decisions and take actions during normal operations;
- An operator’s role when an abnormal operating condition is detected, even if the operator is not the first to detect the condition, including the operator’s responsibility to take specific actions and to communicate with others;
- An operator’s role during an emergency, even if the operator is not the first to detect the emergency, including the operator’s responsibility to take specific actions and to communicate with others; and
- A method of recording operator shift changes and any handover of responsibility between operators.

14.3.2 Process safety management

In each industry, PSM applies to those companies that deal with any of more than 130 specific toxic and reactive chemicals in listed quantities; it also includes
flammable liquids and gases in quantities of 10,000 pounds (4535.9 Kg) or more (US DOL, 2000).

Subject to the rules and procedures set forth in OSHA’s Hazard Communication Standard (29 CFR 1910.1200(i)(1) through 1910.1200(i)(12)), employees and their designated representatives must be given access to trade secret information contained within the process hazard analysis (PHA) and other documents required to be developed by the PSM standard.

The key provision of PSM is PHA—a careful review of what could go wrong and what safeguards must be implemented to prevent releases of hazardous chemicals. Covered employers must identify those processes that pose the greatest risks and begin evaluating those first.

Methodologies allowed are:

- What-if;
- Checklist;
- What-if/checklist;
- Hazard and Operability Study (HAZOP);
- Failure Mode and Effects Analysis;
- Fault Tree Analysis (FTA); or
- An appropriate equivalent methodology.

The PHA shall address:

- The hazards of the process;
- The identification of any previous incident that had a likely potential for catastrophic consequences in the workplace;
- Engineering and administrative controls applicable to the hazards and their interrelationships such as appropriate application of detection methodologies to provide early warning of releases. (Acceptable detection methods might include process monitoring and control instrumentation with alarms, and detection hardware such as hydrocarbon sensors.);
- Consequences of failure of engineering and administrative controls;
- Facility siting;
- Human factors; and
- A qualitative evaluation of a range of the possible safety and health effects of failure of controls on employees in the workplace.

The PHA shall be performed by a team with expertise in engineering and process operations, and the team shall include at least one employee who has experience and knowledge specific to the process being evaluated. Also, one member of the team must be knowledgeable in the specific PHA methodology being used. At least every five years after the completion of the initial PHA, it shall be updated and revalidated by a team meeting these requirements to assure that it is consistent with the current process.

The employer shall establish a system to promptly address the team’s findings and recommendations; assure that the recommendations are resolved in a timely
manner and that the resolution is documented; document what actions are to be taken; complete actions as soon as possible; develop a written schedule of when these actions are to be completed; communicate the actions to operating, maintenance, and other employees whose work assignments are in the process and who may be affected by the recommendations or actions. Employers shall retain PHAs and updates or revalidations for each process covered by this section, as well as the documented resolution of recommendations for the life of the process.

14.3.3 Hazard and operability study

A HAZOP is the predominate methodology used to comply with OSHA’s PSM requirements.

The HAZOP analysis technique uses a systematic process to (1) identify possible deviations from normal operations and (2) ensure that appropriate safeguards are in place to help prevent accidents. The HAZOP technique uses special adjectives (such as “more,” “less,” “no,” etc.) combined with process conditions (such as speed, flow, pressure, etc.) to systematically consider all credible deviations from normal conditions. The adjectives, called guide words, are a unique feature of HAZOP analysis.

The characteristics of HAZOP include:

- A systematic, highly structured assessment relying on guide words and team brainstorming to generate a comprehensive review and ensure that appropriate safeguards against accidents are in place
- Typically performed by a multidisciplinary team
- Applicable to any system or procedure
- Used most as a system-level risk assessment technique
- Generates primarily qualitative results, although some basic quantification is possible.

The most common uses of HAZOP include identifying safety hazards and operability problems of continuous process systems, especially fluid and thermal systems, and reviewing procedures and sequential operations.

There are some limitations of this technique. They are:

1. HAZOP requires a well-defined system or activity. The HAZOP process is a rigorous analysis tool that systematically reviews each part of a system or activity. To apply the HAZOP guide words effectively and to address the potential accidents that can result from the guide word deviations, the analysis team must have access to detailed design and operational information. The process systematically identifies specific engineered safeguards (e.g., instrumentation, alarms, and interlocks) that are defined on detailed engineering drawings.

2. HAZOP is time-consuming. The HAZOP process systematically reviews credible deviations, identifies potential accidents that can result from the deviations, investigates engineering and administrative controls to protect
against the deviations, and generates recommendations for system improvements. This detailed analysis process requires a substantial commitment of time from the facilitator and other subject matter experts, such as crew members, engineering personnel, equipment vendors, etc.

3. HAZOP focuses on one-event causes of deviations. The HAZOP process focuses on identifying single failures that can result in accidents. If the objective of the analysis is to identify all combinations of events that can lead to accidents of interest, more detailed techniques should be used such as FTA.

The procedure for performing a HAZOP analysis consists of the following five steps:

1. Define the system or activity. Specify and clearly define the boundaries for which hazard and operability information is needed.
2. Define the problems of interest for the analysis. These may include health and safety issues, environmental concerns, and so on.
3. Subdivide the system or activity and develop deviations that will be individually analyzed. Once subdivisions are defined, then apply the HAZOP guide words that are appropriate for the specific type of equipment in each section.
4. Conduct HAZOP reviews. Systematically evaluate each deviation for each section of the system or activity. Document recommendations and other information collected during the team meetings, and assign responsibility for resolving team recommendations.
5. Use the results in decision-making. Evaluate the recommendations from the analysis and the benefits they are intended to achieve. The benefits may include improved safety and environmental performance or cost savings. Determine implementation criteria and plans.

14.3.4 Layer of protection analysis

Layer of protection analysis (LOPA) is a methodology for hazard evaluation and risk assessment. On a sliding scale of sophistication and rigor, LOPA lies between the qualitative (characterized by methods such as HAZOP and what-if) and the quantitative (characterized by methods using fault trees and event trees) end of the scale.

LOPA helps the analyst make consistent decisions on the adequacy of the existing or proposed layers of protection against an accident scenario (see Figure 14-1). LOPA results support the following:

- Assigning priorities to recommendations
- Developing safety requirement specifications for safety instrumented systems (SISs), which is a necessary step in complying with ANSI/ISA-84.00.01 (2004).
This decision-making process is ideally suited for coupling with a company’s risk-decision criteria, such as those displayed in a risk matrix. LOPA is a recognized technique for selecting the appropriate safety integrity level of your SIS according to the requirements of standards such as ANSI/ISA-84.00.01.

LOPA is a semiquantitative methodology that can be used to identify safeguards that meet the independent protection layer (IPL) criteria. While IPLs are extrinsic safety systems, they can be active or passive systems, as long as the following criteria are met (Summers, 2002):

- Specificity: The IPL is capable of detecting and preventing or mitigating the consequences of specified, potentially hazardous event(s), such as a runaway reaction, loss of containment, or an explosion.
- Independence: An IPL is independent of all the other protection layers associated with the identified potentially hazardous event. Independence requires that the performance is not affected by the failure of another protection layer or by the conditions that caused another protection layer to fail. Most importantly, the protection layer is independent of the initiating cause.
- Dependability: The protection provided by the IPL reduces the identified risk by a known and specified amount.
- Auditability: The IPL is designed to permit regular periodic validation of the protective function.

Examples of IPLs are:

- Standard operating procedures,
- Basic process control systems,
Alarms with defined operator response,
SISs,
Pressure relief devices,
Blast walls and dikes,
Fire and gas systems, and
Deluge systems.

LOPA should start from the point where the hazards have been identified and it is thus complementary to HAZOP. This use of LOPA often results in a second, in-depth analysis of a hazard scenario by a different team of people, which may challenge the HAZOP team’s understanding of failure events and safeguards (Brennan, 2012).

14.3.5 Fatigue mitigation

Each operating company must implement the following methods to reduce the risk associated with operator fatigue that could inhibit an operator’s ability to carry out the roles and responsibilities the operator has defined:

- Establish shift lengths and schedule rotations that provide operators off-duty time sufficient to achieve 8 h of continuous sleep;
- Educate operators and supervisors in fatigue mitigation strategies and how off-duty activities contribute to fatigue;
- Train operators and supervisors to recognize the effects of fatigue; and
- Establish a maximum limit on operator hours of service, which may provide for an emergency deviation from the maximum limit if necessary for the safe operation of a pipeline facility.

As a result of the U.S. Chemical Safety and Hazard Investigation Board (CSB) investigation of the 2005 BP Texas City incident, the CSB issued several recommendations including the development of an American National Standards Institute (ANSI) Standard that develops Fatigue prevention guidelines for the refining and petrochemical industries that, at a minimum, limit hours and days of work and address shift work.

ANSI RP 755 provides guidance to all stakeholders (e.g., employees, managers, supervisors) on understanding, recognizing, and managing fatigue in the workplace. Owners and operators should establish policies and procedures to meet the purpose of recommended practice (ANSI/API RP-755, 2010).

This document was developed for refineries, petrochemical and chemical operations, NGL plants, and other facilities such as those covered by the OSHA PSM Standard, 29 CFR 1910.119. This document is intended to apply to a workforce that is commuting daily to a job location.

It has been well documented that excess workplace fatigue can be a risk to safe operations. In the past, it was thought that simply placing limits on the hours of service would adequately address the risk of fatigue. However, over the last
several years, a broad international consensus has emerged that the better way to
manage fatigue risk is through a comprehensive fatigue risk management system
(FRMS) that is integrated with other safety management systems as necessary.
ANSI/API RP 755 is based on the FRMS approach.

The FRMS should be based on sound science and recognize operational is-
sues, and includes consultation with key stakeholders in the development and
implementation of the local application of the FRMS. The FRMS should also
include a process to review and enhance the FRMS, as needed, with a goal of
continuous improvement.

The development of the FRMS should address the following subjects which
are discussed in the document:

- Positions in a facility covered by the FRMS
- Roles and responsibilities of those covered by the FRMS
- Staff—workload balance assessments
- Safety Promotion: training, education, and communication
- Work environment
- Individual risk assessment and mitigation
- Incident/near-miss investigations
- Hours of service guidelines
- Callouts
- Exception process
- Periodic review of the FRMS to achieve continuous improvement.

ANSI/API RP 755 provides guidance on the number of consecutive days that can
be worked before a minimum amount of time off is required. The guidance ap-
plies to 8-, 10-, and 12-h-shifts and addresses normal operations, outages, and
extended shifts.

14.3.6 Alarm management

According to 49 CFR Parts 192 and 195, each operating company using a
SCADA system must have a written alarm management plan to provide for
effective operator response to alarms. An operating company’s plan must include
provisions to (ANSI/ISA–18.2, 2009):

- Review SCADA safety-related alarm operations using a process that ensures
  alarms are accurate and support safe pipeline operations;
- Identify at least once each calendar month points affecting safety that have
  been taken off scan in the SCADA host, have had alarms inhibited,
  generated false alarms, or that have had forced or manual values for periods
  of time exceeding that required for associated maintenance or operating
  activities;
- Verify the correct safety-related alarm set point values and alarm descriptions
  at least once each calendar year, but at intervals not to exceed 15 months;
Review the alarm management plan required by this paragraph at least once each calendar year, but at intervals not exceeding 15 months, to determine the effectiveness of the plan;

- Monitor the content and volume of general activity being directed to and required of each operator at least once each calendar year, but at intervals not to exceed 15 months, that will assure controllers have sufficient time to analyze and react to incoming alarms; and
- Address deficiencies identified through the implementation.

ANSI/ISA–18.2 (2009) addresses the development, design, installation, and management of alarm systems in the process industries. Alarm system management includes multiple work processes throughout the alarm system life cycle. This standard defines the terminology and models to develop an alarm system, and it defines the work processes recommended to effectively maintain the alarm system throughout the life cycle.

The stages of the Alarm Management life cycle are:

1. Alarm Philosophy
2. Identification
3. Rationalization
4. Detailed Design
5. Implementation
6. Operation
7. Maintenance
8. Monitoring and Assessment
9. Management of Change
10. Audit.

14.3.6.1 Alarm philosophy

Basic planning is necessary prior to designing a new alarm system or modifying an existing system. Generally, the first step is the development of an alarm philosophy that documents the objectives of the alarm system and the processes to meet those objectives. For new systems the alarm philosophy serves as the basis for the alarm system requirements specification document.

The philosophy starts with the basic definitions and extends them to operational definitions. The definition of alarm priorities, classes, performance metrics, performance limits, and reporting requirements are determined based on the objectives, definitions, and principles. The schemes for presentation of alarm indications in the human-machine interface (HMI), including use of priorities, are also set in the alarm philosophy, which should be consistent with the overall HMI design.

The philosophy specifies the processes used for each of the life cycle stages, such as the threshold for the management of change process and the specific requirements for change and is maintained to ensure consistent alarm management throughout the life cycle of the alarm system.
The development of the alarm system requirements specification is included in the philosophy stage of the life cycle. Most of the specification is system-independent and can be the basis for determining which systems most closely meet the requirements. The specification typically goes into more detail than the alarm philosophy and may provide specific guidance for system design.

### 14.3.6.2 Identification
The identification stage is a collection point for potential alarms proposed by any one of several methods for determining that an alarm may be necessary. These methods are defined outside of this standard so the identification stage is represented as a predefined process in the life cycle. The methods can be formal such as PHA, safety requirements specifications, recommendations from an incident investigation, good manufacturing practice, environmental permits, process and instrumentation diagram (P&ID) development, or operating procedure reviews. Process modifications and operating tests may also generate the need for alarms or modifications. Some alarm changes will be identified from the routine monitoring of alarm system performance. At this stage, the need for an alarm has been identified and it is ready to be rationalized.

### 14.3.6.3 Rationalization
The rationalization stage reconciles the identified need for an alarm or alarm system change with the principles in the alarm philosophy. The steps can be completed in one process or sequentially. The product of rationalization is clear documentation of the alarm, including any advanced alarm techniques, which can be used to complete the design.

Rationalization is the process of applying the requirements for an alarm and generating the supporting documentation such as the basis for the alarm set point, the consequence, and corrective action that can be taken by the operator.

Rationalization includes the prioritization of an alarm based on the method defined in the alarm philosophy. Often priority is based on the consequences of the alarm and the allowable response time.

Rationalization also includes the activity of classification during which an alarm is assigned to one or more classes to designate requirements (e.g., design, testing, training, or reporting requirements). The type of consequences of a rationalized alarm, or other criteria, can be used to separate the alarms into classes as defined in the alarm philosophy.

The rationalization results are documented, typically in the master alarm database (i.e., an approved document or file), which is maintained for the life of the alarm system.

### 14.3.6.4 Detailed design
In the design stage, the alarm attributes are specified and designed based on the requirements determined by rationalization. There are three areas of design: basic alarm design, HMI design, and design of advanced alarming techniques.
The basic design for each alarm follows guidance based on the type of alarm and the specific control system.

The HMI design includes display and annunciation for the alarms, including the indications of alarm priority.

Advanced alarming techniques are additional functions that improve the effectiveness of the alarm system beyond the basic alarm and HMI design. These methods include state-based alarming and dynamic prioritization.

### 14.3.6.5 Implementation

In the implementation stage, the activities necessary to install an alarm or alarm system and bring it to operational status are completed. Implementation of a new alarm or a new alarm system includes the physical and logical installation and functional verification of the system.

Since operators are an essential part of the alarm system, operator training is an important activity during implementation. Testing of new alarms is often an implementation requirement.

The documentation for training, testing, and commissioning may vary with classification as defined in the alarm philosophy.

### 14.3.6.6 Operation

In the operation stage, the alarm or alarm system is active and it performs its intended function. Refresher training on both the alarm philosophy and the purpose of each alarm is included in this stage.

### 14.3.6.7 Maintenance

In the maintenance stage, the alarm or alarm system is not operational but is being tested or repaired. Periodic maintenance (e.g., testing of instruments) is necessary to ensure the alarm system functions as designed.

### 14.3.6.8 Monitoring and assessment

In the monitoring and assessment stage, the overall performance of the alarm system and individual alarms are continuously monitored against the performance goals stated in the alarm philosophy. Monitoring and assessment of the data from the operation stage may trigger maintenance work or identify the need for changes to the alarm system or operating procedures. Monitoring and assessment of the data from the maintenance stage provides an indication of the maintenance efficiency. The overall performance of the alarm system is also monitored and assessed against the goals in the alarm philosophy. Without monitoring an alarm system is likely to degrade.

### 14.3.6.9 Management of change

In the management of change stage, modifications to the alarm system are proposed and approved. The change process should follow each of the life cycle stages from identification to implementation.
14.3.6.10 Audit

In the audit stage, periodic reviews are conducted to maintain the integrity of the alarm system and alarm management processes. Audits of system performance may reveal gaps not apparent from routine monitoring. Execution against the alarm philosophy is audited to identify system improvements, such as modifications to the alarm philosophy.

Audits may also identify the need to increase the discipline of the organization to follow the alarm philosophy.

14.3.7 Training

The objective of training should be achievement of operational excellence by developing and implementing a plan that addresses the reliable operations (see Figure 14-2). Training should close the gap between human performance and system complexity.

Operator training should address four key components. They are:

- Human by reducing the probability of error,
- Equipment by eliminating or reducing deterioration and failure, and extend the mean time between failure,
- Design by minimizing equipment maintenance, increase mean time to repair and ensure equipment performance objectives and standards are achieved, and
- Process by developing tools to reduce variability in plant output.

![Operational Excellence](image)

**Figure 14-2**
Operational excellence key components (Source: GP Strategies Corporation).
Isolated attempts to improve performance in any one of these components will provide benefits, but best results are obtained when all components work well together.

An operating company’s program must provide for training each operator to carry out the roles and responsibilities defined by the operating company. In addition, the training program must include the following elements:

- Responding to abnormal operating conditions likely to occur simultaneously or in sequence;
- Use of a computerized simulator or noncomputerized (tabletop) method for training operators to recognize abnormal operating conditions;
- Training controllers on their responsibilities for communication under the operator’s emergency response procedures;
- Training that will provide a controller a working knowledge of the system, especially during the development of abnormal operating conditions; and
- For operating setups that are periodically, but infrequently used, providing an opportunity for operators to review relevant procedures in advance of their application.

14.4 Maintenance

“Maintenance” is a term generally used to define the routine activities to sustain standards of performance throughout the in-service, or operational, part of the asset life cycle. In doing this, the maintenance policy designer needs to take account of a range of factors. These include the complexities of operating environment, the available resources for performing maintenance, and the ability of the asset to meet its current performance standards.

Cost effectively maximizing the overall equipment effectiveness (OEE) should be the objective of any plant maintenance program. Availability, performance rate (also often described as utilization), and quality rate all factor into OEE. Maximum availability is achieved by reducing planned downtime, setup downtime, and unplanned downtime for each piece of equipment. Maximum performance rate is experienced by minimizing slowdowns and reductions below maximum capacity. Quality rate is maximized when off-specification product is eliminated at maximum yield.

The design and installation of equipment as well as how it is operated and maintained affect the OEE. It measures both efficiency (doing things right) and effectiveness (doing the right things) with the equipment (Mather, 2005).

\[
OEE = A \times PE \times Q
\]  

where A is availability of the machine, PE is performance efficiency, and Q is the quality rate or yield.
Availability is the proportion of time the machine is actually available out of time it should be available.

\[
\text{Availability} = \frac{\text{Planned production time} - \text{unscheduled downtime}}{\text{Planned production time}} \tag{14-2}
\]

\[
\text{Planned production time} = \frac{\text{Gross available production time}}{C_0} \tag{14-3}
\]

Gross available hours for production include 365 days per year, 24 h per day, 7 days per week. Planned downtime includes vacation, holidays, and insufficient raw materials. Availability losses include equipment failures and changeovers indicating situations when the line is not running although it is expected to run.

Performance Efficiency

\[
\text{Performance Efficiency} = \frac{\text{Amount of products actually produced}}{\text{Planned production}} \tag{14-4}
\]

Quality (Yield)

\[
= \frac{\text{Amount of product processed} - \text{Amount of product not recovered}}{\text{Amount of product processed}} \tag{14-5}
\]

For Availability of 95%, Performance Efficiency of 90%, and Quality of 99%, the result is an OEE rating of 85%.

A good maintenance program will focus on the root cause of equipment failure. Otherwise, the maintenance department is spending its time reacting to the symptoms of the problems.

Often before equipment failure we observe poor performance. Prior to poor performance there are signals that the equipment has deteriorated.

Safety and environmental compliance play their part in creating the drive for maintenance activities, particularly given the changing legal and regulatory frameworks around these two areas; in some industries, they are even the principal drivers. However, for most businesses the goal remains that of maximum value from their investment. This means getting the maximum performance possible from the assets, for the least amount spent.

14.4.1 Types of maintenance

14.4.1.1 Breakdown maintenance

In this type of maintenance, no care is taken for the machine, until equipment fails. Repair is then undertaken. This type of maintenance could be used when the equipment failure does not significantly affect the operation or production or generate any significant loss other than repair cost. However, an important aspect is that the failure of a component from a big machine may be injurious to the operator. Hence breakdown maintenance should be avoided.
14.4.1.2 Preventive maintenance
Preventive maintenance is performed periodically (cleaning, inspection, oiling, and retightening). This method retains the healthy condition of equipment and prevents failure by delaying deterioration through inspection or equipment condition diagnosis. It is further divided into periodic maintenance and predictive maintenance.

Condition monitoring plays a dual role in the maintenance process for regulating preventive maintenance applications and, as stress increases, alerting to impending failure.

14.4.1.3 Periodic or time-based maintenance
Periodic maintenance consists of inspecting, servicing, cleaning, and replacing parts to prevent sudden failure and process problems on a predetermined frequency.

14.4.1.4 Corrective maintenance
Corrective maintenance improves equipment and its components so that preventive maintenance can be carried out reliably. Equipment with design weaknesses must be redesigned to improve reliability or maintainability. This happens at the equipment user level.

14.4.1.5 Predictive maintenance
Predictive maintenance is a method in which the service life of important parts is predicted based on inspection or diagnosis in order to use the parts to the limit of their service life. Compared to periodic maintenance, predictive maintenance is condition-based maintenance. It manages trend values, by measuring and analyzing data about deterioration and employs a surveillance system, designed to monitor conditions through an online system.

Most recently, predictive maintenance (also known as condition monitoring) has been leading the way to additional savings over preventive maintenance. The use of real time or portable instruments, such as vibration monitors, thermography, ferrography, and so on, has been effective at recognizing the symptoms of impending machine failure. The major benefit is the availability of an earlier warning, from a few hours to a few days, which reduces the number of breakdown “catastrophic” failures.

Predictive maintenance is usually implemented concurrently with preventive maintenance and targets both the warning signs of impending failure and the recognition of small failures that begin the chain reaction that leads to big failures (i.e., damage control).

Predictive maintenance tasks are established to try to detect the warning signs that indicate the onset of failure, thus allowing for actions to be taken to avoid the failure.

14.4.1.6 Proactive “life extension” maintenance
Proactive maintenance has now received worldwide attention as the single most important means of achieving savings unsurpassed by conventional maintenance
techniques. The approach supplants the maintenance philosophy of “failure reactive” with “failure proactive” by avoiding the underlying conditions that lead to machine faults and degradation. Unlike predictive/preventive maintenance, proactive maintenance commissions corrective actions aimed at failure root causes, not just symptoms. Its central theme is to extend the life of mechanical machinery as opposed to (1) making repairs when often nothing is broken, (2) accommodating failure as routine and normal, or (3) preempting crises failure maintenance in favor of scheduled failure maintenance.

14.4.2 Enterprise asset management systems

Enterprise asset management (EAM) systems provide the ability to capture, manipulate, and analyze, historical failure data. The benefits are the ability to highlight the causes for poor performing assets, provide the volume and quality of information for determining how best to manage the assets, and informing decisions regarding end-of-life and other investment points.

The implementation of these products, when bought for these reasons, often focuses on optimizing processes to capture the dynamic data on asset failures, which is then used throughout the system. Maintain, repair, and operate style inventory management algorithms, for example, use this information as one of the key inputs to determine minimum stocking levels, reorder points, and the corresponding reorder quantities.

As a result, analysts often find themselves recommending and analyzing activities of not only maintenance, but also other areas of asset management, namely those of asset modification and operations.

14.4.3 Reliability centered maintenance

If an asset management program is aimed at maximum cost-effectiveness over an asset’s life, then it must look at the management of critical failures. By definition, this approach is centered on the reliability of the asset or reliability centered.

The role of the maintenance manager can be defined as formulating cost-effective asset management programs, routine activities, and procedures to maintain standards of performance through reducing the likelihood of critical failures to an acceptable level, or eliminating them.

Many EAM implementations are of limited value, because even with well-controlled and precise business processes for capturing data, some of the critical failures that will need to be managed may not yet have occurred. Reliability Centered Maintenance (RCM) facilitates the creation of maintenance programs by analyzing the four fundamental causes of critical failures of assets, namely:

- poor asset selection,
- asset degradation over time,
- poor asset operation, and
- human errors.
The RCM Analyst needs to analyze all of the reasonably likely failure modes in these four areas, to an adequate level of detail. Determining the potential causes for failures in these areas, for a given operating environment, is in part informed by data, but the vast majority of the information will come from other sources.

Sources such as operators’ logs are strong sources for potential signs of failure, as well as for failures often not found in the corporate EAM. Equipment manufacturers’ guides are also powerful sources for gleaning information regarding failure causes and failure rates.

The factors that decide the lengths that an RCM Analyst should go to collect empirical data are driven by a combination of the perceived risk and limitations set on maintenance policy design by commercial pressures. Even when all barriers are removed from the path of the RCM Analyst, they are often confronted or faced with an absence of real operational data on critical failures.

Among the areas where EAM systems provide substantial benefits is through driving out inefficiencies in business processes. Through the capture, storage, manipulation, and display of historical transactional data, companies can take great leaps forward in the level of efficiency with which they execute maintenance programs.

This is where RCM style methodologies contribute to the EAM or computerized maintenance management system. By providing the content that the system needs to manage, they are ensuring that the right job is being executed in the right way.

### 14.5 Troubleshooting

Troubleshooting is a method of finding the cause of a problem and correcting it. The ultimate goal of troubleshooting is to get the equipment back into operation. This is a very important job because the entire production operation may depend on the troubleshooter’s ability to solve the problem quickly and economically, thus returning the equipment to service. Although the actual steps the troubleshooter uses to achieve the ultimate goal may vary, there are a few general guidelines that should be followed. There are often cases where a familiar piece of equipment or system breaks down.

The general guidelines for a good troubleshooter to follow are:

- Use a clear and logical approach
- Work quickly
- Work efficiently
- Work economically
- Work safely and exercise safety precautions.

The secret to success in troubleshooting is gathering as much information as possible about a process operation. The largest source of this information is the daily observations made by the shift operators (Lieberman, 1991). To identify a
malfuction by process equipment, the technician must first understand the normal function of that equipment (Lieberman, 2011).

14.5.1 Troubleshooting steps
A general troubleshooting process consists of the following:

- Verify that a problem actually exists.
- Isolate the cause of the problem.
- Correct the cause of the problem.
- Verify that the problem has been corrected.
- Follow up to prevent future problems.

14.5.1.1 Verify that something is actually wrong
A problem usually is indicated by a change in equipment performance or product quality. Verification of the problem will either provide indications of the cause if a problem actually exists or prevent the troubleshooter from wasting time and effort on nonproblems caused by the operator’s lack of equipment understanding. It should not be accepted that something is wrong without personally verifying the failure. Some steps to take at this stage include:

- Be as specific and defining as possible in stating the problem that is occurring.
- Always check to ensure equipment is lined up for normal operation.
- Analyze the performance of the equipment to make sure it actually has a failure and is not simply reacting to an external condition.
- Determine if the failure is total or if the equipment is operating with degraded performance.

To verify that there actually is a problem, the troubleshooter must use all available means of information. This includes the equipment operator, equipment indications and controls, and technical documentation about the equipment or system. Contacting the equipment operator should be the first action taken. The operator usually can supply many of the details concerning the failure incident. To get the most information, the troubleshooter should ask probing questions. Some examples are:

- What are the operator’s indications of the trouble?
- How did the operator discover the trouble?
- What were the conditions at the time the trouble first occurred?
- Is the trouble constant or intermittent?

14.5.1.2 Identify and locate the cause of the trouble
Trouble is often caused by a change in the system. A thorough understanding of the system, its modes of operation, and how the modes of operation are designed to work, the easier it will be to find the cause of the trouble. This knowledge
allows the troubleshooter to compare normal conditions to actual conditions. Some recommendations to consider include:

- Start the troubleshooting log with as much background information as possible and document each adjustment and its results.
- Note how readings are affected by all modes of operation and switch lineups.
- Be sure to observe all gages, meters, and other indicators as to how they are responding due to the problem.
- Always note if an adjustment has no effect on the symptom; this will help eliminate possible causes later on.
- Determine if the trouble has slowly developed (i.e., drift) or if it is a sudden failure.
- Perform control manipulation with care since detrimental effects can occur to associated equipment or components within the failed equipment.

Troubleshooting should be a series of small logical steps, each one chosen to show a result leading to discovery of the problem or problems. Always use the functional block diagram to ensure all the possible functions are checked. Include functions such as detectors, switches, cables, meters, wiring, connectors, piping, filters, and regulators. Check all pressures, flows, inputs, and outputs associated with the areas of probable faulty functions. If an abnormal reading is obtained, the equipment setup used to obtain the reading and the reading itself should be rechecked.

Some examples of useful graphic documentation are:

- Panel graphics
- Loop diagrams
- Piping and instrumentation diagrams
- Block diagrams
- Wiring diagrams
- Schematic diagrams.

14.5.1.3 Correct the problem

It is very important to correct the cause of the problem, not just the effect or the symptom. This often involves replacing or repairing a part or making adjustments. Never adjust a process or piece of equipment to compensate for a problem and consider the job finished; correct the problem!

14.5.1.4 Verify that the problem has been corrected

Repeating the same check that originally indicated the problem can often do this. If the fault has been corrected, the system should operate properly.

Check all the functions that have been affected by the failure. Although the equipment has been repaired and is now functioning, all operations must be checked and verified. The information obtained in this step can also aid in troubleshooting next time by providing some baseline information.
14.5.1.5 Follow up to prevent further trouble
Determine the underlying cause of the trouble. Suggest a plan to a supervisor that will prevent a future recurrence of this problem.

14.5.2 Troubleshooting documentation
The troubleshooting log provides a valuable source of information from which the troubleshooter can draw on the experience of past troubleshooting efforts to quickly restore the equipment to service. Problems, symptoms, corrective actions, modifications, and preventive maintenance actions all should have entries that can be referenced at a later date. Many companies require their maintenance personnel or engineering staff to maintain historical data on equipment used within their facilities.

It can lead the troubleshooter to the solution to a problem that has not occurred in years and causes troubleshooting efforts to move slowly as the troubleshooter checks every possibility. Additionally, documentation of recurring problems can provide the horsepower needed to get the right part or the engineering solution necessary to not only fix the problem, but also correct it.

The equipment history/troubleshooting log is an ideal place to keep the records necessary to establish and maintain a common problems list. The purpose of the common problems list is to provide the troubleshooter with a ready reference of past problems and their corrective actions. It is from this list that quick fixes can be taken.

If a problem occurs on a regular or routine basis, it should be put on the common problems list. This can be referred to at the beginning of a troubleshooting problem so the quick fixes can be tried. This can save the troubleshooter valuable time when troubleshooting. Troubleshooters or technicians need to be careful of what is placed on the common problems list. If something occurs once, it is not necessarily a common problem. The problem should be listed in the history section and should not be put on the common problems list until it occurs again. This is because the tools used for troubleshooting are only as good as their application. If the common problems list is too long and cumbersome, it cannot be used effectively.

14.5.3 Instrumentation
Instrument and automation technicians are constantly challenged to keep instrumentation loops and input/output (I/O) working at peak efficiency while using the least possible time to do it (Dewey, 2001).

The first indication of a control loop problem often comes from the operator.

The first step is to measure the 4–20 mA signal, either by breaking the loop connecting in series with a digital multimeter (DMM), or by using a Milliamp (mA) clamp meter and verifying the loop current value. If the loop current
measured is not as expected, there are three likely causes: broken/disconnected/shorted wires, a bad loop power supply, or faulty instrumentation.

If no problem is found in the wires, use a DMM to check the loop power supply. If the power supply shows no output, use the 24-V loop power function of the meter to substitute for it; if the loop then works properly, the source of the problem is obvious.

If the wiring and the power supply both check out, it is time to check the transmitter. If you have a loop calibrator, process calibrator, or multifunction clamp-on meter, use its mA simulate mode to substitute for the transmitter. If the loop performs as requested, the problem lies with the transmitter; if not, it is elsewhere.

If a final control element (valve positioner, etc.) is suspected, use the mA source/simulate mode on a DMM to feed a signal into it while watching the local indicator for a response.

If the problem is not a dead loop but an inaccurate one, likely possibilities include a bad I/O card on the PLC or DCS, or a bad final control element (current to pressure, I/P, on a valve positioner, etc.). It is usually best to start by doing a field check of the transmitter, local or remote indicator or final control element.

For a final control element, use a clamp-on meter to measure loop current and compare the value to the local position indicator on the valve or other final control element. Relay that information to the operator to verify findings.

In the case of a measurement loop, use the clamp meter to measure loop current, then check with the operator to see how well the value indicated on the control panel agrees with the actual loop current. This will give a quick check on the PLC or DCS I/O card that handles that particular loop. It is also possible to use the meter’s mA source/simulate mode to send a known signal to the control room. As before, compare the value as read by the operator to the actual current in the loop.

Some loops show random fluctuations or intermittent faults that tend not to happen while a technician is watching. The solution here is to use a clamp meter with a scaled mA output. In this mode, the meter measures the current in the loop without breaking the circuit, and produces an identical and isolated mA output. Feed that output to a DMM with a logging function; by allowing the DMM to record over time, any disturbance will be recorded.

The mA process clamp meter can be used as an accurate signal source to check the operation of I/O cards on PLCs and distributed process control systems (DCSs). For 4–20 mA input cards, disconnect the process loop and use the meter’s mA source mode to feed in a known signal value (4.0 mA for zero, 12 mA for 50 percent—using the meter’s 25 percent step function, and 20.0 mA for 100 percent) and compare it to the value shown on the operator’s readout. Voltage input cards (1–5 V or 0–10 V) are checked in a similar way, using the meter’s voltage source function.

As part of preventive maintenance programs, mA clamp meters can be used for periodic in-field checks of electronic valve positioners. The general method
is to set the meter to the 4–20 mA source/simulate mode and connect it to the input terminals of the valve positioner. Set the meter to output 4 mA and wait for the positioner to settle; then vary the current in small increments between 4.0 mA and approximately 3.9 mA, while feeling the valve stem with your free hand to check for any sign of movement. Adjust for zero movement between these two current settings by using the zero adjustment on the positioner.

Next, increase and decrease current from 4 mA to approximately 4.1 mA. Insure that the valve stem just begins movement above the ~4.1 mA setting and fully closed at 4 mA. Span can be checked similarly, by setting the meter at 20 mA, ~19.9 mA, and ~20.1 mA, and linearity can be checked by using the meter’s 25 percent step function.

To check a loop isolator, apply an mA input signal to the device and measure its 4–20 mA output using the clamp-on current measuring function.

Variable frequency drives are used to power motors, blowers, and fans in process applications as well as conveyor systems and machine tools. Control inputs are generally voltage (1–5 V or 0–10 V) or current (4–20 mA). An mA process clamp meter can feed in a signal to simulate a normal input while the technician observes the result.

While not classified as loop calibrators, today’s mA process clamp meters boast accuracies of 0.2 percent, and can be used for quick calibration checks, while cutting down on the number of instruments needed.

For example, checking a process transmitter on the bench normally requires (aside from a pump and separate pressure standard) a loop power supply and an instrument for reading the transmitter’s 4–20 mA output. But with today’s mA process clamp meters, it is possible to both power the transmitter and read the output.

14.5.4 Process troubleshooting

Process troubleshooting includes (Lynch and Pittman, 2000; Lieberman, 2008):

■ modeling the original plant design,
■ simulating the expected plant performance for the current inlet conditions and pressure profile,
■ collecting and reconciling field data,
■ developing a model matching the reconciled field data,
■ analyze any process equipment problems, and
■ make changes in the plant to achieve acceptable plant performance and verify the performance change with the models.

The most effective method to resolving inconsistencies in the data is to close the heat and material balance across each piece of equipment. Data reconciliation methods will consider redundant data, gross errors in the data, and expected accuracy of each measurement.
14.6 Turnarounds

Plant turnarounds are an important aspect of operating a gas plant. A turnaround is a scheduled period when the plant is shutdown. These efforts should be well planned to determine the duration, cost, and expected benefits in improving plant reliability, efficiency, and safety. Ideally, a turnaround should result in the plant returning to peak performance levels when it comes back online and should allow it to function efficiently until the next shutdown.

Typical operations eased during a turnaround and the focus of plant personnel shifts to repairs, maintenance, cleaning, and inspection. Plant personnel are usually supplemented by many outside contractors working around the clock.

Detailed planning and coordinated execution are the most critical elements of executing a successful turnaround. While speed is extremely important, safety as always is the highest priority. Many resources used by outside contractors may be inexperienced with the hazards found in a gas plant. Specific safety training and safety monitoring is essential to assure a turnaround with no accidents. Due to the extended working hours, worker fatigue should be monitored as part of the safety program.

Some equipment repair issues cannot be resolved because they cannot be addressed while the plant is operational. When these issues affect the overall asset utilization, then a turnaround is warranted to fix the equipment or preclude a more serious safety or efficiency issue. Other maintenance issues often best addressed during a turnaround are overhauls, process piping modifications, process equipment internals, molecular sieve replacement, sulfur reactor catalyst, among others. Sometimes safety regulations or warranty requirements might dictate a turnaround.

Turnarounds are extremely expensive especially in the upstream oil and gas industry. In addition to the expenses of executing the maintenance activities, there is a loss of production and the profits from the sale of the plant’s products during the period. Supplies, rentals, and overtime are some expenses incurred in addition to the contractors’ fees. Often turnarounds will be planned when production demands may be seasonally low and preferred contractor crews are available to limit production impacts as well as assure safe and effective execution.

Typically, a turnaround manager will be assigned and all of the plant personnel will participate by contributing their expertise of administration, operations, maintenance, engineering, quality assurance as well as health, safety, and environment.

Regardless of the short-term expense, a safe and effective turnaround will prove very beneficial in the long term.

References


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15.1 Introduction

Modeling is an accepted design and performance improvement tool of gas processing and transmission facilities. The use of steady-state models is universally accepted in all stages of the design and operation of gas processing plants. Forms of dynamic simulation have been used for many years, but rigorous first principles dynamic simulation has been confined to use by specialists and control engineers where using models based on transfer functions are incapable of representing the nonlinearities and discontinuities in systems. Only since the late 1990s has dynamic simulation become a more generally accepted tool to be used by process engineers and control engineers alike. The software available today enables process engineers with some process control knowledge and control engineers with some process knowledge to build dynamic models fairly easily. The limitation in using dynamic simulation is no longer that the difficulty in configuration, but rather that the implementation time for a dynamic model is in the order of two to four times as long as the time needed to implement a steady-state model. Frequently, a consultant is engaged to develop the model and one or more engineers of an operating company or engineering company would use the model to run the needed studies.

This chapter discusses the areas of application of dynamic process modeling and modeling considerations, both general and for specific equipment, frequently used in gas processing plants. The use of dynamic models in specific gas processing units will be analyzed. Some case studies will be presented to illustrate the use and impact of dynamic simulation on the design and operation of gas processing and transmission plants. Finally, the use of dynamic simulation as part of operator training simulators is discussed.

15.2 Areas of application of dynamic simulation

The areas of application have been divided into two large groups. The plant design group highlights applications that are most frequently used by engineering
companies whereas the plant operation group is mostly used by operating companies.

15.2.1 Plant design

Dynamic models have several applications in plant design including controllability and operability, safety analysis, start-up procedure definition, control system definition, control and safety system checkout, operator training, and advanced process control (APC) model development. Quite often the dynamic simulation will detect serious design flaws that would delay plant start-up and add significant cost to the project to correct after start-up.

It is important to realize that a dynamic model can be reused for various applications in the design of a plant. The dynamic model will evolve as the design and the project evolves. The most detailed and rigorous model will be used just prior to plant commissioning and beyond (Brown and Hyde, 2001).

15.2.1.1 Controllability and operability

Decisions taken very early in the design phase of a new plant or a revamp of a plant can have a significant impact on the controllability or operability of that plant. If the design calculations only use steady-state process simulation, the decision to employ a novel design is often a trade-off between the advantage the novel design brings and the potential controllability and operability issues. A dynamic model will shed more light on the problems that can be expected and lets control engineers devise adapted control strategies to mitigate or remove controllability problems. The use of dynamic simulation will increase the adoption of novel designs and ultimately the efficiency of new or revamped plants.

As these issues need to be analyzed in the early development of a process, the models will necessarily be simpler than the models used when the design has been completed. At this stage, concepts are tested and one would not expect quantitative results from a dynamic model but rather indications of process stability or control feasibility.

15.2.1.2 Safety analysis

A lot of effort goes into ensuring a plant will be safe for the operators and the environment. Huge liabilities are associated with the safety of a plant. Some dynamic models are standard practice for any design but surprisingly in other areas the use of dynamic models is still limited.

For virtually any plant operating at high pressure, depressuring studies will be a standard part of the engineering phase. Depressuring studies are performed to analyze the behavior of pressure and temperature when the pressure of a plant that has been stopped is let down. The depressuring study defines the necessary flaring capacity that is needed and the results may also impact the choice of materials of vessels. Typically, if the pressure letdown of a vessel results in too low temperatures, carbon steel will require replacement with stainless steel to prevent the metal becoming too brittle. The other class of depressuring studies is related to the consequences of a fire in the area where a vessel is located. The main goal
here is to determine the minimum vent rate that is needed to keep the vessel pressure under control and to bring down the vessel pressure to prevent a failure of the vessel wall as temperature rises.

All plants have emergency shutdown systems (ESD). The design of a system to safely stop the operation of part of or a complete plant can be very complex and it is often difficult to foresee all the consequences of everything that happens during a shutdown. A dynamic simulation model of the plant is an invaluable tool to properly set up an ESD system. Far too often a dynamic model is only used after an incident has occurred to analyze the exact causes. The use of such a model during the design phase might have prevented the occurrence of the incident all together.

Modeling the behavior of the plant under emergency shutdown conditions requires the model to be quite detailed and the simulation is usually more challenging than other applications. Under ESD conditions, much of the equipment is shutdown, many flows stop, and these values of zero for flows can be problematic.

Another use is to verify the adequacy of the pressure relief system.

15.2.1.3 Start-up procedure definition
Modeling the start-up procedure also requires a detailed model. Although generally the dry start condition also involves stopped equipment and zero flows, it tends to be somewhat easier to model than an ESD scenario.

The use of a dynamic model to verify the start-up procedures of a plant can reduce the commissioning time by weeks. This exercise consists of adding the start-up logic to the model and to run this start-up logic while observing the behavior of the plant model. When problems occur, the model can be stopped and the start-up logic can be reviewed and rerun. Not only does the start-up procedure become streamlined, but also the engineers who have worked on it acquire a detailed understanding of the behavior of the plant, which allows them to make better decisions during plant commissioning and subsequent operation of the plant.

15.2.1.4 Control system and safety integrity system checkout
Control modifications that are required for start-up and proper plant operation can be validated with the dynamic model and then implemented on the plant. Invensys clients who have taken advantage of this type of testing and prestart-up training opportunity have experienced nearly flawless first-time start-ups.

The elimination of extra start-up days through control system checkout using a fully rigorous dynamic model is often, in itself, a common justification for the dynamic model purchase.

The purpose of the control system checkout is to verify that all the cabling connecting the control system to the plant and the control system configuration is correct. Signals from a dynamic model can replace the plant signals. This will help tremendously in verifying the logical connections inside the control system. If a wrong measurement is routed to a particular controller, this will be seen quite readily as the dynamic model provides realistic numbers for these. It is much
easier to discern an erroneous number among realistic numbers than to match quasirandom numbers.

Control system checkout allows the dynamic simulator to be used to:

- pretune control loops,
- test motor start/stop logic,
- validate permissive logic,
- evaluate controls stability,
- check graphics displays,
- implement checkout controls on the plant, and
- test shutdown systems and logic sequences.

If a dynamic model is not available, then a simple “tie-back” model can be used for control system checkout. For some control systems, a virtual control processor with identical functions as an online control processor is available for control system checkout.

15.2.1.5 Operator training

A classical use of dynamic simulation has been for operator training systems. Nowadays, this is just one of the applications that can be built on top of a detailed dynamic model. In most new projects, an operator training system is becoming a standard requirement and it is yet another driver to start using dynamic simulation early in plant design as part of the earlier work can be reused in an operator training system.

Dynamic simulation for operator training provides the following benefits:

- Improve understanding of general plant theory and concepts
- Increase knowledge of plant systems and their function and interaction with other systems
- Enhance understanding of plant control theory and operation
- Gain operating experience, confidence, and accuracy in normal and abnormal plant operations
- Provide practice following specific operating procedures
- Demonstrate recovery from various upsets and malfunctions.

Besides the dynamic model, operator training systems are comprised of various other parts:

- The operator stations that mimic the real control system operator stations or that serve as spare operator consoles.
- An instructor station to allow the instructor to monitor the student’s progress and to let him or her introduce a selection of failures or other problems the operator may encounter in the real plant.
- Possibly an automated system to assign a score to the performance of the operator and/or to let the operator run predefined training scenarios.
- Software and hardware for the communication between the various modules.
15.2.1.6 Advanced process control

APC and in particular multivariable predictive control (MPC) normally requires access to plant data and step test results from the operating plant. Hence, MPC is usually only implemented once the plant has been commissioned. With a dynamic model available, this is no longer a limitation and the necessary information can be obtained from the step tests on that model. This part will be covered in more detail in the Plant Operation part (15.2.2.6).

15.2.2 Plant operation

A dynamic model also has many uses in plant operation including troubleshooting, plant performance enhancement, incident analysis, operator decision support, operator training, and APC.

15.2.2.1 Troubleshooting

Issues in the control or operability of the plant can be resolved easier, safer, and with no loss of production using a dynamic model. A dynamic model can be exercised at will where the engineer has very little freedom to test the real plant. Maintaining production on specification will almost always override the need for testing to solve a problem. A solution conceived by the plant engineer is not necessarily correct and implementing an untested solution may lead to unsafe operation.

With a dynamic model, the worst that can happen is that the model fails. The engineer can test a large range of operating conditions to ensure that the implemented solution will apply to all normal and abnormal situations.

15.2.2.2 Plant performance enhancement

Many plants now have a goal of no automatic controllers in manual. The reason that controllers are put in manual is often related to the trust operators have in controllers. There are generally two possible reasons for distrust. Either the operator does not understand how the controller will cope with upsets or the controller has shown in the past that it is incapable of dealing with those upsets.

Even without a full operator training system, a dynamic model can be used to show the operators how a series of typical upsets will be handled by the control system. Some of the responses may appear illogical to the operators at first and with a model, the operator will not feel the pressure to act prematurely in order to ensure product quality. This can help to instill more confidence in the automated control and let it run in automatic mode.

Of course, the distrust of the operator may be well founded, but in this case the dynamic model can be used as described in Section 15.2.2.1 to improve the controller behavior and subsequently illustrate to the operator that the problem has been solved to restore trust.
\section*{15.2.2.3 Incident analysis}

Although this is by no means the best use of a dynamic model, it is often the first step toward using dynamic simulation. After an incident, there is always the need to know why it happened. If the incident resulted in damages, there will be legal requirements to determine the root cause of the incident. A dynamic process simulation model will often be used in this analysis to determine the sequence of events on the process side that led up to the incident and how adequate the emergency shutdown procedures were to mitigate the consequences of the incident.

\section*{15.2.2.4 Operator decision support}

Operator decision support is an emerging use of dynamic simulation models. In this type of application, the dynamic simulation model is run in real time and it is receiving the same input signals as the real plant. It is impossible to cover the entire plant with instrumentation to provide all the information one would like to obtain. The real-time model provides the operators and engineers with simulated measurements throughout the entire plant to better appreciate current operation. Typical parts of a plant that do not have all the instrumentation desired are long pipelines and high temperature outlets of reactors or furnaces. A second use of the online model is its predictive capability. Assuming the dynamic simulation model is fast enough, it can be used to predict events minutes or even hours ahead of the actual event. This information can be used to improve the handling of the event and to keep the plant operating within specifications.

\section*{15.2.2.5 Operator training}

It is important for the operators to keep their knowledge of the plant operation up to date. Especially with highly automated plants, it is important that operators are confronted with abnormal situations using the simulator. New operators will also benefit from the use of an operator training system.

If is therefore very important to keep the operator training system that was installed as part of the plant commissioning up to date. This means that any change to the control system screens and systems must also be made to the operator training system and that any change made to the plant must be made in the dynamic process model as well.

\section*{15.2.2.6 Advanced process control}

The implementation of an APC system requires a significant investment and such a decision is not taken lightly. A dynamic simulation model can assist in determining the relevance of an APC implementation and it can help to streamline the implementation itself.

It is fairly straightforward to run the necessary step tests for the implementation of a Multivariable Predictive Controller (MPC) on a dynamic process simulation. The results can be used to design the MPC and to run the MPC on the dynamic model. A comparison of the plant performance using the existing
control system and with the MPC can provide the necessary information to decide if an MPC implementation is an attractive investment. Running step tests on a model has a number of advantages over step tests on the real plant (Alsop and Ferrer, 2006):

- No disturbance of the plant operation.
- Step tests can use a broader range of conditions.
- Step tests do not depend on the availability of plant personnel, plant incidents, and other events that are not compatible with step tests on the plant.
- The dynamic model does not suffer from valves that get stuck and other incidents that make life difficult when performing step tests.
- The dynamic model can be run faster than real time and hence step test that would otherwise take days can be run in an hour or less.

When the final MPC controller has been designed and implemented, it can first be put online using the dynamic simulation model. This setup can be used to discover a significant part of the practical problems that would otherwise only surface during the commissioning of the MPC on the real plant. Although the MPC models should be verified for actual plant operation, this accelerates MPC commissioning and lowers the risk of production loss that may be experienced during commissioning of the MPC (Valappil et al., 2006).

15.3 Modeling considerations

15.3.1 Level of detail in the model

The level of detail required for a dynamic simulation model is very dependent on the application. Often a model will contain components that are modeled in great detail (high fidelity) while other components only capture the overall dynamic behavior.

Taking a gas compressor as an example, an initial application is to use the model to analyze the behavior of the antisurge control logic. In this case, it will be important to properly model all the gas volumes in the main gas lines and also in the antisurge system. The control logic used in the model will be an exact replication of the commercial system that will be installed. As compressor surge is a very fast phenomenon, these controllers have sampling times on the order of 50 ms. The model must run using a time step that is capable of capturing these phenomena and hence have a time step that is even smaller than 50 ms. As a consequence, the model may run slower than real time. But, as in this case the time span of interest is a few minutes at the most, the slower model performance is not really a problem.

A second application for this dynamic model of a compressor station may be for operator training. It is not relevant to model events that happen as fast as surge control, since it is impossible for an operator to respond during the event. The step size and hence the speed of the model can be increased. On the
other hand, it will be important to include manually operated purge valves in the dynamic model to allow the operator to perform all actions he deems necessary. This is a bit of detail that is of no use for the antisurge application.

The level of detail should be assessed based on the objectives for the model. This assessment is not a global assessment for the complete model, but the assessment should consider the objectives for each section of the plant down to each piece of equipment.

15.3.2 Model speed

The speed of a model is most frequently expressed as the real-time factor of the model. This is the ratio of the simulated time divided by the real time. The speed requirements vary depending on the application. For an operator training simulator, it is clear that the model needs to be capable of running at least in real time (real-time factor \( = 1 \)). Quite often the real-time factor should be higher, up to 10 times real time. This allows the operator to fast forward through periods of stabilization of the process, for example.

For an engineering study, the important factor is the total amount of time it takes to study an event. Ideally, that time would be 10 min or less. This means that 3 h events should have a real-time factor of 18 or higher. If the event to be studied only lasts for 1 min, then a real-time factor of 0.1 is acceptable.

The model speed is mainly affected by the following factors:

- Time step of the integrator
- Complexity of the model
- Number of components used to represent the fluids and the complexity of the thermodynamic model.

As the same factors will also affect the accuracy of the model, a balance must be found between speed and accuracy.

15.3.3 Equipment specific considerations

The following sections recommend the information to consider when modeling various pieces of equipment. Depending on the available modeling tool, the recommended level of detail may differ. Recommendations only apply to the main aspects of the model.

15.3.3.1 Valves

The minimum requirement to model a valve properly is to use the correct \( C_V \) value\(^1\) and the type of valve characteristic. For some studies, it is important to

\(^1\) \( C_V \) is the valve coefficient that is defined by convention in field or imperial units as the number of US gallons that will pass through a control valve in 1 min, when the pressure differential across the valve is 1 psi.
capture the dynamic behavior of the valve. For example, an emergency shutdown valve requires a certain time to close. For safety studies, it is important to consider the time it takes to close. Most plants will have one or more check valves. It is important to include these valves in the model, in particularly when the model will be used to run scenarios far away from normal operation.

15.3.3.2 Turbomachinery
The use of high-fidelity turbomachinery control integrated dynamic simulation models to check out the process design and control configuration for turbomachinery applications offers numerous benefits throughout the process life cycle. The primary areas of benefit include cost savings, time savings, improved safety, and increased availability (Willetts and Nair, 2010).

For pumps, compressors, and expanders, it is best to always use the performance curves of the equipment. If this information is not available, it is relatively simple to create a generic performance curve starting from the normal operating point of the equipment. This information is then complemented with either the speed or with the absorbed power. For most motor-driven equipment, a speed specification will be the best option except when studying start-up and shutdown phenomena. For equipment driven by a gas turbine, a specification of the absorbed power is usually a better choice.

If the study concerns the start-up or shutdown of the equipment, it will be necessary to include details such as the rotational inertia, the friction losses, and the dynamics of the driver (e.g., an electrical motor) in the model.

15.3.3.3 Piping equipment
The level of detail for modeling the piping depends on the application. For process piping, it is quite often sufficient to model the pressure loss. Most of the time the volume of the piping is negligible compared to the volume of the process equipment.

A notable exception is the modeling of compressor antisurge systems. An accurate representation of the system volume is crucial to obtain correct results.

For transport pipelines, the model should be more detailed as the expected results may include information like the time lag of a product in the pipeline, the evolution of the temperature and the multiphase behavior of the fluid. The required information includes the pipeline elevation profile, the pipeline diameter, pipe schedule, insulation, and environment.

15.3.3.4 Columns
Distillation models should properly reflect the hold-up volumes of both the liquid and the vapor phase. A significant difference between trayed columns and packed columns is that the much smaller liquid hold-up of packed columns will reduce the response time of the column compositional perturbations.

It is customary to use a reduced number of theoretical stages to model distillation columns in steady-state simulations. If the same approach is used in a
dynamic simulation, the tray or packing characteristics will be modified to use the correct hold-up volumes for liquid and vapor. Another approach is to use tray efficiencies and keep the number of trays used equal to the number of real trays in the column. Keep in mind that the tray efficiency and the column overall efficiency are not the same.

15.3.3.5 Heat exchangers
The level of detail for a heat exchanger will strongly depend on the role of the heat exchanger and the phenomena to be studied. For example, if the exchanger serves to cool down a condensate stream before it proceeds to storage and the focus of the study is on the equipment upstream of this exchanger, then it may be enough to use a model that simply assumes the exchanger is always cooling the fluid to the required temperature. At the opposite side of the spectrum is a plate fin heat exchanger in a natural gas liquefaction plant. In this case the exchanger is the heart of the plant and the model needs to accurately represent the construction of the exchanger and take into account elements like the heat capacity of the metal and the dynamics of the metal temperature. The model will need to provide information like the temperature and pressure profiles inside the exchanger.

15.3.3.6 Control systems
Contrary to steady-state simulation, the modeling of the control equipment is crucial to the success of a dynamic simulation model. Quite often the control strategy and controller tuning is the final objective of the dynamic simulation, but without proper configuration of the control system the model will quickly end up in totally abnormal operating conditions.

For regular PID controllers, the main points to consider are: correct direction of the action (reverse or direct) taken; realistic tuning constants and proper span of the instrumentation. Once the simulation model has reached relatively stable conditions, attention can focus on a high-fidelity representation of the control systems.

The high-fidelity representation can come in various forms. For an operator training system, most of the control system vendors will be able to provide software that emulates the control system. The model itself is then only used to represent the noncontrol equipment. The control system emulation will receive the plant measurements from the model similar to how the real control system receives measurements from the plant and sends signals to the valve positioners according to the control algorithms defined in the control system.

The verification of an antisurge controller for a compressor is also an area where a high-fidelity representation of the particular controller is crucial. The representation can be built by combining blocks that are part of the dynamic simulator, writing a custom model for the controller, linking the model to an emulation program or even linking the model to the controller hardware.
15.4 Control of equipment and process systems

This section describes some typical applications of dynamic simulation in the various processes that are employed in gas transmission and processing. The applications are a combination of the particular equipment used in the process.

15.4.1 Gas gathering and transmission

The key equipment in gas gathering and transmission is pipelines, valves, and compressors. The range of applications in this area is very wide and includes the following:

- assessing the risk of condensate accumulation and associated slug sizes,
- analyzing line packing capacity,
- evaluating procedures for safe pipeline shutdowns,
- studying pipeline depressuring, and
- evaluating compressor station antisurge controls.

15.4.2 Gas treating

Columns are the main equipment used for the absorption of CO$_2$ and H$_2$S using amine solutions and regeneration of the amine. A dynamic model proves useful to show how the amine flow rate, lean amine quality, lean amine temperature, regenerator pressure, and reboiler duty among other variables affect the product gas quality especially as feed rates and acid gas content varies.

15.4.3 Sulfur recovery

The performance of a sulfur recovery unit is mainly governed by the operation of the various reactors in the process. A key factor in the reactor performance is the correct air–gas ratio of the reactor feed. If the acid gas feed is unstable, dynamic modeling can be used to select the best control strategy to cope with these fluctuations and to improve the controller tuning (Young et al., 2001).

15.4.4 Gas dehydration

The glycol gas dehydration process is very similar to the acid gas removal process. Fluctuations in water content of the feed gas and gas flow rate affect the product gas quality. The control strategy insures quality of the product by selecting the appropriate glycol flow rate and by maintaining the quality of the lean glycol.

For dehydration processes using molecular sieve beds, the application of dynamic simulation is similar to an application for plant start-up. The model includes the logic that is driving the bed switching and regeneration cycles. Once implemented in the model, the logic can be tested by running it on the model and by tracking critical operating parameters of the molecular sieve unit.
15.4.5 Liquids recovery, natural gas liquefaction

A cold box is a key piece of equipment in these processes. The streams that exchange heat in the cold box create a multitude of thermal loops in the process, which make it more difficult to control. A detailed dynamic model of the overall process including a detailed model of the cold box will provide an understanding of the severity of the interactions created by these thermal loops. A control strategy designed to cope with these interactions can be tested thoroughly (Valappil et al., 2005).

From the perspective of pressures and flows, the operation of turboexpanders is important. The efficiency of a turboexpander drops quickly as one deviates from the design conditions and the impact of a temporary deviation of the operating conditions on the process dynamics is difficult to understand. A dynamic model will aid in understanding the behavior and selecting the correct control structure and controller tuning to cope with a transient deviation from design conditions.

Also, propane refrigeration systems are often included comprising compressors, condensers, chillers, and separators.

15.4.6 NGL fractionation

NGL fractionation comprises a series of distillation columns. As the stripping and rectification are interactive and affect both top and bottom product purities, the control of the columns is not straightforward. A dynamic model will provide the capability to select the best control strategy given the particular column operation and specifications and given the expected disturbances in the feeds.

15.5 Case study I: Analysis of a fuel gas system start-up

This study analyses the start-up philosophy of the fuel gas system of a Latin American offshore platform. The fuel gas system flow diagram is shown in Figure 15-1. It is thought from steady-state analysis that the system can be started up using cold pipeline gas without preheating (Wassenhove, 2003).

15.5.1 Introduction

Under normal operation the fuel gas used by the turbogenerators (TGs) comes from the platform main compression system at 180 kg/cm² and 38 °C. At the system inlet, the pressure is reduced to 100 kg/cm² and the temperature is lowered to 17 °C. The fuel gas is then preheated in P-513502 to 60 °C with hot water and another pressure reduction to 45 kg/cm² through a Joule–Thompson (JT) valve resulting in a temperature of around 16 °C. After this last pressure reduction, the mixture of condensate and gas is sent to a condensate vessel drum (V-513501). The condensate returns to the process and the gas from the vessel
proceeds through a Gas–Gas Heater Exchanger (P-513501). This fuel gas from P-513501 at an average temperature of 43°C is reheated with hot water in heat exchanger P-513504 to around 63°C and sent to the turbogenerators.

The objective of the system is to produce a fuel gas stream with a defined rate, a defined pressure and a temperature at least 20°C above the dew point. This is a minimum value demanded by the turbogenerator vendor. Fuel gas exit temperature must also be maintained above 0°C to meet material temperature limitations.

During the start-up/restart of the platform, there is no gas source on the platform but cold pipeline gas can be imported at 5°C. Also, there is no hot water available for the preheaters. One solution is to start the turbogenerators with diesel. This solution is undesirable since diesel is imported onto the platform, the turbogenerators require adaptation to cope with diesel feed and there are several other problems leading to production inefficiency and unnecessary cost.

The objective of the dynamic simulation is to study the exact behavior of the fuel gas system under the start-up condition of cold pipeline import gas and the circulation of the seawater at 25°C to heat the gas.

15.5.2 Steady-state analysis

The steady-state results clearly show that thermodynamically and theoretically both normal operating and cold start scenarios result in gas feeds to the
turbogenerators that are well above the dew point approach limitation of 20 °C. Hence, it seems feasible to provide fuel gas with a temperature at 20 °C above dew point. However, this does not account for the transients encountered during start-up. The question can only be answered conclusively by a dynamic analysis.

15.5.3 Dynamic analysis

In the first start-up scenario water is circulating at 25 °C before any gas is fed. Once the water flow is stable, cold pipeline gas is brought on stream. The trends from the model displaying the system temperatures for this case are shown in Figure 15-2.

As the cold import gas hits the warmer exchangers, the gas is heated and some of the heavier components flash off in the condensate drum and return to the gas–gas chiller. The dew point of the fuel gas then rises for about 4 min until eventually the chilling duty of the recycle and JT effect in valve V2 stabilize. The temperature of the fuel gas outlet remains relatively constant as the heat exchange in P-513504 is established. Hence there is a closer approach between the fuel gas temperature and the dew point, reaching 24 °C at 4.1 min. This is too close to the limit to accept without further study.

The second start-up scenario assumes that the cold pipeline gas flow is established first and that the water system is brought on line afterward. The trends of the system temperatures in this case are shown in Figure 15-3.

Due to the JT effect and cold duty on the gas-to-gas exchanger, the gas dew point does not initially increase but maintains a difference from stream 12 temperature. In this mode they would eventually equate. Also the fuel gas exit
temperature is decreasing and within 4 min it will reach the material temperature limit of 0 °C. On start-up of the water system the temperature recovers after 10 min and the fuel gas always maintains more than a 40 °C difference from the dew point.

### 15.5.4 Conclusion

A dynamic model clearly shows potential problems with all start-up modes of the platform fuel gas system with minimum approaches to dew point or minimum material temperatures. However, the dynamic model also demonstrates a combination of procedures, starting with no water flow and then increasing flow quickly, that could maintain all proper flow rates and temperatures for fuel gas start-up without diesel, thus saving millions of dollars.

### 15.6 Case study II: Online dynamic model of a trunk line

The 235-mile Northern Offshore Gas Transport (NOGAT) trunk line is located in the Dutch part of the North Sea. NOGAT connects eight offshore platforms to onshore gas processing facilities near Den Helder, the Netherlands.

**FIGURE 15-3**

Import pipeline gas on, bring up water system temperatures.
platform delivers both gas and condensate to the pipeline, so the line operates in the two-phase region; Two of the platforms sit on oil fields and the offgas from the oil stabilization units is compressed and delivered to the pipeline. The total capacity of the gas transportation system is about 22 million m$^3$/day, associated with 750 m$^3$ of condensate/day (La Rivière and Rodriguez, 2005).

The onshore facilities include a 1000 m$^3$ slug catcher, condensate stabilization units that remove volatile components from the trunk line produced liquids and a series of low-temperature separation (LTS) units to dry the sales gas prior to delivery to the distribution network. Figure 15-4 shows a representation of the system.

There are two major challenges to operating the system:

1. Controlling the sales gas quality in terms of its Wobbe index.$^2$ The different platforms produce different gas qualities and quantities that are fed into the line at different locations. Therefore, the quality of the gas that travels through the trunk line varies along its length and also with time. In spite of

FIGURE 15-4
NOGAT trunk line system (La Rivière and Rodriguez, 2005).

$^2$The Wobbe index is a measure used to compare the equivalent thermal content characteristics of different gases. It is defined as volumetric high heating value divided by the square root of relative density to air.
this, the sales gas quality must stay within the contractual limits at all times (i.e., Wobbe index values between 49 and 54).

2. Controlling the condensate inventory of the trunk line. The amount of condensate retained inside the line builds up during periods of low-gas demand, particularly during trunk line depressuring. The available slug catchers and condensate stabilization unit capacities limit the production ramp-up speed and force the scheduling of periodic clean-up cycles to keep the trunk line liquids from building-up above critical limits.

Neither the Wobbe index nor the condensate hold-up profiles along the length of the trunk line are directly measurable through field instrumentation.

A system was constructed around a supervisory application, which uses dynamic models as the main simulation engine, and data historians to access the current operating conditions. The supervisory application handles the necessary data transfer between the historical databases and simulation models. Two dynamic models are integrated in the application. These are the so-called “online model” to monitor the present time conditions in the trunk line, and the “predictive model” to perform what-if analyses and determine the consequences of programmed or unexpected variations in operating conditions (typically gas supply flow rates).

Gas and condensate sources are modeled by means of standard process streams. While gas flow rates are automatically retrieved from the data historians. Condensate flow rates are calculated using average condensate/gas ratio (CGR) values for each platform. Gas and condensate compositions are set as fixed parameters. The multiphase pipelines are modeled using a proprietary model that operates as a unit operation inside the dynamic simulator. The trunk line temperature profile is assumed to be equal to the ambient temperature profile (simulation runs revealed the adequacy of such an assumption). Ambient temperature profiles were obtained from a North Sea temperature map that takes into account depth and date.

The physical properties of the gas condensate mixtures are calculated using proprietary correlations and databases. Although the available chromatographic analysis provided information for up to 49 chemical species, the components slate was simplified by lumping the C6 and heavier components into a smaller slate to achieve the desired model calculation speed.

Even on today’s powerful desktop computers, calculation speed is still a limiting factor for dynamic two-phase flow combined with rigorous physical property calculations. It is therefore necessary to choose a balanced solution between accuracy and speed of calculation by optimizing the number of trunk line cells, time step size, and number of chemical components. Dedicated simulation runs were performed to select the most suitable values for these parameters. The online trunk line model was given a mathematical grid of 84 cells, and it was able to run as fast as 40 times real time in a single mathematical processor. The predictive model is a copy of the online model, which uses a grid
containing fewer cells than the online model (32) and performs its calculations 250 times faster than real time.

While the predictive model to some extent sacrifices accuracy, predictions are always found to be influenced more by the estimates and forecasts of future platform flows provided by the user as simulation inputs than by model inaccuracies.

Platform discharge pressures and gas delivery flow are computed by the simulation models and stored by the supervisory application in the data historian. Space discrete information, such as pressure, temperature, flow rates, velocities, and hold-up profiles are archived in a dedicated database. The historian user interface for the system shows actual conditions versus model-calculated data. Predictive what-if or look ahead studies are triggered from the system graphical interfaces. The predictive model retains the process and trunk line internal conditions to serve as an initial state for simulations. The user is given the option to edit individual platform flows and onshore landing pressure or flow. The time horizon of the prediction is also an input value.

Simulation results provided by the online and predictive parts of the tool are available to the operators by means of graphical user interfaces (GUIs) that were constructed using historian tools. Predictive results are shown as a series of tables and charts that track the evolution of line conditions (i.e., Wobbe index, compositions, condensate hold-up, velocities, flows, and pressure) along the trunk line distance and as a function of time. Figure 15-5 shows a comparison between the observed and the simulated Wobbe index of the gas at the terminal station.

**FIGURE 15-5**
Comparison between the real and the simulated Wobbe index (La Rivière and Rodriguez, 2005).
Connecting the process engineering models to online measurement data stored in historians has been successful for the support of operational decisions. The described computer program was installed in the control room of the onshore processing facilities in March 2003 and the operators successfully utilize the system for predicting and managing the gas quality. The calibrated models developed for the online system are now also used by process engineers for offline analysis of the clean-out cycles for managing the large amount of condensate, which develops inside the trunk line, for example. Potential slug catcher overflow was prevented by a correct adjustment of platform flows. The model centric approach of the system and the clear split between application components and simulation model also facilitates model maintenance and upgrades. The speed of calculation is more than sufficient to keep the online model synchronized with real-time events. Similarly, the predictive case’s time response is satisfactory for the end-users. The accuracy of the calculated results for gas quality is more than sufficient to support the necessary operational decisions. Although the accuracy of the results regarding liquid inventories is not clear at present, the availability of these data is already a major improvement with respect to the previous situation when information on this aspect of the process was completely missing.

References
16.1 Introduction

Gas processing operations constantly experience changing conditions due to varying contracts, feed rates, feed compositions, and pricing. In order to capture the maximum entitlement measured in profits, these operations are prime candidates for real-time optimization (RTO) (Bullin and Hall, 2000). RTO allows operating facilities the ability to respond efficiently and effectively to the constantly changing conditions of feed rates and composition, equipment condition, and dynamic processing economics. In fact, world-class gas processing operations have learned how to optimize in real time to return maximum value to their stakeholders. Applications of RTO of gas processing facilities have recently been adopted. Advances in computer power, robust modeling approaches, and the availability of real-time pricing have enabled this technology. An online optimization model also provides a continuously current model for accurate simulations required for off-line evaluations and studies. Equipment conditions including fouling factors for heat exchangers and deviation from efficiencies predicted by head curves for rotating equipment are tracked over time. The impact of additional streams under different contractual terms can be evaluated with the most up to date process model available.

The objective of this chapter is to introduce the concepts of RTO and describe the considerations for successful application in the gas processing industry.

16.2 Real-time optimization

Real-time optimization refers to the online economic optimization of a process plant, or a section of a process plant. An opportunity for implementing RTO exists when the following criteria are met:

- Adjustable optimization variables (degrees of freedom) exist after higher priority safety, product quality, and production rate objectives have been achieved.
- The profit changes significantly as values of the optimization variables are changed.
- Disturbances occur frequently enough for real-time adjustments to be required.
Determining the proper values for the optimization variables is too complex to be achieved by selecting from several standard operating procedures. Real-time, online-adaptive control of processing systems is possible when the control algorithms include the ability to build multidimensional response surfaces that represent the process being controlled. These response surfaces, or knowledge capes, change constantly as processing conditions, process inputs, and system parameters change, providing a real-time basis for process control and optimization.

RTO applications have continued to develop in their formative years with 100 or so worldwide large-scale processing applications. RTO systems are frequently layered on top of an Advanced Process Control (APC) system as shown in Figure 16-1, producing economic benefits using highly detailed thermodynamic, kinetic, and process models, as well as nonlinear optimization. While an APC system typically pushes material and energy balances to increase feed and

![FIGURE 16-1](image)

Advanced process control with optimization solution architecture.
preferred products with some elements of linear optimization, RTO systems can trade yield, recovery, and efficiency among disparate pieces of equipment.

Processors frequently use RTO applications for off-line studies because they provide a valuable resource for debottlenecking and evaluating changes in feed, catalyst, equipment configuration, operating modes, and chemical costs. Processing RTO has been hampered by the lack of reactor models for major processing units, property estimation techniques for hydrocarbon streams, and the availability of equipment models. Continued technology advancement has removed many of these hurdles, however, and the number of reported RTO successes continues to grow.

RTO systems perform the following main functions:

- **Steady-State detection**—Monitor the plant’s operation and determine if the plant is sufficiently steady for optimization using steady-state models.
- **Reconciliation/Parameter estimation**—Collect operating conditions and reconcile the plant-wide model determining the value of the parameters that represent the current state of the plant.
- **Optimization**—Collect the present operating limits (constraints) imposed and solve the optimization problem to find the set of operating conditions that result in the most profitable operation.
- **Update set points**—Implement the results by downloading the optimized set points to the historian for use by the control system.

A good RTO system utilizes the best process engineering technology and operates on a continuous basis. The system constantly solves the appropriate optimization problem for the plant in its present state of performance and as presently constrained.

A typical system consists of an efficient (fast) equation solver/optimizer “engine,” coupled with robust, detailed, mechanistic (not correlation-based) equipment models, and a complete graphical interface that contains a Real-Time Scheduling (RTS) system and an External Data Interface (EDI) to the process computer. Three primary components of a fully integrated graphical interface are shown in Figures 16-2–16-4.

The RTO model is composed of separate models for each major piece of equipment. These separate models are integrated and are solved simultaneously. The simultaneous solution (rather than sequential) approach allows for solution of large-scale, highly integrated problems that would be difficult or impossible to solve using sequential techniques offered by many flow sheet vendors.

A RTO system will determine the plant optimum operation in terms of a set of optimized set points. These will then be implemented via the control system.

### 16.2.1 Physical properties

All of the process models for a rigorous RTO system use mixture physical properties, such as enthalpy, K-values, compressibility, vapor pressure, and entropy. Equations of state, such as Soave–Redlich–Kwong (SRK) or Peng–Robinson (PR)
FIGURE 16-2
Real-time optimization model interface.

FIGURE 16-3
Real-time optimization EDI interface.
are used for fugacities, enthalpy, entropy, and compressibility of the hydrocarbon streams. The enthalpy datum is based on methods such as the enthalpy of formation from the elements at absolute zero temperature. This allows the enthalpy routines to calculate heats of reaction as well as sensible heat changes. Steam and water properties are calculated using routines based on standards such as the National Institute of Standards and Technology.

16.2.2 Optimization models

Models used in the optimization system must be robust, easily specified (in an automated manner) to represent changing plant situations and must solve efficiently. These models must be able to fit observed operating conditions, have sufficient fidelity to predict the interactions among independent and dependent variables, and represent operating limits or constraints. State of the art models meet these requirements by using residual format \(0 = f(x)\), open equations, fundamental, mechanistic relationships, and by incorporating meaningful parameters within the models. These state of the art systems provide the highly efficient equation solver/optimizer, and the interface functionality that automates sensing the plant’s operating conditions and situations, and automate posing and solving the appropriate parameter and optimization cases.
Most plant models are standard models. Sometimes, custom models are created specifically for the equipment of a unit. All the models use thermodynamic property routines for enthalpy, vapor–liquid equilibrium, and entropy information.

Rotating equipment models such as compressor, pump, engine, gas turbine, and steam turbine models contain, along with all the thermodynamic relationships, the expected performance relationships for the specific equipment modeled.

16.2.2.1 Optimization objective function
The objective function maximized by a high-level optimization system is the net plant profit. This is calculated as product values minus feed costs minus utility costs, i.e., the \( P - F - U \) form. When appropriately constrained, this objective function solves either the “maximize profit” or “minimize operating cost” optimization problem. Economic values are required for each of the products, feeds, and utilities. The value of each stream is derived from the composition-weighted sum of its components. Economic values for feeds, products, and utilities in the optimization system are reset on a regular basis for best performance.

16.2.2.2 Custom models
Often custom models must be incorporated into a standard optimization package to predict proprietary processes and solvents. The custom model may be incorporated with special properties packages or integrated into the system in a semiopen approach where the iteration criterion is handled by the optimization system, but the actual kinetics or thermodynamic equations remain the same as the off-line custom model. Proprietary gas-sweetening solvent formulations may be the most common example of a custom model in the natural gas industry.

16.2.2.3 Fractionators
Fractionators are modeled using tray-to-tray distillation. Heating and cooling effects as well as product qualities are considered. Column pressure is typically a key optimization parameter. Temperature measurements are used to determine heat-transfer coefficients for condensers and reboilers.

16.2.2.4 Absorbers and strippers
These units will be modeled using tray-to-tray distillation, if required. Component splitters may be used to simplify the flow sheet where acceptable when these units have little effect on the optimum.

16.2.2.5 Compression model
The main optimization set point variables for the compressors are typically suction or discharge pressures. Important capacity constraints are maximum and minimum speed for the driver; maximum current for electric motors and maximum power for steam turbines, gas turbines, and engines. Sometimes, maximum torque will be considered for engines.
A multistage compressor model consists of models for a series of compressor stages, interstage coolers, and adiabatic flashes. Drivers are included for each compressor machine. For each single compressor stage, the inlet charge gas conditions (pressure, temperature, flow rate, and composition) and the discharge pressure specification are used with the manufacturer’s compressor-performance curves to predict the outlet temperature and compressor speed. The power required for the compression is calculated from the inlet and outlet conditions.

The first step in the development of the compression model is to fit the manufacturer’s performance curves for polytropic head and efficiency to polynomials in suction volume and speed. The equations for each compressor stage (or wheel, if wheel information is available) take the form:

\[
E_p = \frac{A}{C_3}N^2 + \frac{B}{C_3}N + \frac{C}{C_3}N_{Vs} + \frac{D}{C_3}V_s + \frac{E}{C_3}V_s^2 + F \tag{16-1}
\]

\[
W_p = \frac{a}{C_3}N^2 + bN + \frac{c}{C_3}N_{Vs} + \frac{d}{C_3}V_s + \frac{e}{C_3}V_s^2 + f \tag{16-2}
\]

where,\( V_s \) is suction volume flow rate, \( N \) is compressor speed, \( E_p \) is stage or wheel polytropic efficiency, \( W_p \) is stage or wheel polytropic head, and “A” through “F” and “a” through “f” are correlation constants.

The polytropic head change across the stage or wheel can also be calculated from the integral of \( VdP \) from suction pressure to discharge pressure. This integration can be performed by substituting \( V = Z*R*T/P \) and integrating by finite difference approximation:

\[
W_p = \frac{R}{C_3} \ln\left(\frac{P_d}{P_s}\right) \times \left(\frac{(Z_s*T_s) + (Z_d*T_d)}{2}\right) \tag{16-3}
\]

where, \( R \) is gas constant, \( P_d \) is discharge pressure, \( P_s \) is suction pressure, \( Z_s \) is compressibility at suction conditions, \( T_s \) is absolute suction temperature, \( Z_d \) is compressibility at discharge conditions, and \( T_d \) is absolute discharge temperature.

For simplicity, the above equations are formulated in terms of a single integration step between suction and discharge, but in the actual implementation, each stage or wheel is to be divided into at least five sections to describe the true profile. The enthalpy change across the stage or wheel can be calculated from the inlet and outlet conditions, the polytropic head change, and the polytropic efficiency.

This analysis results in three simultaneous equations and three unknowns for each integration step. The unknowns are the discharge temperature, discharge pressure, and enthalpy change for each step except the last. The known discharge pressure for the last step is related to the speed of the machine. In state of the art approaches, all of the integration steps are solved simultaneously.

Measured discharge pressures, speeds, and temperatures are used in the parameter estimation run to update the intercept terms in the polynomials used to represent the manufacturer’s polytropic head and efficiency curves. These parameters represent the differences between the actual performance and expected/design performance. As compressors foul, these parameters show increasing deviation from expected performance. It is this difference that has
significant meaning since an absolute calculation of efficiency at any moment in time can vary with feed rate and several other factors, which dilute the meaning of the value. By showing a difference from design, we get a true measure of the equipment performance and how it degrades over time.

The discharge flow from each stage is sent through a heat exchanger model coupled with an adiabatic flash. The heat transfer coefficient for each exchanger is based on the measured suction temperature to the next stage, corrected for addition of any recycle streams. Suction and/or discharge flows are measured to fit heat loss terms in the interstage flash drums.

### 16.2.2.6 Expander model

The actual work produced and actual efficiency are the key values to determine for the expander. The amount of work produced is simply the difference in enthalpy between inlet and outlet conditions multiplied by the mass flow rate. Enthalpies are determined from the observed stream composition, temperature, and pressure.

The ideal enthalpy at outlet conditions is required to determine the actual efficiency. Since the turbo expander process is an isentropic process at 100% efficiency, stream entropy at inlet conditions is required for the efficiency calculation. Referring to Figure 16-5, the steps to determine actual efficiency include (Simms, 2009):

1. Calculate the stream entropy and enthalpy at inlet conditions.
2. Calculate the ideal outlet temperature by determining the temperature at the observed outlet pressure for the stream entropy at inlet conditions.
3. Calculate the ideal enthalpy for the stream at ideal outlet temperature and observed outlet pressure.

![Expander thermodynamics](Simms, 2009)
4. Calculate the ideal enthalpy change from the difference between ideal enthalpy and inlet enthalpy ($\Delta h_{\text{ideal}}$), which is also defined as isentropic enthalpy change.

5. Calculate the actual enthalpy at outlet conditions from the stream composition and observed outlet temperature and pressure.

6. Calculate the actual enthalpy change from the difference between actual inlet and outlet enthalpy ($\Delta h_{\text{actual}}$).

7. Calculate the ratio of actual enthalpy change to ideal enthalpy change to determine the actual efficiency.

The deviation from the performance curve for the current expander performance can be used to estimate the work produced at different outlet pressures.

Expander performance curves are typically given as efficiency versus velocity ratio and efficiency versus velocity ratio (Jumonville, 2010). These performance curves can be regressed similar to Eqn (16-1) for the compressor model for the velocity ratio defined as expander wheel tip speed ($U$) divided by isentropic spouting velocity ($C_o$) as well as specific speed ($N_s$).

$$U = D \times N/229$$  \hspace{1cm} (16-4)

where $D$ is expander wheel diameter, inches; and $N$ is speed of expander wheel, rpm.

$$C_o = 223.8 \times \sqrt{\Delta h_{\text{ideal}}}$$  \hspace{1cm} (16-5)

Specific speed is defined as:

$$N_s = N \times \sqrt{\frac{V_o}{\Delta h_{\text{ideal}}^{0.75}}}$$  \hspace{1cm} (16-6)

where $V_o$ is actual volumetric flow at outlet conditions, cubic feet per second.

The deviation from the head curve for each of these performance factors can be used to predict the performance at other conditions to determine optimum usage of the turbo expander. Finally, compare compressor work to expander work to determine bearing losses.

In order to determine the expander speed to achieve a desired outlet pressure for given inlet conditions (pressure, temperature, volume, and stream composition) and expander wheel diameter, use the following procedure:

1. Calculate mass flow rate from inlet volume and stream composition.
2. Determine inlet entropy from given inlet conditions.
3. Determine inlet enthalpy from given inlet conditions.
4. Determine outlet temperature at constant entropy found in step 2.
5. Calculate change in enthalpy at constant entropy ($\Delta h_{\text{ideal}}$).
6. Calculate isentropic spouting velocity from Eqn (16-5).
7. Assume speed.
8. Calculate expander wheel tip speed from Eqn (16-4).
9. Calculate velocity ratio from isentropic spouting velocity found in step 6 and expander wheel tip speed found in step 8.
10. Determine expander efficiency according to biased performance curve at velocity ratio found in step 9.
11. Calculate actual change in enthalpy from $\Delta h_{\text{ideal}}$ found in step 5 and expander efficiency found in step 10.
12. Calculate outlet enthalpy from inlet enthalpy found in step 4 and actual change in enthalpy found in step 11.
13. Calculate actual outlet temperature from outlet enthalpy found in step 12, desired outlet pressure, and stream composition.
14. Calculate actual volumetric flow at outlet conditions.
15. Calculate specific speed from Eqn (16-6).
16. Determine expander efficiency according to biased performance curve at specific speed found in step 15.
17. If efficiency determined from velocity ratio from step 10 and efficiency determined from specific speed found in step 15 is equal (within desired tolerance), then calculate expander work from mass flow rate and actual change in enthalpy.
18. If efficiency determined from velocity ratio from step 10 and efficiency determined from specific speed found in step 15 is not equal (outside desired tolerance), then return to step 7.

With the above procedure, the expander outlet pressure can be optimized considering the other processing equipment.

16.2.2.7 Distillation calculations
Standard tray-to-tray distillation models are used for distillation calculations in an optimization system. The “actual” number of trays is used wherever possible and performance is adjusted via efficiency. This allows the model to more accurately represent the plant in a way that is understandable to a plant operator.

All distillation models predict column-loading constraints accurately as targets are changed. Condenser and reboiler duties are also calculated for predicting utility requirements and exchanger limitations.

Tray-to-tray distillation method
An equation-based tray-to-tray distillation method is based on mass, heat, and vapor–liquid equilibrium balances on each physical tray. Figure 16-6 depicts a typical distillation column tray.

Component mole balances for each component $i$ on tray $j$ are:

$$ F_i Z_{ij} + L_{i+1} X_{(i+1)j} + V_{(i-1)j} Y_{(i-1)j} - (L_i + L P_i) X_{ij} - (V_i + V P_i) Y_{ij} = 0 \quad (16-7) $$

The overall mole balance is:

$$ F_i + L_{i+1} + V_{i-1} - (L_i + L P_i) - (V_i - V P_i) = 0 \quad (16-8) $$
The vapor–liquid equilibrium definition is written in terms of the liquid mole fractions and K-values:

\[ Y_{ij} = K_{ij}(X_{ij}, T_{jl}, P_i) \cdot X_{ij} \]  

(16-9)

where, \( T_{jl} \) is the liquid temperature on tray \( j \).

The requirement that the mole fractions balance is expressed as:

\[ \sum [X_{ij}] - \sum [K_{ij}(X_{ij}, T_{jl}, P_i) \cdot X_{ij}] = 0 \]  

(16-10)

The Murphree vapor tray efficiency, which accounts for differences between the equilibrium vapor composition \([K_{ij}(X_{ij}, T_{jl}, P_i) \cdot X_{ij}]\) and the actual mixed vapor composition \([Y_{ij}]\) leaving the tray, is an important adjustable parameter. It is defined as:

\[ E_{ij} = (Y_{ij} - Y_{ij-1})/(K_{ij}(X_{ij}, T_{jl}, P_i) \cdot X_{ij} - Y_{ij-1}) \]  

(16-11)

The vapor and liquid enthalpies are defined as:

\[ H_{Vi} - HV(Y_{ij}, T_{jv}, P_j) = 0 \]  

(16-12)

\[ HL_i - HL(X_{ij}, T_{jl}, P_j) = 0 \]  

(16-13)

where, \( T_{jv} \) is the vapor temperature on tray \( j \).

The overall tray heat balance is:

\[ F_i HF_i + L_{i+1} HL_{i+1} + V_{i-1} HV_{i-1} - (L_i + LP_i) HL_i - (V_i + VP_i) HV_i + Q_i = 0 \]  

(16-14)

The liquid and vapor flow rates on the capacity limiting trays are used in conjunction with tray loading calculations to predict column pressure drops. The vapor area factor in the loading correlation is parameterized to enable matching calculated column differential pressure against the measured differential pressure. Upper limits are placed on this calculated differential pressure during the optimization case to ensure avoidance of flooding.

Condenser and reboiler limitations are handled by placing constraints on the minimum approach temperature in these exchangers or on the maximum heat...
transfer area in the case of variable level exchangers. Heat transfer coefficients are calculated for the exchangers in the parameter case.

**Demethanizer**

Set point variables for a cryogenic demethanizer are typically the bottoms methane or carbon dioxide content and the overhead pressure. Constraint variables include overhead condensing capability and minimum column temperatures. The expansion impacts on the demethanizer and the overhead ethane losses are key profit variables.

The demethanizer is modeled with a tray-to-tray distillation model. Rigorous K-Values for the demethanizer column are recommended due to nonlinear vapor–liquid equilibrium relationships. Overall heat transfer coefficients for the reboilers and feed chillers are calculated from temperature and flow data.

**Deethanizer**

The set point variables in the deethanizer are the overhead pressure, the overhead propane content, and the bottoms ethane content. Any cooling medium (such as refrigeration, cooling water, or air) and heat medium (such as steam or hot oil) requirements of the deethanizer and the propane loss in the overhead are the major profit considerations.

The deethanizer is modeled with a tray-to-tray distillation model. Parameters include column pressure, column pressure drop, propane content in the deethanizer overhead, reflux flows, deethanizer bottoms ethane content, and bottoms draw rate. Overall heat transfer coefficients are calculated for the exchangers.

**Depropanizer**

The set point variables for the depropanizer are the overhead butane content and column pressure. Key constraint variables are the propane in the debutanizer overhead, the overhead exchanger capacities, and column pressure drop (flood- ing). Any cooling and heating medium usage requirements of the depropanizer are the major profit considerations.

The depropanizer is modeled with tray-to-tray distillation models. Parameters considered include column feed temperature, column pressure, column pressure drop, butane content in the tops, reflux flow, bottoms propane content, and bottoms draw rate. Overall heat transfer coefficients are calculated for the reboiler and condenser. A column capacity factor is also parameterized.

**Debutanizer**

The set point variables for the debutanizer are the overhead pentanes and heavier content as well as column pressure. Key constraint variables are the propane in the debutanizer overhead, the overhead exchanger capacities, and column pressure drop (flooding). Any heating and cooling medium usage requirements of the debutanizer are the major profit considerations.
The debutanizer is modeled with a tray-to-tray distillation model. Parameter considered include column feed temperature, column pressure, column pressure drop, pentanes and heavier content in the overhead, reflux flow, bottoms butanes content, and bottoms draw rate. Overall heat transfer coefficients are calculated for the reboiler and condenser. A column capacity factor is also parameterized.

Butanes splitter
The set point variables for the butanes splitter is the column pressure, normal butane in overhead, and isobutane in the bottoms. The constraint variables of interest are reboiler and condenser loading and product specifications. The profit variables of interest are the isobutane losses in the bottoms normal butane stream.

The butanes splitter is modeled with a tray-to-tray distillation model. The parameters considered include column pressure, column differential pressure, normal butane in the isobutane product, bottoms isobutane concentration, reflux flow rate, bottoms flow rate, product draws, bottom reboiler flow rate. Heat transfer coefficients for the exchangers are calculated.

Refrigeration models
The main set point variable for refrigeration machines is the first-stage suction pressure. Refrigeration system models relate refrigeration heat loads to compressor power. The compressor portions of the refrigeration models use the same basic equations as compressor models discussed above. Refrigeration systems should use the appropriate composition and will have a component mixture for any makeup gas.

The measured compressor suction flows and the heat exchange duties calculated by the individual unit models are used to determine the total refrigeration loads. The refrigerant vapor flows generated by these loads are calculated based on the enthalpy difference between each refrigerant level. Exchanger models of the refrigerant condensers are used to predict compressor discharge pressures.

Demethanizer feed chilling models
The demethanizer feed chilling system is modeled as a network of heat exchangers and flash drums. These models are used to predict the flow rates and compositions of the demethanizer feeds. The effects of changing demethanizer system pressures and flow rates are predicted.

The demethanizer feed drum temperatures and feed flow rates are measured to fit the fractions of heat removed from the feed gas by each exchanger section. Flow and temperature measurements on the cold-stream side allow the fraction of feed gas heat rejected to each stream to be estimated. The two sides of the feed exchangers are coupled through an overall heat balance. The inlet and outlet temperatures from the refrigerant and other process exchangers are used to fit overall heat transfer factors in the parameter case. A pressure drop model for the feed gas path is also included.
16.2.2.8 **Steam and cooling water system models**

Heat and material balance models of the steam system are developed. These models include detailed representations of the boilers.

The cooling water system will be modeled with heat exchangers, mixers, and splitters to allow for constraining the cooling water temperature and its effects on the operation of distillation columns and compressors.

16.2.2.9 **Turbines**

A turbine model, as shown in Figure 16-7, is used for steam turbines (back-pressure, condensing, or extraction/condensing) or any expander in which the performance relationship can be expressed using the following equation:

\[
\text{Design Power} = A + B \times (\text{Mass Flow}) + C \times (\text{Mass Flow})^2 + D \times (\text{Mass Flow})^3
\]  

(16-15)

Back-pressure and condensing turbine expected performances are usually presented as essentially linear relationships between power and steam flow. Extracting/condensing turbine expected performance relationships are typically presented as power versus throttle steam flow, at various extraction steam flows. This kind of performance “map” can be separated into two relationships of the form above, one representing the extraction section, and the other the condensing section. An extraction/condensing turbine can be thought of as two turbines in series, with part of the extraction section flow going to the condensing section.

---

**FIGURE 16-7**

Schematic of turbine model.
Design power refers to expected power from performance “maps” that are at specific design inlet pressure, inlet temperature, and exhaust pressure. Expected power is the power expected from a turbine operating at other than design conditions. The design power is adjusted by the “power factor,” as illustrated:

\[
\text{Expected Power} = \text{Design Power} \times \text{Power Factor} \quad (16-16)
\]

\[
\text{Power Factor} = \frac{\Delta \text{Isentropic Enthalpy at actual conditions}}{\Delta \text{Isentropic Enthalpy at design conditions}} \quad (16-17)
\]

\[
\text{Brake (shaft) Power} = \text{Expected Power} + \text{Power Bias} \quad (16-18)
\]

Design power and expected power are equal if actual expanding fluid (i.e., steam) conditions are the same as design conditions. The power bias can be parameterized using the exhaust temperature for steady-state turbines or the extraction section of an extraction/condensing turbine. In both of these cases, the exhaust steam is superheated (single phase). For condensing turbines or the condensing section of an extraction/condensing turbine, the measured steam flow is calculated since the temperature of the two-phase exhaust cannot be used. The fraction vapor of the two-phase exhaust is determined by energy balance. The total energy demand can be determined from the compressor (or other driven power consumer). The power extracted from the extraction section can be determined from the throttle steam flow and the inlet and outlet (single phase) conditions. The power extracted from the condensing section is just the difference between the total demand and the extraction section power.

The Power Loss is taken from the steam and affects its outlet conditions, but is not transferred to the shaft, or brake power.

To calculate the change of enthalpies needed for the Power Factor and for the energy balance, fluid conditions are calculated for each turbine section at the inlet (throttle), outlet (exhaust), and at inlet entropy and outlet pressure for both actual and design conditions. The nomenclature associated with the exhaust fluid conditions required for the enthalpy change calculations is:

- \( \text{Exh IDM} \): Exhaust Isentropic at Design inlet conditions for vapor/liquid Mixture.
- \( \text{Exh IDV} \): Exhaust Isentropic at Design inlet conditions for Vapor.
- \( \text{Exh IDL} \): Exhaust Isentropic at Design inlet conditions for Liquid.
- \( \text{Exh IAM} \): Exhaust Isentropic at Actual inlet conditions for vapor/liquid Mixture.
- \( \text{Exh IAV} \): Exhaust Isentropic at Actual inlet conditions for Vapor.
- \( \text{Exh IAL} \): Exhaust Isentropic at Actual inlet conditions for Liquid.

For condensing turbines or the condensing section of extraction/condensing turbines, the exhaust pressure is used to specify the model. The parameter that this measurement updates is the condenser heat transfer coefficient. The exhaust pressure is free to move in the optimization cases, since the actual pressure moves as the condenser calculates the pressure required to condense the steam sent to the condensing turbine or condensing section.
The maximum mass flow at reference (usually design) conditions is used to predict the maximum mass flow at actual conditions as an additional constraint on the turbine performance. Sonic flow relationships are used for this prediction. This is the maximum flow through the inlet nozzles when the inlet steam chest valves are wide open. The nozzle area is not needed if the maximum flow at a set of reference conditions is known. Vendors usually list the maximum flow, and not the nozzle area.

The model also includes the gland steam flow, which is used to counterbalance the axial thrust on the turbine shaft, and which is lost through labyrinth seals. The gland steam does not contribute to the shaft power.

Typically, turbine performance is not considered to be a function of speed. Consequently, speed is a variable that has no effect on the solution. However, turbine performance is typically a very weak function of speed over a wide range, and its effect on performance is not typically presented on the expected performance “maps.” The possibility of adding the weak effect of speed on performance will be considered when the model is being built.

16.2.3 Plant model integration

After the plant section models have been developed, they are integrated into the overall plant model. All of the interconnecting streams are specified and checked and a consistent set of variable specifications developed. At this stage, the overall validity of the plant model is checked using off-line plant data.

Reconciliation/parameter and optimization cases are run and the results are checked for accuracy and reasonableness. A material balance model is included in the plant model integration work to confirm an absolute model material balance closure. This material balance will include a furnace area balance, a recovery area balance, and an overall balance. In addition, the plant integration allows the objective function to be tested and validated with connections to all feed, product, and utility variables.

It is important to have the engineers who will be responsible for commissioning the optimization system involved in the project during plant model integration. A thorough understanding of the plant model is imperative for a smooth implementation of the online system.

16.2.3.1 Model fidelity and measurement errors

In an optimization system as described here, neither the models nor the measurements need to be absolutely perfect for the system to work well and to deliver significant improvement in profitability. An online optimization system continuously receives feedback from real-time measurements. The model parameters are updated prior to each optimization so that the models fit the plant and the optimum set points calculated are valid and can be confidently implemented. Without this constant feedback of plant measurements and regular updating of the plant model, the optimization solution might not be feasible.
The fidelity of the models and the accuracy of the measurements are reflected in trends of the parameters. During commissioning of the optimization system, the best available measurements are identified by analyzing many parameter cases, running with real-time data, and prior to closing the loop.

If, for example, significant heat balance discrepancies exist between process side and fuel/flue gas side measurements in a furnace, that discrepancy can be handled by determining the bias required on the fuel gas flow measurement(s) to satisfy the heat balance. That bias could be a parameter that is updated prior to each optimization. The variation of the bias over time (its trend) would be monitored. If this parameter varies significantly, the furnace model and other measurements used by it would be investigated thoroughly. Alternate measurements used to “drive” the parameter case solution would be investigated as well. The outcome of this analysis is that the best available measurements are selected and the model relationships are thoroughly investigated to ensure that all significant relationships are included. This analysis is a standard and required step in building the optimization system.

Validity checking is an integral part and is built into an online optimization system. It is used to screen out gross errors. If alternate measurements are available, the validity checkers can use these when primary measurements are unavailable or bad. Generic validity checking takes care of common errors, while custom validity checking can respond to site-specific situations. Measurements can be designated as critical, so if they are unavailable, the optimization cycle will be directed to monitor for steady state, and will only complete its cycle when the measurement becomes available. Validity checking has several features designed to keep the online service factor of the optimization system high in the face of imperfect measurements. Another feature of the online system is that the measurements used to drive the solution of the parameter case are averages over a specified time window (usually 1 h) so that measurement noise is suppressed.

The better the measurements and the better the models, the better an RTO can fully exploit the process equipment and consequently the more potential profit is realized.

Processing RTO in the future will likely include wider applications driven by demonstrated benefits, reduced implementation costs, and acceptance as a best practice. RTO applications are also becoming tightly intertwined with economic planning systems, where real-time pricing and contractual considerations are available.

Evolving technologies changing the value proposition for refining RTO include:

- Detailed kinetic models for all major processing units and configurations, proven by reported applications.
- Optimization technology improvements that incorporate robust solvers, integer variables, and the capability to handle increasing problem sizes.
Today’s technology can handle applications with several hundred thousand equations; a typical refining application has 100 measurements, more than 100,000 variables, and 25 outputs.

- Greater integration with higher level systems including shared models and reconciled measurement data.
- Multiunit optimization that leverages shared resources between process units and continues to lead toward rigorous refinery-wide optimization.
- Computing technology improvements, which have already shifted RTO from minicomputers to personal computers, and that will allow more solutions/day and more complex formulations. Solver and computing improvements will eventually lead to true dynamic optimization.
- Application and model building tools, operating graphical user interfaces, and sustained performance technologies that will lower cost, improve benefits, and remove other hurdles.

### 16.3 RTO project considerations

The steps of a real-time project implementation include:

- Front-end engineering design
- Flow sheet development
- Model testing and tuning
- Online open loop testing
- Online closed loop testing
- Application sustainment.

During front-end design, the project objectives should be clearly defined including:

- Process envelope—The system boundaries shall be determined including the process equipment included within the boundary. Various equipment lineups that will be considered for optimization are identified. Also, the modeling methodology decisions occur at this stage.
- Economic parameters and objective function—The profit function requires definition including the specific economic parameters that contribute to the profit function.
- Source of process and economic information—Process and economic information will be available from various sources. Process information is typically acquired from a process historian, while economic information will come from various commercial sources. Care should be taken to determine a source that is current and updated frequently.
- Set points to be generated from the optimizer—Process set points are critical for effective optimization. These set points must be selected where process control action will be attained accurately. APC strategies are usually best for implementing set points from a real-time analyzer.
Metrics required for success—Metrics to track optimizer online time as well as optimizer effectiveness should be determined. An owner of these metrics should be identified to take responsibility for the success and sustainability of the optimizer.

Flow sheet development is obviously a critical step in the implementation of a RTO project. The flow sheet should be developed by subsections of the plant and imported into the main application flow sheet as block diagrams. Redundant streams and equipment used in the initialization phase are removed from the flow sheet, and the sections of the plant are connected together at the block boundaries to fully integrate the subsections into a single flow sheet. Unit and flow sheet customizations are then added to improve the model’s representation of the process.

Model testing and tuning includes reconciling the imported data in off-line mode and testing the real-time sequencing. The standard data reconciliation report is used to identify the measurements with the worst mismatch, and corrective action is taken. The model is then tested by importing multiple sets of data via the Electronic Data Interface before placing the model online. Optimization cases are run and verified. Steady-state detection is fine tuned. Initial online runs are observed to verify transfer of optimal set points to the advanced control and distributed control systems.

Sustainment of the application is critical to the long-term success. An engineer should be assigned responsibility for maintaining the application and working with operations to resolve any concerns immediately. Metrics should be tracked and reported to measure the success. These metrics may include:

- The time on optimization as a percentage of time on control, averaged over all optimizable controllers.
- The time the optimizable controllers had optimization set points rejected as a percentage of the time on control.
- The time the optimizable controllers ran at expired optimization set points as a percentage of the time on control.

Optimization set points should expire after a predetermined amount of time, reflecting the nominal period for which a single optimization solution is valid. A standard stream-factor tracking program will calculate and report these statistics.

16.4 Example of RTO

The Gassled joint venture operated by Gassco and supported by Statoil has applied RTO at their Kårstø gas processing plant in Norway (Kovach et al., 2010). The model is called the Plant Production Performance Model or 3PM.

Kårstø is the largest Natural Gas Liquids (NGLs) recovery plant in Europe. The Gassled owners have first rights to book capacity. Spare capacity is available for any other qualified shipper and subject to published tariffs for transporting and processing gas.
Gassled has a flexible gathering network connecting the respective producers and processing terminals allowing gas streams from several fields to be routed to different destinations. Mixing of gas streams provides sales gas quality with respect to Gross Calorific Value (GCV), Wobbe Index (WI), and CO₂.

Rich gas processing capacity at the Kårstø plant depends on several variables and constraints. One of the most significant variables is the feed gas composition. Simulations demonstrate that the rich gas processing capacity may be significantly lower than the nominal design capacity if the feed has a high NGL content.

New expansion to increase the capacity of the Kårstø facilities has increased the complexity of the Kårstø facilities. The need for an online model to determine plant capacity was recognized to enable the plant to operate with a high degree of capacity utilization, realizing that production regularity is an inherent property of the throughput obligations. Underutilization of capacity sacrifices processing fees. The RTO model allows precise and reliable capacity predictions, while reducing the work required to determine the capacity. Operating set points required to reach the predicted capacity are generated as well as information for maintenance planning and infrastructure development.

16.4.1 Process description

Figure 16-8 shows a simplified process diagram of the Kårstø gas processing plant. As can be seen, rich gas enters the plant and is preconditioned by removal of H₂S and mercury. The gas is then preheated prior to dehydration and NGL is extracted from the rich gas. The fractionation facilities produce raw ethane, stabilized condensate, propane, \( n \)-butane, \( i \)-butane, and naphtha. CO₂-rich ethane extracted from each train is routed for purification of ethane.

The processing facilities include 26 distillation columns. Steam is used as the heating medium for the reboilers on the distillation columns and to power turbines. Steam boilers and waste heat recovery systems on gas turbines generate the steam. Three levels of steam are employed. Additional utilities used are sea water for cooling and propane for refrigeration.

Sales gas is exported to two high-pressure subsea pipelines operated at pressures up to 189 barg. The export compression facilities include four compressor manifolds operated at different suction and discharge pressures. The compressors are driven by both gas turbines and electric motors.

In the Kårstø plant, the extraction trains may be bypassed. This allows more gas to be produced and processed. However, the bypass quantities are limited to provide sales gas quality to comply with the GCV, WI, and CO₂ specifications.

16.4.2 Plant operation

Effective field production and development of new offshore fields require high capacity utilization of the Kårstø plant. By squeezing the throughput margins, the plant performance becomes increasingly sensitive to variations in feed gas
FIGURE 16-8
Process flow overview of Kårstø gas processing plant.
composition and the plant constraints become more evident. Operating toward the plant limits challenges plant regularity. A conceptual illustration of the intrinsic relation between these two characteristic parameters is shown in Figure 16-9.

Several parameters impact the rich gas capacity at Kårstø plant. Some of the most important are: rich gas composition, NGL recovery, product purity, and CO2 recovery. Production rates and the fluid composition determine the rich gas composition to the Kårstø plant. Significant and sometimes rapid changes in the feed composition must be handled continuously. Changes in feed composition may move the capacity constraint from one facility to another. For example, a richer composition of the feed gas may cause the plant to reach a constraint in the NGL fractionation. A leaner gas could hit a constraint in the extraction facilities. Crossover piping allows several combinations of routing of the gas streams.

The NGL extraction trains at Kårstø plant are designed to provide about 90% recovery of NGL. However, recovery of NGL products may be limited for some feed compositions by the fractionation capacity. When the production objective is to maximize rich gas throughput, the NGL recovery can be decreased to process more gas until reaching a constraint in the extraction facilities, compression, preconditioning, or other parts of the plant. Specifications for the liquid products, defined in the transportation and processing agreements, require a minimum purity.

The carbon dioxide concentration of the sales gas is a contractual constraint with a limit of 2.5% mole. The CO2 removal capacity of the plant is not a fixed. Feedstock composition and gas rate are the major factors that affect CO2 removal capacity. Optimum CO2 removal can be affected by the routing of gas and liquids through the plant as well. Rich gas can be split between two treating systems or bypassed and delivered to two export pipelines. Also the GCV is controlled
between an upper and lower constraint. Routing rich gas through one of the bypass connections may be employed to increase plant throughput, but this rate is then limited by the upper GCV constraint. To avoid breaching the lower GCV constraint, extracted ethane can be reinjected in the sales gas. Furthermore, the sales gas streams can be mixed at the compressor manifolds by utilizing the downstream crossover. In addition to increased capacity utilization, the crossover pipelines allow enhanced flexibility of the operations with respect to handling of feedstock variations. It is possible to mix the various feed streams to optimize NGL recovery, CO₂ extraction, and quality of the products.

16.4.3 Production objectives

The primary production objective is typically plant throughput and to deliver sales gas and products within the specifications. However, the production objectives include: maximum daily throughput, high annual capacity utilization, optimum NGL production, and optimum fuel gas consumption.

Most shippers request high production at Kårstø throughout the year, taking into account the seasonal swing in gas demand. This means high demand for processing capacity at Kårstø throughout the year.

The operator is responsible for coordination of the yearly maintenance planning of all the installations connected to the gas transport infrastructure. A primary objective is to obtain a total plan where the availability of gas for deliveries to the market is maximized. Depending on the extent of yearly maintenance at Kårstø as well as at the upstream fields, this could put restrictions on production from certain fields or allow accelerated production of more NGL- or CO₂-rich gas from other fields. Maximizing CO₂ production from CO₂-rich fields could imply postponed or reduced investments in future CO₂ removal capacity, to meet sales specifications.

In periods when processing demand is below the plant capacity (when Kårstø has no bottleneck effect on the offshore production), the primary operational objective is to maximize NGL recovery in order to provide for increased value creation for the shippers. For the NGL products, this means achieving the minimum product purity.

The Kårstø plant is a large energy consumer, and optimizing energy consumption is an important objective. However, optimizing energy consumption should not compromise the primary production objectives, related to the value creation at Kårstø.

16.4.4 Project drivers

Value generated by RTO for the Kårstø plant operations comes from the following:

- Increased utilization of the plant capacity, by introducing RTO
- Improved quality of the capacity figures issued for booking
- Reduced time needed to prepare for the booking process
- Improved position in the business development process.

On a daily basis, the feed stream compositions will vary. Processing feeds of variable composition requires the plant control system to give fast and accurate responses, in order to maintain production targets. Skilled operators have a basic understanding of the plant operational characteristics and learn how to respond to feed disturbances. However, in transition periods the plant capacity will not be fully utilized. Also, the operator may not push the plant to its full capacity, or they may choose a suboptimal routing of the gas through the processing facilities. On a regular basis, the RTO determines optimized set points for the advanced control system, thereby ensuring a rapid and smooth transition period to new optimal plant conditions.

An illustration of the possible benefits with the 3PM employed for RTO to handle feed composition variations is shown in Figure 16-10. In this example, the primary operational objective is to achieve maximum plant throughput. The upper curve illustrates achieved production over a day when the operator employs the 3PM. The lower curve illustrates production with no 3PM implemented. It should be recognized that the upper curve assumes a shorter transition period following changes in the feed conditions. The 3PM model enables an increase of the plant throughput by 1%.

Plant regularity has a very high focus, and a regularity target of at least 98.5% is set for the Kårstø plant. However, when failure occurs, emphasis is put on maintaining the highest possible service degree to limit upstream consequences to oil production and minimize the consequences for the gas customers. In situations where equipment exhibit underperformance like fouling in heat exchangers or degrading of compressors, this is revealed by the optimization
model and appropriate actions to correct the problem could be executed. If equipment fails to work or is temporarily out of service, the optimization model automatically computes new set points of the APC to minimize the consequences.

The operator of the plant is responsible for issuing capacity figures for the booking processes. Often, there are requests for processing capacity beyond current capacity at the plant. This puts pressure on the capacity margins of the plant. With the optimization model, which closely mimics the real operation performance, Gassco can provide enhanced confidence to reduce the uncertainty margins.

Extraction of NGL from the rich gas and subsequent fractionation into commercial products adds value to the shippers. Some shippers limit their gas deliveries based on their booked fractionation capacity, while shippers with leaner gas are limited by their booked extraction capacity. Any free processing capacity is identified and can be available to the shippers, allowing them to optimize their petroleum portfolio. This allows accelerated production of NGL-rich gas to maximize revenue for the shippers.

Boosting processing capacity within the limits of the facilities is the primary objective. The online model improves planning with a higher accuracy than the current process simulation models. Improved quality and accuracy of the planning tool results in a higher confidence in the calculation of the maximum plant capacity, denoted Available Technical Capacity (ATC). The capacity committable on a long-term basis is denoted Maximum Available Capacity (MAC). This allows for an operational margin for daily operations as shown in Figure 16-11.

Reduced uncertainty due to the capability of the optimization model may narrow the operational margin and thereby increase the MAC. The biannual

![3PM Capacity potential](image)

**FIGURE 16-11**

Increased available technical capacity and maximum available capacity.
capacity estimation process is demanding and time-consuming. Performance of existing equipment and systems that limit capacity utilization of the plant requires evaluation. The optimization model allows this estimation process to run more efficiently.

16.4.5 Features of the optimization model

A single model that is applied for various purposes minimizes the inconsistency between various simulation results and the actual operation of the plant. A high degree of confidence in the planning scenarios and simulation results is achieved as the same model is applied for optimization of the plant in real time. With the increased complexity of the plant, and thereby the increase in number of opportunities for routing/processing the gas at Kårstø, a model of this type provides a means of improving the plant performance by optimizing the operation by use of a mathematical model to compute the optimum operation point and routing.

16.4.5.1 Implementation and usage of the model

The model is built in a commercially available equation-based modeling tool from developed specifically for real-time online optimization in the hydrocarbon and chemical industry. The model is a single model, used for all purposes of the 3PM project, i.e., RTO, simulation, and planning. The setup of the system is described briefly here. The online model reads automatically plant measurement tags, and based upon these readings, a steady-state test is performed. With the steady plant, a data reconciliation with the actual plant data is performed. Afterward, the reconciled model is subject to optimization with a given objective function.

The online system is located at the Kårstø gas processing facility. The off-line system is located at both Kårstø and the Gassco. The model includes the four gas separation trains (serving the Åsgard and Statpipe rich gas pipelines), the Sleipner condensate train, and simplified representations of the utility (steam and refrigeration) systems. This results in the following unit counts:

- Distillation Columns – 26
- Single Compressors – 18
- Paired expander–compressors – 13
- Process measurements – 1600
- Controllers (regulatory and advanced control, 175 and 110 variables, respectively)
- Associated heat exchangers, drums, motors, furnaces, pumps, pipes, and valves

The output of the optimization is a set of set points to the advanced control system. The online model is anticipated to run in intervals of 2–4 h, depending on solution time, performance, and actual implementation and usage in the daily routines.
The planning interface is a simplified representation of the plant in an external Graphical User Interface (GUI), where one has the opportunity to generate new feeds by mixing different field flows and compositions, to test out the performance in simulation as well as optimization mode. The simplified representation of the plant consists in principle of a block diagram, where the blocks represent sections of the plant. The planner then has the option to switch off one or more blocks, and check the capacity (performance) of the plant with this very simplified model user interface. The fact that the planner has a simplified overall GUI for the model of the plant, with a number of sections representing logical groups of unit operations in the plant, makes it easy to do scenario analysis for both new fields (new compositions) and for shutdown and maintenance. In each case, it becomes easier for the planning department to give a precise answer to the capacity of any feed with a given layout of the plant.

Simulation is done in the core model GUI, which is also used in the RTO mode. Further, new plant developments may be simulated with the model in order to get a better prediction of throughput capacity than a conventional simulation would give.

The online model is located on a computer in the plant control room. It is running in a fully automated fashion, on a given interval basis. The model immediately sends the solution to a central storage computer. From here, the model can then be accessed by planners and off-line simulation engineers, who copy the solved models to their own computers, respectively. This facilitates a common arena for operations as well as the planning and engineering departments.

The online model produces results to various types of end users. In addition to generating the actual set points to the advanced control system (and the operators), the model generates a set of reports with different types of information every time it is solved. The type of information to be generated is mass and energy balances, production figures, economical performance, utility consumption, capacity utilization of each unit, shadow prices, and equipment monitoring.

16.4.5.2 Modeling and optimization strategy

The model for the Kårstø plant is a plant-wide model, comprising all processing facilities, as well as the steam boilers and other utilities. The model is a single plant model, based on open equations. The individual unit operations are modeled by use of standard library models. In the cases of rotating equipment and important valves, Statoil and original vendor performance curves are used. The plant instruments that the model reads and applies are carefully chosen, in order to ensure that sufficient and trustworthy signals form the basis of the data reconciliation, and subsequently the optimization. All alternative operation scenarios in terms of equipment failure, maintenance, etc., are dealt with by lineups defined by a set of macros.
The various crossovers that are normally in use in the plant are treated as continuous variables in order to avoid integer variables in the optimization problem. Only equipment that plays an active role in the plant are modeled, and the set of compounds used are as comprehensive as necessary to achieve representative results. This is in order to maintain a fairly reasonable size (and solution time) of the model.

The model (and the planning facility) comprises a set of objective functions from which one can be applied according to the desired operation mode. The objective functions for the optimizer reflect operation modes that are, or may be, relevant to operation of the Kårstø plant. The objective functions are predefined and can be chosen arbitrarily by the system administrator. Typical examples of objective functions are to maximize sales gas production or to maximize liquid NGL production. For all cases, the general product qualities are modeled as constraints together with the relevant operational limitations.

There are over 100 flags to indicate the routings and equipment status in the plant. While the physical realities prevent this from being a true combinatorial problem, in practice, there are easily over a 100 possible plant configurations. This poses challenges in both making sure that the appropriate equipment is on or off in the model, but more importantly, in creating a good starting point for the equation-based solver.

In addition, the CO2 removal and ethane recovery unit can operate near the CO2/C2 azeotrope and some portions of the unit approach the critical point of the mixture. As with most modern gas plants, it is heavily heat integrated with many heat-pumped columns, cold boxes, and paired expander–compressors.

16.4.5.3 Online usage
The online system follows the sequence of detecting steady-state, checking data consistency, reconciling the data, calculating the optimal set points, and sending the optimized set points to the controllers.

16.4.5.4 Quantifying measurement errors
The 3PM model imports over 1600 process measurements, which are used in data reconciliation. Of these, several hundred were identified as key process variables and have results from the data reconciliation runs written back to the database. The data exported for these points consists of the sample value used in the data reconciliation, the reconciled value, and the measurement offset.

Some nonmeasured variables such as compressor efficiencies are also saved. The scan and reconciled values are accessible on operator process displays. The same operator displays have links to preconfigured trends of the measurement offsets. This gives operations a quick way of evaluating the accuracy of their process measurements and seeing inferred values (e.g., efficiencies and unmeasured temperatures) in a familiar format.

Data reconciliation detects and runs when the plant is steady and puts the data in a readable format that supports scanning the data over long time periods. Any
trended offset, which is not distributed around zero, and which has a consistent bias, may be in error.

A suspected point can be immediately checked. Quantifying measurement errors can also help improve the accuracy of other off-line simulation tools. Any process variable that is used as a specification in a process simulator needs to be accurate. Detecting a bias can significantly improve the accuracy of the simulation results.

In addition to highlighting possible measurement errors, an understanding of the behavior of unmeasured variables such as intermediate temperatures or efficiencies can have a positive impact on operations. Data reconciliation results have highlighted differences in efficiencies between parallel compressor trains and furnaces.

16.4.5.5 Off-line usage

The off-line system consists of a web-based interface, an SQL data repository for storing cases, and rigorously validated models from the online system. The interface is designed to facilitate queued simulation and optimization runs of the most common plant configurations using various combinations of flows from the 30+ fields, which feed the plant. Planning personnel have the option to run the model either through the standard web-based simulation interface or through the more detailed model builder interface. The latter is used to explore special situations, which are not supported by the web-based interface. This requires a more detailed knowledge of the model and the simulation software. The model only runs in this mode to conduct special studies with high value.

16.4.5.6 Use for planners

The user of the planning system can use two strategies to create a starting point. In the first, the user starts with an online model taken from the fully operational plant, then specifying the sections that are to be turned off and the feed rates. In the second, they begin with an online model, which very closely resembles the plant configuration they wish to model.

For general studies, the first method is simpler and will generate acceptable results. For very specific conditions such as modeling CO₂ constraints, the second method will generate the most accurate results because the model tuning parameters will have been reconciled for the exact conditions of interest.

The second method is feasible because the online system archives valid models to a separate directory. In this way, a library of models for all process operating modes is created automatically.

To locate the time of the desired operating mode, the user can either look at trends of the appropriate equipment flags or use the operator display to identify the time and date that the appropriate model was created. Once the date is determined, the user can then copy the model from the library and move it into the planning system for further study.
References


Jumonville, J., October 4–7, 2010. Tutorial on cryogenic turboexpanders. Paper Presented at the 39th Turbomachinery Symposium, Houston, TX, USA.


17.1 Introduction

Maximizing the return on gas plant assets becomes increasingly difficult because of the rising cost of energy in some cases, and the increased demand for operations agility in most cases. On top of these demands, there is the constant need to increase the availability and utilization of these plants. It is worthwhile to consider techniques and the profit improvement of some of the world’s best gas plants. This analysis identifies some key manufacturing and business strategies that appear to be necessary and are feasible for many facilities to achieve and sustain satisfactory performance. The essential nature of operating a gas plant is dealing with change. Disturbances such as slugging, trips of gathering compressors, weather changes, and changes in market demands make it challenging to operate reliably let alone profitably. A stabilization strategy is necessary but insufficient for most gas plants; it is necessary to sustain a degree of flexibility. A useful objective can be tracking desired changes and withstanding undesired changes in a manner that is safe, environmentally sound, and profitable (DeVries et al., 2001).

The main challenges to gas plant profitability are as follows:

- Continuous energy, yield, and throughput inefficiency due to providing operating margin for upsets and plant swings.
- Continuous energy, yield, and throughput inefficiency due to not operating at optimum conditions.
- Energy, yield, and throughput inefficiency due to the consumption during plant swings.
- Labor costs to operate and support a gas plant—a poorer performing plant requires more personnel with a higher average salary.
- Plant integrity.
- Process and equipment reliability.
- Poor throughput due to low availability.
- Maintenance.
- Safety.

Poorer performance compounds the cost challenges in most gas plants. To ensure these problems are avoided, it is essential to develop a fundamental understanding of all the technical factors impacting the performance of the plant, how
they interact, and how they manifest themselves in business performance. This can only be achieved through the application of an integrated approach utilizing the diverse skills of a multidisciplined team of engineers combined with the application of a robust performance-modeling tool. Technical assessments allow an understanding of technical risks to which the facility is exposed ranging from deterioration of equipment due to exposure to corrosive environments to poor process efficiency as a result of poor design. Integration of expertise within a single team, which covers the broad range of technical fields involved in the operation of the gas processing plant, ensures that all the technical risks and interactions are identified.

Other business questions that asset managers are asking themselves (Howell, 2004) are given below:

- Are the assets performing to plan, and how do we know?
- Are we choosing the optimal plans for developing the assets over their lifetime?
- Are we achieving the targeted Return on Capital Employed (ROCE) for the assets?
- Are we meeting all of the ever-growing Health, Safety, and Environment (HSE) guidelines?
- Can we forecast reliably, allocate with confidence, or optimize with knowledge?
- Do we derive enough value from the simulation and engineering model investments?
- How effective is the organization at capital avoidance?
- Are we drowning in data or are we knee-deep in knowledge?

Another area of opportunity is optimization of the process to maximize capacity or yields and minimize energy consumption, while maintaining product qualities. Many tools have become available as powerful, inexpensive computing power has become available to run complex software quickly and reliably.

This chapter describes a vision of the integrated gas plant of the future and methods to identify solutions for attaining operational goals to maximize asset values.

17.2 The performance strategy—integrated gas plant

Successful gas plants have found that a combination of techniques is necessary, which are as follows:

1. A strategy to influence organizational behavior
2. A strategy to integrate information
3. An operations strategy that uses remote operations and support of unmanned plants
4. Process performance monitoring
5. Asset management

Merely adding technology without developing a new organizational behavior and operations strategy has often reduced gas plant performance instead of improving it.

The revolution in digital technologies could well transform the industry. Achieving the vision of the integrated gas plant of the future will require more than new technologies alone. It will require the alignment of strategy structure, culture, systems, business processes and, perhaps most important, the behaviors of people. Visionary companies who truly want to capture the “digital value” will need to create a climate for change, and then maintain strong leadership through the change and employ the skills and techniques from vendors to provide the technologies.

Gas plant operators are looking to integrate global operations and the energy supply chain into a cohesive picture. A global enterprise resource planning (ERP) system gives companies the resources they need to better balance supply with demand, reduce inefficiencies and redundancies, and lower the total cost of information technology infrastructures. The challenge is to develop overlay solutions with more domain content that can improve the knowledge inside the enterprise systems, which producers depend on most for operations, planning, project management, workflow, document management, executive information and decision support, scheduling, database management, data warehousing, and much more.

The industry must capitalize on the opportunities provided by ever more capable and cost-effective digital technology.

17.3 Strategies for organizational behavior and information

Understanding and managing organizational behavior is an important element of effective operations. The effective management and coupling of technology to people is a crucial element in ensuring that the plant’s culture will change to support modern manufacturing strategies.

Four typical organizational cultures have been observed (Neumann, 1999). These cultures vary from cultures very resistant to change to cultures that change for the sake of change.

17.4 Organizational behavior model

It is worthwhile to review recent modeling attempts for organizational behavior. Two models, each focusing on different aspects of the organization, deserve exploration. Both attempt to explain the influence of information and management
activities on organizational behavior. The first model, as shown in Figure 17-1, attempts to identify and provide a structure for the various major influences on the organization.

The above model supports the following observations:

1. Unless constrained by their situation, people will improve their behavior if they have better information on which to act.
2. Behavior is constrained by their capability to perform. This constraint is tied to the span of their control, their associated physical assets, etc., and what the military calls “readiness.”
3. The classical personal hierarchy of needs (Mazlow, 1968) that ranges from “survival,” “hunger,” “need to belong” applies to organizations as well.

17.4.1 Information quality

The natural gas processing industry is unusual in its high degree of dependence on information technology in order to meet its business goals. With the enormous quantities of data that it generates and processes, an edge is gained by ensuring the quality of the data and using the information intelligently. Alongside this problem, the industry is addressing the changes required due to new and evolving working practices that the popularity of interdisciplinary asset teams has brought about.

This organizational model explains in part why people and organizations behave in a rebellious manner in spite of receiving “better” information. For example, operators and their supervisors are penalized for “tripping” units and major equipment; however, e-business decisions might require them to operate the plant in a manner that would increase the likelihood of trips. Another example is where a plant or unit’s performance is broadcast to a team of marketing and other plants—while the plant or unit is chronically “underperforming” based on the other organizations’ expectations. This motivates the plant’s managers to hide information and resist plans for performance improvement.
This model also points to a key issue in managing technology. Poor-quality information will be discarded, and information that is harder to use will not be used. This points to the need to define the quality of information. Suggested attributes of information quality include the following:

- Available (is the information chain broken?)
- Timely
- In the right context
- Accurate
- To the right place
- Easy to understand.

It may be helpful to imagine a fuel gauge in the dashboard of a car. If the fuel gauge is working 75% of the time, the driver will not trust it—the gauge will be ignored. In process plants, information has to be much more reliable than that to be trusted.

Gas plants struggle to convert supply-chain schedules into actual plans. Too often, the marketing department does not have appropriate feedback that equipment has failed or that a current swing in recovery mode is taking longer than expected to complete. Conversely, the operations department often lacks the tools to plan which resources and lineups to use, while avoiding poor utilization of equipment. This gap between strategic and tactical scheduling is sometimes referred to as the “operations wall.”

Much attention is given to networking information and knowledge software. So much attention is given to making as many components as possible accessible via the Web, to support Virtual Private Networks (VPNs) and restructuring knowledge software using Application Service Providers (ASPs). This work is worthwhile, but the underlying information reliability issues are chronic and damage the credibility of major network and software installations.

The old expression “if garbage goes into the computer, then garbage will come out” is especially true for modern information technology strategies that end up with more than seven layers of components that process information. It is extremely important to note the following:

1. Much of the information used to enhance and change plant performance, as well as support supply-chain management decisions, is based on sensors that degrade in accuracy and “fail” due to stresses or material that reduces their ability to produce a reliable measurement. A modern hydrocarbon-processing plant uses hundreds of these to support advanced control and advanced information software. Simply connecting digital networks to these sensors does not improve information reliability if the sensors do not have enough “intelligence” to detect and possibly compensate for failure or inability to measure.
2. Most components and software packages process online diagnostic information that gives some indication of the reliability of the information.
However, most installations do not integrate these diagnostics with the calculations. If the information reliability of each of seven components from the sensor to the top e-business software is 95%, then, on average, the overall information reliability is no better than 69%. Raising the component reliability to 99% still only yields a maximum overall information reliability of 93%—unacceptable for supply-chain management.

Therefore, it is important to use as much “intelligence” as possible at all levels, and maintain the integration of online quality statuses to maximize the quality and minimize the time to repair the failures.

17.4.2 Perception of information

Tables of menus and numbers do not help when a downsized team is asked to open multiple applications and “mine” for information. Most information displays are essentially number tables or bar graphs superimposed on plant flow sheets. Several plant examples exist to show how this interferes with improved performance.

17.4.2.1 Two-dimensional curves and plots

As an example, compressors can be the main constraining factor in major hydrocarbon-processing facilities, and it is important to maintain the flow through each stage of compression sufficiently high to avoid surge. Too low a pressure increase is inefficient and limits production throughput. The most efficient is at higher pressure increases, but yields operation closer to surge. Operators are normally given a pair of numbers or bar graphs for pressure and flow, and they need to recall from experience where the “operating point” is compared to the “surge line.” To make matters worse, the “surge line” changes position as the gas’ characteristics change. To solve this, plot the moving surge line and the operating point in real time. This is suitable solution for distillation and reactor temperature profiles as well (Figure 17-2).

17.4.2.2 Prediction trends

The same operators who avoid trips tend to shutoff advanced control software when the software is driving the plant toward undesirable states. But often the software is only using a spike to accelerate the transition to best performance, and the top of the spike still keeps the plant in the desirable state. Prediction trends allow operators to see the remainder of the curve, so that if the curve looks safe, they will allow the software to continue with the spike.

17.4.2.3 Dynamic performance measures

Operators and their supervisors are given performance information and targets, but often the information is not “actionable,” i.e., they cannot directly influence the information (Figure 17-3). Furthermore, as manufacturing strategies change,
they receive e-mails with a set of numbers. Ideally, the senior operators and their chain of management are seeing the targets and acceptable bands along with each actionable performance measure. In this way, coordination of a change in manufacturing tactic is faster and easier. A common example is changing from maximum throughput (at the expense of efficiency) to maximum efficiency (at the expense of throughout) to take advantage of market opportunities and minimize costs when the demand is reduced.

### 17.4.2.4 Performance messages

Operators who monitor long batches, especially batches that will cover a shift change, tend to have lower productivity, especially during times that the main support staff is not available. Ingredients that exhibit quality problems can be used in applications such as in-line blending of fuels, and batch blending with online analyzers to support a performance model with a set of messages and procedures that allow operators to modify the batch during the run, and avoid a rework.
17.4.3 Capability to perform

When the organization can more accurately perceive their performance and the distance away from targets, they can develop a culture of learning how to improve. This is different than monthly accounting reports that show shadow costs and prices. This requires a condensed set of reliable information that supports factors that they can change. Examples include reliability, yield, quality, throughput, and often it might be expressed in a different way. For example, reliability may be more “actionable” when expressed as “cycle time,” “turnaround time,” etc. Cutting the set-up time from an overall 5% to 2.5% is actually cutting the time by 50%; this is far more vivid that showing availability changing from 95% to 97.5%.

Research has been conducted on complex operations where a unit is several steps away from the true customer and attempts to establish internal costs and prices of feedstock, utilities, and internal products have had insufficient credibility or even ability to effect a change. Some attempts at activity-based costing have tried to increase the percentage of direct costs, with the result that the cost per unit of product is so weighted with costs outside of the unit, that the unit manager cannot effect a useful improvement, or is motivated to maximize throughput. Maximizing throughput is the only degree of freedom because the “costs” are relatively fixed. This is incompatible with modern manufacturing strategies that require precise, timely changes in prioritization of throughput, efficiency, and specific formulations to attract and retain key customers and more lucrative long-term contracts.

A key element of an organization’s capability to perform is its ability to handle both planned and unplanned disturbances. Modern manufacturing strategies become a nightmare if the organization and facilities cannot cope with the accelerated pace of change brought on by the tight coupling with suppliers and customers. The following model, as shown in Figure 17-4, helps to describe different levels of readiness.

![Levels of readiness (DeVries et al., 2001).](image-url)
Containment is the lowest level of readiness, and at this level, unplanned disturbances stop production or cause product rework, but the damage is “contained”—no injury, equipment damage, or environmental release. Unplanned disturbances include operator errors, equipment failures, and deviations in feedstock quality as well as other factors. This level of readiness can be achieved with the following conventional information strategies:

- Loose or no integration between components, from sensors to supply-chain software
- Minimal integration of diagnostics of information quality
- Minimal performance information
- Minimal coordination between units or assets to withstand disturbances
- Reactive or scheduled maintenance strategies.

Prevention is a level of readiness where unplanned disturbances rarely affect production availability or product quality. However, this operation cannot consistently support planned disturbances to production rates, feedstock quality, or changes in yield or product mix. It is usually a case of stabilizing the plant as much as possible in order to achieve this level of performance. This level of readiness requires more advanced information strategies, which are as follows:

- Moderate integration between components and a reliable connection to supply-chain software
- Good management of information quality
- Good coordination between units and assets
- Scheduled maintenance at all levels
- Better performance information.

Agility is a level of readiness where the plant can consistently support wider and faster changes in feedstock quality, production rates, and output mix without reducing production availability or quality. This operation consistently achieves “prevention” readiness so that it can outperform other plants because of its agility. This level of readiness requires the most advanced information strategies, which are as follows:

- The tightest integration of all levels of software and sensors
- Good models that drive performance information and online operations advice
- A reliability-centered maintenance strategy for information technology
- Good coordination with knowledge workers, who will likely come from key suppliers of catalysts, process licensors, information and automation technologies, and workers at other sites of the company.

There is a tendency in many operations to try to jump from “containment” to an “agility” level of readiness by adding more information technology, but with the characteristics of “containment” (minimal integration, etc.). This is potentially
disastrous. Supporting modern manufacturing strategies can mean that the quality of integration is as important as or more important than the quality of the information components themselves.

Due to the complexity of the interactions between technical factors affecting asset performance and the difficulty in converting technical understanding into a business context, the traditional approach has been to try and simplify the problem. Application of benchmarking, debottlenecking studies, maintenance and integrity criticality reviews have all been applied to enhance process plant performance.

Each of these approaches tends to focus on a particular element of an asset's operation. As a result, enhancement decisions are made without a full understanding of the impact on the asset's performance and the overall business impact. This can result in opportunities being missed, capital being poorly invested, and delivery of short-term benefits, which are unsustainable. In the worst cases, this can result in a net reduction in plant performance and increased life cycle costs.

17.4.4 Organizational hierarchy of needs

Attempts to evolve plant culture by transplanting methods and equipment have not only failed but actually eroded performance. Key observations include the following:

1. Personnel may not associate performance or knowledge with increased security, wealth, and sense of belonging or esteem. They observe promotion and compensation practices and evaluate the effectiveness of improving their situation against new expectations for better performance.

2. Risk-adverse cultures dread the concept of visual management, benchmarking, or any other strategy that is designed to help a broader teamwork together to continually improve performance. Information that shows performance can be used to support rivalries rather than inspire teams to improve.

3. Organizations and personnel worry about surviving—will the plant be shut down? Will the staffing be reduced? Efforts to change behavior using information have to be coupled to the hierarchy of needs, and then the information can evolve as the organization evolves (and performance improves).

Research on dynamic performance measures has uncovered an effective model of organizational behavior that is especially appropriate for manufacturing. Two of the key issues that have been addressed are overcoming over 85 years of traditional organization structure that tries to isolate units and departments, and the 20 years of cost and performance accounting.

The challenges with cost accounting are assigning costs in the correct manner. A plant manager or operations manager might face an insurmountable fixed cost. This motivates him/her to maximize production, which might be opposite to the
current manufacturing requirements. The poor credibility of internal prices for utilities, feedstock, and products also has motivated middle management to drive performance in the wrong way. These point to the need to redefine key performance indicators (KPIs). The group of research efforts to deal with this is called dynamic performance measures. The US Department of Labor and many large corporations embrace dynamic performance measures.

Vollman et al. (1998) have developed a model that associates the manager’s strategy and actions to their measures of performance. This combination is called the Vollman Triangle, as shown in Figure 17-5.

The manager’s subordinates will develop their own triangles to support their supervisor’s measures, within the constraints of their capability to perform, etc., as outlined in Figure 17-6.

This occurs at all levels of the organization. The goal is to ensure that a small group of measures (no more than 4), which can be consistently affected by that level of operations, are maintained with suitable quality. If this structure is
maintained, then it is easier for top management and modern manufacturing strategies to effect change.

The “actionable” measures tend to conflict with each other. Higher throughput or quality often comes at the expense of efficiency, or each has a different optimum. Each person learns how these interact. Discussions on performance improvement become more effective, because personnel can now describe this behavior. The discussion is vastly different than reviewing monthly reports. As far as operations are concerned, something that occurred several days or weeks ago is ancient history.

17.4.5 Behavior
The most important variables for performance are the work culture, job security, and career mobility. In many parts of the world, knowledge, experience, or skill is not sufficient to achieve promotion. Many cultures do not embrace an openness of sharing performance information, visual management techniques, etc. Therefore, the information strategy needs to be adjusted to reflect the current human resources strategy, and ideally both will evolve in phases to achieve world-class performance.

Another strong issue is traditions at the plant. It is often much easier to initiate new teamwork and performance initiatives to a new plant with a new organization. Nevertheless, the right information strategy becomes a catalyst of change. Everyone can see the performance against the current targets, and everyone can understand the faster change of priorities—such as changing from maximum production to minimum, where maximum efficiency is desired.

17.5 The successful information strategy
Technology makes it easier to measure variables such as pressure, speed, weight, flow, etc. But many operations achieve profitability with very complex equipment, and the key characteristics are properties, not basic measurements. A notable example is found in oil refining. The basic manufacturing strategy is to improve yield by converting less valuable components from crude oil (molecules that are too short or too long) into valuable components (medium-length molecules). However, the essential measurement to determine the proportion of different molecules is chronically unreliable. New technology now exists to reliably indicate a property called “carbon aromaticity,” which helps the operations team ensure that the process will be effective and that the feedstock will not degrade the expensive catalyst materials and erode production.

Reliable property measurement is the key. This also helps to evolve the plant culture from thinking about pressures and flows to thinking about properties. The information strategy becomes the following:

1. Assess the organization’s readiness to support the desired manufacturing strategy and use of information to evolve the plant culture.
2. Depending on the organization’s hierarchy of needs, develop a phased evolution of the information strategy.

3. Develop appropriate measures that will directly support the actions needed to implement the strategy at all levels.

4. Develop the appropriate information to maximize its quality, reliability, and perception.

5. Manage the quality and reliability of information.

Data integration and visualization will continue to be a key area of focus throughout the next few years in digitizing the asset. Many new applications will bring knowledge-gleaning ability from the data collected.

17.6 The impact of living with information technology

If the organization invests in “intelligent” components that enable advanced maintenance strategies such as reliability-centered maintenance or performance-monitoring centers, then the organization is committing to the advanced strategy. Otherwise the flood of extra information will be disruptive. Intelligent sensors deliver up to 10 times as much information as conventional ones; a typical downsized organization, with as little as one-fifth the staffing of 10 years ago, would have to deal with 50 times the volume of information—to much. The information is only useful if it is coupled with appropriate software and methods to use the information to optimize the maintenance and maximize overall information and plant availability.

If the organization invests in supply-chain management technology, then the organization is committing to an “agility” level of operations readiness and a team-based performance culture. The organization is also committing to business measurement changes—cost per unit of product gets in the way of profitability measures of reducing procurement costs or increasing return on sales using modern manufacturing strategies.

The organization develops a new language and starts to use new terminology. Furthermore, managers start focusing on managing the higher levels of information. The following model, as shown in Figure 17-7, suggests a new terminology for dealing with information.

This terminology can be defined as follows:

- **Data**—the raw information from sensors and personnel keyboard entry.
- **Information**—the validated data, using diagnostics and techniques to enhance the quality and reliability of information.
- **Knowledge**—a comparison of information to targets and constraints. This answers the set of “how are we doing?” questions.
- **Wisdom**—guidance for best practices, customer satisfaction, supply-chain management, and any other performance guidelines.

Many organizations have managers spending much of their time dealing with the “Information” function, rather than focusing on “Knowledge” and “Wisdom.”
It is extremely important to maximize the quality and type of information to ensure that management can support the “agility” demands of modern manufacturing strategies.

17.7 Vision of the modern plant operation

There are four different groups of activities that require consistent information for most effective operations. These include the following:

- Operational decisions on an hourly or more frequent basis by operators, shift supervisors, or engineers.
- Tactical decisions on a 1 to 30-day basis by shift supervisors, engineers, purchasing, trading, and accounting.
- Strategic decisions made on a one-month to five-year basis by plant management, purchasing, and accounting.
- Monitoring on at least an hourly, weekly, and monthly basis by all parties involved.

Operational decisions include determination of set points for the equipment and switching equipment on or off. These decisions need to consider current pricing for commodities such as fuel and electricity, equipment availability, and environmental constraints. Furthermore, environmental constraints might be accumulative.

Tactical decisions include maintenance scheduling, demand forecasting, production planning, and emissions forecasting and trading. Strategic decisions require evaluation of future investment, budgeting, and long-term contract negotiations with suppliers and customers. Monitoring requirements include
tracking plan versus target versus actual performance such as process energy use, cost accounting based on real costs, emissions accounting and performance monitoring of utilities equipment.

Given that the quality and perception of information has to be improved in order to be used effectively to change an organization’s culture, there are five key issues highlighted by the above model that must be addressed:

1. Capability to perform
2. Operations readiness
3. Organizational hierarchy of needs
4. Establishing “actionable” information to act as a catalyst of change and
5. Measuring the right things.

Technology can be the most effective tool to evolve plant culture. The speed of distributing reliable, timely, and easy-to-use performance and supply information changes the level of discussion and the dynamic behavior of the organization. But the organization needs to have the right set of performance measures, a persistent effort to enhance operations readiness, an appropriate human resources strategy, and above all, appropriate quality of information. The impact of these on changing a plant culture can be dramatic—as soon as the team realizes the ability and need to improve, they learn quickly how to evolve their work methods to achieve it.

Modern manufacturing strategies are viable, and information technology strategies can transform the plant culture to achieve manufacturing success. The evidence of a change in the level of dialog and team dynamics confirms the success. But several key strategies must be consistently maintained in order to sustain this success.

17.8 Operations strategy

Major installations have been able to consolidate up to 10 control rooms and improve plant output by up to 15%. The goal was not to reduce personnel—the goal was to improve flexibility because it is easier for a small team to drive change than a large one. This has been consistently proven in remote jungle sites in Southeast Asia, remote mountain sites in the United States and Canada, and remote sites in Northern Africa. There are several techniques necessary to make this feasible:

- Remote, secure, online computer access to all control and monitoring equipment.
- Advanced alarm management that dynamically filters alarms during upsets, so that “alarm showers” are avoided that blur the visibility to key cause and effect alarms.
- Stabilization techniques to help units withstand upstream and downstream disturbances.
Facility simulation has developed quickly over the last several decades. The integrated oil and gas field with production facilities is a complicated system with a high degree of interaction and dynamics that make it impossible for the human mind to control and optimize both technical and business parameters. Integrated discipline workflow and the development of model-based asset management are necessary to effectively deal with these complexities.

Model-based asset management techniques will begin to play a major role in the integrated gas plant of the future in bringing the predictive power of the production engineering toolset to the real-time data platforms. Process and production simulation models will move from the domain of the engineering expert to be used by Managers, Operators, Business Development, Contracts, and Finance. The analysis and prediction of near-real time and future asset performance becomes a reality in the world of model-based asset management. For this, the complete integrated asset will be modeled dynamically in real time for both slow rigorous and fast proxy loop modes.

Facility simulation has come of age. The ability to predict very complex facility simulations in both steady state and dynamic mode has been achieved by continued integration of thermodynamic methods, hydraulic simulations, and unit operations. Facility simulation has progressed such that dynamic start up and shut down simulations have become routine and trusted among the engineering community. One can begin to think of dynamic simulation as a “Virtual Plant” for operational, advanced process control and business analysis by a number of different departments.

Facilities simulation, control, optimization, operator training, collaborative engineering and planning disciplines have a rather fragmented application of a number of associated technologies. To overcome this fragmentation, the first step is the adoption of Integrated Asset Models that use comprehensive existing applications that are integrated together by a common “glue” layer allowing a full model of the entire operation.

New concepts are appearing from the software community in the form of workflow solutions that enable assets to be modeled from a suite of selected software adaptors and can bring data from disperse and third party tools into one common environment, creating the Integrated Asset Model. One can then apply engineering and business applications to identify asset-wide improvement opportunities.

A portal environment can be built that will allow multiple asset models to be visualized, compared, and analyzed with distributed data sharing and events management, via proven and accepted enterprise platform message bus technologies. Key visualization and workflow technologies required include look forward analytics, production scorecard, workflow management, production
reporting, capital planning, and scheduling that can be performed on a uniform, company-wide, basis and allow for rapid, informed decision-making.

17.10 Optimization

Maximizing the profit contribution of natural gas processing facilities is challenging given the fluctuating economics, changing ambient conditions, and feed variations that processors must address.

New dynamic markets for gas components lead to a need for stronger analytical capabilities of decision support tools. Also the supply situation becomes more variable, as the gas companies respond to market opportunities. The result is a more rapidly changing environment and as a consequence processing plants needs to be reconfigured more often. There is a need for better understanding on how to make plant-wide production plans and implement these through process management.

Decision support tools have to combine both optimization and simulation capabilities in order to analyze the consequence of different scenarios. Advanced modeling and optimization is needed in order to address the new following challenges in plant control and optimization systems, which have impact on the optimal design of gas processing plants:

- Many of the processing plants are short of capacity. It is more important than before to maximize flow through the plants or the profit from final products.
- New opportunities exist because of recent advances in modern control technology. For example, Model predictive control lifts the level of automation and gives better opportunities for process optimization systems at higher levels.
- Integration of tasks and systems in operation support centers. E-fields give a new perspective on operation of the oil and gas production, and this mode of operation must be prepared for in the gas processing plants.
- The Man-Technology-Organization (MTO) perspective, which provides relevant information for the personnel involved.
- The use of models for planning and process control plays a central role.
- Modeling the dynamics of the markets and incorporate it in contingent plans. The plans should reflect capacities and production possibilities of the plant. This again depends on the design and operation of the plant control system.
- Structuring of the information flow between layers and systems in the process control hierarchy, the real-time optimization (RTO) level, and the planning and scheduling levels.
- Self-regulating and robust systems, which optimizes the other parameters.
- Limitations imposed by process design.
17.10.1 Tools for optimization

Many of the tools mentioned in Chapter 15 could become integral elements of the operations strategy for maximizing profitability. A steady-state process model is necessary to determine the capability and current performance of the operation. Ideally, a dynamic model would be available to train operators and investigate control options. These models can then be used in a real-time mode to report the capability of the plant under different conditions.

Predictive equipment models can be used to determine when maintenance will yield long-term benefits that more than offset the short-term costs. These models can be an extension of existing steady-state models. These steady-state models can also be key elements of an online optimization strategy.

Real-time control models can be constructed from perturbations of a high-fidelity dynamic model. It is necessary to validate these models constructed from models against actual plant data. Often the control models require detuning to provide adequate and robust control.

17.10.2 Optimization alternatives

Some optimization alternatives for natural gas processing plants include the following:

- Advanced regulatory control
- Multivariable predictive control
- Neural network controllers
- Off-line process simulators
- Online sequential simulation
- Online equation-based optimization
- Linear programs
- Web-based optimization.

Advanced regulatory control and multivariable predictive control are discussed in Chapter 15.

Neural network-based controllers are similar to multivariable controls except that they gather plant data from the DCS and use the data to “learn” the process. Neural network controllers are said to handle nonlinearities better than multivariable controllers and are less expensive to commission and maintain.

Neural network-based models are only valid within the range of data in which they were trained. Conditions outside the range of data that were used to train a neural network model may be suspect. For example, if a neural network model was trained on data collected while a Joule–Thomson (JT) valve was closed, then the predictions from the model are invalid when the JT valve is open.

Refining and chemical companies have attempted using neural networks for control many times over the last 10 years. The technology has not proven to be viable compared to the other approaches such as advanced regulatory control and multivariable control.
Off-line process simulators are used to develop a rigorous steady-state or dynamic model of the process. They are used by process engineering personnel to design and troubleshoot processes. Off-line simulations allow for what-if case studies to evaluate process enhancements and expansion opportunities.

Off-line simulators are typically not used to support daily operational decisions. They must be updated and calibrated to actual plant conditions for every use. They are not as robust as equation-based optimizers and can have difficulty in converging large problems reliably and quickly.

A few of the off-line simulation companies offer an inexpensive, sequential-based optimization system. The optimizer is based on a rigorous steady-state model of the process and is typically less expensive than equation-based systems. They leverage the work done to develop the off-line model for online purposes. The extended convergence times inherent in these systems bring into question the robustness of the technology. The sequential nature of the solving technology also can limit the scope of the system. These systems require hardware and software to be purchased, installed, commissioned, and maintained onsite and require specialized resources to support them.

Equation-based optimizers use a rigorous steady-state model of the process as the basis for optimization and include an automatic calibration of the model with each optimization run. The equation-based solving technology allows optimizers to execute quickly and robustly making them viable for larger scale problems i.e., multiphase load optimization for plants on a common gathering system.

Equation-based optimizers require a hardware platform, a costly software component, and highly specialized engineering services to install, commission, and maintain the technology. Closed-loop implementation requires a multivariable controller to be installed to effectively achieve the optimal targets.

Online equation-based optimizers, when coupled with a multivariable controller, represent the standard in optimization technology for refining and petrochemical industries. Most refiners and petrochemical companies are rapidly deploying this technology to improve the profit contribution of their larger scale processing facilities. Unfortunately for gas processors, this technology is justifiable only for very large gas processing facilities and is not scalable across their asset base.

Linear programs are used for evaluating feed and supply chain options. Linear programs are an off-line tool that allows for what-if case studies and evaluation of supply-chain alternatives. They are relatively inexpensive. Linear programs provide a linear representation of the plant process and do not provide guidance for operators (Roop et al., 2002).

Web-based optimization has been applied to cost-effectively supply equation-based optimization to the natural gas processing industry. However, the time lag in collecting data, calculating the optimum, and providing advice to the operators to implement the advice may not be fast enough to keep up with the constantly changing conditions experienced in natural gas processing plants.

Another option is online performance-monitoring tools that predict the optimum operating point under all conditions (Ralph et al., 2011). These tools
usually provide dashboards and graphical indicators to show the operator the gap between current plant performance and optimum performance. They may or may not provide advice on what parameters to be changed to reach optimum performance.

17.11 Industrial relevance

In the upcoming decade, there will be large investments in gas production, and in facilities for transportation and processing. Optimal utilization of all these facilities is vital in order to maximize the value from produced natural gas. Advanced process control and operation has raised the level of automation in the process industry significantly in the last decades. For example, in the refinery industry, methods like Model-based Predictive Control (MPC) and RTO have become widely used (Qin and Badgwell, 2003). The focus on these technologies has given large benefits to the industry in form of increased throughput and more robust operation. The improvements obtained by the use of better control and decision support tools ends up directly on the bottom line for the operating companies. The substance of these tools is in fact software realization of process knowledge, control, and optimization methods. MPC and RTO are now more or less off the shelf products, although for complex processes the adoption of this technology requires specialist competence. However, there are significant potentials for further improvements in this area. One challenge for the gas processing plants is being able to quickly adapt the plant operation to dynamic changes in the markets, thus the plant flexibility and ability to perform rapid production changes becomes more important. It is also required to know accurately the plant production capability, both on very short term (today–tomorrow) and on weekly, monthly, and even longer horizons. And this requires use of advanced optimization tools and efficient process calculations. It is important to consider these issues related to dynamic operation also in the design of new processing plants, and for modification projects. Investments in this type of project have typically short payback time (Moen, 2004).

Combination of individual units into an integrated plant gives a large-scale control problem that is more than just the sum of the units. Cross-connections, bypasses, and recycling of streams give more flexibility, but at the same time, the operation becomes significantly more complicated, and it is almost impossible to utilize the full potential of a complex plant without computer-based decision support tools. Thus, there is a need for development of new decision support tools that combines optimization technology, realize process calculation models at a suitable level of speed and accuracy, and structure the information flow, both from the process measurements and deduced variables, and from the support tool down to the manipulated variables in the control system. It is also needed to develop further the methodology related to plant-wide control in this context (Skogestad, 2004).
17.12 The technology integration challenge

There are three likely scenarios of the deployment of the integrated gas plant:

*Scenario 1: Business as Usual*—Digital technology, information, data, and models are used in an incremental way to reduce costs, increase recovery, and improve production, but no fundamental changes are made in business models, competitive strategies, or structural relationships.

*Scenario 2: Visionary*—Those who can best adopt and apply digital technologies, and concepts, will use them to gain significant competitive advantage. This will involve significant investment in software and IT technologies along with culture and management change. The industry will need to demand leaders in the highly technical processes and modeling software contribute new and innovative solutions.

*Scenario 3: Symbiotic Relationships*—Those who use the availability of turnkey solutions to optimize production and leverage into larger industry positions. This will involve the use of third-party technology consultants and solution partners in an unprecedented way.

For most companies, an evolution approach that blends all three scenarios may well be the chosen route.

The foundation of the integrated gas plant of the future is engineering simulation, with integrated asset models and portfolio views of the business built on top. At the heart of the digital revolution in the upstream energy industry is a shift from historic, calendar-based, serial processes to real-time, parallel processes for finding, developing, and producing oil and gas assets. Real-time data streams, combined with breakthrough software applications and ever-faster computers, are allowing the creation of dynamic, fast-feedback models. These dynamic models, running in conjunction with remote sensors, intelligent wells, and automated production and facility controls, will allow operators to visualize, like never before, what is happening in the facility and accurately predict what needs to happen next to maximize production and efficiently manage field development.

17.13 Scientific approach

Optimal operation of gas processing plants is a challenging multidisciplinary task. Large-scale process optimization is challenging in itself. Thus, when we also will consider dynamic conditions in the market and on the supply side, the operation will most certainly run into problems that must be solved. Some will arise from the size of the problem, some from complex process behavior and from requirements to solution of complex optimization problems.

The starting point in this project is the need for decision support as seen from the personnel in a plant operating company. This defines a set of tasks that requires optimization calculations, process calculations, and measurement data
handling. The personnel in question can be plant operators, production planners, sales personnel, maintenance planners, process engineers, managements, etc. Experience from other applications like gas transportation will be utilized (Rømo et al., 2003).

The inclusion of market factors, capacity planning, and scheduling shall be focused, as this sets new requirements to vertical integration of process control and the optimization layers. We may in fact have several such layers, where, for example, the classical RTO is just one element. In planning and capacity assessment, the RTO layer may be accessed by superior layers in order to compute the optimal process targets over a certain horizon.

The requirements to process calculations for each type of task shall be classified. This may result in a set of optimization problems with different properties and requirements to solution approach and to the underlying process calculations and data handling, e.g., one approach from the planning side is to start with empty or extremely simple process models and to refine the models based on the requirements to the planning.

Segmentation into suitable process sections and control hierarchies are central issues. Here, we can apply methods from plant-wide control in order to structure the control of the plant units in a way so that the influence from unknown disturbances and model uncertainties are minimized (Skogestad, 2004). A very important output in the first phase is to define high-level targets for the process control. The next important issue is to develop methods to select the variables that should be exchanged between the optimization and the process control layers. This is a control structure design task where the focus is on selecting the variables that are best suited for set point control in order to fulfill the process optimization targets in presence of unknown disturbances and model parameters and measurement errors. Segmentation of the control into suitable sections and layers is also a part of this task.

Recent advances in process control technology also give a new perspective. For example, with an active MPC controller, information about active and inactive process constraints is high-level information that can be exchanged with the optimization layer, instead of representing the constraint equations at the optimization layer.

Efficient use of models is a so wide area that this issue can be subject to extended research in separate programs. For example, in process design, it is industrial practice to use quite detailed process models, including rigorous thermodynamics and representation of detailed phenomena within each process unit. In operator training simulators, detailed dynamic models are used, but these are rarely the same models as used in design, and the built-in process knowledge in form of model configurations and parameters is usually not interchangeable because of different modeling approaches and different model data representation. The models used in MPCs are normally captured from experiments on the process itself and are not connected to the other two types of models. For RTO, steady-state models are normally used, and in some cases, model
tools with rigorous models are used there too. For capacity assessment, correct representation of potential bottlenecks is important.

17.14 Other miscellaneous initiatives

Maintenance Management, Field Information Handling, Work Process Optimization, Compensation Design, and Procurement Initiatives are several of the current gas processing management initiatives.

In Field Information projects, companies considering upgrade should understand that technical support must also be upgraded, and that care should be taken to select systems with an eye toward ensuring ongoing availability of support over a reasonable period.

In an effort to determine how they are doing against the competition as well as discovering new areas for potential, some processors have become involved in industry benchmarking activities. Benchmarking tools with a reasonable level of analytical content provides benchmarking against a select peer group as well as individual analysis of various cost components. Most organizations that persevere in the benchmarking process and are diligent to follow up on findings testify that benchmarking is a useful tool.

A company that fully utilizes second-wave technologies to streamline its back-office and process support technologies could reap a reduction in selling, general, and administrative (SG&A) costs in the range of 8–10%. For a typical, large firm, SG&A costs represent an estimated 10% of the total enterprise costs. Thus, a 10% reduction in SG&A outlays would reduce overall corporate spending by one percent; a major gain, given that these savings would drop to the bottom line.

17.15 Conclusion

The goal of an integrated operations environment is to enable direct translation of management strategies into manufacturing performance. The vision is that

- Utilization of raw materials is optimal
- Overall margin and yield of product(s) is maximized
- Planning, operational, and monitoring cycles are fully integrated
- Identification and correction of problems occurs rapidly
- Operational (short- and long-term) factors are fully understood
- The workforce is well informed and aligned for a common purpose.

The most effective system is an integrated platform for computing and information processing at the production level. It is built around the premise that information is not to be isolated and that better information, when made widely available, will help people operate the facility closer to the optimum. A key
principle is to empower everyone to maximize the value of their activities. The production management system provides the tools to help personnel do their job better.

A fully functional system will enable the quality cycle of planning, measuring, analyzing, correcting, and then planning again. To make improvements, the staff must be able to see and measure progress.

The integrated production/management system provides the data and the means to analyze situations, define solutions, and track progress. It is an integrated platform of computers, networks, and applications. It brings together the many individual automated systems that exist today and fills any gaps to bring the overall system to a high level of performance. The production/management system spans the gulf between process control and corporate business systems to support the day-to-day operations.

To achieve these goals, the production management system must:

- Be a single, comprehensive source of real-time and business data addressing all operations and available to all appropriate personnel. This means providing long-term storage of all data (e.g., historical process, laboratory, plan, production, and shipment data), merging of these data, and retrieving data.
- Provide information-retrieval tools for the full range of users. This usually means highly graphical tools that provide ease of use and a consistent look and feel to minimize the burden of finding and accessing information.
- Integrate a wide range of computer systems and applications. No one system or set of tools will provide all the functionality needed. Instead, the production management system should allow for the use of the best products from different vendors.
- Provide standard screens and reports that focus attention on problems and opportunities. The system should report by exception, highlighting the unusual, the exceptions, and the opportunities. It should compare actual results with the established plans and economic KPIs.
- Maximize information content, while minimizing data volume. This is achieved by the use of performance indices and other numerical, measurable indicators, and the presentation of this information in graphical form whenever possible.
- Present operational data in economic terms whenever possible. Opportunities, problems, and deviations from an operating plan should be prioritized based on their impact on overall profitability, and where possible, include an indication as to the action or follow-up activity that alleviates the deviation.
- Provide analytical tools that enable users to explore and pursue their own ideas. Much of the value of integrated operations comes not from presenting data about current operations, but from people looking for ways to improve current operations.
Facilitate plant-wide communications and workflow. For instance, plans and economic KPIs set in the planning group should flow automatically to operations to both help operate the plant. Beyond the manufacturing issues, implementing a project with the scale and complexity of an integrated production/management system creates several management of technology issues.

These create the need to

- Balance the selection of individual applications with the need to integrate applications across departmental boundaries.
- Provide a single, accessible, look, and feel. This is particularly important for users accessing data that originate in systems that belong to other departments. “Single pane of glass.”
- Employ the latest proven information system technology as it becomes available and at the pace the operator can assimilate and manage.

Integration is a true example of the total being greater than the sum of the parts. A gas plant can profoundly affect the nature, quality, and profitability of its operations throughout the life of the gas plant with a truly integrated production management system.

Several operations have adopted some or all of the discussed strategies to improve their performance. A couple of these plants are discussed in Kennedy et al. (2002). Other examples include gas processing operations in Tunisia, Norway, Nigeria, and Indonesia.

The future seems to belong to those who will be able to mix vision, intelligence, and understanding of human nature, technology, and the processing business into a formula for success in the new world of natural gas gathering and processing!

References


Neumann, R.W., March 1–3, 1999. Managing your gas processing plant fundamentalist, fashionable, farsighted or fantastic! Paper Presented at the 78th GPA Annual Convention, Nashville, TN, USA.


Ralph, B., Hughes, J., Sigal, R., Morrison, R., Poe, W., April 3–6, 2011. Gas plant process manager application at Williams Mobile Bay Plant. Paper Presented at the 90th GPA Annual Convention, San Antonio, TX, USA.


Roop, M., Leger, J., Hendon, S., March 11–13, 2002. Alternatives to optimize gas processing operations. Paper Presented at the 81st GPA Annual Convention, Dallas, TX, USA.


18.1 Introduction

Project management is the application of knowledge, skills, tools, and techniques to project activities in order to meet or exceed stakeholder needs and expectations of a project. The project manager sometimes referred to as the project coordinator or leader coordinates project activities on a day-to-day basis. This is an ongoing challenge that requires an understanding of the broader contextual environment of the project and the ability to balance conflicting demands between (1) available resources and expectations (especially with respect to quality, time, and cost); (2) differing stakeholder priorities; (3) identified needs and project scope; and (4) quality and quantity of the project’s deliverables. Project management for engineering and construction projects requires the application of principles and techniques of project management from the feasibility study through design and construction to completion. Good project management during the early stages of project development greatly influences the achievement of quality, cost, and schedule.

There must be a particular focus on predictability, transparency, and reliability, including managing the costs associated with these projects. With more joint venture projects, there is often the need to manage these relationships and preserve reputations.

This chapter covers many aspects of managing capital projects in the gas processing business. For the most part, best practice management for gas plant projects follows generic project management principles applicable to most industrial engineering and construction projects. This chapter reviews many of the standard and accepted practices that lead to successful installations as well as some of the unique considerations for gas plant projects, which arise from relatively complex processes employed in typically remote locations.

18.2 Project management overview

One or more parties can perform the design and/or completion of a project. Regardless of the method that is used to handle a project, the management of a project generally follows these steps:

■ **Step 1**: Project Definition—Determine the conceptual configurations and components to meet the intended use.
Step 2: Project Scope—Identify the tasks that must be performed to fulfill the project definition. Also clarify what the project does not include.

Step 3: Project Budgeting—Define the permissible budget plus contingencies to match the project definition and scope.

Step 4: Project Planning—Determine the strategy and tasks to accomplish the work.

Step 5: Project Scheduling—Formalize the product of planning.

Step 6: Project Execution and Tracking—Complete project tasks and measure work, time, and costs that are expended to ensure that the project is progressing as planned.

Step 7: Project Close-out—Final testing, inspection, and payment upon owner satisfaction.

Successful projects require effective management, which means: (1) clear objectives, (2) a good project plan, (3) excellent communication, (4) a controlled scope, and (5) stakeholder support. Project management in today’s organizations demands multiskilled persons who can handle and manage far more than their predecessors and requires competencies that span all of the critical management fields.

18.3 Industry perspective

Gas plant project management includes the planning, design, engineering, construction, and commissioning of the plant. Key elements are covered under the general headings of Engineering, Procurement, and Construction (EPC). Today in many large organizations there has been a trend away from the owner company performing the whole project management function toward the delegation of EPC in part or whole to engineering consultant organizations. At the same time, smaller companies have almost always subcontracted EPC activities. Most companies, however, specify and procure the major equipment themselves (or at least oversee those activities) to assure technical compatibility and adequate lead times for delivery, and so on. For example, the compressors and drivers will be specified and selected by the owner company as a priority.

A successful project in the gas processing industry is not only one that is profitable, but one that leads to the safe, reliable, predictable, stable, and environmentally friendly operational characteristics.

Gas processing is a service to the oil and gas production business. Oil and gas operations desire to produce into a system that has high availability and can produce saleable product in a safe, quality-consistent and environmental friendly manner. Gas processors must, of course, provide this service in a profitable manner. Flexibility to operate in various modes to respond to the markets and provide various processing alternatives should provide competitive advantages for a plant.

In many cases, the gas processing facilities are owned by the oil and gas producers as first facilities to enable the production of oil and gas and secondly as
a value added operation. In some cases, particularly the present state of the 
business in the United States, independent gas processors compete for gas that 
can be produced into a number of gas gathering systems. Several types of 
contracts exist, but the prevalent contract type is a “percent of proceeds” 
arrangement. Processors, who can offer the greatest revenue to the oil and gas 
producers, have an advantage in any case. The key is recovery of the greatest 
percentage of feedstock at the highest availability and at the lowest cost. 

Adoption of integrated plant design and engineering is allowing gas 
processors, licensors, as well as engineering and construction firms to streamline 
workflows to improve efficiency and execute projects faster (Mullick and Dhole, 
2007). Integrated models and tool sets allow rapid learning and ensure reuse of 
information and knowledge throughout the life cycle of the operation. 

Engineering and construction firms are also are utilizing their information 
technology infrastructure to support collaborative engineering environments to 
manage and execute projects around the clock and across the globe. This global 
execution capability allows companies to fully utilize available talent in a cost-
effective manner and to improve project schedules. 

Rapid deployment of new engineering tools across the organization through a 
“virtualized” environment is another emerging trend.

18.4 The project management process

At the onset of a project, the owner company will undertake the required economic 
and business analysis regarding new or expanded facilities in order to receive board 
approval and budget allocation. Based on this approval and funding, the project 
definition will be refined. The owner company will initiate the project and set out 
design objectives, usually embodied in the Design Basis Memorandum (DBM), 
which will lay down the operating parameters and any key design guidelines and 
specifications. The owner company will solicit bids from EPC contractors and from 
those will select the successful bidder. In some cases, owner companies will have a 
partnership with an EPC contractor for certain types and sizes of projects. Pricing 
will be prenegotiated according to a range of possible contract models (fixed price, 
cost-plus, risk-sharing, etc.) or performed on a reimbursable basis. Both parties, 
owner and contractor, will set up teams to do the work. The EPC company may be 
asked to be completely responsible for all aspects of engineering, procurement, and 
construction or may only be required to do engineering and some procurement with 
the construction carried out by another company. Estimates and schedules will be 
set up by the EPC company in consultation with the owner. 

Consultants and contract personnel will fill areas where the EPC company 
lacks resources or expertise. The owner also embeds staff into the contractor to 
ensure oversight of the management process and also may include specialist 
engineering personnel to monitor progress against plan including quality assur-
ance as well as training of the owner’s staff on major projects.
For more complex, large projects that involve elements of innovation or requirements to build a gas facility in a region or country where such projects have not previously been conducted, it is not unusual for the process to include feasibility studies and front-end engineering and design (FEED) studies conducted prior to awarding the EPC contract. The feasibility and FEED studies are usually conducted by competent engineering consulting companies, capable themselves of conducting the EPC work. The deliverable from a FEED study may form the basis of a competitive tender for the EPC contract in which a number of prequalified contractors may compete.

A successful project requires the owner and EPC companies to work very closely together. Typically, the owner will bring in operations staff at an early stage in the project to ensure that these staff members contribute to the project specifications and review deliverables as they unfold. In this way, commissioning and operation proceed without serious problems. Companies with experienced staff often prefer to have the operating people closely integrated with the construction work from the outset. Similarly with the controlling people it is essential that the owner’s control philosophy is conveyed to the engineers of the contractor and that the machinery suppliers are brought into the loop so that engineering specifications and machine products reflect these perspectives. The owner company will undertake the required economic and business analysis regarding new or expanded facilities to receive Board approval and budget allocation. From this the project will be defined.

18.4.1 Defining business and project objectives

The first step in the project management process is to align the business and project objectives. A project can be installed on time and on budget, but if it does not meet the defined business objects, then the project cannot be deemed a success. Some of the questions that the business owners must be asked by the project team include:

- How much gas is available for processing (ultimate reserves and daily deliverable quantities)?
- What is the market demand for gas and gas products that can be met by this project?
- What is a realistic gas production schedule?
- What are the production pressure, temperature, and composition of the gas?
- How will the gas pressure, temperature, and composition change over time?
- What products can be sold and at what price?
- What are the product specifications?
- How will the products be delivered to market?
- What are the local environmental policies?
- What are the local safety policies?
- What infrastructure such as roads, bridges, loading and unloading facilities, personnel housing, etc. are required?
- What is the skill level of available operations and maintenance personnel?
Since most gas processing plants are services to the oil and gas producers, then collaboration between reservoir and production engineers and gas marketers to obtain answers to these questions is imperative. In cases where processors compete, then collaboration with those responsible for obtaining the processing and sales contracts (e.g., economists, lawyers, and negotiators) can be critical.

18.4.1.1 The project charter
The project charter is a document that demonstrates management support for the project. In particular, it authorizes the project manager to lead the project and allocate resources as required. It simply states the name and purpose of the project, the project manager’s name, and a statement of support by management. Senior managers of the responsible organization and the partner organizations sign it. The project charter should be distributed widely—to anyone with an interest in the project. This will help build momentum, encourage questions and concerns early in the project’s evolution, reinforce the project manager’s authority, and possibly draw other interested and valuable stakeholders into the project.

The project owner may be a joint venture of oil and gas companies, with one designated as project operator. The project charter is then usually signed off by all joint venture partners together with an authority for expenditure (AFE) approving the project budget and/or initial stages of expenditure.

18.4.1.2 Project team roles and responsibilities
Project team size and makeup is dependent on the complexity of the project, however the basic composition of the project team and their responsibilities is recommended for all projects including:

1. Project Manager

The project manager is responsible for project development, developing schedule, budget, and deliverables definitions; evaluation of alternatives; determining return on investment; adherence to company policies; obtaining funding; acquiring internal and external project resources; contractor selection; maintaining project schedules and budgets; evaluating quality of project deliverables as they evolve; identifying and mitigating downside risks; identifying and exploiting upside opportunities; reporting to business owners; and creating project close-out reports.

2. Business Owner Representative

The business owner representative is responsible for assuring that the project adheres to business objectives as objectives may change or require alteration during the project.

3. Plant Manager

A plant manager should be appointed as early as possible to address operability and maintainability issues.
4. Project Engineer/Construction Engineer/Start-up Engineer

A project/construction/start-up engineer can be one role on smaller projects and multiple roles in larger projects. This engineer (or engineers) is responsible for technical specifications for contract bidding purposes, technical evaluation of contractor bids, owner’s representative during construction, management of construction inspectors, turnover of facility to operations, training of operators, determination of plant performance, and identification of any project deficiencies.

5. Purchasing Representative

The purchasing representative is responsible for commercial evaluation of contractor bids and negotiation of contract.

6. Process Engineer

A process engineer is recommended for evaluating alternative processing schemes during project development, assistance with technical specifications and evaluation of contractor bids, and assistance with operator training and start-up issues.

7. Environmental Engineer

An environmental engineer is recommended to review and provide advice on environmental issues encountered during the project including technical specifications and obtaining environmental permits.

8. Safety Engineer

A safety engineer is recommended to review and advise on safety issues encountered during the project including technical specifications and participating on hazard analysis evaluations.

9. Production or Reservoir Engineer

A production or reservoir engineer is recommended to be available to evaluate any oil and gas production issues that may be encountered during the project.

10. Facilities Planner

For larger projects a facilities planner should be available to assist with project economics and to serve as a liaison for economic premises and marketing issues.

18.4.2 Contracting strategy

There are several alternative contracting stages and strategies. The first stage of contracting may be a front-end engineering design. With this approach, a contract is entered based on the design objectives for an engineering contractor to evaluate process and construction alternatives as well as develop technical specifications
for the project. In some cases, the owner’s engineers may accomplish the front-end engineering design tasks. The second stage of contracting is for EPC services. Either stage can be contracted as a lump sum, fixed fee price also known as a turn-key project or on a reimbursable basis also known as a time and expense contract. In some cases more complex contracts, such as risk-sharing or gain-sharing, will divide risks and rewards more evenly between the contractors and project owners.

18.4.3 Conceptual estimates and schedules

Most operating companies have developed estimating tools for budgeting of plants similar to those they are currently operating. Many operating companies have the capabilities and resources to evaluate alternative process and mechanical designs with budgetary or conceptual level estimates. These estimates typically have an accuracy of ±30–40%.

Under other circumstances such as when proprietary processes are in use, unique locations are to be selected, or if there is a lack of available resources, an engineering firm may be hired to evaluate alternatives and determine budgetary estimates. After evaluation and selection of a conceptual process and mechanical design, the operating or engineering company will undertake a front-end engineering design. The detailed specifications and request for proposals will be the deliverable from the front-end engineering design.

Conceptual estimates and schedules should take the following into consideration:

- Location
- Operators and Operability
- Constructability
- Special Materials.

The availability of fresh water and electricity are considerations in determining location. Port facilities, roadways, and waterways are another consideration. A qualified and available work force is always a consideration when determining location. In some locations qualified operating personnel are difficult to find, so inexperienced and poorly educated operators may be hired. In order to overcome their lack of qualifications, intensive training is required. Generally, it is good to include in the project training using high fidelity simulators particularly where inexperienced operators are to be hired. In addition, plants with novel processes with which even experienced and highly educated operators are not familiar should include additional training provisions. Such training will impact the project’s cost and schedule.

Regardless of schedule, the project team’s capability to construct the facility must be addressed. For instance, vessels of large diameter and height will require shop facilities that have appropriate size capacity, as well as trucking and rail facilities that accommodate the finished products. In some cases, the vessel may
require field fabrication or multiple vessels will be needed if shop fabricated. Alternatives for prime mover drivers such as electric motors, steam turbines, gas turbines, and gas engines may be influenced by the availability of infrastructure to support these devices. If electrical service is not provided by a utility, then generation or cogeneration facilities may be required. These must be addressed in the project cost estimates and schedules.

Special materials are often required in gas processing plant construction due to components such as hydrogen sulfide, carbon dioxide, mercury, water, etc. The availability of these materials and their delivery should be considered. Sometimes cladding or linings may be alternatives to expensive and scarce alloys. In addition, approved welding procedures may not be available or the work force may not have the expertise to perform certain procedures. Addressing such obstacles must be part of the project plans.

During the proposal solicitation and award of the construction contract, the cost estimates and project schedule for the prime contractor will be focused on construction activities and therefore fairly detailed and inclusive. However, the overall project schedule from an operating company point of view must consider nonconstruction activities such as permits, licenses, and other government requirements; staffing, accounting, other internal issues as well as contracts with suppliers and customers to name a few.

As project definition improves, the uncertainties associated with cost estimates should decrease to a funding level of accuracy of approximately $+15\%$/$-10\%$ with a 10% contingency identified. A probabilistic approach to cost estimating identifying percentiles (P90, P50, and P10) is also widely used to illustrate cost uncertainties (McIntire, 2001).

18.4.3.1 Hazards and operability analysis
A hazards and operability (HAZOP) analysis or equivalent is good practice even when not a statutory requirement. Such an analysis will most likely recommend the addition or deletion of valves, lines, instrumentation, and equipment needed for safe and reliable operation.

Figure 18-1 illustrates a stages and gates approach to oil and gas facilities project management that emphasizes the importance of the planning stages (feasibility, FEED) leading into EPC contracting, construction, and fabrication activity, on to HAZOP and ultimately plant commissioning. To move from one stage to another requires a gate to be passed where decisions and approvals have to be made associated with funding, technical design, and project priority. Such approvals are usually structured in the form of authorities for expenditure (AFEs) to be signed off by the project owners (and often other stakeholders—e.g., government authorities) as positive approval to proceed under stages of a project budget. Although the diagram for simplicity suggests a linear process proceeding from one stage to another, in practice there are often loops and feedbacks to the work of earlier stages that require adjustments to design and so on.
18.4.3.2 Scheduling and cost estimating software

Software exists to assist with both conceptual cost and time estimates as well as detailed estimates and complex project networks involving the optimization of project networks with critical path analysis. Most major EPC contractors have their own custom tools. Smaller contractors and operating companies may use products supplied by a vendor specializing in these tools. For larger projects it is increasingly common for Monte Carlo simulation analysis to be used in conjunction with critical path identification to yield probabilistic estimates of cost and time associated with each project activity and for the project as a whole (Wood, 2001).

18.4.4 Project execution planning

A project must be planned and tracked against the plan to assure successful execution. A project plan sets the ground rules and states them in a clear fashion. The project plan helps to control and measure progress and helps to deal with any changes that may occur. Previous experience is the best guide for determining the necessary tasks and the time to complete them. Many engineering and operating companies maintain databases that include previous project plans with actual time to completion and costs. To be able to benefit from such an approach requires good quality record keeping both during a project and following its completion.
Although, no two projects are identical no matter how similar they appear, these databases of past experience contain valuable information on which to plan. It is necessary to understand any unique requirements that previous projects met and how the current project compares. Some dissimilarities may include:

- location which impacts the government regulatory bodies, remoteness, cost of labor, etc.,
- makeup of the project team, including expertise, time available, geographic dispersion of team members, cultural differences, organizational affiliations/loyalties,
- project scope, and
- current economic conditions, which affects inflation and employee availability.

The project plan should be relevant, understandable, and complete, and reflect the size and complexity of the unique project. The project plan should include the following elements (Hauge and Cramer, 2002):

- A project charter
- A project timeline
- A responsibility matrix
- A project plan budget
- Major milestones with target dates
- A risk management strategy.

### 18.4.5 Pre-project planning measurement

The project objectives, or the measure for project success or failure, are often defined in terms of cost, schedule, and technical performance. In order to serve as a baseline for project execution, measurements should be in place to identify target completion dates, budgets, and expected technical performance. These measures should be included in a system that allows tracking of actual, target, and projected dates and costs with variances highlighted. The technical expectations should be tracked as well and checked for compliance as the project proceeds.

### 18.4.6 The responsibility matrix

Projects are a collaborative effort between a number of individuals and organizations working together toward a common goal. Managing a diverse team, often spread over several locations (and countries), can present some special challenges. A responsibility matrix is a valuable project management tool to help meet these challenges. The matrix ensures that someone accepts responsibility for each major project activity. It also encourages accountability. The responsibility matrix should correspond with the project timeline. An example is shown in Table 18-1.
The left-hand column lists all the required tasks for your project, while the team members (e.g., project manager, project engineer, safety engineer, plant manager, purchasing agent, etc.) are listed across the top. A code is entered in each cell that represents that team member’s involvement in the task in that row. For example, choose codes appropriate to the project; the key is to clearly identify who has a role in every activity, who is accountable and who must sign off. Make sure the matrix is included in the project plan so that every participant is clearly aware of his or her responsibilities.

When developing the construction-phase-site staffing plan, the contractor’s site organization should be analyzed for strengths and weakness that can affect the success of the project. In some cases, it is necessary to supplement the contractor’s organization with project sponsor’s own personnel or specialist independent contractors. The need for these adjustments generally is not known until a vulnerability assessment is performed on the contractor. Normally this is done during the evaluation phase in advance of the final project investment decision and award of the main EPC contracts. Upon completion of such an assessment, project sponsor staffing adjustments should be made to supplement and leverage the contractor’s weakness and strengths, respectively. This approach helps to reduce project execution risk and improve constructability performance (Wood et al., 2008a).

For project contracts such as reimbursable contracts where the project sponsor is bearing most of the risk exposure, the project sponsor’s site teams are commonly developed and aligned to match the structure and function of the contractor’s site team. For lump sum and unit price contracts the project sponsor and contractor’s site teams generally do not match.

A well-disciplined and managed team can make the difference to a project. Figure 18-2 highlights that the time it takes to build a highly performing cohesive project team can be crucial, as project teams do not have time to waste. A focus on early team building initiatives can pay off as it is difficult to assemble a group of skilled staff and expect them to perform as a team without progressing through

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**Table 18-1 Typical Responsibility Matrix**

<table>
<thead>
<tr>
<th>Task</th>
<th>Contractor</th>
<th>Owner/Operator</th>
<th>1</th>
<th>2</th>
<th>3</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 A</td>
<td>S</td>
<td>P</td>
<td>–</td>
<td>–</td>
<td></td>
</tr>
<tr>
<td>2 –</td>
<td>A</td>
<td>S</td>
<td>P</td>
<td>I</td>
<td></td>
</tr>
<tr>
<td>3 –</td>
<td>A</td>
<td>S</td>
<td>P</td>
<td></td>
<td></td>
</tr>
<tr>
<td>4 A</td>
<td>S</td>
<td>P</td>
<td>–</td>
<td>–</td>
<td></td>
</tr>
<tr>
<td>5 A</td>
<td>S</td>
<td>P</td>
<td>I</td>
<td>I</td>
<td></td>
</tr>
<tr>
<td>6 A</td>
<td>S</td>
<td>P</td>
<td>–</td>
<td>I</td>
<td></td>
</tr>
<tr>
<td>7 –</td>
<td>A</td>
<td>S</td>
<td>–</td>
<td>P</td>
<td></td>
</tr>
</tbody>
</table>

S, Sign-off; A, accountable; P, primary responsible; I, input.
team building and integration phases. This becomes even more important in multicultural teams (Safakish and Wood, 2011).

In order to maximize team performance and motivation, team building and development must be given time and priority as well as careful project management attention at an early phase in the execution of any significant project. This is especially the case for mega-scale facilities projects with activities conducted at multiple sites in several countries separated by large time differences and cultural barriers (Safakish and Wood, 2010).

Key team attributes to achieve targets by overcoming obstacles:

- well-trained,
- highly-motivated,
- integrated,
- collaborative,
- led by focused project plan,
- strong leadership,
- shared values through buy-in and alignment, and
- adequately resourced.

18.5 Project controls

The two main elements of a project plan are timeline and cost control. A timeline assures that the project is scheduled properly to meet anticipated and promised dates. For projects that involve several parallel or overlapping activities the
timeline becomes a network. Cost control assures that the project meets its budget.

A checklist approach has been proposed to provide a structured process for incorporating construction and other discipline knowledge that should enhance project safety, quality, schedule, and cost and risk management objectives (Wood et al., 2008a).

This checklist considers 10 key facets of project execution:

1. Schedule
2. Design factors: simplicity, standardization, and ease of construction
3. Facility layouts and arrangements
4. Safety, environment, security, community and regulation
5. Specifications
6. Plans and logistics
7. Cost estimates
8. Construction execution issues
9. Modularization
10. Risk mitigation and opportunity exploitation.

18.5.1 Project timeline

By dividing a project into the individual tasks required for completion, the project timeline (Greer, 2002):

- Provides a detailed view of the project’s scope
- Allows monitoring of what has been completed and what remains to be done
- Allows tracking of labor, time, and costs for each task
- Allows assigning of responsibility for specific tasks to team members
- Allows team members to understand how they fit into the “big picture”.

This timeline can take on a variety of formats and philosophies. The most prevalent philosophy is to determine the date for project completion and work backwards to identify key dates when certain milestones require completion. In practice, the project completion date is first determined through iterations of what is possible going forward with contingencies and identifying an end date. This usually involves establishing a critical path of activities which must be completed on time for a project not to fall behind its scheduled end date. If this end date is not acceptable, then acceleration plans should be explored. Most methods of acceleration require additional expense to accomplish objectives. These methods may include parallel tasking, overtime, and contractor bonuses for meeting an aggressive date to name a few.

Once the project milestones are set, then subtasks and assignments are identified. On large projects, certain tasks and subtasks will be assigned to an assistant project manager. The overall project manager will become responsible for coordinating the activities of the assistant project managers. The milestones, major tasks, and subtasks are commonly shown in a Gantt chart. This type of
presentation presents dependencies in a graphical display such as predecessors (tasks that must be completed prior to commencing the next task) and successors (tasks that cannot be started until a previous task is completed). Milestones are significant events in a project, usually the completion of a major deliverable or tied to vendor or contractor payments. A very good method to analyze subtasks is to consult with an expert on accomplishing the subject task. And the very best way is to receive an estimate with commitment from the person responsible for executing the task. This is where contractors and their bids contribute to the best task analysis possible.

Note, as the project progresses, there may be tasks that were not foreseeable in the original plan, or tasks that are added to enhance the overall project outcome. The impact of the additional tasks, delays, or even acceleration requires consideration of the impact on both the time schedule and resources. When changes to the schedule are warranted and feasible, the project manager should get a written agreement for the revised plan from all the key stakeholders in the project. A regular update of the time schedule is recommended with as part of a routine project status report. The time period for such reporting may be weekly or monthly depending upon the project or the stage of the project.

### 18.5.2 Risk management

Risk is inherent in all projects. In project management terms, “risk” refers to an uncertain event or condition that has a cause and, that if it occurs, has a positive or negative effect on a project’s objectives, and one or more consequences on project cost, schedule, or quality. As discussed in the previous sections, the measure for project success or failure is defined in terms of cost, schedule, and technical performance. Project risk management is intended to increase the likelihood of attaining these project objectives by providing a systematic approach for analyzing, controlling, and documenting identified threats and opportunities both during the planning and execution of a project. The application of project risk management will vary from the operator (owner) or the contractor side. The term risk management is used to lump together different activities. These activities may be divided into the following (Hauge and Cramer, 2002):

1. Activities related to the day-to-day identification, assessment, and control of uncertainties, i.e., risk management activities related to understanding and controlling the most important risks threatening the achievement of well-defined project objectives. This type of risk management may be based on a qualitative approach where each risk is assessed separately.
2. Activities related to the periodic assessment of achieving project objectives, i.e., assessing the probability of achieving well-defined project objectives with respect to schedule, budget, or performance. The periodic assessment must use a quantitative approach based on the aggregation of most critical uncertainties.
3. Activities related to the ranking of a set of alternative decision options/system solutions, i.e., ranking different alternatives with respect to their desirability measured in terms of the corresponding project objectives. Such ranking is typically performed at major decision gates during the conceptual design stage.

A good risk management system identifies, assesses, and controls uncertainties during all phases of the project. This allows the project team to mitigate risks and take advantage of opportunities. The selected system should properly assess all project risks, causes, and consequences including (Wood et al., 2008b):

- Technical and technological
- Financing and funding
- Joint venture and alliance
- Contractor and vendor performance
- Contract
- Health, safety, and security
- Community and public relations
- Environmental
- Statutory planning and regulation
- Capital and operations costs
- Construction, installation, and assembling schedules
- Completion and commissioning.

The above risks should be addressed holistically to include all stakeholders, involved parties, all phases and types of project exposure and uncertainties. The system should be developed to properly address all actions necessary to fulfill the requirements of any and all contractual obligations, and to incorporate the results from all risk assessments into the requisite project plans.

Risk management is a key to success for project execution, but is often constrained by inadequate work processes and tools. An overall understanding of the different risk factors and how these affect the defined project performance goals are critical for successful project management and decision-making. Project risk management is a systematic approach for analyzing and managing threats and opportunities associated with a specific project and will increase the likelihood of attaining project objectives. The usage of project risk management will also enhance the understanding of major risk drivers and how these affect the project objectives. Through this insight, the decision-makers can develop suitable risk strategies and action plans to manage and mitigate potential project threats and exploit potential project opportunities. Project risk management is based upon a number of different analysis techniques. The choice between these techniques is dependent on the quality of the information available and what kind of decisions project risk management should support. Day-to-day usage of project risk management is typically based on using risk matrices, accounting for both threats and opportunities. With sufficient uncertainty information, the
project risk management analysis can be extended to provide more direct decision support through probabilistic cost–benefit analyses.

Gas processing projects are often characterized by large investments, tight time schedules, and the introduction of technology or construction practices under unproven conditions. These challenges can result in a high-risk exposure but also opportunities that bring great rewards.

18.5.2.1 Project risk management methodology

It is important to perform risk management in a structured manner. Indeed transparent risk management frameworks are becoming statutory requirements for many companies obliging them to demonstrate how they are managing risks throughout their organizations (enterprise risk management—ERM) or on an enterprise-wide basis (e.g., the COSO framework in the US). It is important to ensure that the project risk management methodology is consistent with ERM frameworks (Wood and Randall, 2005). The project risk management process is often broken down into the following five general steps (Hauge and Cramer, 2002):

1. Initiation and focusing—initiate risk management process including identifying project objectives. The initiation should also assign personnel to the main risk management roles such as risk manager.
2. Uncertainty identification—identify risks affecting the project objectives and assign responsibility for assessing and mitigating each risk.
3. Risk analysis—assess the probability of occurring and the corresponding objective consequences for each risk, given that the risk occurs. Based on the risk assessment, classify each risk in terms of criticality.
4. Action planning—identify risk mitigating actions so that the most critical risks are mitigated. Assign responsibility and due dates for each action.
5. Monitoring and control—review and, if necessary, update risk assessments and corresponding action plans once new and relevant information becomes available.

The “Initiation and focusing” step is normally performed once at the start of the project whereas the four other steps are performed in an iterative manner. The initiation of project risk management in projects has the following set of goals (Hauge and Cramer, 2002):

- Identify, assess, and control risks that threaten the achievement of the defined project objectives, like schedule, cost targets, and performance of project delivery. These risk management activities should support the day-to-day management of the project as well as contribute to efficient decision-making at important decision points.
- Develop and implement a framework, processes, and procedures that ensure the initiation and execution of risk management activities throughout the project.
- Adapt the framework, processes, and procedures so that the interaction with other project processes flow in a seamless and logical manner.
The project risk management process should be assisted by a set of tools that supports these processes.

### 18.5.2.2 Risk response planning

A risk response plan can help maximize the probability and consequences of positive events and minimize the probability and consequences of events adverse to the project objectives. It identifies the risks that might affect the project, determines their effect on the project, and includes responses to each risk. The first step in creating a risk response plan is to identify risks, which might affect the project. The project team members should collaborate referring to the project charter, project timeline, and budget to identify potential risks. Those involved in the project can often identify risks on the basis of experience. Common sources of risk include:

- Technical risks—such as unproven technology
- Project management risks—such as a poor allocation of time or resources
- Organizational risks—such as resource conflicts with other activities
- External risks—such as changing priorities in partner or contractor organizations
- Construction risks—such as labor shortages or stoppages and weather.

### 18.5.2.3 Developing risk response strategies

There is no preparation for mitigating all possible risks, but risks with high probability and high impact are likely to merit immediate action. The effectiveness of planning determines whether risk increases or decreases for the project’s objectives. Several risk response strategies are available (HRDC, 2003):

- **Avoidance**—changing the project plan to eliminate the risk or protect the objectives from its impact. An example of avoidance is using a familiar technology instead of an innovative one.
- **Transference**—shifting the management and consequence of the risk to a third party. Risk transfer almost always involves payment of a premium to the party taking on the risk. An example of transference is using a fixed-price contract.
- **Mitigation**—reducing the probability and/or consequences of an adverse risk event to an acceptable threshold. Taking early action is more effective than trying to repair the consequences after it has occurred. An example of mitigation is seeking additional project partners to increase the financial resources of the project.
- **Acceptance**—deciding not to change the project plan to deal with a risk. Passive acceptance requires no action. Active acceptance may include developing contingency plans for action should the risk occur. An example of active acceptance is creating a list of alternative vendors those can supply materials with little notice.

Since not all risks will be evident at the outset of the project, periodic risk reviews should be scheduled at project team meetings. It is also important not to view risk
with a negative mind set. In addition to the downside consequences associated with many risks lie opportunities that should be identified and appropriate strategies developed to exploit them where possible. Risks that do occur should be documented, along with their response strategies in a risk register that assigns responsibilities for specific risks.

**18.5.2.4 Qualitative project risk management**

The routine day-to-day identification, assessment, and control of project risks are similar to hazard and operability identification techniques. The identification of risk consists of collecting and examining information on potential events that may influence the achievement of the project objectives. Each such event is categorized as a risk or an opportunity. The identification of these events should involve expertise from all main project competencies to reduce the possibility of important risks being overlooked. These risks will normally be prioritized so that only the most likely and consequential risks will be entered into a formal risk management process. The prioritization of risk should only be performed after thorough assessments and discussions among the project team. New information could mean that risks those have been previously determined as lower likelihood must be inserted into the risk management process.

The assessment of each risk or opportunity is made in terms of scores for probability of occurrence and for consequence, given that it occurs, for each project objective. Based on the probability and consequence scores, the criticality of each risk with respect to the achieving the project objectives can be assessed. Typically, a classification consisting of several possible probability scores and several possible consequence scores leads to several different classes of criticality, e.g., “critical,” “significant” and “negligible.” **Figure 18-3** is an example of how risk can be classified in relation to probability, project cost (economy), project effort or duration (time), and project performance.

The control of each risk event is normally based on its risk classification. A risk that is classified as “critical” will normally result in actions being identified in order to reduce the risk classification to either “significant” or “negligible.”

**FIGURE 18-3**

Risk matrices (Hauge and Cramer, 2002).
The risk reduction can be caused by either preventive measures (reducing the probability that the event will occur), or corrective measures (lessening the consequences of the event) or both.

18.5.2.5 Quantitative project risk management assessment
A periodic assessment of the probability of meeting project objectives must be based on quantitative calculation of the aggregated effect of the most important risks on the project objectives. The aggregation must also take into account sequences of scenarios and risk events, as well as the project structures given by the budget, schedule, and operability. There exist several methodologies with supporting tools that can be used for risk assessment of the total budgets and schedules. The challenge is to apply a methodology and find a tool that supports this methodology so that the integration of risks in the different domains can take place, their mutual independence can be represented, and their aggregate effect on the project objectives can be assessed. Fortunately, the usage of influence diagrams enables such a methodology and there exists several influence diagram tools. In an influence diagram each risk is represented as a symbol (or “node”) in a graphical diagram. The diagram represents the structural relationship between the different risks and their aggregate effect on the project objectives. In this manner, influence diagrams are well suited to represent risk scenarios. The mathematics used for assessing the aggregate effect is hidden away “behind” the diagram. In this way the influence diagram also represent a methodology to split a risk management model into two: (1) the structural relationship between the various risks and (2) the mathematics of the risk model such as probabilistic distribution functions for the risks. Other quantification methods to identify criticality and rank risks include Monte Carlo simulations and presentation through “Tornado diagrams.” Tornado diagrams (an example shown in Figure 18-4) help identify which input parameters, if they were to change, would have most consequential impacts on the analysis. They help to establish materiality of potential outcomes and to illustrate how the project will be impacted by changes, in order or significance, of those selected inputs. This provides valuable insight into which parameters might warrant further investigation to determine how changes would impact the objectives.

18.5.2.6 Risk process modeling
The model of the general risk management process is by no means complete. The two most important omissions are (Hauge and Cramer, 2002):

- No direct representation of the interaction with external organizations and their processes.
- No direct representation of the interaction with other internal processes.

The interaction with other processes should be designed so that the (new) risk management process is integrated as seamlessly as possible into the existing organizations and its already existing other processes (e.g., ERM framework).
Any required modifications should be as small as possible. Hence, this risk management process is integrated into the organization by having the already existing weekly management meeting also assessing the weekly risk report. Hence, no new forum for management review of risk is established, only an additional item is added to an existing agenda.

Since different parties in the risk management process have different interest, these interests should correspond to different views into the risk management process. Risk mitigation (and/or exploitation) strategies themselves involve costs to instigate and may lead to secondary risks. It is important to reconcile such costs and secondary risks with the risk management objectives, materiality of the risks being addressed, and the project budget and resources available.

18.5.2.7 Project risk management in interaction with other management processes

It is important to note that risk management has major similarities with other common management processes. Examples of such processes are (HRDC, 2003):

- Management of project changes
- Management of public permits
- Management of health, safety, and environmental issues
- Management of decision gates.

Similarities exist in the identification of items, assessment of their criticality, identification of corresponding mitigating actions, and follow-up of criticality.
assessment and corresponding mitigation plans. These similarities should be exploited when setting up the risk management processes and establishing the tools for the management of these processes.

The risk management process will remain constant over the different project phases throughout the life cycle of a construction project. The different risk management techniques for assessing day-to-day risks, calculating the ability to meet defined project objectives and ranking different decision alternatives will also remain the same. In the ranking of different investment opportunities for a gas processing project, a number of issues will in general need to be looked into:

- Revenue
- Costs (costs of different activities: capital expenditures—CAPEX, operational costs—OPEX)
- Schedule (of project tasks and completion of milestones)
- Taxes and depreciation
- Health, safety, and environment (meeting regulations and company requirements)
- Structural reliability (design that meets requirements)
- On-stream factor (design that meets availability requirements).

What will vary, however, over the life cycle is the quality of the available risk-related information, the kind of competence that is needed to compile and prioritize this information and the kind of decisions that are supported by the risk management activities. A believable risk management process must be conducted by personnel with domain knowledge of the project phase in which decisions are to be made. Since the required competence will vary with the project phase, it is unlikely that the same person can fill the risk manager role throughout the project. Whether one or more risk managers or a team are involved in managing risks, it is crucial that analysis, actions, and outcomes are documented in a risk register and widely communicated both within the project team and to project stakeholders.

18.5.2.8 Other risk mitigation concepts

Other methods of risk mitigation include (Hauge and Cramer, 2002):

- Cost overrun protection—Cost overrun insurance can be purchased. This insurance is most often used for infrastructure like bridges, islands, etc. that may be required to access or locate the facility.
- Regulatory risk—A wide-ranging political or regulatory risk form of insurance is available. This is designed to provide an indemnity in the event that any changes occur to the regulatory requirements or political stance perceived at conception of the project during construction and into commercial operations.
- Revenue stream stabilization—In addition to the historic insurance market places, capital markets are available that enable industry to transfer risks to
financial vehicles and through structures different from traditional insurance policies. This market convergence has produced a wide range of creatively devised financial products, which can be applied to uninsurable risks. Hedging the sale prices for specific volumes of products is now widely used, particularly in the initial years of plant production prior to project payback, to help reduce project finance risks.

### Blended risk solution
These are integrated risk solutions, which will result in a more comprehensive package of protection than the traditional set of policies providing limited but specific coverage addressing different areas of risk.

#### 18.6 Quality assurance

An important part of defining the result and performance of the project is the specification of its quality-related features, which the project must then aim to deliver. Quality assurance has been an issue at the forefront of organizational concerns for decades. The development of quality conscious construction practices has been identified as being of the utmost importance in gaining and retaining a competitive edge. In the context of a project that aims to deliver a complex result, the quality aspects of that result will need to be planned, designed, aspired, and monitored. Quality assurance is a term used to incorporate the quality policy, quality management, and quality control functions, which combine to assure that the end result will be consistently achieved to the required condition. Its aim is to attain and assure quality through the adoption of a cost-effective quality control system and through external inspections and audits. Quality planning is an integral part of the planning activity. It manifests itself in the descriptions and in the scheduling of quality-related activities. The results of the quality planning activities are reflected in the resource and technical plans, at each level of the project. Quality control is concerned with ensuring that the required qualities are built into all of the tasks throughout their development life cycles. Quality control utilizes measurable quality criteria and is exercised via quality reviews, project reviews, and by the testing of products. Quality assurance requires agreement on the level of quality controls to be adopted, both specifically relating to the project and to the overall organizational policy. It is important that all three interests represented by the project owner are taken into account when deciding the mechanisms to be adopted.

The task descriptions should describe the purpose, form, and components of a task. It should also list, or refer to, the quality criteria applicable to that task. Task descriptions should be created as part of the planning process, to shadow the identification of the tasks that are required by the project. Each task description may either apply to a specific item, or to all the tasks of a given type. The component tasks of a complex task may be described in separate descriptions, giving rise to a hierarchy of task descriptions for that task.
Quality criteria should be used to define the characteristics of a task in terms that are quantifiable, and therefore allow it to be measured at various points in its development life cycle, if required. The criteria effectively define quality in the context of the product and are used as a benchmark against which to measure the finished task. Quality criteria should be established by considering what the important characteristics of a result or task are in satisfying the need that it addresses, and they should always be stated objectively. Subjective or descriptive criteria such as “quick response” or “maintainable” are unsatisfactory—as they do not permit meaningful measurement.

Quality planning should ensure that all quality-related activities are planned and incorporated into the project schedule. The tasks required to ensure the quality of the delivered result are often overlooked, with the result that the project schedule fails to represent quality-related work. This can have serious consequences for either: the quality levels achieved, the overall budget, or both.

Quality control is concerned with ensuring that the required qualities are built into all of the tasks throughout the construction cycle. It defines the method of inspection, in-process inspection, and final inspection to determine if the result has met its quality specification. Quality control utilizes measurable quality criteria and is exercised via change control, quality reviews, project reviews, and by the testing of products.

Reviews should be scheduled prior to key decision dates and important milestones such as shipment of rotating equipment and major equipment. For instance, modifications to a gas turbine are more easily made in the shop prior to shipment rather than in the field after transit. Modifications required that could have been found with shop inspections can cause significant delays.

Many of the top engineering consulting companies operate integrated quality, health, safety and environmental management systems which they apply generically across their operations. This is appropriate as it makes clear to their staff that all four of its components are important to ensure successful outcomes and that quality, health, safety and environmental management systems all influence each other as well as the budgetary, schedule and risk issues that drive project decisions. The adage that one should expect what you inspect applies to gas processing project management. Management and technical peer reviews are good practice. Inspection reports monitor not only quality but progress and should be used liberally for best results.

18.7 Commissioning and start-up
Commissioning and start-up of a new facility or unit is a very important phase of any processing project. These activities can be very expensive from a standpoint of project costs as well as deferral of operating revenues, if delays are encountered. For this reason, it is good practice to assign a start-up engineer to plan and
coordinate these activities. Operators and maintenance personnel should be hired well before the start-up date and trained on the equipment and process basics. Familiarity with the particular control system can be accomplished through the use of operator training simulators. These simulators can pay off easily by reducing start-up time. More detailed training on particular aspects of the equipment and process can be given to selected individuals. Indeed many now argue that it is important to have operations and maintenance personnel involved in the design and engineering phase of a project as it is much easier to sort out operational control logistics at the planning stage than later on.

Thorough checkout of the equipment should be conducted by the owner and operator of the facility. Punch lists of all deficiencies should be prepared, reviewed with the constructor, and updated on a daily basis. Purging of the equipment is quite important for safe start-up. Process gas or an inert gas, such as nitrogen can be introduced to remove pockets of oxygen, which could lead to explosions. All high point vents should be opened and checked with a portable oxygen analyzer. These vents should only be closed when oxygen is no longer detected. Excess water and other liquids should also be removed from the system by opening low point drains until purge gas escapes through the drains.

Start-up should be a joint effort between the constructor, operator, and process designers. The constructor should be advised of any deficiencies in the equipment and instrumentation as they are identified by operations and have personnel readily available to resolve the issues. It is not uncommon for construction contracts to include a retention portion of the plant cost that the owner will withhold for a period (e.g., six months or one year) following start-up to motivate the contractor to deal promptly with any teething problems.

18.8 Operate and evaluate

The final phase of a processing plant project is continuous operation and evaluation of the project results including plant capability over a finite period of time. Before the plant is turned over to operation, the performance of the plant should be measured. Sometimes this is required by the contractor or process licensor to meet performance guarantees. The capability of each process and equipment should be calculated and become the baseline for operation. At this point you may discard the design capability as this was used only as a basis for sizing equipment. Many times equipment is sized with a contingency giving better than design capability. Other times, the equipment may not be capable of design. For example, a difference in the actual inlet gas composition from the design basis is a common culprit.

After detailed measurement of the plant’s capabilities and deficiencies, these items should be well documented. The capability information is valuable should the plant require expansion in the future. Deficiencies become the basis for possible debottleneck and retrofit projects.
Capturing and transferring highly detailed and specific knowledge of a new plant, its machinery, suppliers, and commercial contracts is important for ensuring reliability, safety, and sustained plant performance.

Training and retaining operations staff is very important and in remote facilities where local skills may be in short supply is not an easy task. An operator training simulator (OTS) can also be developed and maintained for training of new plant operators and analyzing plant performance under specific conditions relevant to planned operations. The OTS may be based on a dynamic simulation developed during the plant design. The long-term benefits of investing in an operator training simulator according to KBR (2012) are:

- Operator best practices are reinforced through refresher training.
- Workforce flexibility is increased through cross-trained operators.
- An opportunity to test trial modifications to the plant’s control and automation systems and process configurations is provided.
- Development and refinement of operating procedures is facilitated.
- An OTS ensures plant operators receive plant specific and realistic hands-on training ahead of plant start-up and throughout plant operation.

18.9 Project closeout

In addition to the evaluation activities mentioned in the previous section, the project activities should be reviewed. This review should comprise:

- What worked well?
- What didn’t work well?
- What were the actual costs?
- What was the actual schedule?
- What assumptions need revision?
- What risks materialised and required attention? (as documented in the project risk register)
- Did risk mitigation strategies employed achieve their objectives?
- What are the project economics as constructed and operated?

As a result of this review or project postmortem, the project manager should write a project closeout report that includes the results of the project closeout review and recommendations for future similar projects.

18.10 Conclusion

Planning is critical to project success. Detailed, systematic, team-involved plans are the foundation for such success. When events cause a change to the plan, project managers must make a new one to reflect the changes. So continuous planning is a requirement of project management. Project managers must focus
on three dimensions of project success. Project success means completing all project deliverables on time, within budget, and to a level of quality that is acceptable to sponsors and stakeholders. The project manager must keep the team’s attention focused on achieving these broad goals and the stakeholders aligned to the project objectives. It is essential that the project team be comprised of all key disciplines that create or use the deliverables. The responsibilities of all team members should be clearly defined. Project managers must feel, and transmit to their team members, a sense of urgency. Because projects are endeavors with limited time, money, and other resources available, they must be kept moving toward completion. Since most team members have lots of other priorities, it’s up to the project manager to keep their attention on project deliverables and deadlines. Regular status checks, meetings, and reminders are essential. All project deliverables and activities must be visualized and communicated in vivid detail. The project manager and project team must create a picture of the finished deliverables in the minds of everyone involved so that all effort is focused in the same direction. Avoid vague descriptions and make sure everyone understands what the final product will be.

Projects require clear approvals and sign-off by sponsors. Clear approval points, accompanied by formal sign-off by sponsors and key stakeholders, should be demarcation points in the evolution of project deliverables. Anyone who has the power to reject or to demand revision of deliverables after they are complete must be required to examine and approve them as they are being built.

Risk management is an essential responsibility of project management. All risks should be identified and a contingency plan should accompany all critical risks.

A gas processing project is not complete until the plant, unit, or equipment is placed in service. An evaluation of the operability should be a deliverable upon completion of the project.

References


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## Conversion Factors

### Length

<table>
<thead>
<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 m</td>
<td>= 39.37 inch = 10^6 μm = 10^10 Å</td>
</tr>
<tr>
<td>1 inch</td>
<td>= 2.54 cm</td>
</tr>
<tr>
<td>1 ft</td>
<td>= 30.48 cm = 0.3048 m</td>
</tr>
<tr>
<td>1 mile</td>
<td>= 5280 ft = 1760 yards = 1609.344 m</td>
</tr>
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</table>

### Energy

<table>
<thead>
<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 J</td>
<td>= 1 W s = 1 kg m^2/s = 1 N m^3</td>
</tr>
<tr>
<td></td>
<td>= 10^7 dyne cm = 10^7 erg</td>
</tr>
<tr>
<td>1 Btu</td>
<td>= 778 ft lbf = 252 cal</td>
</tr>
<tr>
<td></td>
<td>= 1055 J = 10.41 lit atm</td>
</tr>
<tr>
<td>1 HP h</td>
<td>= 2545 Btu</td>
</tr>
<tr>
<td></td>
<td>= 3412 Btu = 1.341 HP h</td>
</tr>
</tbody>
</table>

### Mass

<table>
<thead>
<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 lbm</td>
<td>= 453.6 g = 0.4536 kg = 7000 grain</td>
</tr>
<tr>
<td>1 kg</td>
<td>= 2.2046 lbm</td>
</tr>
<tr>
<td>1 slug</td>
<td>= 1 lbf s^2/ft = 32.174 lbm</td>
</tr>
<tr>
<td>1 US ton</td>
<td>= 2000 lbm (also called short ton)</td>
</tr>
<tr>
<td>1 long ton</td>
<td>= 2240 lbm (also called British ton)</td>
</tr>
<tr>
<td>1 ton</td>
<td>= 1000 kg (also called metric ton)</td>
</tr>
</tbody>
</table>

### Power

<table>
<thead>
<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 HP</td>
<td>= 550 ft lbf/s = 33000 ft lbf/min</td>
</tr>
<tr>
<td></td>
<td>= 746 W = 0.746 kW</td>
</tr>
</tbody>
</table>

### Gas Constant

<table>
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<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>R</td>
<td>= 1.9859 Btu/lbmole</td>
</tr>
<tr>
<td>°R</td>
<td>= 1.9859 cal/gmole °K</td>
</tr>
<tr>
<td></td>
<td>= 0.73024 atm ft^3/lbmole °R</td>
</tr>
<tr>
<td></td>
<td>= 1545.3 ft lbf/lbmole °R</td>
</tr>
<tr>
<td></td>
<td>= 10.732 psia ft^3/lbmole °R</td>
</tr>
<tr>
<td></td>
<td>= 0.082057 lit atm/gmole °K</td>
</tr>
<tr>
<td></td>
<td>= 82.057 atm cm^3/gmole °K</td>
</tr>
<tr>
<td></td>
<td>= 8314.5 Pa m^3/kgmole</td>
</tr>
<tr>
<td></td>
<td>= 8.3145 kJ/kgmole °K</td>
</tr>
<tr>
<td>°K</td>
<td>= 82.057 atm cm^3/gmole °K</td>
</tr>
<tr>
<td></td>
<td>= 8.3145 kJ/kgmole °K</td>
</tr>
</tbody>
</table>

### Force

<table>
<thead>
<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 lbf</td>
<td>= 4.448 N = 4.448 × 10^5 dyn</td>
</tr>
<tr>
<td></td>
<td>= 32.174 poundals = 32.174 lbm ft/s^2</td>
</tr>
<tr>
<td></td>
<td>= 1 lbw</td>
</tr>
</tbody>
</table>

### Pressure

<table>
<thead>
<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 atm</td>
<td>= 14.696 psia = 2116 lbf/ft^2</td>
</tr>
<tr>
<td></td>
<td>= 29.92 in Hg</td>
</tr>
<tr>
<td></td>
<td>= 760 mm Hg = 760 Torr</td>
</tr>
<tr>
<td></td>
<td>= 1.013 bar</td>
</tr>
<tr>
<td></td>
<td>= 33.9 ft H_2O = 1.013 × 10^5 Pa</td>
</tr>
<tr>
<td></td>
<td>= 101.3 kPa</td>
</tr>
<tr>
<td>1 Pa</td>
<td>= 1 N/m^2 = 10^{-5} bars</td>
</tr>
</tbody>
</table>

### Dynamic Viscosity

<table>
<thead>
<tr>
<th>Unit</th>
<th>Conversion</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 cp</td>
<td>= 0.01 Poise = 0.01 g/cm s</td>
</tr>
<tr>
<td></td>
<td>= 0.01 (dyne s)/cm^2</td>
</tr>
<tr>
<td></td>
<td>= 0.001 kg/m s = 0.001 Pa s</td>
</tr>
<tr>
<td></td>
<td>= 0.001 N s/m^2</td>
</tr>
<tr>
<td></td>
<td>= 2.42 lbf/ft h = 0.0752 slug/ft h</td>
</tr>
<tr>
<td></td>
<td>= 6.72 × 10^{-4} lbfm/ft s</td>
</tr>
<tr>
<td></td>
<td>= 2.09 × 10^{-5} lbf s/ft^2</td>
</tr>
<tr>
<td>1 Pa s</td>
<td>= 0.0209 lbf s/ft^2</td>
</tr>
<tr>
<td></td>
<td>= 0.672 lbm/ft s</td>
</tr>
</tbody>
</table>

---

*Continued*
---cont’d

<table>
<thead>
<tr>
<th>Volume</th>
</tr>
</thead>
<tbody>
<tr>
<td>$1 \text{ ft}^3 = 7.4805 \text{ US gal} = 6.23 \text{ Imperial gal}$</td>
</tr>
<tr>
<td>$= 28.317 \text{ lit}$</td>
</tr>
<tr>
<td>$1 \text{ m}^3 = 1000 \text{ lit} = 264.2 \text{ US gal} = 35.31 \text{ ft}^3$</td>
</tr>
<tr>
<td>$1 \text{ bbl} = 42 \text{ US gal (oil)} = 5.615 \text{ ft}^3$</td>
</tr>
<tr>
<td>$1 \text{ lit} = 1000 \text{ cc}$</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Kinematic Viscosity</th>
</tr>
</thead>
<tbody>
<tr>
<td>$1 \text{ St} = 1 \text{ cm}^2/\text{s} = 0.0001 \text{ m}^2/\text{s}$</td>
</tr>
<tr>
<td>$1 \text{ ft}^2/\text{s} = 929 \text{ St} = 0.0929 \text{ m}^2/\text{s}$</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Force-mass conversion factor</th>
</tr>
</thead>
<tbody>
<tr>
<td>$g_r = 1 \text{ kg m/s}^2$ $N = 1 \text{ g cm/s}^2$ dyne</td>
</tr>
<tr>
<td>$= 32.174 \text{ lbf ft/s}^2$ $\text{lbf} = 1 \text{ slug ft/s}^2$ $\text{lbf}$</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Acceleration due to gravity</th>
</tr>
</thead>
<tbody>
<tr>
<td>$g = 32.2 \text{ ft/s}^2 = 9.81 \text{ m/s}^2 = 981 \text{ cm/s}^2$</td>
</tr>
<tr>
<td>(varies very slightly with longitude and elevation)</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Density</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water $= 62.43 \text{ lbf/ft}^3 = 1000 \text{ kg/m}^3$</td>
</tr>
<tr>
<td>$= 1 \text{ g/cm}^3$</td>
</tr>
<tr>
<td>$= 8.346 \text{ lbm/US gal}$</td>
</tr>
<tr>
<td>Mercury $= 13.6 \text{ g/cm}^3$</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Temperature</th>
</tr>
</thead>
<tbody>
<tr>
<td>$^\circ \text{F} = 1.8 (^\circ \text{C}) + 32$</td>
</tr>
<tr>
<td>$^\circ \text{R} = ^\circ \text{F} + 459.67 = 1.8 (^\circ \text{K})$</td>
</tr>
</tbody>
</table>
Standard Gas Conditions

Gas volumes given are for ideal gas only.

1. Normal: continental and scientific applications
   0 °C, 1.01325 bar, gas volume = 22.4136 m³/kg-mol
   32 °F, 14.696 psia, gas volume = 359.031 ft³/lb-mol
   (Nm³ based upon these conditions)

2. Standard (scf): US engineering applications
   15.5556 °C, 1.01325 bar, gas volume = 23.6900 m³/kg-mol
   60 °F, 14.696 psia, gas volume = 379.49 ft³/lb-mol
   (Scf = standard cubic foot based upon these conditions)

   volume = 23.6444 m³/kg-mol (Sm³ based upon these conditions)
### Table 3-1 Properties of Hydrocarbons and Common Gases (GPSA, 2004)

<table>
<thead>
<tr>
<th>Component</th>
<th>Formula</th>
<th>Mol Wt</th>
<th>Boiling Point °F (1 atm)</th>
<th>Vapor Press. @ 100 °F, Psia</th>
<th>Critical Press. Psia</th>
<th>Critical Temp. °F</th>
<th>Liquid Spec. Grav. 60/60 °F</th>
<th>Volume Ratio, Scf gas per gal liquid</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane</td>
<td>CH₄</td>
<td>16.043</td>
<td>–258.73</td>
<td>(5000)²</td>
<td>666.4</td>
<td>–116.67</td>
<td>0.3³</td>
<td>59.135³</td>
</tr>
<tr>
<td>Ethane</td>
<td>C₂H₆</td>
<td>30.070</td>
<td>–127.49</td>
<td>(800)²</td>
<td>706.5</td>
<td>89.92</td>
<td>0.35619³</td>
<td>37.476³</td>
</tr>
<tr>
<td>Propane</td>
<td>C₃H₈</td>
<td>44.097</td>
<td>–43.75</td>
<td>188.64</td>
<td>616.0</td>
<td>206.06</td>
<td>0.50699⁵</td>
<td>36.375⁵</td>
</tr>
<tr>
<td>i-Butane</td>
<td>C₄H₁₀</td>
<td>58.123</td>
<td>10.78</td>
<td>72.581</td>
<td>527.9</td>
<td>274.46</td>
<td>0.56287⁵</td>
<td>30.639⁵</td>
</tr>
<tr>
<td>n-Butane</td>
<td>C₄H₁₀</td>
<td>58.123</td>
<td>31.08</td>
<td>51.706</td>
<td>550.6</td>
<td>305.62</td>
<td>0.58401¹</td>
<td>31.790⁵</td>
</tr>
<tr>
<td>i-Pentane</td>
<td>C₅H₁₂</td>
<td>72.150</td>
<td>82.12</td>
<td>20.445</td>
<td>490.4</td>
<td>369.10</td>
<td>0.62470</td>
<td>27.393</td>
</tr>
<tr>
<td>n-Pentane</td>
<td>C₅H₁₂</td>
<td>72.150</td>
<td>96.92</td>
<td>15.574</td>
<td>488.6</td>
<td>385.8</td>
<td>0.63112</td>
<td>27.674</td>
</tr>
<tr>
<td>n-Hexane</td>
<td>C₆H₁₄</td>
<td>86.177</td>
<td>155.72</td>
<td>4.960</td>
<td>436.9</td>
<td>453.6</td>
<td>0.66383</td>
<td>24.371</td>
</tr>
<tr>
<td>n-Heptane</td>
<td>C₇H₁₆</td>
<td>100.204</td>
<td>209.16</td>
<td>1.620</td>
<td>396.8</td>
<td>512.7</td>
<td>0.68820</td>
<td>21.729</td>
</tr>
<tr>
<td>n-Octane</td>
<td>C₈H₁₈</td>
<td>114.231</td>
<td>258.21</td>
<td>0.537</td>
<td>360.7</td>
<td>564.2</td>
<td>0.70696</td>
<td>19.580</td>
</tr>
<tr>
<td>n-Decane</td>
<td>C₁₀H₂₂</td>
<td>142.285</td>
<td>345.48</td>
<td>0.061</td>
<td>305.2</td>
<td>652.0</td>
<td>0.73421</td>
<td>16.326</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>N₂</td>
<td>28.013</td>
<td>–320.45</td>
<td>–</td>
<td>493.1</td>
<td>–232.51</td>
<td>0.80940¹</td>
<td>91.413¹</td>
</tr>
<tr>
<td>Oxygen</td>
<td>O₂</td>
<td>31.999</td>
<td>–297.33</td>
<td>–</td>
<td>731.4</td>
<td>–181.43</td>
<td>1.1421¹</td>
<td>112.93¹</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>CO₂</td>
<td>44.010</td>
<td>–109.26⁵</td>
<td>–</td>
<td>1071</td>
<td>87.91</td>
<td>0.81802²</td>
<td>58.807²</td>
</tr>
<tr>
<td>Hydrogen sulfide</td>
<td>H₂S</td>
<td>34.08</td>
<td>–76.50</td>
<td>394.59</td>
<td>1300</td>
<td>212.45</td>
<td>0.80144³</td>
<td>74.401³</td>
</tr>
<tr>
<td>Water</td>
<td>H₂O</td>
<td>18.0115</td>
<td>212.00</td>
<td>0.950</td>
<td>3198</td>
<td>705.16</td>
<td>1.00000³</td>
<td>175.62³</td>
</tr>
<tr>
<td>Air Mixture</td>
<td></td>
<td>28.9625</td>
<td>–317.8</td>
<td>–</td>
<td>546.9</td>
<td>–221.31</td>
<td>0.87976³</td>
<td>95.557³</td>
</tr>
</tbody>
</table>

### Continued
### Table 3-1  Properties of Hydrocarbons and Common Gases (GPSA, 2004)—Cont’d

<table>
<thead>
<tr>
<th>Component</th>
<th>Acentric Factor, ( u )</th>
<th>Heating Value @ 60 °F 1 atm, Btu/scf&lt;sup&gt;d&lt;/sup&gt;</th>
<th>Freezing Point @ 1 atm, °F</th>
<th>Cont’d</th>
</tr>
</thead>
<tbody>
<tr>
<td>i-Butane</td>
<td>0.1852</td>
<td>Lower 1.8 Upper 8.5 Net 3000.4 Gross 3251.9</td>
<td>-255.82</td>
<td>157.23</td>
</tr>
<tr>
<td>n-Butane</td>
<td>0.1995</td>
<td>Lower 1.5 Upper 9.0 Net 3010.8 Gross 3262.3</td>
<td>-217.05</td>
<td>165.93</td>
</tr>
<tr>
<td>i-Pentane</td>
<td>0.2280</td>
<td>Lower 1.3 Upper 8.0 Net 3699.0 Gross 4000.9</td>
<td>-255.82</td>
<td>147.12</td>
</tr>
<tr>
<td>n-Pentane</td>
<td>0.2514</td>
<td>Lower 1.4 Upper 8.3 Net 3706.9 Gross 4008.9</td>
<td>-217.05</td>
<td>153.57</td>
</tr>
<tr>
<td>n-Hexane</td>
<td>0.2994</td>
<td>Lower 1.1 Upper 7.7 Net 4403.8 Gross 4755.9</td>
<td>-139.58</td>
<td>143.94</td>
</tr>
<tr>
<td>n-Heptane</td>
<td>0.3494</td>
<td>Lower 1.0 Upper 7.0 Net 5100.8 Gross 5502.5</td>
<td>-131.05</td>
<td>136.00</td>
</tr>
<tr>
<td>n-Octane</td>
<td>0.3977</td>
<td>Lower 0.8 Upper 6.5 Net 5761.1 Gross 6248.9</td>
<td>-70.18</td>
<td>129.52</td>
</tr>
<tr>
<td>n-Decane</td>
<td>0.4898</td>
<td>Lower 0.7 Upper 5.4 Net 7189.6 Gross 7742.9</td>
<td>-21.36</td>
<td>119.65</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.0372</td>
<td></td>
<td></td>
<td>-346.00&lt;sup&gt;a&lt;/sup&gt;</td>
</tr>
<tr>
<td>Oxygen</td>
<td>0.0216</td>
<td></td>
<td></td>
<td>-361.82&lt;sup&gt;a&lt;/sup&gt;</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>0.2667</td>
<td></td>
<td></td>
<td>-69.83</td>
</tr>
<tr>
<td>Hydrogen sulfide</td>
<td>0.0948</td>
<td>Lower 4.3 Upper 45.5 Net 586.8 Gross 637.1</td>
<td>-121.88&lt;sup&gt;a&lt;/sup&gt;</td>
<td>235.63</td>
</tr>
<tr>
<td>Water</td>
<td>0.3442</td>
<td></td>
<td></td>
<td>32.00</td>
</tr>
<tr>
<td>Air</td>
<td></td>
<td></td>
<td></td>
<td>88.20</td>
</tr>
</tbody>
</table>

<sup>a</sup>Above critical point, extrapolated, or estimated.

<sup>b</sup>Sublimation point.

<sup>c</sup>At saturation pressure, 60 °F.

<sup>d</sup>At normal boiling point.

<sup>e</sup>Gas at 60 °F, liquid at normal boiling point.

<sup>f</sup>At the triple point pressure.

### Table 3-2  Approximate Ratio of Specific Heats (“k”-values) for Various Gases<sup>b</sup>

<table>
<thead>
<tr>
<th>Gas</th>
<th>Symbol</th>
<th>Mol. Wt</th>
<th>k. @ 14.7 Psia</th>
<th>Density @ 14.7 psi &amp; 60 °F Lbs./cu. ft</th>
</tr>
</thead>
<tbody>
<tr>
<td>Monatomic</td>
<td>He, Kr, Ne, Hg</td>
<td>–</td>
<td>1.67</td>
<td>–</td>
</tr>
<tr>
<td>Most diatomic</td>
<td>O&lt;sub&gt;2&lt;/sub&gt;, N&lt;sub&gt;2&lt;/sub&gt;, H&lt;sub&gt;2&lt;/sub&gt;, etc.</td>
<td>–</td>
<td>1.4</td>
<td>–</td>
</tr>
<tr>
<td>Acetylene</td>
<td>C&lt;sub&gt;2&lt;/sub&gt;H&lt;sub&gt;2&lt;/sub&gt;</td>
<td>26.03</td>
<td>1.3</td>
<td>1.22</td>
</tr>
<tr>
<td>Air</td>
<td>–</td>
<td></td>
<td>28.97</td>
<td>1.406</td>
</tr>
<tr>
<td>Ammonia</td>
<td>NH&lt;sub&gt;3&lt;/sub&gt;</td>
<td>17.03</td>
<td>1.317</td>
<td>1.29</td>
</tr>
<tr>
<td>Argon</td>
<td>A</td>
<td>–</td>
<td>1.667</td>
<td>–</td>
</tr>
<tr>
<td>Benzene</td>
<td>C&lt;sub&gt;6&lt;/sub&gt;H&lt;sub&gt;6&lt;/sub&gt;</td>
<td>78.0</td>
<td>1.08</td>
<td>1.09</td>
</tr>
<tr>
<td>Butane</td>
<td>C&lt;sub&gt;4&lt;/sub&gt;H&lt;sub&gt;10&lt;/sub&gt;</td>
<td>58.1</td>
<td>1.11</td>
<td>1.08</td>
</tr>
<tr>
<td>Isobutane</td>
<td>C&lt;sub&gt;4&lt;/sub&gt;H&lt;sub&gt;10&lt;/sub&gt;</td>
<td>58.1</td>
<td>1.11</td>
<td>1.08</td>
</tr>
<tr>
<td>Butylene</td>
<td>C&lt;sub&gt;4&lt;/sub&gt;H&lt;sub&gt;8&lt;/sub&gt;</td>
<td>56.1</td>
<td>1.1</td>
<td>1.09</td>
</tr>
<tr>
<td>Iso-Butene</td>
<td>C&lt;sub&gt;4&lt;/sub&gt;H&lt;sub&gt;8&lt;/sub&gt;</td>
<td>56.1</td>
<td>1.1</td>
<td>1.09</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>CO&lt;sub&gt;2&lt;/sub&gt;</td>
<td>44.0</td>
<td>1.3</td>
<td>1.27</td>
</tr>
<tr>
<td>Carbon monoxide</td>
<td>CO</td>
<td>28.0</td>
<td>1.4</td>
<td>1.4</td>
</tr>
<tr>
<td>Carbon tetrachloride</td>
<td>CCl&lt;sub&gt;4&lt;/sub&gt;</td>
<td>153.8</td>
<td>1.18</td>
<td>–</td>
</tr>
<tr>
<td>Chlorine</td>
<td>Cl&lt;sub&gt;2&lt;/sub&gt;</td>
<td>70.9</td>
<td>1.33</td>
<td>–</td>
</tr>
</tbody>
</table>
### Table 3-2 Approximate Ratio of Specific Heats (“k”-values) for Various Gases—Cont’d

<table>
<thead>
<tr>
<th>Gas</th>
<th>Symbol</th>
<th>Mol. Wt</th>
<th>k. @ 14.7 Psia</th>
<th>Density @ 14.7 psi &amp; 60 °F Lbs./cu. ft</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>60 °F</td>
<td>150 °F</td>
</tr>
<tr>
<td>Dichlorodifluoromethane</td>
<td>CCl2F2</td>
<td>120.9</td>
<td>1.13</td>
<td>–</td>
</tr>
<tr>
<td>Dichloromethane</td>
<td>CH2Cl2</td>
<td>84.9</td>
<td>1.13</td>
<td>–</td>
</tr>
<tr>
<td>Ethane</td>
<td>C2H6</td>
<td>30.0</td>
<td>1.22</td>
<td>1.17</td>
</tr>
<tr>
<td>Ethylene</td>
<td>C2H4</td>
<td>28.1</td>
<td>1.25</td>
<td>1.21</td>
</tr>
<tr>
<td>Ethyl chloride</td>
<td>C2H5Cl</td>
<td>64.5</td>
<td>1.13</td>
<td>–</td>
</tr>
<tr>
<td>Flue gas</td>
<td>–</td>
<td>–</td>
<td>1.4</td>
<td>–</td>
</tr>
<tr>
<td>Helium</td>
<td>He</td>
<td>4.0</td>
<td>1.667</td>
<td>–</td>
</tr>
<tr>
<td>Hexane</td>
<td>C6H14</td>
<td>88.1</td>
<td>1.08</td>
<td>1.05</td>
</tr>
<tr>
<td>Heptane</td>
<td>C7H16</td>
<td>100.2</td>
<td>–</td>
<td>1.04</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>H2</td>
<td>2.01</td>
<td>1.41</td>
<td>1.40</td>
</tr>
<tr>
<td>Hydrogen chloride</td>
<td>HCl</td>
<td>36.5</td>
<td>1.48</td>
<td>–</td>
</tr>
<tr>
<td>Hydrogen sulfide</td>
<td>H2S</td>
<td>34.1</td>
<td>1.30</td>
<td>1.31</td>
</tr>
<tr>
<td>Methane</td>
<td>CH4</td>
<td>16.03</td>
<td>1.316</td>
<td>1.28</td>
</tr>
<tr>
<td>Methyl chloride</td>
<td>CH3Cl</td>
<td>50.5</td>
<td>–</td>
<td>–</td>
</tr>
<tr>
<td>Natural gas (approx)</td>
<td>–</td>
<td>19.5</td>
<td>1.27</td>
<td>–</td>
</tr>
<tr>
<td>Nitric oxide</td>
<td>NO</td>
<td>30.0</td>
<td>1.40</td>
<td>–</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>N2</td>
<td>28.0</td>
<td>1.41</td>
<td>1.40</td>
</tr>
<tr>
<td>Nitrous oxide</td>
<td>N2O</td>
<td>44.0</td>
<td>1.311</td>
<td>–</td>
</tr>
<tr>
<td>Oxygen</td>
<td>O2</td>
<td>32.0</td>
<td>1.4</td>
<td>1.39</td>
</tr>
<tr>
<td>Pentane</td>
<td>C5H12</td>
<td>72.1</td>
<td>1.06</td>
<td>1.06</td>
</tr>
<tr>
<td>Propane</td>
<td>C3H8</td>
<td>44.1</td>
<td>1.15</td>
<td>1.11</td>
</tr>
<tr>
<td>Propylene</td>
<td>C3H6</td>
<td>42.0</td>
<td>1.16</td>
<td>–</td>
</tr>
<tr>
<td>Sulfur dioxide</td>
<td>SO2</td>
<td>64.1</td>
<td>1.256</td>
<td>–</td>
</tr>
<tr>
<td>Water Vapor (steam)</td>
<td>H2O</td>
<td>18.0</td>
<td>1.33^a</td>
<td>1.32</td>
</tr>
</tbody>
</table>

^aAt 212 °F.

*Compiled from: “Plain Talks on Air and Gas Compression, Fourth of Series,” Worthington Corp. and “Reciprocating Compressor Calculation Data Cooper–Bessemer Corp, (1956),” by permission.

### Table 3-3 Physical Properties of Selected Amines

<table>
<thead>
<tr>
<th>Compound</th>
<th>Monoethanol-amine</th>
<th>Diethanol-amine</th>
<th>Methyl diethanol-amine</th>
<th>Diglycol-amine</th>
<th>Diisopropanol-amine</th>
</tr>
</thead>
<tbody>
<tr>
<td>Formula</td>
<td>HOC₃H₇NH₂</td>
<td>(HOC₂H₅)₂NH</td>
<td>(HOC₂H₅)₂NHCH₃</td>
<td>HOC₃H₇NH₂</td>
<td>(HOC₃H₇)₂NH</td>
</tr>
<tr>
<td>Molecular wt</td>
<td>61.08</td>
<td>105.14</td>
<td>119.16</td>
<td>105.14</td>
<td>133.19</td>
</tr>
<tr>
<td>Boiling point @ 760 mmHg, °C</td>
<td>170.5</td>
<td>269</td>
<td>247</td>
<td>221</td>
<td>248.7</td>
</tr>
<tr>
<td>Freezing point, °C</td>
<td>10.5</td>
<td>28.0</td>
<td>–23</td>
<td>–12.5</td>
<td>42</td>
</tr>
</tbody>
</table>

Continued
<table>
<thead>
<tr>
<th>Compound</th>
<th>Monoethanol-amine</th>
<th>Diethanol-amine</th>
<th>Methyl-diethanol-amine</th>
<th>Diglycol-amine</th>
<th>Diisopropanol-amine</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure, kPa (abs)</td>
<td>5985</td>
<td>3273</td>
<td>3772</td>
<td>3770</td>
<td></td>
</tr>
<tr>
<td>Temperature, °C</td>
<td>350</td>
<td>442.1</td>
<td>402.6</td>
<td>399.2</td>
<td></td>
</tr>
<tr>
<td>Density @ 20 °C, gr/cc.</td>
<td>1.018</td>
<td>1.095</td>
<td>1.058 @ 15.6 °C</td>
<td>0.999 @ 30 °C</td>
<td></td>
</tr>
<tr>
<td>Weight, kg/m³</td>
<td>1016 @ 15.6 °C</td>
<td>1089 @ 15.6 °C</td>
<td>1040</td>
<td>1057 @ 15.6 °C</td>
<td></td>
</tr>
<tr>
<td>Relative density</td>
<td>1.0179</td>
<td>1.0919 (30 °C/20 °C)</td>
<td>1.0418</td>
<td>1.0572</td>
<td>0.989 @ 45 °C/20 °C</td>
</tr>
<tr>
<td>Specific heat @ 15.6 °C, kJ/kg °C</td>
<td>2.55 @ 20 °C</td>
<td>2.51</td>
<td>2.24</td>
<td>2.39</td>
<td>2.89 @ 30 °C</td>
</tr>
<tr>
<td>Thermal conductivity</td>
<td>0.256</td>
<td>0.220</td>
<td>0.275</td>
<td>0.209</td>
<td>–</td>
</tr>
<tr>
<td>Latent heat of vaporization, kJ/kg</td>
<td>826 @ 760 mmHg</td>
<td>670 @ 73 mmHg</td>
<td>476</td>
<td>510 @ 760 mmHg</td>
<td>430 @ 760 mmHg</td>
</tr>
<tr>
<td>Heat of reaction, kJ/kg of acid gas</td>
<td>H₂S</td>
<td>1568</td>
<td>–</td>
<td>–</td>
<td>–</td>
</tr>
<tr>
<td>Viscosity, mPa.sec</td>
<td>24.1 @ 20 °C</td>
<td>350 @ 20 °C</td>
<td>1.3 x 10⁻⁶ m²/s @ 10 °C</td>
<td>40 @ 16 °C</td>
<td>870 @ 30 °C</td>
</tr>
<tr>
<td></td>
<td></td>
<td>(at 90% wt solution)</td>
<td></td>
<td>198 @ 45 °C</td>
<td>86 @ 54 °C</td>
</tr>
<tr>
<td>Refractive index, N_d 20 °C</td>
<td>1.4539</td>
<td>1.4776</td>
<td>1.469</td>
<td>1.4598</td>
<td>1.4542 @ 45 °C</td>
</tr>
<tr>
<td>Flash point, COC, °C</td>
<td>93</td>
<td>138</td>
<td>129.4</td>
<td>127</td>
<td>124</td>
</tr>
</tbody>
</table>

### Table 3-4  Physical Properties of Selected Glycols and Methanol (GPSA, 2004)

<table>
<thead>
<tr>
<th>Compound</th>
<th>Ethylene Glycol</th>
<th>Diethylene Glycol</th>
<th>Triethylene Glycol</th>
<th>Tetraethylene Glycol</th>
<th>Methanol</th>
</tr>
</thead>
<tbody>
<tr>
<td>Formula</td>
<td>$C_2H_6O_2$</td>
<td>$C_4H_{10}O_3$</td>
<td>$C_6H_{14}O_4$</td>
<td>$C_6H_{18}O_5$</td>
<td>CH$_3$OH</td>
</tr>
<tr>
<td>Molecular mass</td>
<td>62.1</td>
<td>106.1</td>
<td>150.2</td>
<td>194.2</td>
<td>32.04</td>
</tr>
<tr>
<td>Boiling point@ 760 mmHg, °C</td>
<td>197.3</td>
<td>244.8</td>
<td>285.5</td>
<td>314</td>
<td>64.5</td>
</tr>
<tr>
<td>Vapor pressure @ 77 °F (25 °C), mmHg</td>
<td>0.12</td>
<td>&lt;0.01</td>
<td>&lt;0.01</td>
<td>&lt;0.01</td>
<td>120</td>
</tr>
<tr>
<td>Density @ 77 °F (25 °C), kg/m$^3$</td>
<td>1110</td>
<td>1113</td>
<td>1119</td>
<td>1120</td>
<td>790</td>
</tr>
<tr>
<td>Freezing point, °C</td>
<td>−13</td>
<td>−8</td>
<td>−7</td>
<td>−5.5</td>
<td>−97.8</td>
</tr>
<tr>
<td>Pour point, °C</td>
<td>−</td>
<td>−54</td>
<td>−58</td>
<td>−41</td>
<td></td>
</tr>
<tr>
<td>Viscosity in Centipoise @ 77 °F (25 °C)</td>
<td>16.5</td>
<td>28.2</td>
<td>37.3</td>
<td>44.6</td>
<td>0.52</td>
</tr>
<tr>
<td>@ 140 °F (60 °C)</td>
<td>4.68</td>
<td>6.99</td>
<td>8.77</td>
<td>10.2</td>
<td></td>
</tr>
<tr>
<td>Specific heat @ 77 °F (25 °C), kJ/(kg K)</td>
<td>2.43</td>
<td>2.30</td>
<td>2.22</td>
<td>2.18</td>
<td>2.52</td>
</tr>
<tr>
<td>Flash point, °C Pensky Martens Closed Cup (PMCC) tester</td>
<td>116</td>
<td>124</td>
<td>177</td>
<td>204</td>
<td>12</td>
</tr>
</tbody>
</table>

**Note:** These properties are laboratory results on pure compounds or typical of the products, but should not be confused with, or regarded as, specifications.

$^a$Glycols decompose at temperatures below their atmospheric boiling point. Approximate decomposition temperatures are:

- Ethylene glycol 165 °C
- Diethylene glycol 164 °C
- Triethylene glycol 207 °C
- Tetraethylene glycol 238 °C
Index

Note: Page numbers followed by “f” and “t” indicate figures and tables respectively

A
3PM. See Plant Production Performance Model
AAs. See Anti-agglomerants
Absolute open flow, 20
Absorber, 235–236, 492
side cooler, 192
Absorption, 182, 419–420
Acid gas
  enrichment, 308
  feed drums, 325–326
  fuel burner, 308
  treating system, 426
Acid gas enrichment unit (AGEU), 308
Acid gas removal unit (AGRU), 127, 171, 318
Acoustical treatment, 400–401
Activated alumina, 242–243
Adsorbent, 336, 420
  selection, 238–243
Adsorption, 182
  capacity, 238
  isotherm, 237–238
  principle, 243–244
Advanced process control system (APC system), 132–133, 468, 471–473, 488–489, 488f
Advanced regulatory control, 427
Aerosols, 150–151, 151f
AFE. See Authority for expenditure
AGEU. See Acid gas enrichment unit
AGRU. See Acid gas removal unit
Alarm management, 449–450
  alarm philosophy, 450–451
  audit, 453
  detailed design, 451–452
  identification, 451
  implementation, 452
  maintenance, 452
  management of change, 452
  monitoring and assessment, 452
  operation stage, 452
  rationalization, 451
Alkanolamine solvents, 183–195
  amine processes, 186–191
  amine unit operating problems, 194–195
  special design considerations, 191–194
American National Standards Institute (ANSI), 448
ANSI RP 755 Standard, 448
ANSI/API RP 755 Standard, 449
Amine
  processes, 186–191, 296–298
  double absorption process, 190–191
  two-stage absorption process, 189–190
  unit operating problems, 194–195
Annular flow, 44–45
Annular mist. See Annular flow
Annular velocity (V_{ann}), 155–156. See also
  Media velocity (V_{med})
ANSI. See American National Standards Institute
Anti-agglomerants (AAs), 80–82
APC system. See Advanced process control system
Application Service Provider (ASP), 521
Aquifers, 34
Aquisulf process, 322
ASP. See Application Service Provider
Asset
  management, 425–426
  values, 518
ATC. See Available Technical Capacity
Authority for expenditure (AFE), 547, 550
Automation, 413
  applications, 424–433
Available Technical Capacity (ATC), 511

B
BAHX. See Brazed aluminum heat exchanger
Baillie and Wichert method, 74–75, 77f
Baselining, 432
Bed refluxing, 256
Beggs and Brill method, 52–56
Benfield process, 195
Bernoulli’s equation, 384
Bio-SR process, 314
Brake horsepower (BHP), 366
Brazed aluminum heat exchanger (BAHX), 270–271, 278–279, 338
Breakdown maintenance, 455
British thermal unit (Btu), 6
Bubble flow, 45
Bulk modulus of elasticity. See Isothermal gas compressibility
Business  
   owner representative, 547
   and project objectives, 546–548
Butanes splitter, 499

C
Calcium chloride, 258
Capital expenditure (CAPEX), 397, 563
Carbon aromaticity, 528
Carbon dioxide (CO$_2$), 11–12, 181
   removal for NGL, 129
Carbon disulfide (CS$_2$), 181
Carbonyl sulfide (COS), 181, 242, 297–298
Cascade  
   flash separation, 170–171
   refrigeration, 268–270
Casinghead gas, 3–4
Catacarb process, 195
Catalytic oxidation, 318
   sub-dew-point processes, 318
   SuperClaus process, 318
Cathodic protection, 89
Caustic processes, 294–295
Caustics, 258
   processes, 294–295
Centrifugal compressors, 268, 351–353, 352f, 363, 369–376, 378, 422–423. See also Reciprocating compressors  
   performance map, 370f  
   recycle system layout for, 372f
Centrifugal pumps, 423
Centrifugal separators, 145–146
CFD. See Computational fluid dynamics
CFR. See Code of Federal Regulations
CGR. See Condensate/gas ratio
Checklist approach, 555
Chemical absorption processes, 183–195. See also Physical solvent processes  
   alkanolamine solvents, 183–195
   potassium carbonate solution, 195
Chemical inhibition, 80–86  
   inhibitor types, 80–83  
   injection system design, 85–86  
   prediction of inhibitor requirements, 83–85
Chemisorption, 237
Chemsweet$^\text{®}$ process, 213
Churn (transition) flow, 45
Clathrate, 68–69
Claus  
   converters, 306–307  
   process, 418  
   reaction, 303
Clinsulf process, 311
CNG. See Compressed natural gas
Coal-bed methane. See Natural gas from coal
Code of Federal Regulations (CFR), 442–443
Cold box, 478
Coldfinger process, 230
Columns, 475–476
Combustion air blowers, 326
Commercial software programs, 78–79
Commissioning, 437–442  
   control systems testing, 438–439  
   mechanical completion, 437–438  
   performance testing, 442  
   precommissioning, 437–438  
   process commissioning, 440–442  
   start-up procedures, 440
Communications medium, 415
Compressed natural gas (CNG), 24–27
Compression, 349  
   model, 492–494  
   power, 408  
   calculation, 366–367
Compression ratio (CR), 362–364
Compressor stations, 394  
   acoustical treatment, 400–401  
   arrangements, 399  
   facilities, 395–399  
   reliability and availability, 401–402  
   spacing, 404–408  
   control, 399–400
Compressors(s), 396  
   control, 367–374  
   drivers, 396–397  
   performance maps, 375–376  
   selection, 354–355
Computational fluid dynamics (CFD), 163
Computational pipeline monitoring. See Internally based methods
Condensate  
   effluent treatment, 176–178  
   hydrotreating, 174–175  
   production, 169–170  
   stabilizer case study, 434–436  
   storage, 178–180
tank design considerations, 178–179
tank emission control, 180
Condensate stabilization, 127, 169–174, 420–421
cascade flash separation process, 170–171
design considerations, 173–174
stabilizer column pressure, 173
stabilizer system control, 174
distillation process, 171–172
condensate and LPG production, 172, 173f
condensate production only, 171–172
operating problems, 174
Condensate/gas ratio (CGR), 483
Condition monitoring. See Predictive maintenance
Containment, 525
Continuous monitoring systems, 114
Contractual agreements, 133–135
flat fees contracts, 134
keep-whole contracts, 134
percentage of proceeds contracts, 134
processing fee contracts, 135
Control
functions, 399
methods, 109–111
system, 476
checkout, 469–470
testing, 438–439
Control room, 416
alarm management, 449–453
fatigue mitigation, 448–449
HAZOP, 445–446
LOPA, 446–448, 447f
management, 442–443
operational excellence key components, 453f
process safety management, 443–445
roles and responsibilities, 443
training, 453–454
Conventional demethanizer, 285
Conventional gas, 3, 17–20. See also
Unconventional gas
completion, 19
drilling, 18–19
electric logs, 18f
exploration, 17
production, 19–20
seismic survey, 17f
Conveying formed sulfur, 324
Cooling water system model, 500
Corrective maintenance, 456
Corrosion, 86–90
cathodic protection, 89
choice of corrosion-resistant metals, 87
inhibitors, 87–89
monitoring, 90
protective coatings, 89–90
Corrosion resistant alloy (CRA), 87–88
CR. See Compression ratio
CRA. See Corrosion resistant alloy
Cricondenbar, 4–5
Cricondentherm, 4–5
Cryogenic fractionation, 218
Cryogenic nitrogen rejection process, 338
double-column nitrogen rejection process, 340–342, 342f
process selection, 344–345
single-column nitrogen rejection process
classical design, 338–339, 339f
modified design, 339–340, 340f
two-column nitrogen rejection process, 343–344, 343f
Cryogenic process, 335–336
Cryogenic recovery, 421
CrystaSulf™ process, 313–314
CSB. See U.S. Chemical Safety and Hazard Investigation Board
Custom models, 492
Cyclone separators. See Centrifugal separators
D
D’GAASS process, 321–322
DAP. See Double absorption process
Darcy’s law, 19
Data, 529
historians, 424
DBM. See Design Basis Memorandum
DCS. See Distributed control system
DEA. See Diethanolamine
Debutanizer, 498–499
Deep hydrocarbon dew pointing, 273–274
Deethanizer, 498
Dehydration, 130
Deliverability equation, 19–20
Demethanizer, 422, 498
feed chilling models, 499
Demisting device, 139
“Dense phase” compressor, 362
DEPG. See Dimethyl ether of polyethylene glycol
Depleted reservoirs, 34
Depropanizer, 498
Design Basis Memorandum (DBM), 545
Design power, 501
Desuperheating station (DS), 268
DFM. See Drift-flux model
DGA. See Diglycolamine
Diethanolamine (DEA), 183–184
Diethylene glycol, 227
Differential pressure (DP), 157
Digital multimeter (DMM), 461–462
Digital technology, 399–400
Diglycolamine (DGA), 183–184
Diisopropanolamine (DIPA), 183, 185
Dimethyl ether (DME), 27
Dimethyl ether of polyethylene glycol (DEPG), 197, 202–207
carbon capture process, 205
hydrocarbon dew point control, 207
landfill gas, 207
mercaptan removal, 207
process, 203–205
DIPA. See Diisopropanolamine
Direct oxidation processes, 309–311
Clinsulf process, 311
Selectox process, 310–311
Dispersed bubble flow, 42
Distillation
calculations, 496–499
process, 171–172
condensate and LPG production, 172, 173f
condensate production only, 171–172
towers, 425
Distributed control system (DCS), 132–133, 413–416, 439, 462
DME. See Dimethyl ether
DMM. See Digital multimeter
Doctor sweetening process, 294–295
Double absorption process (DAP), 190–191, 319
Double-column nitrogen rejection process,
340–342, 342f. See also Single-column nitrogen rejection process; Two-column nitrogen rejection process
DP. See Differential pressure
Drift-flux model (DFM), 57–59
DRIZO™ process, 229
Dry-bed processes, 182
DS. See Desuperheating station
Dual column reflux process, 281–282
Dynamic analysis, 480–481
Dynamic models, 468
Dynamic simulation, 467
areas of application, 467
plant design, 468–471
plant operation, 471–473
equipment and process system control, 477–478
specific considerations, 474–476
fuel gas system start-up analysis, 478–481
in gas processing plants, 467
level of detail in model, 473–474
model speed, 474
trunk line online dynamic model, 481–485

E
EAM system. See Enterprise asset management system
EDI. See External Data Interface
EIA. See Energy Information Administration
Elemental sulfur, 301
Emergency shutdowns (ESD), 371, 373, 398, 469
Emulsions, 157–158
Energy Information Administration (EIA), 16
Engineering, Procurement, and Construction (EPC), 544
Enterprise asset management system (EAM system), 457
Enterprise resource planning (ERP), 519
Enterprise risk management (ERM), 558
Environmental engineer, 548
EOS. See Equation of state
EPC. See Engineering, Procurement, and Construction
Equation of state (EOS), 8, 59
Equilibrium zone (EZ), 244
ERM. See Enterprise risk management
ERP. See Enterprise resource planning
ESD. See Emergency shutdowns
EUR. See Expected ultimate recovery
Expander model, 494–496
Expander thermodynamics, 494f
Expected ultimate recovery (EUR), 20
External Data Interface (EDI), 489
EZ. See Equilibrium zone

F
Fatigue mitigation, 448–449
Fatigue risk management system (FRMS), 448–449
Fault Tree Analysis (FTA), 444
Feed
contaminants, 346–347
gas, 191, 338
characteristics, 345
cooling system, 268
preheating, 307
FEED. See Front-end engineering and design
FEF. See Front-end furnace
Ferrox process, 211
Fiber-reinforced plastic (FRP), 27
Filters, 402
Finger-type slug catcher, 149f
Fischer–Tropsch process (FT process), 27
Fixed fee price, 548–549
Flare system, 399
Flat fees contracts, 134
Flexsorb, 316–317
FLEXSORB SE process, 186
Floating liquefied natural gas (FLNG), 26
Flow assurance, 68
risk management, 111–114
Flow pattern maps, 46–48
Flow sheet development, 505
FLUOR Solvent process, 198–200
Fluor twin column high absorption process, 285
Fluor twin-reflux absorption process, 285–287
Foaming, 195, 347
Formed sulfur storage, 324
Fractionators, 492
Friction factor model. See Homogeneous flow approaches
FRMS. See Fatigue risk management system
Front-end engineering and design (FEED), 546
Front-end furnace (FEF), 309
FRP. See Fiber-reinforced plastic
FT process. See Fischer–Tropsch process
FTA. See Fault Tree Analysis
Fuel gas system, 478, 479f
start-up analysis, 478–479
dynamic analysis, 480–481
dynamic model, 481
steady-state analysis, 479–480

G
Gas. See also Natural gas
compressibility factor, 8–12
compressors, 425
condensate “A”, 92–93
condensate “B”, 93
dehydration, 419–420, 477
unit, 127
density, 13
flow fundamentals, 383
friction factor correlations, 385–388
general flow equation, 384–385
simplified flow equations, 388–390
formation volume factor, 12–13
gathering
system, 417–418
and transmission, 477
gravity method, 75–78
hydrates, 3–4, 68–86
locus for natural gas components, 70, 71f
prediction of hydrate formation conditions, 70–79
prevention techniques, 79–86
piping, 401
plant profitability, 518
specific gravity, 8
temperature profile prediction, 390–392
transmission pipelines
compressor station, 394f
transient flow in, 392–393
transmission system, 394
treating, 418, 477
viscosity, 14–15
Gas processing, 544
modeling, 467
and optimization strategy, 513–514
operations, 487
plants, 467
model speed, 474
Gas coalescers, 150–157, 152f. See also
High-efficiency liquid–liquid coalescers
aerosols, 150–151, 151f
applications, 157
construction/operation principles, 151–154
modeling liquid, 154–156
performance/operational limits, 157
effect of surface treatment, 153f
Gas compression thermodynamics, 355
basic relations, 356–357
isentropic model, 357–360
polytropic model, 360–361
real gas behavior, 361–362
Gas deviation factor. See Gas—compressibility factor
Gas horsepower (GHP), 366
Gas plant assets profitability, 517
challenges, 517
industrial relevance, 536
information hierarchy, 530f
integrated gas plant, 518–519
impact of living with information technology, 529–530
miscellaneous initiatives, 539
model-based asset management, 532
operations strategy, 531
optimization, 533
alternatives, 534–536
tools for, 534
Gas plant assets profitability (Continued)
organizational behavior model, 519–528
scientific approach, 537–539
strategies for organizational behavior and
information, 519
successful information strategy, 528–529
technology integration challenge, 537
vision of modern plant operation, 530–531
Gas plant project management, 543–544.
See also Integrated gas plant; Natural gas
commissioning and start-up, 566
industry perspective, 544–545
operation and evaluation, 566–567
process, 545
conceptual estimates and schedules, 549–551
contracting strategy, 548–549
defining business and project objectives,
546–548
pre-project planning measurement, 552
project execution planning, 551–552
responsibility matrix, 552–554, 553t
team integration phases, 554f
project closeout, 567
project controls, 554
project timeline, 555–556
risk management, 556–564
quality assurance, 564–565
Gas processing plant automation, 413
applications, 424–433
centrifugal
compressors, 422–423
pumps, 423
condensate stabilizer case study, 433–436
early methods, 413–414
gas dehydration, 419–420
gas gathering system, 417–418
gas treating system, 418
liquids recovery, 420–422
microprocessor-based automation, 414–416
NGL fractionation, 422
reciprocating pumps, 423
sulfur recovery, 418–419
utilities, 423–424
Gas processing plant operations, 437
commissioning and start-up
control systems testing, 438–439
mechanical completion, 437–438
performance testing, 442
precommissioning, 437–438
process commissioning, 440–442
start-up procedures, 440
control room management, 442–454
maintenance, 454–458
plant turnovers, 464
troubleshooting, 458–463
Gas subcooled process (GSP), 282–283
Gas-condensate
flow regimes, 49
wax deposition envelope, 92–95
Gas-to-liquids (GTL), 24–25, 27–28
Gas-to-solids technologies (GTS technologies),
24–25, 28–29
Gas-to-wire technologies (GTW technologies),
24–25, 29–30
Gas-well gas, 3–4
Gas–liquid cylindrical cyclone separator (GLCC
separator), 145–146, 146f
Gas–oil ratio streams (GOR streams), 141
GCV. See Gross Calorific Value
GHP. See Gas horsepower
GLCC separator. See Gas–liquid cylindrical
cyclone separator
Glycol
circulate rate, 232–233
dehydration, 226–237
conventional TEG dehydration process,
227–228
enhanced TEG dehydration process, 229–230
future technology developments, 237
glycol injection process, 230–232
operational problems, 235–237
TEG unit design considerations, 232–234
gas dehydration process, 477
injection process, 230–232
purity, 234
GOR streams. See Gas–oil ratio streams
Granulated sulfur, 324
Graphical user interface (GUI), 484, 513
Gravity segregation, 137
Gravity separators, 137–144
design considerations, 144
general description, 138–140
horizontal three-phase separator, 139f
selection, 140–142
theory, 142–144
vertical three-phase separator, 140f
Gross Calorific Value (GCV), 506
GSP. See Gas subcooled process
GTL. See Gas-to-liquids
GTS technologies. See Gas-to-solids
technologies
GTW technologies. See Gas-to-wire
technologies
GUI. See Graphical user interface
Index 587

Hagen–Poiseuille equation, 386
Hazard and Operability Study (HAZOP), 444–446
analysis, 550
HCN. See Hydrogen cyanide
HE. See Hydrate envelope
Health, Safety, and Environment (HSE), 518
Heat exchangers, 476
Heaters, 403
Heating value
natural gas, 387
of sales gas, 386
Hexane plus (C₆₊), 5–6
High-efficiency liquid. See Gas coalescers
High-efficiency liquid–liquid coalescers, 157–162, 160f
applications, 162
emulsions, 157–158
limitations, 162
liquid–liquid coalescer performance, 161–162
mechanism of operation, 158–161
coalesced droplets separation, 160–161
coalescence, 159–160
solid separation /fluid preconditioning, 158–159
principles and materials of construction, 158
High-pressure (HP), 126
column, 343
services, 126
High-voltage direct current (HVDC), 24–25
Higher heating value, 6
HMI. See Human Machine Interface
Holdup, 38–39
Homogeneous flow approaches, 50–56
Beggs and Brill method, 52–56
Lockhart and Martinelli method, 51–52
Horizontal flow regimes, 42–44. See also Vertical flow regimes
annular flow, 44
disperssed bubble flow, 42
plug (elongated bubble) flow, 42
slug flow, 43–44
stratified (smooth and wavy) flow, 43
Horizontal separators, 141
Horizontal three-phase separator, 139f
Horsepower (HP), 356
Hot gas bypassing, 307
Hot pot process, 195
HP. See High-pressure; Horsepower
HSE. See Health, Safety, and Environment
Human Machine Interface (HMI), 414
HVDC. See High-voltage direct current
Hydrate curve, 70
formation, 347
prevention techniques, 79–86
chemical inhibition, 80–86
thermal methods, 79–80
Hydrate envelope (HE), 95
Hydrocarbon dew pointing control, 207
deep, 273–274
gas plant with, 125–128, 126f
acid gas removal unit, 127
condensate stabilization, 127
gas compression and transmission, 128
gas dehydration unit, 127
inlet facility, 125–126
nitrogen rejection, 128
sulfur recovery and handling unit, 127
with Joule–Thomson cooling, 271–272
with propane refrigeration, 272
Hydrocarbon removal processes, 287–289
membrane separation, 287–289
solid-bed adsorption process, 287
Twister® supersonic separation, 289
Hydrodynamic slugging, 102–103
Hydrogen cyanide (HCN), 202
Hydrogen sulfide (H₂S), 11–12, 181, 301
scavenger, 311–312
liquid injection scavenger, 311–312
selective H₂S removal section, 316–317
Hydrogenation section, 315–316
Hydrostatic tests, 438
I/O. See Input–output
IFPEX-1®, 258
IGCC. See Integrated gasification combined cycle
IGT. See Institute of Gas Technology
In situ velocity. See Phase velocity
Incident analysis, 472
Inclined flow regimes, 45–46
Independent protection layer (IPL), 447
Industrial relevance, 536
Inflow performance relationship (IPR), 22
Information, 529
Inlet diverters, 142
Inlet facility, 125–126
Input–output (I/O), 415–416, 461
Institute of Gas Technology (IGT), 388
Insulation, 346
### Integrated gas plant, 518–519
Integrated gasification combined cycle (IGCC), 203
Intelligent pigs, 66
Internally based methods, 65–66
Interstage cooling, 364
IPL. See Independent protection layer
IPR. See Inflow performance relationship
Iron sponge process, 210–211
Isentropic model, 357–360
Isothermal gas compressibility, 13–14

### J
Joule–Thomson (JT) coefficient, 15–16, 390–391
cooling, 335–336
effect, 392
valve, 421, 478–479, 534

### K
K-factor method, 70–74
Kärstø gas processing plant, 505, 507f
implementation and usage, 512–513
large energy consumer, 509
modeling and optimization strategy, 513–514
off-line usage, 515
online usage, 514
optimization model features, 512
plant operation, 506–509
plant-wide model, 513
process description, 506
production objectives, 509
project drivers, 509–512
quantifying measurement errors, 514–515
use for planners, 515
Keep-whole contracts, 134
Kennelly equation, 391
Kerogen, 2–3
Key performance indicator (KPI), 526–527
Kinetic hydrate inhibitor (KHI), 80–82
KSB. See Souders and Brown
design coefficient

### L
Landfill gas, 207
Layer of protection analysis (LOPA), 446–448, 447f
LDHI. See Low Dosage Hydrate Inhibitor
Leak detection, 65–66
Lean acid gas operations, 307–308
acid gas enrichment, 308
acid gas/natural gas fuel burner, 308
feed preheating, 307
hot gas bypassing, 307
Lean amine feed locations, 192
Lean oil absorption process, 279
Leveraging automation, 430
automation upgrade master plans, 430
benefits, 432–433
Linde’s Clausulf process, 309
Line sizing criteria, 403–404
Linear programming technique (LP technique), 427
Liquefied natural gas (LNG), 24–26, 124–125
Liquefied petroleum gas (LPG), 124, 171
Liquid enthalpy, 497
injection scavenger, 311–312
recovery, 420–422, 478
LNG. See Liquefied natural gas
Load control, 380
Lockhart and Martinelli method, 51–52
LOPA. See Layer of protection analysis
Low Dosage Hydrate Inhibitor (LDHI), 80–82
Low-temperature separation (LTS), 482
LP technique. See Linear programming technique
LPG. See Liquefied petroleum gas
LTS. See Low-temperature separation

### M
mA. See Milliamp
MAC. See Maximum Available Capacity
Maintenance, gas processing plant, 454
EAM system, 457
RCM, 457–458
types, 455–457
Man-Technology-Organization (MTO), 533
MAOP. See Maximum allowable operating pressure
Mass transfer zone (MTZ), 244
Material balance equation, 19–20
model, 502
Maximum allowable operating pressure (MAOP), 405
Maximum Available Capacity (MAC), 511
MDEA. See Methyl-diethanolamine
MEA. See Monoethanolamine
Mechanical completion, 437–438
Mechanistic models, 56
Media velocity ($V_{med}$), 154–155. See also Annular velocity ($V_{ann}$)
effect of system conditions on, 155
See Monoethylene glycol
Membrane separation, 287–289
Membrane systems, 214–217
membrane process, 216–217
advantages, 215
disadvantages, 215–216
pretreatment system, 217
Membranes and Twister technology, 258
MeOH. See Methanol
Mercaptan, 7, 181
removal, 207
Mercury, 223
Mercury removal
natural gas hydration, 260–262
nonregenerative mercury sorbents, 260–261
process selection considerations, 261–262
regenerative mercury adsorbents, 261
NG, 130
Merox® process, 295
Metal oxides (MO), 211
Metering stations, 402–403. See also
Compressor stations
Metering system, 403
Methanol (MeOH), 197, 207–208
refrigeration, 258
Methyl diethanolamine (MDEA), 183, 185
Microbiological treatment processes, 218–219, 314
Microprocessor-based automation, 414–416
Milliamp (mA), 461–462
Minimum housing diameter determination, 156
Mixed refrigerants, 270–271
Mixture
density, 40
enthalpy, 41
pressure drop, 41
velocity, 38
viscosity, 40–41
MO. See Metal oxides
Model-based asset management, 532
Model-based Predictive Control (MPC), 536
Modern NGL recovery processes, 281–287.
See also Turboexpander NGL recovery processes
dual column reflux process, 281–282
Fluor twin column high absorption process, 285
Fluor twin-reflux absorption process, 285–287
GSP, 282–283
Ortlhoff single-column overhead recycle process, 283
residue gas recycle, 283–285
Modified Claus process, 303–309
lean acid gas operations, 307–308
acid gas enrichment, 308
acid gas/natural gas fuel burner, 308
feed preheating, 307
hot gas bypassing, 307
oxygen enrichment, 308–309
process description, 303–307
catalytic section, 305–306
Claus burner performance, 306
COS and CS2 destruction, 306–307
thermal section, 303–305
Molecular Gate™ adsorbent, 336
Molecular Gate™ adsorption-based technology, 336
Molecular sieves, 239–242
technology, 295–296
Molten sulfur handling system, 323
Monoethanolamine (MEA), 183–184
Monoethylene glycol (MEG), 126, 170, 227
regeneration and reclaiming, 176–177, 176f
Moody correlation, 386
Moody friction factor, 387f, 388
MPC. See Model-based Predictive Control;
Multivariable Predictive Controller
MTO. See Man-Technology-Organization
MTZ. See Mass transfer zone
Multiphase flow assurance, 68–114
corrosion, 86–90
flow assurance risk management, 111–114
gas hydrates, 68–86
slugging, 102–111
wax, 90–102
Multiphase gas flow, 59–60
Multiphase pipeline operations, 65–68
leak detection, 65–66
pigging, 66–68
pipeline depressurization, 66
“Multiphase riser base lift” technique, 108
Multiphase transportation technology, 37
Multistage separation, 145
Multivariable predictive control (MVPC), 427–428, 471
Multivariable Predictive Controller (MPC), 472–473
Multivariable Predictive Controller (MPC), 472–473
Murphree vapor tray efficiency, 497
MVPC. See Multivariable predictive control
N
N-methyl-2-pyrrolidone (NMP), 197, 208
Natural gas, 1, 383. See also Gas plant project management; Gas processing plant automation; Integrated gas plant
Natural gas (Continued)

chemical and physical properties, 6–15, 7t from coal, 3
comparison of methods, 30–33
composition and classification, 3–4
dehydration, 223
glycol dehydration, 226–237
mercury removal, 260–262
process selection, 259–260
processes, 258
solid-bed dehydration, 237–258
water content determination, 224, 226f
exploration and production, 16–23
heaters, 403
history, 1–2
origin and sources, 2–3
phase behavior, 4–6
processing, 33
contractual agreements, 133–135
industry, 520
objectives, 123–124
pipeline gas specifications, 124t
plant configurations, 124–131
route, 131
support systems, 132–133
reserves, 16
thermodynamic properties, 15–16
transportation, 24–33
treating, 181
chemical absorption processes, 183–195
cryogenic fractionation, 218
membrane systems, 214–217
microbiological treatment processes, 218–219
mixed physical and chemical absorption processes, 208–210
physical solvent processes, 196–208
processes, 182
selection, 219–220
solid bed absorption processes, 210–214
specifications, 181
Natural gas compression
antisurge and recycle system, 371f
centrifugal compressors, 351–353, 352f
comparison between compressors, 353–354
compressor
control, 367–374
design, 364
performance maps, 375–376
power calculation, 366–367
selection, 354–355

CR, 362–364
example for operating compressor in pipeline system, 376–380
gas compression thermodynamics, 355–362
reciprocating compressors, 350–351, 350f stages, 364–366
Natural gas fuel burner, 308
Natural gas hydrates (NGHs), 28
Natural gas liquid (NGL), 124, 223, 265, 442, 478, 505, 508
CO2 removal, 129
dehydration, 130
desorption, 292–293, 422, 478
column design and operation, 293
gas plant for, 128–131
hydrocarbon dew point
deep, 273–274
with Joule–Thomson cooling, 271–272
with propane refrigeration, 272
hydrocarbon removal processes, 287–289
liquid product processing, 293–298
dehydration, 298
NGL contaminant treating, 293–298
mercury removal, 130
modern NGL recovery processes, 281–287
recovery, 130–131, 265
refrigeration processes, 266–271
selection, 289–290
technology development, 290
turboexpander NGL recovery processes, 274–279
unit design considerations, 291
unit operating problems, 291
Near-infrared equipment (NIR equipment), 93
Netural network-based controllers, 534
models, 534
NGHs. See Natural gas hydrates
NGL. See Natural gas liquid
NIR equipment. See Near-infrared equipment
Nitrogen, 335
cryogenic nitrogen rejection, 338–345
design considerations, 345–346
NRU integration, 336–338
operating problems, 346–347
rejection, 128
cryogenic process, 335–336
noncryogenic process, 336
options, 335
safety, 347
Nitrogen rejection unit (NRU), 335, 434
with helium production, 338f
integration, 336–338
for sales gas production, 337f
NMP. See N-methyl-2-pyrrolidone
NOGAT. See Northern Offshore Gas Transport Noise, 400–401
Noncryogenic nitrogen separation process, 336
Noncryogenic process, 336. See also Cryogenic nitrogen rejection process
Nonregenerative mercury sorbents, 260–261
Northern Offshore Gas Transport (NOGAT), 481–482
trunk line system, 482f
NRU. See Nitrogen rejection unit

O
Object Linking and Embedding for Process Control (OPC), 416
OEE. See Overall equipment effectiveness
Off-line
simulators, 535
system, 515
usage, 515
Oil well gas. See Casinghead gas
Online model, 483, 513
Online-adaptive control, 488
OPC. See Object Linking and Embedding for Process Control
Operating expenditure (OPEX), 397
Operational decisions, 530
Operationally induced slugging, 111
“Operations wall”, 521
Operator decision support, 472
Operator training, 472
Operator training simulator (OTS), 567
OPEX. See Operating expenditure
Optimization, 429–430
gas plant assets, 533
alternatives, 534–536
tools for, 534
objective function, 492
Organizational behavior model, 519–520
behavior, 528
capability to perform, 524–526
information quality, 520–522
levels of readiness, 524f
organizational hierarchy of needs, 526–528
perception of information, 522–523
Vollman Triangle, 527f
Ortloff single-column overhead recycle process, 283
OSHA. See U.S. Department of Labor
Occupational Safety & Health Administration
OTS. See Operator training simulator
Overall equipment effectiveness (OEE), 454
Oxygen enrichment, 308–309

P
Panhandle
“A” equation, 389–390
“B” equation, 389–390
PC. See Propylene carbonate
Peng–Robinson equations of state (PR equations of state), 489–491
Performance management, 425–426
Performance testing, 442
Periodic maintenance, 456
PHA. See Process hazard analysis
Phase separation, 137
centrifugal separators, 145–146
gas coalescers, 150–157, 152f
gravity separators, 137–144
high-efficiency liquid–liquid coalescers, 157–162
multistage separation, 145
practical design, 162–165
analysis, 164–165
case study, 163
modified situation, 165, 165f
objective and methodology, 164
situation, 164, 164f
slug catchers, 148–150
Twister™ supersonic separator, 146–147
Phase velocity, 39
PHMSA. See U.S. Department of Transportation Pipeline and Hazardous Materials Safety Administration
Physical absorption, 418
Physical adsorption, 237–238
Physical solvent processes, 196–208
DEPG, 202–207
methanol (rectisol process), 207–208
NMP, 208
PC, 198–202
PID. See Proportional, integral, and derivative
Pigging, 66–68, 67f
Pipe-type slug catcher, 148–149
Pipelines, 25
depressurization, 66
operations, 409–410
Piping, 325
equipment, 475
and valves, 398
Planning interface, 513

Plant
  design, 468
  APC, 471
  control system checkout, 469–470
  controllability, 468
  operability, 468
  operator training, 470
  safety analysis, 468–469
  safety integrity system checkout, 469–470
  start-up procedure, 469
  manager, 547–548
  model integration, 502
  measurement errors, 502–504
  model fidelity, 502–504
  operation, 471
    APC system, 472–473
    incident analysis, 472
    operator decision support, 472
    operator training, 472
    plant performance enhancement, 471
    troubleshooting, 471
  optimum operation, 489
  turnarounds, 464
  utility system, 426

Plant Production Performance Model (3PM), 505, 514

PLC. See Programmable logic controller

Plug (elongated bubble) flow, 42

Polytropic compression, 356

Polytropic model, 360–361

Potassium carbonate solution, 195

Power factor, 501

Power loss, 501

PR equations of state. See Peng–Robinson equations of state

Pre-project planning measurement, 552

Precommissioning, 437–438

Predictive maintenance method, 456

Pressure
  control, 346
  ratio, 363
  reduction and regulation system, 403
  relief system, 398

Pressure swing adsorption (PSA), 252, 336

Pressure–volume diagram (PV diagram), 350–351, 356f

Prevention, 525

Preventive maintenance, 456

Prilled sulfur, 324

Primary thermogenic gas, 2–3

Proactive “life extension” maintenance, 456–457

Process commissioning, 440–442

Process engineer, 548

Process hazard analysis (PHA), 444

Process safety management (PSM), 437

Processing fee contracts, 135

Production engineer, 548

Programmable logic controller (PLC), 400, 414, 439

Project
  charter, 547
  execution planning, 551–552
  management, 543
  manager, 543, 547
  risk management methodology, 558–559
  team roles and responsibilities, 547–548
  timeline, 555–556

Project coordinator. See Project—manager

Propane refrigeration, 266–268

Proportional, integral, and derivative (PID), 413–414

Proprietary solvents, 316–317

Propylene carbonate (PC), 197–202

FLUOR solvent unit, 198–200

innovations in FLUOR solvent process, 201–202

Protective coatings, 89–90

Proved reserves, 16–17

PSA. See Pressure swing adsorption

PSM. See Process safety management

PURASPECTM JM technology, 211

PURASPEC™/C212 materials, 260–261

Purchasing representative, 548

“Purisol” process, 208

PV diagram. See Pressure–volume diagram

Pyrobitumen, 2–3

Q

Qualitative project risk management, 560–561

Quality
  assurance, 564–565
  control, 565
  planning, 564

Quantitative project risk management assessment, 561

“Quick cycle” units, 242–243

R

Rationalization, 451

Raw gas transmission
  multiphase flow
assurance, 68–114

design parameters, 49–60
regimes, 42–49
terminology, 37–41
multiphase pipeline operations, 65–68
temperature profile prediction, 60–64
velocity criteria for sizing, 64

RCM. See Reliability Centered Maintenance

Re. See Reynolds numbers

Real gas behavior, 361–362

Real-time control model, 534
Real-time optimization (RTO), 487
APC system, 488–489, 488f
EDI interface, 490f
functions, 489
Kärstø gas processing plant, 505–515
level, 533
model interface, 490f
optimization models, 491–502
physical properties, 489–491
plant model integration, 502–504
process plant online economic optimization, 487–488
project considerations, 504–505
RTS interface, 491f

Real-Time Scheduling system (RTS system), 489

Reboiler, 236–237
duties, 345–346
hydraulics, 346
Reciprocating compressors, 350–351, 350f, 363, 368, 375, 378–380. See also Centrifugal compressors
Reciprocating pumps, 423
Redox process, 312
Reducing gas generator (RGG), 315
Reduction, 402–403
hydrogenation section, 315–316
processes, 314–317
selective H₂S removal section, 316–317
Reflux

duties, 345–346
exchanger, 274
Refrigeration, 421
models, 499
processes, 266–271
cascade refrigeration, 268–270
mixed refrigerants, 270–271
propane refrigeration, 266–268
system, 423–424
unit, 266–268

Regenerator, 236

Regulatory risk, 563–564

Reid Vapor Pressure (RVP), 169–170, 433

Reliability Centered Maintenance (RCM), 457–458

Remote control panel, 415
Remote Transmission Unit (RTU), 415
Reservoir engineer, 548

Residue gas recycle, 283–285

Responsibility matrix, 552–554, 553t

Retrograde condensation, 4–5

Return on Capital Employed (ROCE), 518

Revenue stream stabilization, 564

Reynolds numbers (Re), 156, 385, 388

RGG. See Reducing gas generator

Rhombic sulfur, 301

Riser base gas injection, 108–109

Riser-induced (severe) slugging, 103–111

mechanism, 105–106

prevention and control, 107–111
control methods, 109–111
riser base gas injection, 108–109
topside choking, 109
stability analysis, 107

Risk management, 556–557

developing risk response strategies, 559–560

interaction with other management processes, 562–563

project risk management methodology, 558–559

qualitative project risk management, 560–561

quantitative project risk management assessment, 561

risk matrices, 560f

risk mitigation concepts, 563–564

risk process modeling, 561–562

risk response planning, 559

ROCE. See Return on Capital Employed

Rod loading, 417

Rotating equipment models, 492

RTO. See Real-time optimization

RTS system. See Real-Time Scheduling system

RTU. See Remote Transmission Unit

RVP. See Reid Vapor Pressure

Ryan–Holmes process, 218

S

S3. See Slug suppression system

Safety

analysis, 468–469

integrity system checkout, 469–470
Safety (Continued)
systems, 133
Safety instrumented system (SIS), 446
Sales gas, 506
carbon dioxide concentration, 508–509
heating value, 386
NRU for, 337f
Sales gas transmission, 33
compressor stations, 394–402
design considerations, 403
compression power, 408
compressor station spacing, 404–408
gas flow fundamentals, 383–390
gas temperature profile prediction, 390–392
pipeline operations, 409–410
reduction and metering stations, 402–403
transient flow in gas transmission pipelines, 392–393
Salt caverns, 34–35
SCADA system. See Supervisory control and data acquisition system
SCORE. See Single-column overhead recycle process
SCOT. See Shell Claus Off-gas Treating
Scrubbers, 138, 396
Secondary thermogenic gas, 2–3
SEEHT. See Skin effect electrical heat tracing
Selectox process, 310–311
Selling, general, and administrative (SG&A), 539
Shell Claus Off-gas Treating (SCOT), 314–315
Short cycle units, 242–243
Shutdown system, 398
Silica gels, 242
Single-column nitrogen rejection process. See also Double-column nitrogen rejection process; Two-column nitrogen rejection process
classical design, 338–339, 339f
modified design, 339–340
Single-column overhead recycle process (SCORE), 283
Single-phase flow approaches, 50
SIS. See Safety instrumented system
Skin effect electrical heat tracing (SEEHT), 323
Slating, 323
Slip, 39
law, 58
velocity, 39
Slug catchers, 148–150
flow, 43–45
Slug suppression system (S3), 110
Slugging, 102–111
hydrodynamic slugging, 102–103
operationally induced slugging, 111
riser-induced (severe) slugging, 103–111
terrain-induced slugging, 103
Slurry processes, 213
Chemsweet® process, 213
Sulfa-Check® process, 213
Small-and medium-scale processes, 311–314
CrystaSulf™ process, 313–314
H2S scavenger, 311–312
microbiological treatment processes, 314
Redox process, 312
SNPA. See Societe Nationale des Petroles d’Aquitaine
SO2 scrubbing processes, 317
Soave–Redlich–Kwong equations of state (SRK equations of state), 489–491
Societe Nationale des Petroles d’Aquitaine (SNPA), 185
Solid bed absorption processes, 210–213
iron sponge process, 210–211
PURASPECTM technology, 211
slurry processes, 213
zinc oxide process, 211
Solid desiccants properties, 243, 243t
Solid-bed adsorption process, 213–214, 287
Solid-bed dehydration, 237–258
adsorbent selection, 238–243
adsorption capacity, 238
adsorption technology, 243–247
adsorption principle, 243–244
solid-bed design considerations, 244–247
operation, 247–251
2 + 2 mode of operation, 250
3 + 1 mode of operation, 247–250
mode operations, 250–251
operational problems, 253–258
unit design considerations, 251–253
Solidification. See Wax precipitation
Solution-diffusion type, 214
Sorbead™ gel, 242
Souders and Brown design coefficient (KSB), 143
Sour gas, 181
water stripping, 177–178, 177f
Spare capacity, 505
Specific heat, 15
Specific speed, 495
SRK equations of state. See Soave–Redlich–Kwong equations of state
SRU. See Sulfur recovery unit
Stabilizer
column pressure, 173
system control, 174
Standards and protocols, 416
Start-up procedures, 440, 469
Statistical analysis, 432
Statistical process control, 426–427
Steady-state
analysis, 479–480
detection, 489
model, 467
process model, 534
three-phase flow, 56–57
two-phase flow, 50–56
homogeneous flow approaches, 50–56
mechanistic models, 56
single-phase flow approaches, 50
Steam boiler system, 426
Steam water system model, 500
Sterically hindered amines, 186
Stratified (smooth and wavy) flow, 43
Sulfrex® process, 295
Sulfur
condensers, 327
degassing, 320–322
disposal by acid gas injection, 332–333
forming, 323–324
handling unit, 127
pit, 327
properties, 301–302
recovery, 302–314, 418–419, 477
selection, 331–332
storage and handling, 322–324
Sulfur recovery unit (SRU), 127, 303
design considerations, 324–327
acid gas feed drums, 325–326
combustion air blowers, 326
main burner and reaction furnace, 326
piping, 325
waste heat boiler, 326–327
operation problems, 327–331
carbon deposits, 329
catalyst support screens, 330
combustion air control, 330–331
excessive COS and CS₂, 329
leakage of reheat exchanger, 329
proper air ratio, 327–328
reactor activity, 328
reactor pressure drop, 329
steam heater, 330
water vapor and carbon dioxide, 330
Super-compressibility factor, 8, 13–14
SuperClaus process, 318
Superficial velocity, 38
Supervisory control and data acquisition system (SCADA system), 65–66, 132–133, 395, 415, 443
Support systems, 132–133
process control, 132–133
safety systems, 133
utility and off-site, 132
Surfactants, 157–158
Sweet gas, 181
T
Tactical decisions, 530–531
Tail gas cleanup, 314–319
catalytic oxidation, 318
reduction processes, 314–317
SO₂ scrubbing processes, 317
tail gas treating configurations, 318–319
integration with AGEU, 319
integration with AGRU, 319
Tail gas treating unit (TGTU), 305
Tank design considerations, 178–179
Tank emission control, 180
TCHAP. See Twin column high absorption process
TEG. See Triethylene glycol
Temperature control, 346
Temperature swing adsorption (TSA), 252
Terrain-induced slugging, 103
Tetraethylene glycol, 227
TFM. See Two-fluid model
TGs. See Turbogenerators
TGTU. See Tail gas treating unit
Thermal methods, 79–80
Thermodynamics
expander, 494f
gas compression, 355
basic relations, 356–357
isentropic model, 357–360
polytropic model, 360–361
real gas behavior, 361–362
inhibitors, 80–83
second law, 359–360
Thiol. See Mercaptan
THIOPAQ™ process, 218, 219f, 314
Three-phase flow regimes, 48–49. See also Two-phase flow regimes
“Tie-back” model, 470
Time and expense contract, 548–549
Time-based maintenance, 456
Titanium oxide (TiO₂), 306–307
Topside choking, 109
TPC. See Tubing performance curve
Transient flow
in gas transmission pipelines, 392–393
multiphase flow, 57–59
DFM, 58–59
TFM, 57–58
Transportation tariffs, 32f
TRAP. See Twin reflux absorption process
Tray-to-tray distillation models, 496–498
Triazine, 312
Triethylene glycol (TEG), 130, 227
unit design considerations, 232–234
Troubleshooting, 458
documentation, 461
instrumentation, 461–463
process troubleshooting, 463
steps, 459–461
Troubleshooting, 471
True Vapor Pressure (TVP), 169–170
Trunk line online dynamic model, 481–485
TSA. See Temperature swing adsorption
Tubing performance curve (TPC), 22–23
Turbine model, 500–502, 500f
Turboexpander NGL recovery processes,
274–279
BAHXs, 278–279
lean oil absorption process, 279
Turboexpander process. See Cryogenic recovery
Turbogenerators (TGs), 478–479
Turbomachinery control, 475
Turn-key project. See Fixed fee price
Turndown, 371
TVP. See True Vapor Pressure
Twin column high absorption process (TCHAP), 285
Twin reflux absorption process (TRAP), 286
Twister™
separation, 289
supersonic separator, 146–147
Two-column nitrogen rejection process,
343–344, 343f. See also Double-column nitrogen rejection process; Single-column nitrogen rejection process
Two-fluid model (TFM), 56–58
Two-phase flow regimes, 42–48
flow pattern maps, 46–48
horizontal flow regimes, 42–44
inclined flow regimes, 45–46
vertical flow regimes, 44–45
Two-stage absorption process, 189–190

U
U.S. Chemical Safety and Hazard Investigation Board (CSB), 448
U.S. Department of Labor Occupational Safety & Health Administration (OSHA), 437
U.S. Department of Transportation Pipeline and Hazardous Materials Safety Administration (PHMSA), 443
Unconventional gas. See also Conventional gas
completion, 21
drilling, 21
exploration, 20
production, 21–22
resources, 3
Underground gas storage, 33–34
aquifers, 34
depressed reservoirs, 34
salt caverns, 34–35

V
Validity checking, 503
Valves, 474–475
Vapor enthalpy, 497
Vapor-solid equilibrium constants, 71–72
carbon dioxide and hydrogen sulfide, 76f
isobutane, 74f
methane and ethane, 72f
for n-butane, 75f
propane, 73f
Vapor–liquid equilibrium definition, 497
Variable speed planetary gear (VSPG), 396–397
Venting system, 398
Vertical flow regimes, 44–45. See also
Horizontal flow regimes
annular flow, 45
bubble flow, 45
churn (transition) flow, 45
slug flow, 45
Vertical separators, 141–142
Vertical three-phase separator, 140f
Vessel-type slug catchers, 148–149
Vetrocoke process, 195
Virtual Private Network (VPN), 521
Void fraction, 39
Volatile organic compound (VOC), 227–228
Vollman Triangle, 527, 527f
Volume Optimized Transport and Storage (VOTRANS™), 27

W
Waste heat boiler (WHB), 303–304, 326–327
Water
content determination, 224, 226f
refluxing, 256
wash trays, 191
water-washing process, 138–139
Wax, 90–102
controlled production of wax deposits, 101–102
deposition, 90–95
gas-condensate wax deposition envelope, 92–95
inhibition/prevention, 98–100
problems identification, 97–98
remediation, 101
WDE, 91–92
formation in multiphase gas-condensate pipelines, 95–102
precipitation, 90–91
slush, 99
Wax deposition envelope (WDE), 91–92
Wax precipitation, 90–91
Web-based optimization, 535
Well deliverability, 22–23, 23f
Wet-prilled sulfur, 324
Weymouth equation, 388–390
WHB. See Waste heat boiler
Wobbe Index (WI), 6, 482–483, 506
Wobbe number, 6

Y
Y-Grade, 265, 266t

Z
Z-factor, 8–11
Zeolites. See Molecular sieves
Zinc oxide process, 211