A Universal Methodology to Develop, Test, and Calibrate a Carbon Dioxide Enhanced Oil Recovery and Storage Capacity Estimate

Duane B. Myers,[1] Carrie M. Beitler,[1] Ray W. McKaskle,[1] and Scott M. Frailey[2]

[1]Trimeric Corporation, Buda, Texas

[2]Illinois State Geological Survey, Prairie Research Institute, University of Illinois at Urbana-Champaign, Champaign, Illinois

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Front cover: *Typical horizontal free water knockout vessel. The horizontal slug catcher and demulsifier vessels are similar in appearance. Photograph courtesy of Denbury Onshore.*

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[1]Trimeric Corporation, Buda, Texas

[2]Illinois State Geological Survey, Prairie Research Institute, University of Illinois
at Urbana-Champaign, Champaign, Illinois

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ILLINOIS STATE GEOLOGICAL SURVEY
Prairie Research Institute
University of Illinois at Urbana-Champaign
615 E. Peabody Drive
Champaign, Illinois 61820-6918
http://www.isgs.illinois.edu

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Principal Investigator: Sallie Greenberg
Illinois State Geological Survey
Prairie Research Institute
University of Illinois
(217) 244-4068
sallieg@illinois.edu

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Trimeric Corporation

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EXECUTIVE SUMMARY

The Midwest Geological Sequestration Consortium is leading a program to demonstrate the feasibility of carbon dioxide (CO_2) capture and storage, particularly in the Illinois Basin. One potential storage method uses CO_2 for enhanced oil recovery (EOR) by injecting it in producing oil reservoirs whose production rates have been diminished by conventional means (e.g., waterflooding). A fraction of the CO_2 that is injected returns to the surface with the produced oil and is captured and compressed for reinjection. Trimeric, working with the Midwest Geological Sequestration Consortium, has developed conceptual process designs and estimated the costs for a variety of EOR surface processing facility configurations so that the CO_2 accompanying the produced oil can be captured and reinjected. The scope of the facility work included the following major tasks:

- Defining the equipment that would be required for typical facilities;
- Identifying capacity breakpoints in the major equipment (i.e., CO_2 compressor frame sizes);
- Estimating capital and operating costs for the facilities; and
- Evaluating the feasibility and applicability of natural gas liquid (NGL) recovery from the recycled CO_2.

The EOR surface facility configuration cases evaluated were intended to bracket the expected range of field and equipment capacities and conditions that were projected for CO_2 EOR in the Illinois Basin. The conceptual facility designs included the equipment required to separate produced liquids from the CO_2, short-term storage of the produced liquids, and compression of the CO_2 to be reinjected. Cases considered ranged from CO_2 flow rates of 1,180 to 24,780 Sm^3/h (standard cubic meters per hour; 1 to 21 MMscfd [million standard cubic feet per day]), facility inlet pressures of 1,034 and 2,172 kPag (kilopascal gauge; 150 and 315 psig [pounds per square inch gauge]), and facility discharge pressures of 3,448 and 6,895 kPag (500 and 1,000 psig).

The purchased equipment costs for facilities without NGL recovery were estimated to range from approximately \$1 million for the case with a 1,180 Sm^3/h (1 MMscfd) CO_2 rate and 3,448 kPag (500 psig) discharge pressure up to approximately \$5.5 million for the case with a 24,780 Sm^3/h (21 MMscfd) CO_2 rate and 6,895 kPag (1,000 psig) discharge pressure. The estimated total fixed capital investment (FCI) for facilities that require all new infrastructure ranged from approximately \$3 million to \$16.4 million, excluding NGL recovery. The FCI is the total cost for a new facility that requires the installation of basic oil production infrastructure in addition to the EOR equipment.

Options for recovering NGLs were evaluated for feasibility for the EOR surface facility cases, with capacities in the range of 12,390 to 24,780 Sm^3/h (10.5 to 21.0 MMscfd) of CO_2, and NGL recovery from recycled CO_2 was determined to be technically feasible. Natural gas liquid recovery typically is included only for facilities having significantly higher capacities, so the costs were not evaluated for cases having less than 12,390 Sm^3/h (10.5 MMscfd) of CO_2. The total FCI was estimated to be \$6.1 million and \$10 million for NGL recovery for the 12,390 and 24,780 Sm^3/h (10.5 and 21.0 MMscfd) cases, respectively. The NGL costs are based on an idealized case assuming conventional refrigeration with a stabilizer. However, conventional refrigeration with a stabilizer most likely would not be effective for the recovery of NGL from the lean produced gas containing only 0.03 L/m^3 (0.22 GPM [gallons of recoverable hydrocarbons in the gas per thousand standard cubic feet of gas]) of recoverable liquids. The costs for the mechanical refrigeration with a stabilizer-type process are provided as a best case scenario for this evaluation, but a significantly more costly cryogenic process would be required to achieve significant NGL recovery from a gas stream this lean in NGL content.

INTRODUCTION

The Midwest Geological Sequestration Consortium (MGSC), working as one of the Regional Carbon Sequestration Partnerships for the U.S. Department of Energy, has conducted a three-phase program to demonstrate the feasibility of carbon dioxide (CO_2) capture and storage. One of the storage options involves injecting the CO_2 in mature oil fields for enhanced oil recovery (EOR). This report evaluates the design and costs for the surface processing equipment for EOR applied to mature oil fields with characteristics similar to those in the Illinois Basin (ILB), as part of the MGSC's Development Phase (Phase III).

The objective of this report is to provide information and calculation tools to determine the feasibility of implementing CO_2 EOR in the ILB. This evaluation considers the surface process equipment required to compress and dehydrate CO_2 and to separate produced oil, water, and CO_2. The costs for the CO_2 delivery pipeline, injection wells, and production wells are not included, with the exception of unit costs for piping materials that could be used for flowlines to bring produced fluids to the central facility and to deliver CO_2 from the central facility to the injection wells. Field-wide costs are also not part of this report. The process configurations and costs provided in this report are intended as examples that are representative of typical EOR surface facilities, but alternative configurations may be equally feasible or preferable.

ENHANCED OIL RECOVERY SURFACE FACILITY DESIGN BASIS

This section describes the scope of work and assumptions for the surface facility cases evaluated. The cases were intended to bracket the expected range of field and equipment capacities and conditions that could be typical for early-stage CO_2 EOR in the ILB.

Scope of Work

The scope of work for the EOR surface facility evaluation was developed jointly by the Illinois State Geological Survey (ISGS), Advanced Resources International (ARI), and Trimeric. The following list summarizes the scope of work by Trimeric, which is the subject of this report:

- Develop process requirements and configurations, and prepare process flow diagrams for typical EOR surface facilities.
- Determine what equipment is needed, and then size the equipment.
- Define why the equipment is required and discuss other conditions in which some of the equipment may be unnecessary. Develop a "minimum requirement" equipment case.
- Determine the minimum size at which natural gas liquid (NGL) separation is considered feasible and economic.
- Determine the minimum-size facility (to address a minimum-size EOR field-scale project that could be considered).
- Determine the maximum-size facility (to address the feasibility of a large facility at a single, large oil field and the possibility of a central gas-handling facility) for surrounding smaller fields.
- Prepare purchased equipment cost estimates for equipment defined per the previous items.
- Prepare installed equipment cost estimates.
- Provide information needed for any further economic analysis related to the surface processing facilities, including the following:
    - Unit operating costs
        - Electricity (kWh/unit)
        - Include an on-stream factor (percentage of time the facility is running)
        - Include a capacity factor (average percentage of the full production capacity during operations)
        - Cost of chemical treatments (corrosion inhibitors, emulsion breakers, etc.)
        - Number of operators and the cost of labor
        - Maintenance costs (spare parts)
        - Consumable costs (compressor lubrication oil, filters)
    - Annual operating costs
    - Fixed capital investment (FCI) for surface processing facilities.

Description of Cases

A list of cases was developed to cover the range of conditions (i.e., gas production rate, facility inlet pressure, facility outlet pressure, oil rates, and water rates) anticipated in the ILB EOR facilities. Table 1 presents the cases evaluated in this study. Twelve cases were selected. The ISGS provided the facility outlet (injection) pressures of 3,448 and 6,895 kPag (kilopascal gauge; 500 and 1,000 psig [pounds per square inch gauge]) based on the anticipated miscible and immiscible CO_2 flood surface and bottomhole pressure requirements.

The lowest CO_2 production rate, 1,180 Sm^3/h (standard cubic meters per hour; 1 MMscfd [million standard cubic feet per day]) at 1,034 kPag (150 psig) facility inlet pressure, was selected to match the smallest CO_2 compressor that would typically be used for CO_2 EOR. The gas is mostly CO_2 but also contains some hydrocarbons. A 1,180 Sm^3/h (1 MMscfd) compressor may be applicable for a small field with only a few wells. The 2,478 Sm^3/h (2.1 MMscfd) gas rate at 2,172 kPag (315 psig) facility inlet pressure requires approximately the same size compressor as the 1,180 Sm^3/h at 1,034 kPag (1 MMscfd at 150 psig) case because the compressor size scales most closely with the actual volumetric flow rate at the inlet. The actual volumetric flow rate is inversely proportional to changes in the absolute pressure of the feed to the compressor inlet. The temperature of the fluids entering the facility was assumed to be 37.8 °C (100 °F) in all cases. Although actual fluid temperatures coming in from the field may be lower, these facilities typically include heat integration to warm the fluids entering the facility and cool the gas leaving the CO_2 compressors. Details on fluid temperatures are not addressed in this report.

The 2,172 kPag (315 psig) inlet (wellhead) pressures were selected so that a single-stage reciprocating compressor would be adequate to provide the required discharge pressure of either 3,448 or 6,895 kPag (500 or 1,000 psig). The 1,034 kPag (150 psig) inlet pressure will require a single stage of compression to 3,448 kPag (500 psig) and two stages of compression to 6,895 kPag (1,000 psig). Options were evaluated for phasing in compressor trains as the CO_2 production rate

Table 1 Cases for the enhanced oil recovery surface facility study[1]

Case	Peak gas production rate, million Sm^3/h (MMscfd)	Peak gas production rate, m^3/min (acfm)	Peak water production rate, m^3/d (bpd)	Peak oil production rate, m^3/d (bpd)	Facility inlet pressure, kPag (psig)	Facility outlet pressure, kPag (psig)	NGL recovery?	Compression deployment phases, million Sm^3/h (MMscfd)
1	1,180 (1.0)	1.88 (66.5)	76.8 (483)	14 (91)	1,034.2 (150)	3,447.4 (500)	No	Single
2	1,180 (1.0)	1.88 (66.5)	76.8 (483)	14 (91)	1,034.2 (150)	6,895 (1,000)	No	Single
3	2,478 (2.1)	1.77 (62.5)	154 (967)	29.1 (183)	2,172 (315)	3,447.4 (500)	No	Single
4	2,478 (2.1)	1.77 (62.5)	154 (967)	29.1 (183)	2,172 (315)	6,895 (1,000)	No	Single
5	12,390 (10.5)	18.8 (665)	768 (4,830)	145 (912)	1,034.2 (150)	3,447.4 (500)	No	Three: 2,950, 4,720, 4,720 (2.5, 4.0, 4.0)
6	12,390 (10.5)	18.8 (665)	768 (4,830)	145 (912)	1,034.2 (150)	6,895 (1,000)	No	Three: 2,950, 4,720, 4,720 (2.5, 4.0, 4.0)
7	24,780 (21.0)	17.7 (625)	1,537 (9,669)	290.3 (1,826)	2,172 (315)	3,447.4 (500)	No	Three: 4,720, 9,440, 9,440 (4.0, 8.0, 8.0)
8	24,780 (21.0)	17.7 (625)	1,537 (9,669)	290.3 (1,826)	2,172 (315)	6,895 (1,000)	No	Three: 4,720, 9,440, 9,440 (4.0, 8.0, 8.0)
9	12,390 (10.5)	18.8 (665)	768 (4,830)	145 (912)	1,034.2 (150)	3,447.4 (500)	Yes	Three: 2,950, 4,720, 4,720 (2.5, 4.0, 4.0)
10	12,390 (10.5)	18.8 (665)	768 (4,830)	145 (912)	1,034.2 (150)	6,895 (1,000)	Yes	Three: 2,950, 4,720, 4,720 (2.5, 4.0, 4.0)
11	24,780 (21.0)	17.7 (625)	1,537 (9,669)	290.3 (1,826)	2,172 (315)	3,447.4 (500)	Yes	Three: 4,720, 9,440, 9,440 (4.0, 8.0, 8.0)
12	24,780 (21.0)	17.7 (625)	1,537 (9,669)	290.3 (1,826)	2,172 (315)	6,895 (1,000)	Yes	Three: 4,720, 9,440, 9,440 (4.0, 8.0, 8.0)

[1] Sm^3/h, standard cubic meters per hour; MMscfd, million standard cubic feet per day; acfm, actual cubic feet per minute; bpd, barrels per day; kPag, kilopascal gauge; psig, pounds per square inch gauge.

increased over the life of the EOR facility. This means that a small-sized compressor would initially be installed at the site to handle the low CO_2 production rate in the early years of the EOR facility. As the production rate increased over several years, somewhat larger sized compressors and other processing equipment components would be added incrementally until the peak CO_2 flow rate could be processed, compressed, and reinjected. Multiphasing offers several potential advantages, including delaying equipment costs, adjusting compressor needs based on current CO_2 production, and allowing more options for the reuse of smaller compressor trains at other facilities at the end of life of the reservoir of interest.

The economics of NGL recovery from the produced CO_2 were evaluated. Natural gas liquids were considered C_{3+} (propane and heavier) components; methane (C_1) and ethane (C_2) separation were not considered a requirement of this analysis. Natural gas liquids are usually recovered to generate additional revenue; however, in some cases, recovery is also performed to remove hydrocarbons in the produced CO_2 so that they do not adversely affect the CO_2 EOR performance upon reinjection.

Production Rate and Natural Gas Liquid Content Assumptions

The oil and water production rates provided in Table 1 were determined from the CO_2 production rate for each case and the assumed peak oil-to-gas and water-to-gas ratios. The peak production ratios were used for equipment sizing, but it was understood that the ratios could vary throughout the lifetime of the EOR operation and that the ratios would vary from field to field. On the basis of ILB CO_2 EOR reservoir simulation projections and experience from similar projects, the peak oil-to-gas ratio was assumed to be 0.00049 m^3/m^3 (0.087 bbl/Mscf [barrels/thousand standard cubic feet]) and the peak water-to-gas ratio was assumed to be 0.00259 m^3/m^3 (0.461 bbl/Mscf).

Generally speaking, projections of the NGL content in produced gas from CO_2 EOR of ILB oil reservoirs were rather lean compared with that typical of produced natural gas treated for NGL recovery. This was not unexpected for a typical oil field that has been producing with waterflood conditions for many decades because a relatively large fraction of the light hydrocarbons already would have been stripped out of the oil. Additionally, ILB crude oil typically has very low associated gas.

The NGL content in gases is typically characterized in terms of the gallons of recoverable hydrocarbons in the gas per thousand standard cubic feet of gas (GPM). Rich gases usually have values exceeding 0.8 L/m^3 (6 GPM), whereas lean gases have less than 0.2 L/m^3 (1.5 GPM), with a midpoint around 0.4 L/m^3 (3 GPM). It was assumed, based on ILB CO_2 EOR reservoir simulation projections, that the produced gas from the CO_2 EOR would have an NGL content of only 0.03 L/m^3 (0.22 GPM) at the peak of the NGL potential production rate. The characteristics of the produced gas NGLs used for the analysis are summarized in Table 2, with the balance being CO_2.

Process Configurations

A typical EOR surface facility has three primary functions:

1. To separate the produced gas (primarily CO_2 and hydrocarbons) from the produced liquids.
2. To compress the produced gas for reinjection or for distribution in a pipeline.
   a. To remove hydrocarbons to generate revenue or, if necessary, for efficient compression and subsurface operations, depending on the hydrocarbon composition and concentration in the CO_2.
   b. To dehydrate the recycle gas, if necessary, to meet site-specific requirements for reinjection or pipeline specifications for CO_2.
3. To separate produced water and oil, with short-term storage of each of the liquids.
   a. To capture or treat low-pressure gas, if necessary, from flashing gas from liquids during the pressure let-down steps.
   b. To apply chemical treatment to break the oil–water emulsion for improved liquids separation. (Heating instead of or in addition to chemical treatment is used to separate oil and water at some EOR facilities.)

The equipment required to accomplish these three primary surface facility functions varies depending on the properties of the inlet gas, such as pressure in this evaluation as well as composition in other applications, the required gas discharge pressure, and the flow rates of the inlet gas, oil, and water streams. Process flow diagrams for the 12 cases listed in Table 1 are provided in Figures A1–A12 in Appendix A, and the equipment design and cost estimates are described in detail in the following section. Individual facility component costs are listed in tables later in the report so that the impact of removing or adding a particular component on the overall cost of the facility can be evaluated.

Table 2 Assumed peak natural gas liquid (NGL) characteristics of produced gas[1]

Component	Value
Methane + ethane	1.5 mol. %
NGLs	0.7 mol. %
NGLs	0.03 L/m^3 (0.22 GPM)

[1]NGLs refer to propane and heavier hydrocarbons. The NGL content in gases is typically characterized in terms of the gallons of recoverable hydrocarbons in the gas per thousand standard cubic feet of gas (GPM).

EQUIPMENT DESIGN AND COST ANALYSIS SUMMARY

This section describes the general approach used to size and select the surface equipment for the EOR facilities. Included here are the equipment capital costs and the anticipated fluid processing rates for the plants. The economic results from the study are also presented.

Equipment Sizing

The surface equipment for the EOR facilities was sized using different methods, depending on the type of equipment. This section discusses those methods and presents other important design criteria that could potentially affect the cost of the equipment.

Separators

Various separators are used in the EOR surface equipment. The separator types can be described briefly as follows:

- *Slug catcher*. This vessel is used to separate the produced gas from the oil and water from the wells at the inlet to the facility. The gas exits the top of the vessel and flows to the compressor train, whereas the oil and water exit in a combined stream to downstream separation vessels. The slug catcher operates at approximately the pressure of the wellhead (1,034 and 2,172 kPag [150 and 315 psig], depending on the case). The slug catcher is typically a horizontal vessel.
- *Free water knockout*. This horizontal vessel is used to separate the bulk of the water from the oil. The vessel operates at a low pressure of approximately 172 kPag (~25 psig), and some dissolved CO_2 will evolve and be sent to the low-pressure suction scrubber. The free water knockout is typically a horizontal vessel. An example photograph of a horizontal free water knock-out is shown in Figure 1. In many parts of the United States, heat from burning natural gas or electricity, sometimes transferred to the free water knockout via an intermediate heat transfer fluid, is used in the free water knockout to help separate the oil from the water. However, according to a discussion between Trimeric and Ken Hake of Baker-Hughes (personal communication, July 15, 2015), the separation of oil and water by chemical addition is the most common approach in the ILB, and is the one assumed in this report.
- *Demulsifier*. In this vessel, chemicals are added to break any oil–water emulsions to further separate the water and oil. A small amount of CO_2 may evolve from the liquids, and this gas is also sent to the low-pressure suction scrubber. A pressure drop of 6.9 kPa (1 psi) was assumed while transferring the liquids from the free water knockout to the demulsifier. The demulsifier vessel typically has a horizontal orientation, and it is similar in appearance to the horizontal free water knockout vessel shown in Figure 1. Sometimes heat is applied for this type of separation (i.e., heater-treater vessels) when fuel gas, electricity, or some form of waste heat input is available, but discussions with oilfield operators in the ILB suggest that a chemical separation approach is used almost exclusively in ILB oil production facilities.
- *High-pressure suction scrubber*. This vertical vessel is used to prevent any liquids from entering the high-pressure compressor train. Figure 2 shows an example of a vertical vessel used as a compressor suction scrubber and the compressor itself. The suction scrubber is used (1) to remove liquids that may condense in the line coming from the top of the slug catcher and (2) to remove slugs of liquid that could carry over from the slug catcher if unexpectedly high fluid volumes come to the facility. The high-pressure suction scrubber will operate at an inlet gas design pressure of either 1,034 or 2,172 kPag (150 or 315 psig), depending on the case.
- *Low-pressure suction scrubber*. This vertical vessel is used to prevent any liquids from entering the low-pressure compressor train. The low-pressure compressor train is typically added at an EOR facility when enough flash gas is present to justify the cost of the low-pressure train, which is needed to feed these gases to the suction of the high-pressure compression system. The low-pressure suction scrubber operates at a low pressure of 165 kPag (~24 psig).

The vessels were sized using an approach in the literature for three-phase separators (Monnery and Svrcek 1994) and with WinSim Design II software. The estimated size of the separators is preliminary for a feasibility study such as this. Consultation with vendors that specialize in separation equipment would be required for a more detailed design for an actual plant. The separator lengths shown in the tables in this report are from seam to seam (S/S). The cost estimate for the vessels was based on cladded carbon steel to approximate the cost of commercially applied corrosion-resistant coatings on the inside of carbon steel vessels in wet CO_2 service. Cladded or coated carbon steel vessels have a cost between that of regular carbon steel vessels and that of stainless steel vessels. It should also be noted that, for the multiphasing cases with multiple compressors (Cases 5–12), a separate high-pressure suction scrubber was assumed for each compressor train.

Chemical Injection System

Chemicals are added to the inlet of the slug catcher to break any oil–water emulsions and further remove water from the oil. In fact, according to discussions between Trimeric and Ken Hake of Baker-Hughes (personal communication, July 15, 2015), the chemicals might be added further upstream of the facilities discussed in this report to allow them more contact time to mix with the produced fluids. The demulsifier chemical will be added to give 1,000 ppmv of demulsifier concentration in the oil volume, and the water will then be removed from the oil in the demulsifier vessel. The demulsifier concentration used was based on experience with other emulsion-breaking projects. The demulsifier injection pump was sized to transfer the appropriate amount of chemical for each case; the same size chemical injection pump can be used for all the cases because the required flow rates are very low. The demulsifier storage tank was sized to hold a 14-d supply of demulsifier chemical.

Oil Storage Tank

Oil production was assumed to start out initially at low rates, peak, and then steadily decrease until the end of life for

Figure 1 Typical horizontal free water knockout vessel. The horizontal slug catcher and demulsifier vessels are similar in appearance. Photograph courtesy of Denbury Onshore.

Figure 2 Typical vertical vessel high-pressure suction scrubber and compressor. Typical low-pressure suction scrubbers are similar in appearance. Photograph courtesy of Denbury Onshore.

the field. The oil would be stored in tanks until it could be trucked off-site. For the smaller cases (Cases 1–4), a 7-d holding capacity was assumed for peak oil production. For the larger cases (Cases 5–12), a 3-d storage capacity was assumed because the 7-d size and cost of the tanks was unreasonably large. The infrastructure, including roads, oil pipelines, or both, was assumed to be more developed for the larger cases so that a longer duration would not be required to account for inaccessibility during inclement weather. For the purposes of cost estimation, the oil tanks for Cases 5–12 were assumed to be American Petroleum Institute-style steel tanks. The oil tanks for Cases 1–4 were assumed to be portable 33 m^3 (210-barrel) fiberglass tanks. Pipelines are often used for oil product delivery instead of tanks and trucks as the facility size increases beyond the cases considered in this study.

Water Storage Tank

The same holding capacities were assumed for water storage tanks as discussed in the Oil Storage Tank section. The 7-d or 3-d holding capacity for water may not be required if other disposal methods are available (e.g., a disposal well) or used in a waterflood field, so the water storage capacity may be conservatively high for some locations. For the purposes of cost estimation, the water tanks were assumed to be American Petroleum Institute-style tanks. For facilities larger than those considered in this study, pumps are usually installed for on-site water disposal via injection into an underground formation. The use of injection pumps for water disposal reduces the amount of water storage tank volume needed for these larger facilities. Generally, water production decreases during CO_2 EOR, so fields with an existing waterflood may not need new water storage or handling facilities.

CO_2 Compressor Trains

The high-pressure and low-pressure CO_2 compressor trains were modeled with WinSim Design II software using the Peng-Robinson equation of state. The number of stages in the compressor train was determined from the specified inlet and outlet pressures and the need to limit the compressor discharge temperature from each stage to 149 °C (300 °F) or less because of construction material constraints. Interstage cooling was assumed to be 48.9 °C (120 °F), which is a typical design condition for air coolers. A polytropic efficiency of 79.5% was used for CO_2 compression. The horsepower requirement for the CO_2 compressor train design was obtained from the simulation. The construction material for components on the suction side of the compressor cylinders was assumed to be a combination of cladded or coated carbon steel and solid stainless steel. Coated carbon steel or stainless steel is typically used on the suction side, where the gas is cold and water can condense. Carbon steel is used on the discharge side of the compressor cylinders because the discharge is hot, at 149 °C (~300 °F), and therefore well above the water dew point during normal operation. The gas flow and discharge pressure requirements warranted reciprocating compressors for all cases per the guidelines provided by the Gas Processors Suppliers Association (GPSA 2004a). Other types of compressors, such as in-line centrifugal and integral-gear centrifugal compressors, have been used for CO_2 compression in some cases, but the costs are generally comparable to the reciprocating compressor costs for a study at this level of accuracy; reciprocating compressors are commonly used for CO_2 EOR applications.

It was assumed that low-pressure compression trains would be used to send flash gases from the free water knockout and demulsifier to the suction of the high-pressure compression train when the low-pressure flash gas flow rate reached 1,180 Sm^3/h (1 MMscfd), that is, in Cases 7, 8, 11, and 12. At flow rates less than this, the flash gas typically would be vented or sent to a flare.

Cases 5–12 in Table 1 included three phases (multiphasing) of the CO_2 compression train as the gas production rate increased over time. Multiphasing means installing smaller compressors over a longer time period as the CO_2 rate returning from the oil field increases instead of installing one larger compressor at the beginning of the CO_2 EOR operation. If the compressor train is installed with a discharge-to-suction recycle capability, it can compress gas at flow rates as low as approximately 25% of the design gas flow rate. Variable-volume clearance pockets, cylinder head unloading mechanisms, and variable-frequency drives (primarily for smaller units) can also be used to reduce the throughput in these types of reciprocating compressors. The compressor operating costs were based on the peak product throughput; however, the energy efficiency may be lower when the compressors are not fully loaded.

Installation of the CO_2 compression trains was assumed to take place in three phases. Many companies that operate CO_2 EOR facilities in the United States elect to defer the relatively high capital cost of compression equipment purchases until such time as the amount of CO_2 returning with the produced oil and water requires additional CO_2 compression equipment capacity. A small compressor train would first be installed to handle the low gas production rate during initial CO_2 injection. Then, as the produced CO_2 flow rate increased, a somewhat larger second compressor train would be installed, and several years later, as the produced CO_2 flow rate increased further, a third compressor train would be purchased. In Cases 5, 6, 9, and 10, the compressors were sized at 2,950, 4,720, and 4,720 Sm^3/h (2.5, 4.0, and 4.0 MMscfd) each. In Cases 7, 8, 11, and 12, the compressors were sized at 4,720, 9,440, and 9,440 Sm^3/h (4.0, 8.0, and 8.0 MMscfd) for each of the three phases, respectively. A net present value financial analysis was outside the scope of this study, so the economic impact of phasing the compressor installation on a net present value basis was not evaluated.

Dehydration

Costs were included for dehydration of the compressed CO_2 before reinjection. Dehydration would likely be needed if the added costs to use corrosion-resistant materials downstream of the compressors offset the cost of dehydration or if the CO_2 had to go through a common carrier pipeline before it was reused. Without dehydration, the CO_2 from the compressor train could be saturated with water. The temperature of the CO_2 may cool as the gas flows through the underground injection well piping. Water could condense and cause increased corrosion. The injection pressures anticipated for ILB EOR facilities (3,448–6,895 kPag [500–1,000 psig]) are too low to take advantage of the increased water-holding capacity of CO_2 at pressures exceeding 6,895 kPag

(1,000 psig). The possibility of forming CO_2–water solid hydrates may also be an issue that requires the dehydration of CO_2.

As a simplification, it was assumed in all cases that dehydration would take place at the discharge of the compressor train at high pressure. Triethylene glycol (TEG) dehydration could be used for the 3,448 kPag (500 psig) cases; however, TEG losses into the CO_2 stream at 6,895 kPag (1,000 psig) might begin to become detrimental such that glycerol might be required instead. Alternatively, TEG dehydration could be performed between the first and second stages of compression in the cases that had two-stage compressors. In any case, these detailed design decisions are unlikely to affect the cost estimates provided in this early-stage conceptual evaluation. The cost of dehydration is shown separately in Appendix B: Equipment List and Purchased and Installed Costs for Cases 1–12 to show the cost impact of this unit operation and to facilitate the removal of these costs if dehydration is not required.

Buildings

Buildings to house compressors, controls, chemicals, and maintenance equipment were included in the EOR facility. The estimated cost of the building(s) was extrapolated from other recent projects of similar size.

Natural Gas Liquid Recovery

Natural gas liquid recovery was considered in Cases 9–12. Appendix C of this report contains additional details regarding the NGL recovery evaluation. A straight refrigeration-type NGL system with stabilizer was assumed. A stabilizer is a distillation column intended to remove a relatively small amount of light hydrocarbons from the product stream. The assumed NGL recovery was 52%, which is typical for the total C_{3+} components in a straight refrigeration system (Vargas 2010). Peak NGL production occurs at approximately 58% of the peak gas flow rate. The NGL recovery rate was estimated to be constant at 2.6 m^3/d (16.6 bbl/d) for Cases 9 and at 10 and 5.3 m^3/d (33 bbl/d) for Cases 11 and 12 over the 20-yr period of the CO_2 EOR operation.

Trimeric assumed that the recovered NGLs would be trucked off-site for fractionation. On-site fractionation would require the installation of separate fractionation columns to recover ethane, propane, butane, and heavier components at a prohibitive capital expense and utility consumption for reboiler heat and overhead condensation, given the size of the facilities evaluated in this study. If (1) the expected NGL content of the produced gas is higher than assumed for this analysis, (2) more wells are combined to give a higher CO_2 produced gas rate, or (3) an off-site fractionation facility is not within close proximity to the EOR plant, then on-site fractionation may be warranted.

Capital Costs

This section describes the approach used to estimate the purchased and installed costs for the EOR facilities evaluated in this study. The purchased equipment costs were obtained from a combination of vendor quotes and costing software. The In-Plant Cost Estimator software package from AspenTech was used to estimate the purchased equipment costs for some of the chemical process equipment. The In-Plant Cost Estimator costs are from the first quarter of 2011. The purchased costs were adjusted to a May 2012 cost basis (the most recent index available at the time of the initial evaluation for this report) by using published plant cost indices found in the September 2012 issue of the magazine *Chemical Engineering*. The list below shows the source of the purchased equipment costs by equipment type:

- Separators (slug catcher, free water knockout, demulsifier, high- and low-pressure suction scrubbers)—In-Plant Cost Estimator
- Chemical injection pump—In-Plant Cost Estimator
- Chemical injection tank—in-house vendor data for ILB equipment for the small CO_2 flow cases (Cases 1–4) and In-Plant Cost Estimator for all other cases (Monnery and Svrcek 1994)
- Water storage tanks—In-Plant Cost Estimator
- CO_2 compressor trains—In-Plant Cost Estimator (which correlated well with recent actual reciprocating compressor cost data)
- Dehydration—scaled from other CO_2 projects in 2012
- Building—scaled from other CO_2 projects in 2012
- Natural gas liquid recovery—literature data for straight refrigeration type and adjusted for CO_2 operation instead of natural gas (see the Natural Gas Liquid Recovery section; Tannehill 2009)

The installation costs for purchased equipment were estimated using typical factors as a percentage of the purchased equipment cost (Morris and Williams 2001). The sum of the purchased equipment cost and the installation cost is the installed equipment cost. Tables B1–B12 in Appendix B show the equipment sizes and the estimated purchased and installed equipment costs for the 12 cases. Purchased equipment and installed costs for the NGL recovery equipment are not shown in the tables in Appendix B for Cases 9–12 because the reference source reported costs on a fixed capital basis (Tannehill 2009). For this reason, the NGL recovery costs are described in the Fixed Capital Investment section, where the fixed capital costs are summarized for all the cases evaluated in this study.

The total installed costs listed in Tables B1–B12 represent the estimated cost for installing and connecting the particular piece of equipment in an existing facility that already has a basic infrastructure in place (e.g., electrical power, roads, and prepared plot areas). The FCI estimates provided in the Fixed Capital Investment section represent the total costs for a new facility that requires the installation of basic infrastructure in addition to the EOR equipment.

Operating Cost Information

Operating cost information for the 12 cases is provided in Table 3. The information is separated into two categories: variable costs (with the capacity utilization factor) and fixed costs. The operating cost information will be combined with the field-wide operating costs developed by others, so only the anticipated operating rates are shown and not the dollar value of the annual expenses.

As shown in Table 3, a capacity utilization factor of 95% was assumed for the variable costs. The capacity utilization factor takes into account both the on-stream factor, which is the total percentage of time the facility is operating, and the capacity factor, which is the average percentage of the production rate compared with the design production rate. The 95% value was based on data collected by Charles Monson at the ISGS for several facilities in the ILB (Monson 2012). The electricity usage for the major equipment is also shown. Compression power ranged from 85% to 95% of the total electricity demand at the EOR facility. The compression power for cases that also have a low-pressure compression train (Cases 7, 8, 11, and 12) includes the power required for both the high-pressure and low-pressure compression trains. The peak water rate is shown so that disposal costs for on-site or off-site disposal can be estimated. The peak oil rate is given to facilitate the estimation of transportation fees. The total dehydration and NGL recovery operating costs are included so that the operating expenses can be estimated for the entire EOR system when these processes are required for a given facility. The NGL recovery rates in barrels per day (bpd) are provided so that revenue from this product can be estimated. However, it should be noted that the assumed NGL content in the gas is rather lean for NGL recovery purposes (see the Enhanced Oil Recovery Surface Facility Design Basis and the Effect of Natural Gas Liquid Recovery sections).

The fixed costs include an estimate of the number of operators required to run the facility as well as an estimate of the supervisor labor (assumed to be 20% of the operating labor costs). Maintenance expenses are estimated at $40/(hp-yr) based on experience with these types of compressor facilities. The plant operating overhead is assumed to be 75% of the operating and supervisor cost (typical factor). The fixed costs do not include the capacity utilization factor.

Fixed Capital Investment

The purchased equipment costs for the EOR facility (except the NGL portion) were multiplied by a factor of 3 to arrive at the FCI cost. This factor accounts for the costs of items such as purchased equipment installation, instrumentation and controls, piping, electrical systems, engineering and supervision, construction expenses, contractors' fees, and contingency. A multiplier of 3 with the purchased equipment costs is a typical value used for a mix of vendor-provided skid-mounted equipment, on-site assembly (separators, tanks, etc.), and fabrication of interconnecting piping.

As noted in the Capital Costs section, the FCI represents the total cost for a new facility that requires the installation of all basic infrastructure in addition to the EOR equipment. Table 4 summarizes the estimated FCI for the surface equipment for all 12 cases and for NGL recovery for Cases 9–12. Several observations from these cost data are described below.

General Fixed Capital Investment Cost Relationship

As shown in Table 4, the FCI ranged from $2.94 million for Case 1 with only 1,180 Sm^3/h (1 MMscfd) of produced gas to as high as $16.4 million for Cases 8 and 12 with 24,780 Sm^3/h (21 MMscfd) of produced gas (not including NGL recovery).

Figure 3 shows a graph of the FCI over the range of CO_2 flow rates considered in this study, with a trend line and equation so that the FCI can be estimated for other intermediate CO_2 rates. This trend is valid for the water-to-gas, oil-to-gas, and CO_2 compression ratios used in this study. It excludes the cost for NGL recovery.

Effect of Facility Outlet Pressure

A comparison of pairs of cases that differed only in the facility outlet pressures (Cases 1 vs. 2, 3 vs. 4, 5 vs. 6, and 7 vs. 8) shows the impact of compressor discharge pressure on the FCI (Figure 4). The facility outlet pressure was either 3,448 or 6,895 kPag (500 or 1,000 psig), whereas the facility inlet pressure, gas production rate, and peak oil and water production rates were the same for each pair of cases. As shown in Figure 4, the FCI increased in range from 5.7% to 9.3% when the facility outlet pressure increased from 3,448 to 6,895 kPag (500 to 1,000 psig). These percentage increases were calculated by subtracting the FCI for the lower discharge pressure case from the FCI for the higher discharge pressure case and dividing this difference by the FCI for the lower discharge pressure case. The percentage increase is larger for the cases with higher CO_2 flow rates (Cases 5–8) because more gas is compressed to the higher pressure. The cases in this comparison do not have NGL recovery and therefore do not include NGL recovery costs.

Effect of Facility Inlet Pressure

The differences in costs for Cases 5 versus 7 and 6 versus 8 indicate the impact of facility inlet pressure, which is assumed to be the pressure of the produced fluids at the wellhead, on the purchased compressor costs. Only the high-pressure compressor cost is evaluated in this analysis.

Other equipment costs are not included in this comparison because the inlet pressure also affects the costs of other equipment that are not fixed between the cases compared. Furthermore, comparing the total facility FCI, which includes the other equipment in this case, would skew the comparison of the compressor costs versus inlet pressure because changes in the other equipment costs (i.e., separator costs) are related to the inlet pressure (design pressure).

As shown in Table 5, the inlet wellhead pressure affects the compressor cost in two ways: the pressure differential (or more rigorously, the pressure ratio) across the compressor and the mass flow rate. For Cases 5 and 7, the pressure differential is 2,413 and 1,276 kPag (350 and 185 psig), respectively, so the pressure ratio for Case 7 is only 50% that of Case 5. The mass flow rate for Case 7 (43,546 kg/h [96,000 lb/h]) is nearly twice that of Case 5 (22,680 kg/h [50,000 lb/h]). The larger pressure ratio results in a higher compressor power and cost for Case 5 than for Case 7, with the lower pressure ratio, yet higher than the mass flow rate of gas. However, in the comparison with Cases 6 and 8, the pressure ratio across the compressor between the two cases is closer, 5,861 and 4,723 kPag (850 and 685 psig), respectively, because the discharge pressure is higher (6,895 kPag [1,000 psig]). The trade-off between the lower pressure ratio and higher mass flow rate for Case 8 results in a higher power requirement for Case 8 than for Case 6, as shown in Table 5.

Table 3 Operating cost summary

Operating cost information	Unit	Case 1	2	3	4	5	6	7	8	9	10	11	12
Variable costs (includes capacity utilization factor)													
Capacity utilization factor	%	95	95	95	95	95	95	95	95	95	95	95	95
Electricity usage	kW	55	81	37	101	557	843	480	1,114	557	843	480	1,114
Motor efficiency	%	95	95	95	95	95	95	95	95	95	95	95	95
Annual electricity cost	$/yr	45,300	67,700	30,600	83,500	462,100	699,700	398,100	924,200	462,100	699,700	398,100	924,200
Chemical injection	$/gal	10	10	10	10	10	10	10	10	10	10	10	10
Chemical injection rate	gal/day	4	4	8	8	40	40	80	80	40	40	80	80
Annual chemical injection cost	$/yr	14,600	14,600	29,100	29,100	145,600	145,600	291,300	291,300	145,600	145,600	291,300	291,300
Wastewater disposal	$/bbl	1	1	1	1	1	1	1	1	1	1	1	1
Water disposal rate	bbd	459	459	919	919	4,589	4,589	9,186	9,186	4,589	4,589	9,186	9,186
Annual wastewater cost	$/yr	167,500	167,500	335,300	335,300	1,675,000	1,675,000	3,352,900	3,352,900	1,675,000	1,675,000	3,352,900	3,352,900
Dehydration expenses	$/yr	23,200	18,600	35,400	28,300	113,400	90,900	238,300	190,600	113,400	90,900	238,300	190,600
NGL recovery expenses	$/yr	0	0	0	0	0	0	0	0	109,300	109,300	218,500	218,500
NGLs recovered	bpd	0	0	0	0	0	0	0	0	16	16	31	31
Total variable operating costs	$/yr	250,600	268,400	430,400	476,200	2,396,100	2,611,200	4,280,600	4,759,000	2,505,400	2,720,500	4,499,100	4,977,500
Fixed costs													
Operating labor	Full time equivalent	1	1	1	1	1.5	1.5	2	2	1.5	1.5	2	2
Cost of labor	$/h	66	66	66	66	66	66	66	66	66	66	66	66
Operating labor cost	$/yr	164,800	164,800	164,800	164,800	247,200	247,200	329,500	329,500	247,200	247,200	329,500	329,500
Supervisor labor	% of operating labor cost	20	20	20	20	20	20	20	20	20	20	20	20
Supervisor labor cost	$/yr	33,000	33,000	33,000	33,000	49,500	49,500	65,900	65,900	49,500	49,500	65,900	65,900
Compressor maintenance cost factor	$/(hp-yr)	40	40	40	40	40	40	40	40	40	40	40	40
Compressor horsepower	hp	76	114	51	141	785	1,189	676	1,571	785	1,189	676	1,571
Annual compressor maintenance cost	$/yr	3,100	4,600	2,100	5,700	31,400	47,600	27,100	62,900	31,400	47,600	27,100	62,900
Plant operating overhead	% of operating + supervisor cost	75	75	75	75	75	75	75	75	75	75	75	75
Plant operating overhead cost	$/yr	148,400	148,400	148,400	148,400	222,500	222,500	296,600	296,600	222,500	222,500	296,600	296,600
Total fixed operating costs	$/yr	349,300	350,800	348,300	351,900	550,600	566,800	719,100	754,900	550,600	566,800	719,100	754,900
Total operating costs	$/yr	599,900	619,200	778,700	828,100	2,946,700	3,178,000	4,999,700	5,513,900	3,056,000	3,287,300	5,218,200	5,732,400

Table 4 Summary of the total fixed capital investment[1]

		Case											
Parameter	Unit	1	2	3	4	5	6	7	8	9	10	11	12
Actual gas flow	acfm	66.5	66.5	62.5	62.5	665	665	625	625	665	665	625	625
Produced gas flow	MMscfd	1.0	1.0	2.1	2.1	10.5	10.5	21.0	21.0	10.5	10.5	21.0	21.0
Peak water rate	bpd	483	483	967	967	4,830	4,830	9,669	9,669	4,830	4,830	9,669	9,669
Peak oil rate	bpd	91	91	183	183	912	912	1,826	1,826	912	912	1,826	1,826
Inlet pressure	psig	150	150	315	315	150	150	315	315	150	150	315	315
Discharge pressure	psig	500	1,000	500	1,000	500	1,000	500	1,000	500	1,000	500	1,000
Natural gas liquid (NGL) recovery		No	No	No	No	No	No	No	No	Yes	Yes	Yes	Yes
Installation phase(s)		Single	Single	Single	Single	Multiple	Multiple	Multiple	Multiple	Multiple	Multiple	Multiple	Multiple
Equipment cost information													
Purchased equipment costs, May 2012 basis													
Separators and chemical injection	$	152,000	152,000	191,000	191,000	320,000	320,000	581,000	581,000	320,000	320,000	581,000	581,000
Oil tanks	$	17,000	17,000	34,000	34,000	126,000	126,000	253,000	253,000	126,000	126,000	253,000	253,000
Water tanks	$	126,000	126,000	253,000	253,000	385,000	385,000	712,000	712,000	385,000	385,000	712,000	712,000
Compression	$	453,000	532,000	421,000	528,000	1,838,000	2,213,000	2,247,000	2,849,000	1,838,000	2,213,000	2,247,000	2,849,000
Dehydration	$	114,000	91,000	173,000	139,000	455,000	364,000	689,000	551,000	455,000	364,000	689,000	551,000
Building	$	121,000	121,000	142,000	142,000	309,800	309,800	520,000	520,000	309,800	309,800	520,000	520,000
Total purchased equipment costs (PEC)	$	983,000	1,039,000	1,214,000	1,287,000	3,434,000	3,718,000	5,002,000	5,466,000	3,434,000	3,718,000	5,002,000	5,466,000
Total installed equipment cost (TIC)	$	1,504,000	1,579,000	1,954,000	2,055,000	5,383,000	5,787,000	7,949,000	8,609,000	5,383,000	5,787,000	7,949,000	8,609,000
Capital cost information													
Factor for estimating the fixed capital investment (FCI) for the plant from the PEC		3	3	3	3	3	3	3	3	3	3	3	3
FCI excluding NGL recovery	$	2,949,000	3,117,000	3,642,000	3,861,000	10,302,000	11,154,000	15,006,000	16,398,000	10,302,000	11,154,000	15,006,000	16,398,000
FCI for NGL recovery	$	0	0	0	0	0	0	0	0	6,100,000	6,100,000	10,049,000	10,049,000
Total FCI	$	2,949,000	3,117,000	3,642,000	3,861,000	10,302,000	11,154,000	15,006,000	16,398,000	16,402,000	17,254,000	25,055,000	26,447,000

[1]acfm, actual cubic feet per minute; MMscfd, million standard cubic feet per day, bpd, barrels per day; psig, pounds per square inch gauge.

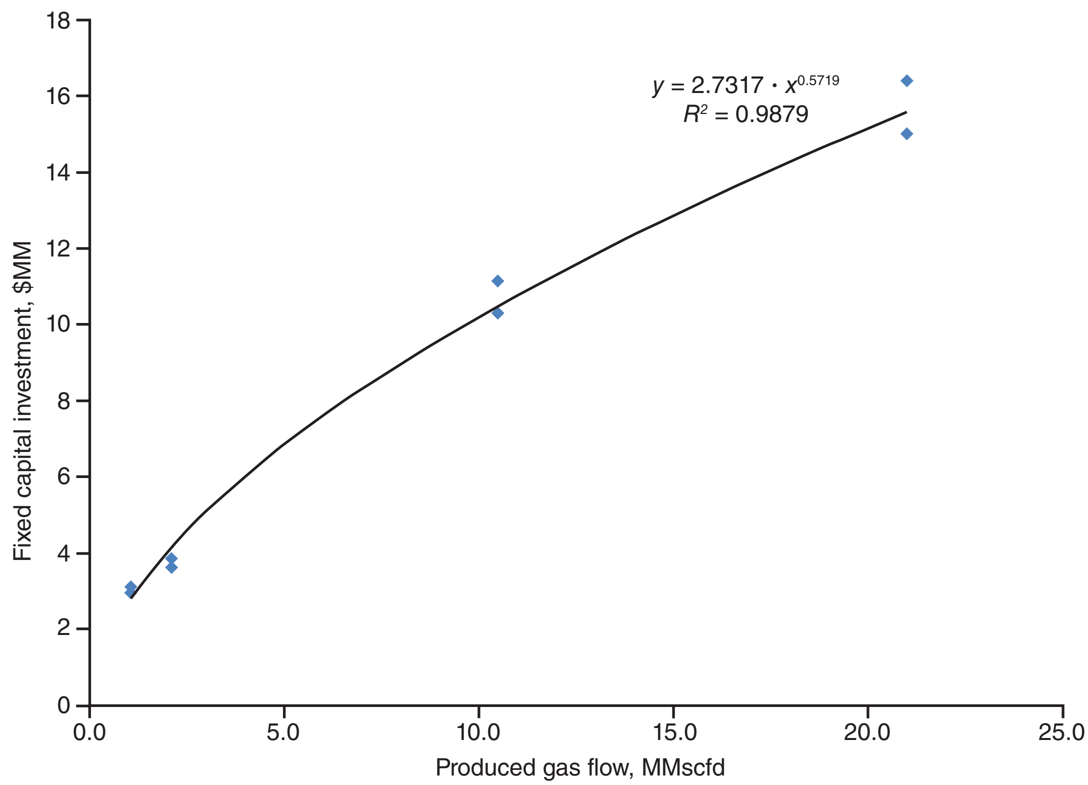

Figure 3 Fixed capital investment as a function of the produced gas flow. MM, million; MMscfd, million standard cubic feet per day.

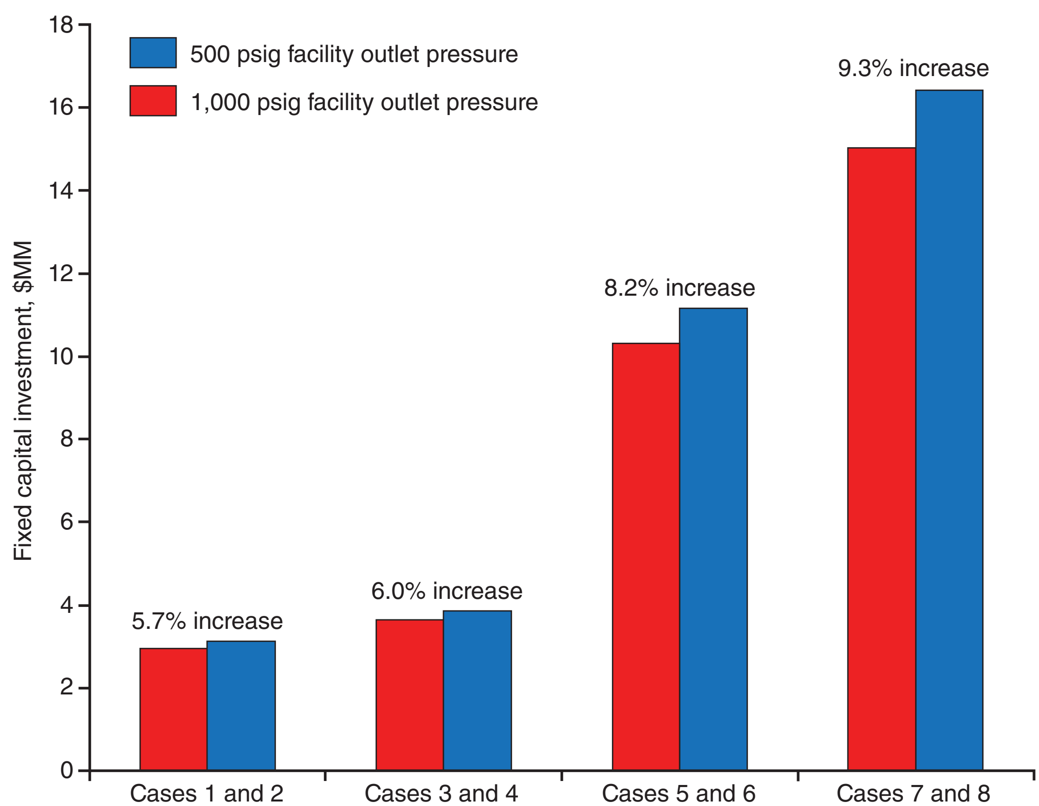

Figure 4 Effect of facility outlet pressure on the fixed capital investment. MM, million; psig, pounds per square inch gauge.

Table 5 Effect of facility inlet pressure on the high-pressure compressor purchased cost

		Discharge pressure comparison			
		3,448 kPag (500 psig)		6,895 kPag (1,000 psig)	
Parameter	Unit	Case 5	Case 7	Case 6	Case 8
Facility inlet pressure	kPag (psig)	1,034 (150)	2,172 (315)	1,034 (150)	2,172 (315)
Facility outlet pressure	kPag (psig)	3,448 (500)	3,448 (500)	6,895 (1,000)	6,895 (1,000)
Pressure differential	kPa (psi)	2,413 (350)	1,276 (185)	5,861 (850)	4,723 (685)
Pressure ratio	dimensionless	3.1	1.6	6.1	3.1
Mass flow rate	kg/h (lb/h)	22,680 (50,000)	43,545 (96,000)	22,680 (50,000)	43,545 (96,000)
Suction actual volumetric flow rate	m^3 (acfm)	18.8 (665)	17.7 (625)	18.8 (665)	17.7 (625)
High-pressure CO_2 compressor power	hp (kW)	785 (586)	512 (382)	1,189 (887)	1,407 (1,050)
Number of compression stages	—	1	1	2	1
High-pressure CO_2 compressor purchased equipment cost	$MM	1.84	1.65	2.21	2.25

[1]kPag, kilopascal gauge; psig, pounds per square inch gauge; kPa, kilopascal; psi, actual pounds per square inch; acfm, actual cubic feet per minute; MM, million.

The cost of the compressor trains for Cases 6 and 8 is similar because the second compression stage required for Case 6 offsets the increased mass flow rate for Case 8, with only a single stage. Similar trends are observed for Cases 1 and 3 and Cases 2 and 4 (not shown).

In this evaluation, no simple correlation was found for the impact of either suction pressure or mass flow rate on the cost of the high-pressure compressor. The compressor mechanical equipment size (and cost) was affected by the suction actual volumetric flow rate. The motor power requirement (and cost) was affected by both the pressure ratio and the mass flow rate.

Effect of Natural Gas Liquid Recovery

Natural gas liquid recovery was included in the economics for Cases 9 through 12. The FCI of NGL recovery was 35% to 40% of the total FCI for the overall EOR surface facility (including NGL recovery). Natural gas liquid recovery is more difficult and expensive with CO_2 applications than is recovery from natural gas because the CO_2 tends to condense with the NGLs at the low temperatures required for the recovery processes. Assuming a typical unit value of $60/bbl for NGLs, the total value of recoverable NGLs in the gas at a peak production rate is $326,000/yr for Cases 9 and 10 and $651,000/yr for Cases 11 and 12, based on a 95% capacity utilization factor. At the time this report was written, spot market NGL prices in dollars per oilfield barrel ($/bbl) were approximately 60% of the West Texas Intermediate price in dollars per oilfield barrel ($/bbl) for crude oil. When the NGL operating expenses are included, as summarized in Table 3, the net annual NGL revenue is $216,700/yr for Cases 9 and 10 and $432,500/yr for Cases 11 and 12 at peak production rates.

On the basis of the FCI values shown in Table 4, the simple payback period for the NGL recovery unit would be approximately 28 yr in Cases 9 and 10 and approximately 23 yr in Cases 11 and 12, which is longer than the expected life of the EOR facility. The actual payback time would most likely be longer because (1) the most favorable capital and operating costs were used for the NGL recovery system, (2) a more expensive cryogenic refrigeration process would most likely be required for significant NGL recovery from a gas stream with an NGL content as low as the gas composition used in this report (see Table 2), and (3) the estimated NGL revenue was based on peak production, but the average revenue over the EOR life span would be lower than the peak revenue. The economics do not appear favorable because the assumed NGL content of the produced gas is rather low. Typically, 0.13–0.2 L/m^3 (1.0–1.5 GPM) of recoverable hydrocarbon NGL per 28.32 m^3 (1,000 scf) of gas is considered the breakpoint for economic recovery. Conditions that would make NGL

Table 6 Comparison of single- and multiphasing purchased compressor costs[1]

Case	Multiphase gas flow capacity per compressor, Sm³/h (MMscfd)	Single-phase gas flow capacity for one compressor, Sm³/h (MMscfd)	Wellhead pressure, kPag (psig)	Injection pressure, kPag (psig)	Multiphase purchased compression, $	Single-phase purchased compression, $
5	2,950, 4,720, 4,720 (2.5, 4.0, 4.0)	12,390 (10.5)	1,034 (150)	3,448 (500)	1,838,000	909,000
6	2,950, 4,720, 4,720 (2.5, 4.0, 4.0)	12,390 (10.5)	1,034 (150)	6,895 (1,000)	2,213,000	1,029,000
7	4,720, 9,440, 9,440 (4.0, 8.0, 8.0)	24,780 (21)	2,172 (315)	3,448 (500)	1,645,000	787,000
8	4,720, 9,440, 9,440 (4.0, 8.0, 8.0)	24,780 (21)	2,172 (315)	6,895 (1,000)	2,248,000	1,152,000

[1]Sm^3/h, standard cubic meters per hour; MMscfd, million standard cubic feet per day; kilopascal gauge; psig, pounds per square inch gauge.

recovery more economically favorable are described in the Natural Gas Liquid Recovery Study section in Appendix C of this report.

Effect of Phasing Compressor Installation

The concept of adding compressors as the produced gas flow rate increases during operation of the EOR flood ("phasing") was also evaluated. Many companies that operate CO_2 EOR facilities elect to defer the relatively high capital cost of compression equipment purchases until such time as the amount of CO_2 returning with the produced oil and water requires additional CO_2 compression equipment capacity. Table 6 shows the comparative purchased equipment costs for installing compressors in multiple phases or installing a single compressor with the aggregate capacity of the multiple compressors for Cases 5 through 8. In all cases, the cost of installing a single high-pressure compressor is less expensive by a factor of approximately two when compared with purchasing three smaller high-pressure compressors that have the same aggregate capacity as the single large compressor. Note that the costs do not include the cost of the low-pressure compressor for Cases 7 and 8. Even though the high-pressure compressor cost comparison does not appear favorable for multiphasing, other factors should be considered. The operability of a smaller compressor would be expected to be better in the early years of EOR, when gas rates are increasing and the smaller compressor is operating closer to its design point, as opposed to a single large compressor operating at a small fraction of its capacity. If a single compressor must be purchased based on projections of future CO_2 production rates but there is a chance that the produced CO_2 may never actually reach that rate, then a phased approach using smaller machines may ultimately be a lower risk, more cost-effective approach. Smaller compressors may also be easier to transfer and reuse at other facilities. The phased approach also allows producers to defer some of the major equipment costs to future years, which could improve the overall economics of the project.

MISCELLANEOUS COST ITEMS

The scope of Trimeric's facility work included estimates of the costs for potential environmental controls and costs for flowlines to and from the EOR surface facilities. These miscellaneous cost items are not included in the total FCI facility cost estimates provided elsewhere in this report.

Environmental Controls

Environmental regulations have not been developed for EOR facilities in Illinois, so the information in this section is intended to provide some guidance on what costs could be encountered for providing air emissions control (of hydrocarbons, CO_2 gases, or both); however, this document is not a recommendation or prediction for what will be required. The EOR surface facilities evaluated have two potential sources of air (gas) emissions: (1) the low-pressure suction scrubber and (2) the oil storage tanks.

The flash gas generated in the low-pressure suction scrubber is compressed in a low-pressure compressor and combined with the inlet gas for the high-pressure compressor train(s) for Cases 7, 8, 11, and 12 because the low-pressure gas flow rate is large enough to justify the installation of a low-pressure compressor. It was assumed for Cases 1–6 and Cases 9 and 10 that the low-pressure gas could be vented to the atmosphere or sent to a flare. However, in Cases 1–6 and Cases 9 and 10, the purchased costs were estimated for low-pressure compressors to feed low-pressure gas to the high-pressure compressor train(s) in order to provide an estimate of the additional purchased equipment costs if environmental regulations required low-pressure compressors when they might not otherwise have been needed. The additional low-pressure compressor costs are summarized in Table 7.

The working losses[1] from the oil storage tanks could increase if the oil production rate increased with the change from waterflood to CO_2 flood. The characteristics of the vapors vented as working losses would depend on the composition and properties of the produced oil, so it is difficult to generalize what vent controls might be required. A flare (with or without an inlet blower) is a typical vapor emissions control device for oil storage tank vents, but proper flare design is critical for smoke-free operation with low-pressure oil storage tank vents. Alternatively, a compressor similar to that used for the low-pressure suction scrubber vent could be used to send the recovered storage tank vapors to the suction of the recycle CO_2 compressor for reinjection.

[1]"Working losses" from oil storage tanks are the vapors that are pushed out of the vent when the liquid level rises during production.

Table 7 Low-pressure flash gas vent recovery compressor purchased equipment cost estimates[1]

Case	Vent flow rate, Sm^3/h (MMscfd)	Vent gas pressure, kPag (psig)	Recovery compressor discharge pressure,[2] kPag (psig)	Cost source	Purchased equipment, cost, $
1	26 (0.022)	165 (24)	1,034 (150)	Scaled from Case 5[3]	69,000
2	26 (0.022)	165 (24)	1,034 (150)	Scaled from Case 5[3]	69,000
3	118 (0.10)	165 (24)	2,172 (315)	Aspen In-Plant Cost Estimator	349,000
4	118 (0.10)	165 (24)	2,172 (315)	Aspen In-Plant Cost Estimator	349,000
5	271 (0.23)	165 (24)	1,034 (150)	Aspen In-Plant Cost Estimator	369,000
6	271 (0.23)	165 (24)	1,034 (150)	Aspen In-Plant Cost Estimator	369,000
9	271 (0.23)	165 (24)	1,034 (150)	Aspen In-Plant Cost Estimator	369,000
10	271 (0.23)	165 (24)	1,034 (150)	Aspen In-Plant Cost Estimator	369,000

[1] Sm^3/h, standard cubic meters per hour; MMscfd, million standard cubic feet per day; kilopascal gauge; psig, pounds per square inch gauge.
[2] Assumes that the recovered vapor will be returned to the high-pressure compressor suction.
[3] The Aspen cost database does not include compressors at this small capacity.

Pipelines

The sizes for flowlines to send the produced fluids to the EOR surface facility were estimated by assuming the typical design velocity for two-phase flow of 9.1 m/s (30 ft/s) or less. The break point diameters to remain at or below the typical design velocity, as shown in Table 8, were 101.6-mm (4-in.) pipe for Cases 1–4 and 203.2-mm (8-in.) pipe for Cases 5–12 (both schedule 40). Separate flowline sizing calculations (not included) would be required for flowlines to deliver CO_2 from the central facility. The estimated purchased costs (using the Aspen In-Plant Cost Estimator) for carbon steel and stainless steel piping material options are summarized in Table 8, including 50.8-mm (2-in.) and 152.4-mm (6-in.) piping as reference pipeline sizes for other flow rates. Costs for both carbon steel and stainless steel piping are provided because it is possible that either could be specified for flowlines going to and from the central facility. Carbon steel is assumed to be more common, given the difference in purchased cost. Other options in these types of applications can include fiberglass and carbon steel with internal corrosion-resistant coatings. A full analysis of the installed cost for the flowlines was not within the scope of Trimeric's facility work for this report because such an analysis involves assumptions about the number of wells and the distances between the wells and surface facilities, which are being estimated by others, as well as site-specific decisions regarding construction materials for the flowlines.

CONCLUSIONS

The primary functions of a CO_2 EOR central facility or CO_2 recycle facility are to (1) separate produced gas (primarily CO_2 with some hydrocarbons) from the produced liquids (oil and water), (2) compress the produced gas for reinjection into the CO_2 EOR flood, and (3) separate the produced oil and water and provide short-term storage of these products. Major central facility components include separators, compressors, and storage tanks.

This report provides a curve and equation that can be used to estimate the FCI for central facilities as a function of CO_2 recycling rates ranging from 1,180 Sm^3/h (1 MMscfd) to 24,780 Sm^3/h (21 MMscfd) for suction pressures of 1,034 kPag (150 psig) or 2,172 kPag (315 psig) and discharge pressures of 3,448 kPag (500 psig) and 6,895 kPag (1,000 psig). These conditions represent ranges that might be expected for any early-phase CO_2 EOR floods in the ILB. Estimates of major operating costs are also provided.

In some cases, NGL recovery from recycled CO_2 is necessary in EOR operations. Most often, NGL recovery from recycled CO_2 is done when justified by the economic value of the recovered NGL products or when necessary to maintain the CO_2–recycled gas composition to sustain a suitable minimum miscibility pressure (MMP) for the CO_2 and oil in the formation. This report examines the types of NGL recovery processes that might be considered for operations in the ILB and concludes that NGL recovery would likely not be economical or necessary to maintain the MMP in initial CO_2 EOR operations in the ILB. This conclusion is based on the historically low amounts of associated hydrocarbon gas encountered in mature waterflood operations in the ILB that might be candidates for CO_2 EOR.

The FCI for this type of CO_2 EOR central facility is typically dominated by compressor costs. Smaller compressors are often installed in multiple phases as the

Table 8 Unit purchased equipment costs for piping[1]

Cases	Pipe size	Carbon steel pipe cost, $/mi	304 stainless steel pipe cost, $/mi
None selected	50.8-mm (2-in.) schedule 40	49,400	80,000
1–4	101.6-mm (4-in.) schedule 40	51,500	162,000
None selected	152.4-mm (6-in.) schedule 40	117,500	323,100
5–12	203.2-mm (8-in.) schedule 40	139,000	641,700

[1]Costs of piping materials can fluctuate significantly. It is necessary to verify current pricing at the beginning of each project.

produced gas rate from the field increases during the CO_2 EOR flood operations in order to improve the economics of the project and provide more operational flexibility (relative to installing a single larger compressor at the beginning of operations). Compressor costs are a function of the suction pressure, discharge pressure, gas composition, and mass flow rate of the gas. These factors can have varying degrees of influence on a case-by-case basis, but the suction pressure (which influences the actual volumetric flow rate of the gas to be compressed) is often an important factor in determining compressor costs.

A subsequent report will provide similar ILB-specific central facility cost estimates for larger recycling facilities with capacities ranging from 59,000 Sm^3/h (50 MMscfd) to 236,000 Sm^3/h (200 MMscfd) and similar operating pressures. The subsequent report will also provide a curve and equation that can be used to estimate the FCI for the central facility over the entire range of CO_2 recycling rates covered in these reports, which is 1,180 Sm^3/h (1 MMscfd) to 236,000 Sm^3/h (200 MMscfd).

ACKNOWLEDGMENTS

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APPENDIX A

Process Flow Diagrams for Cases 1–12

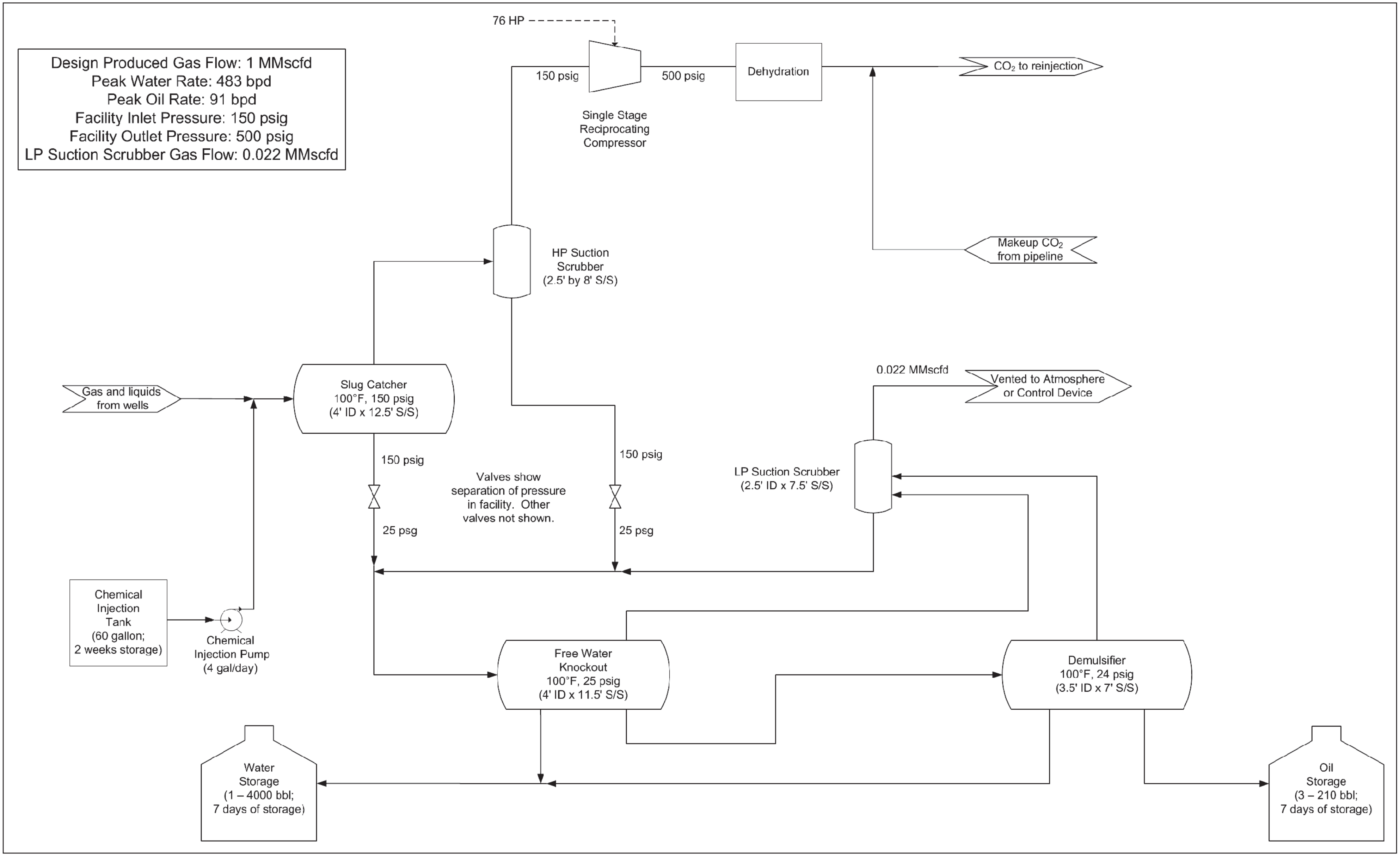

Figure A1 Process flow diagram for Case 1.

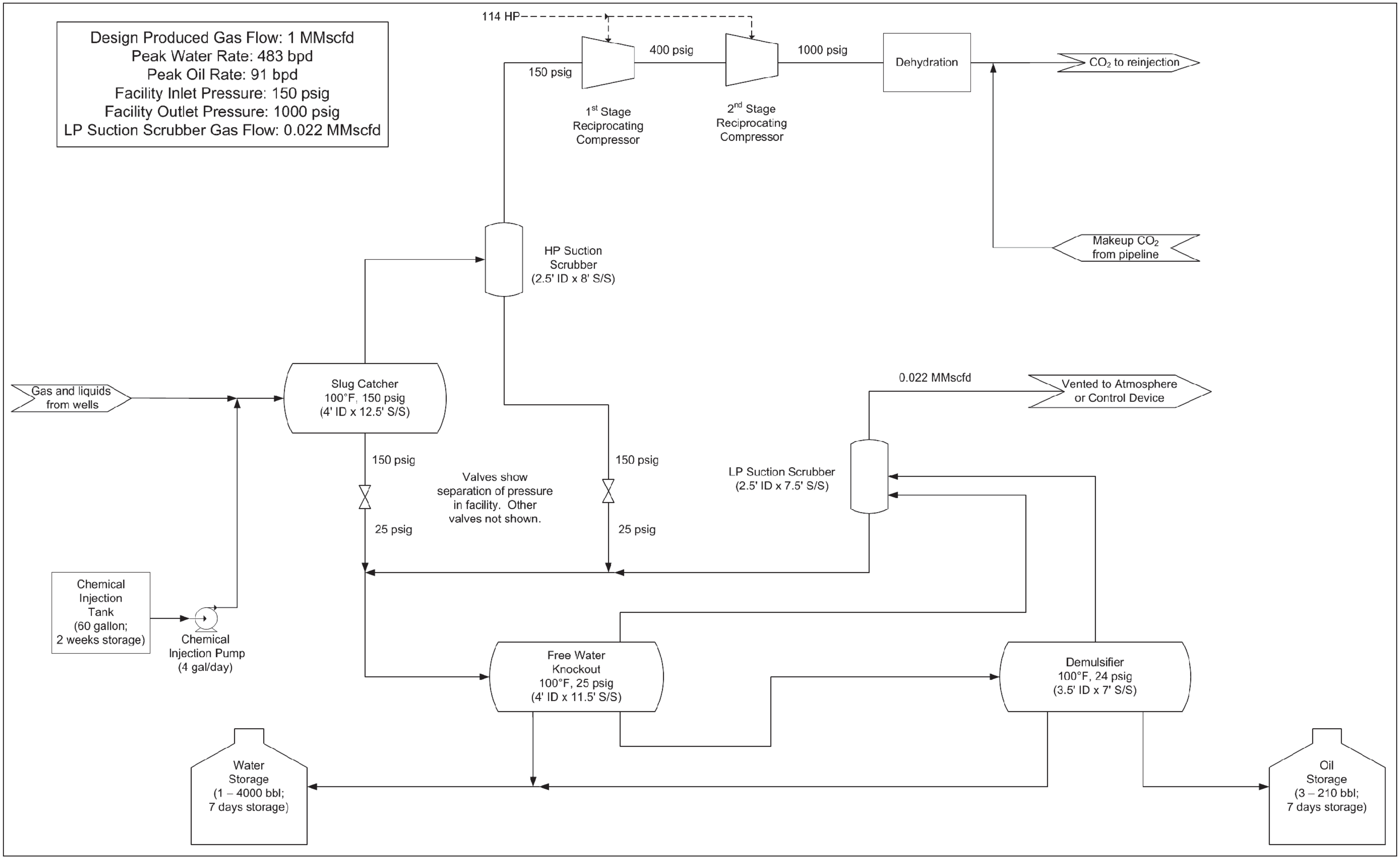

Figure A2 Process flow diagram for Case 2.

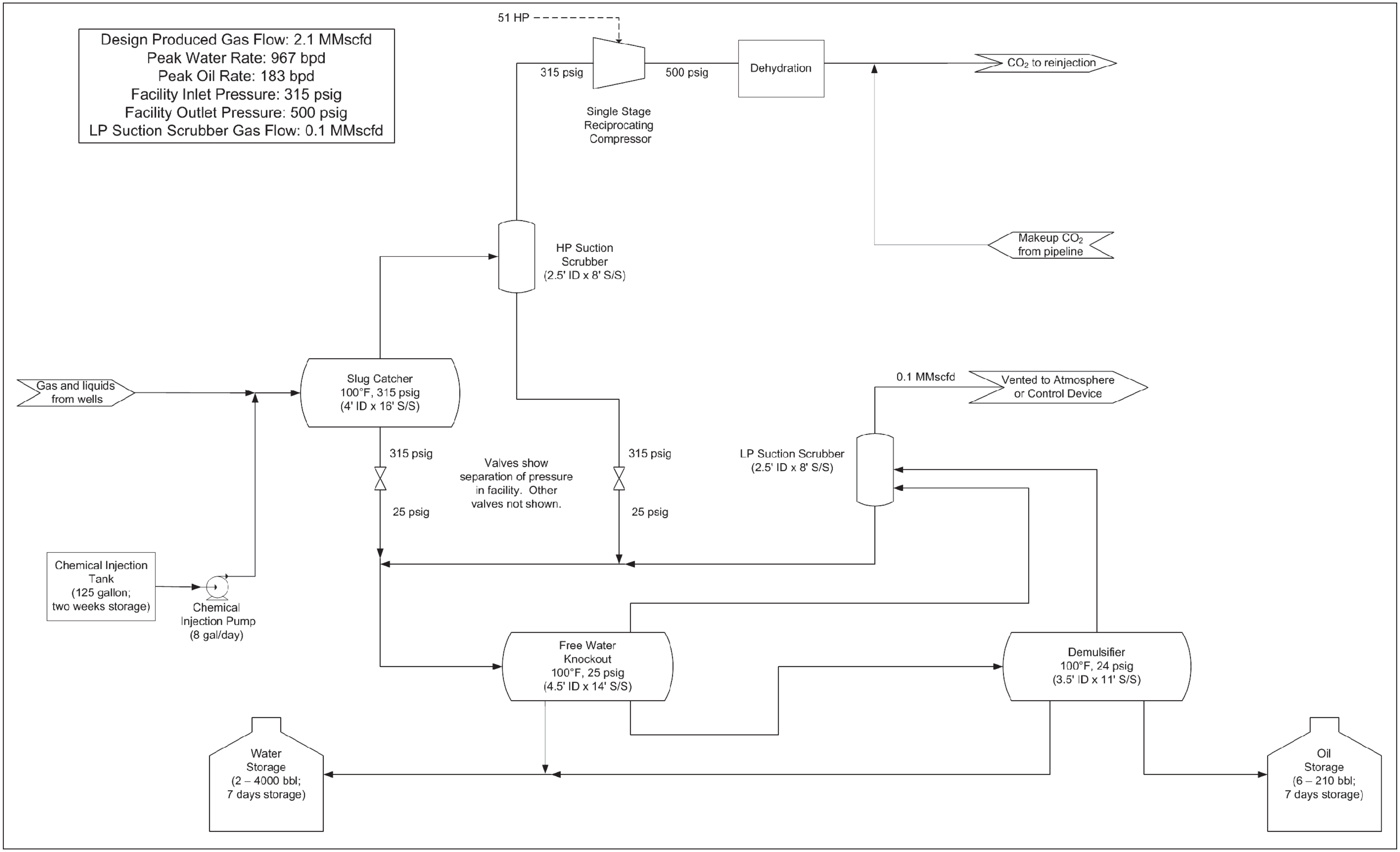

Figure A3 Process flow diagram for Case 3.

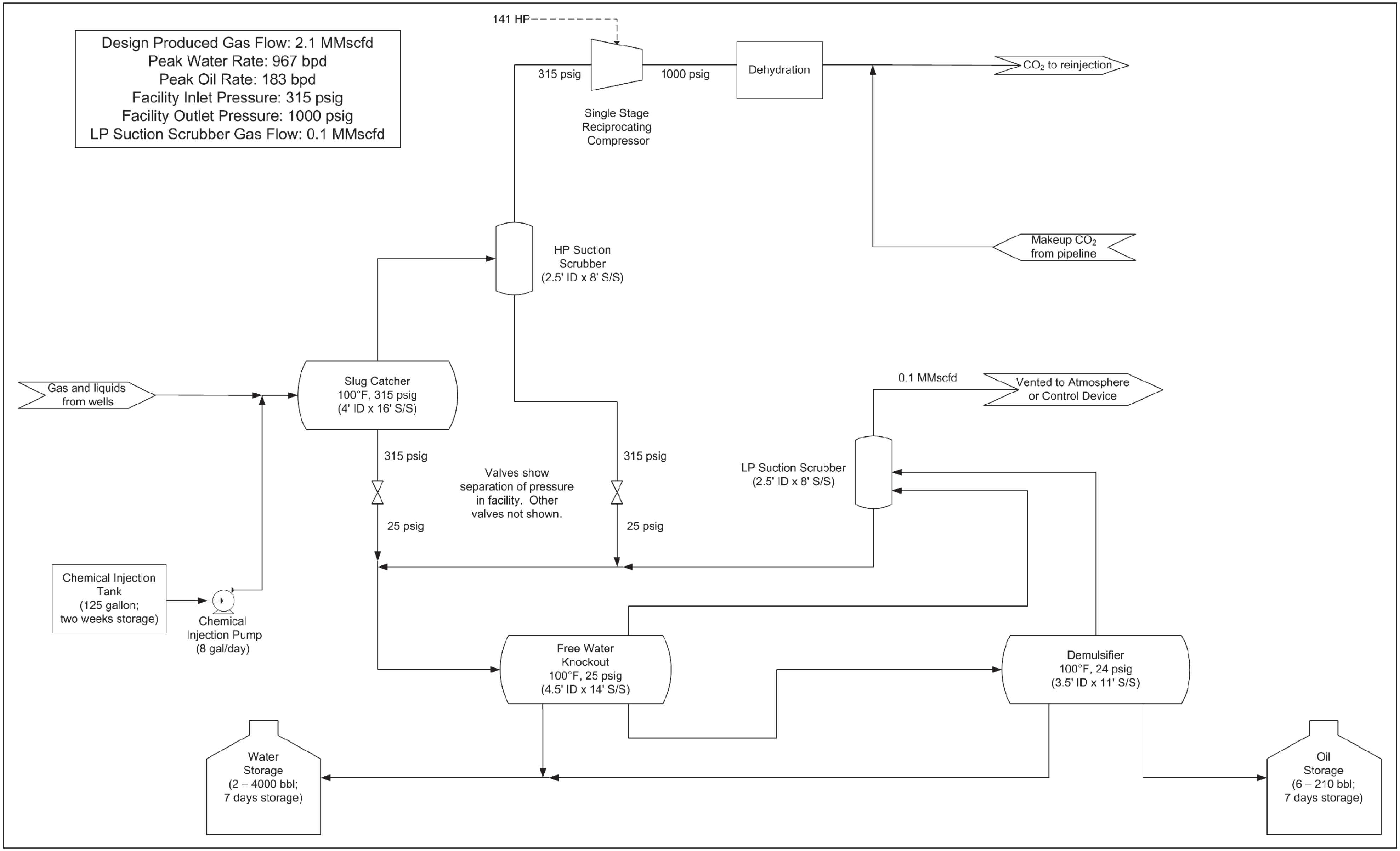

Figure A4 Process flow diagram for Case 4.

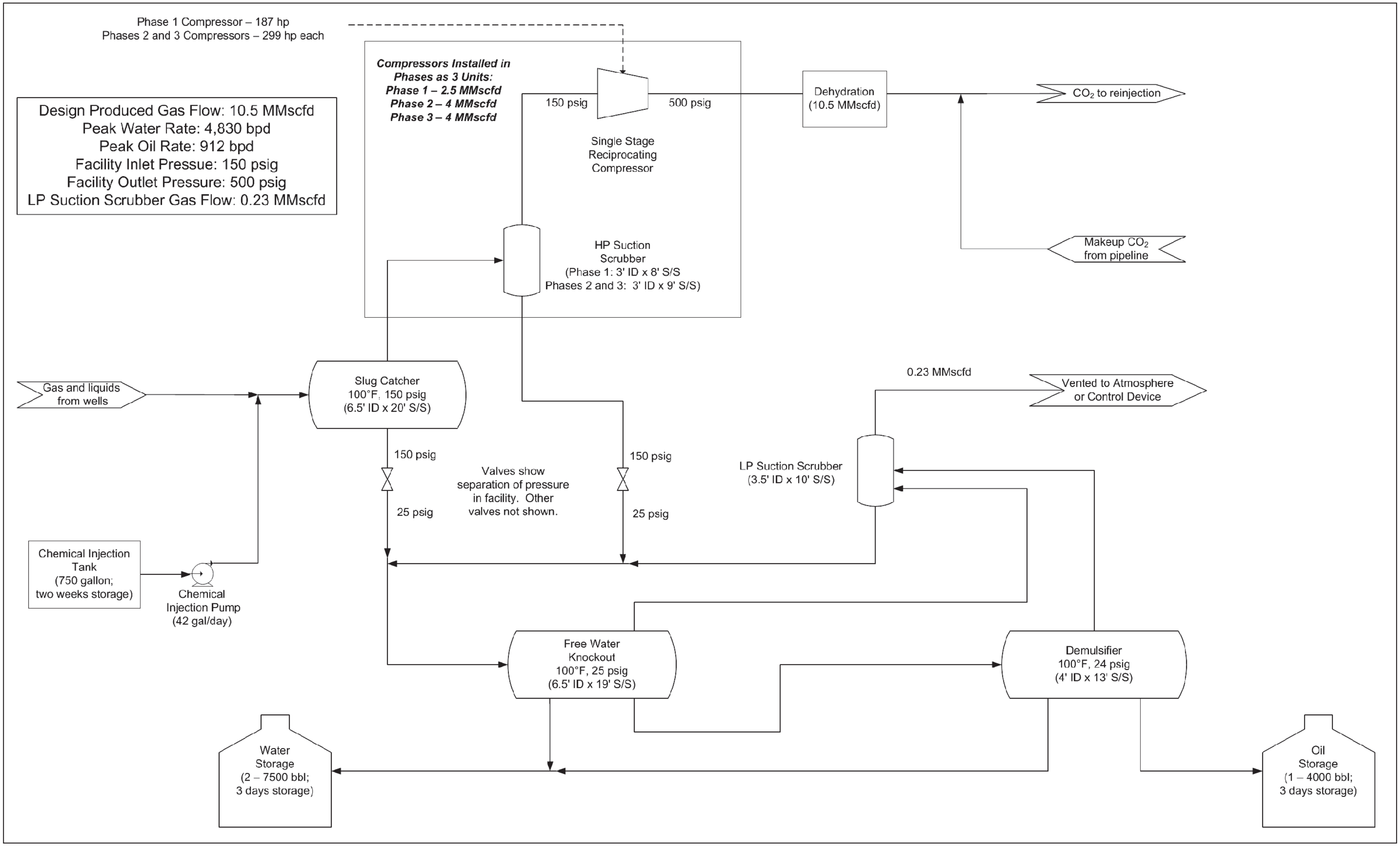

Figure A5 Process flow diagram for Case 5.

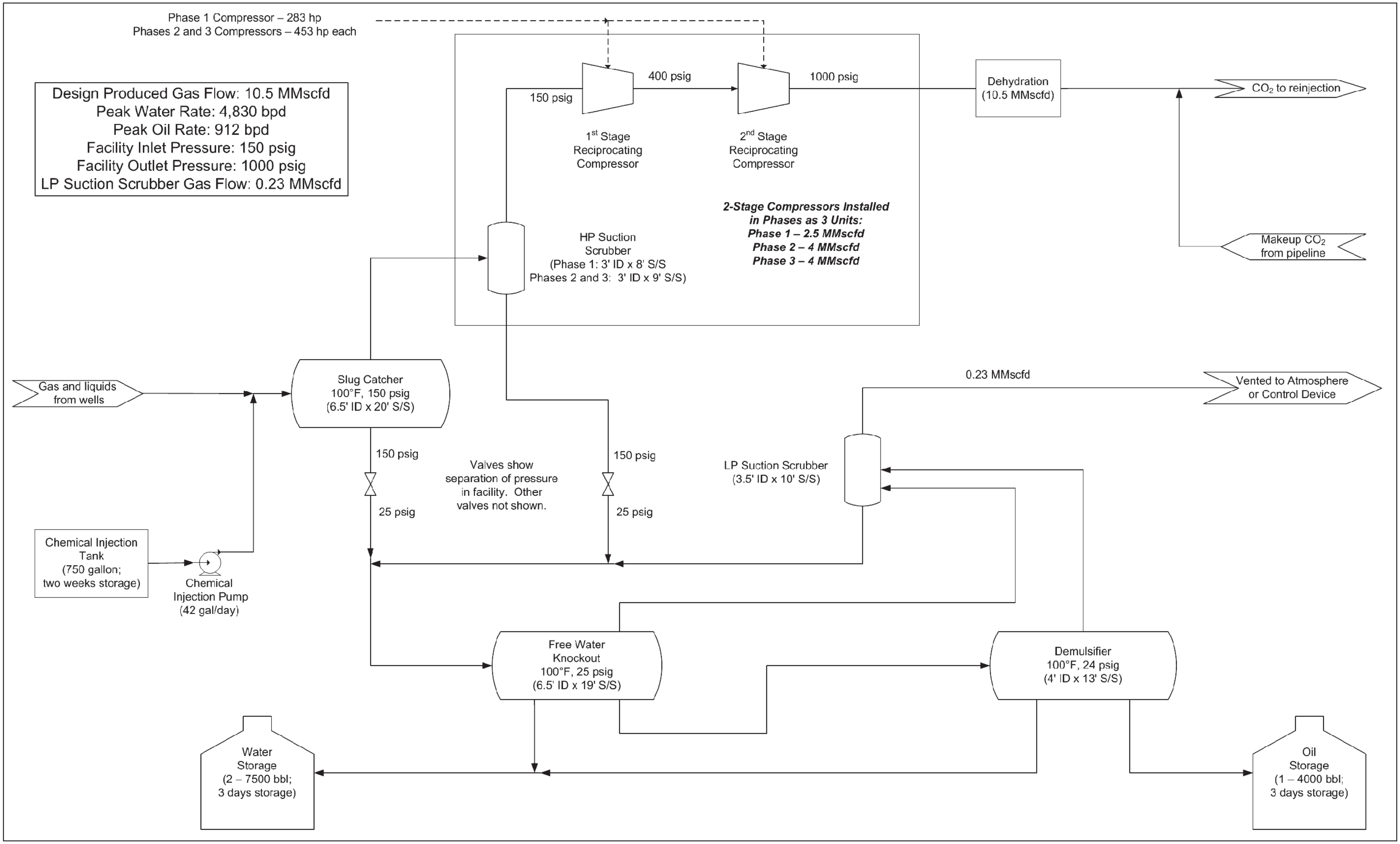

Figure A6 Process flow diagram for Case 6.

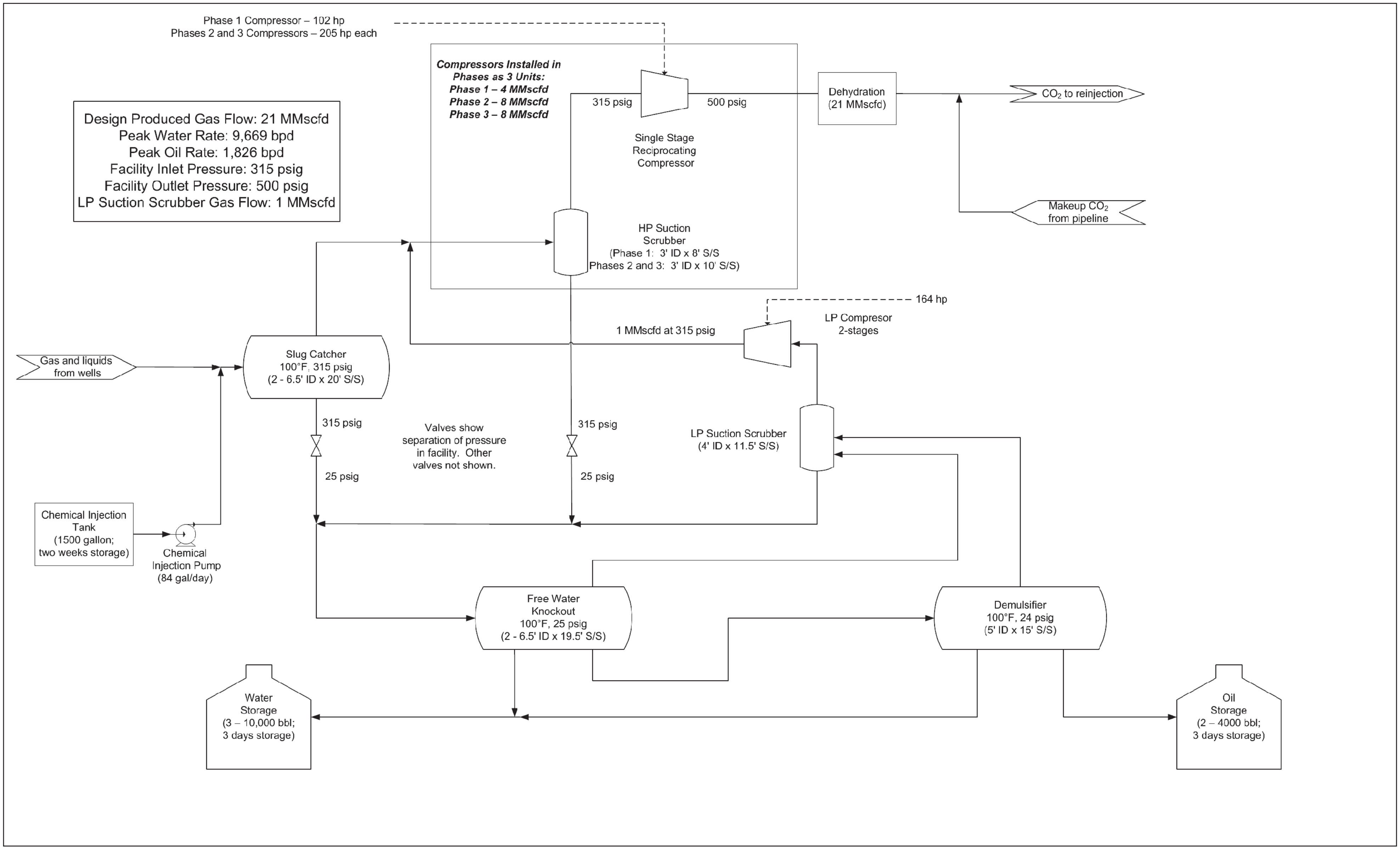

Figure A7 Process flow diagram for Case 7.

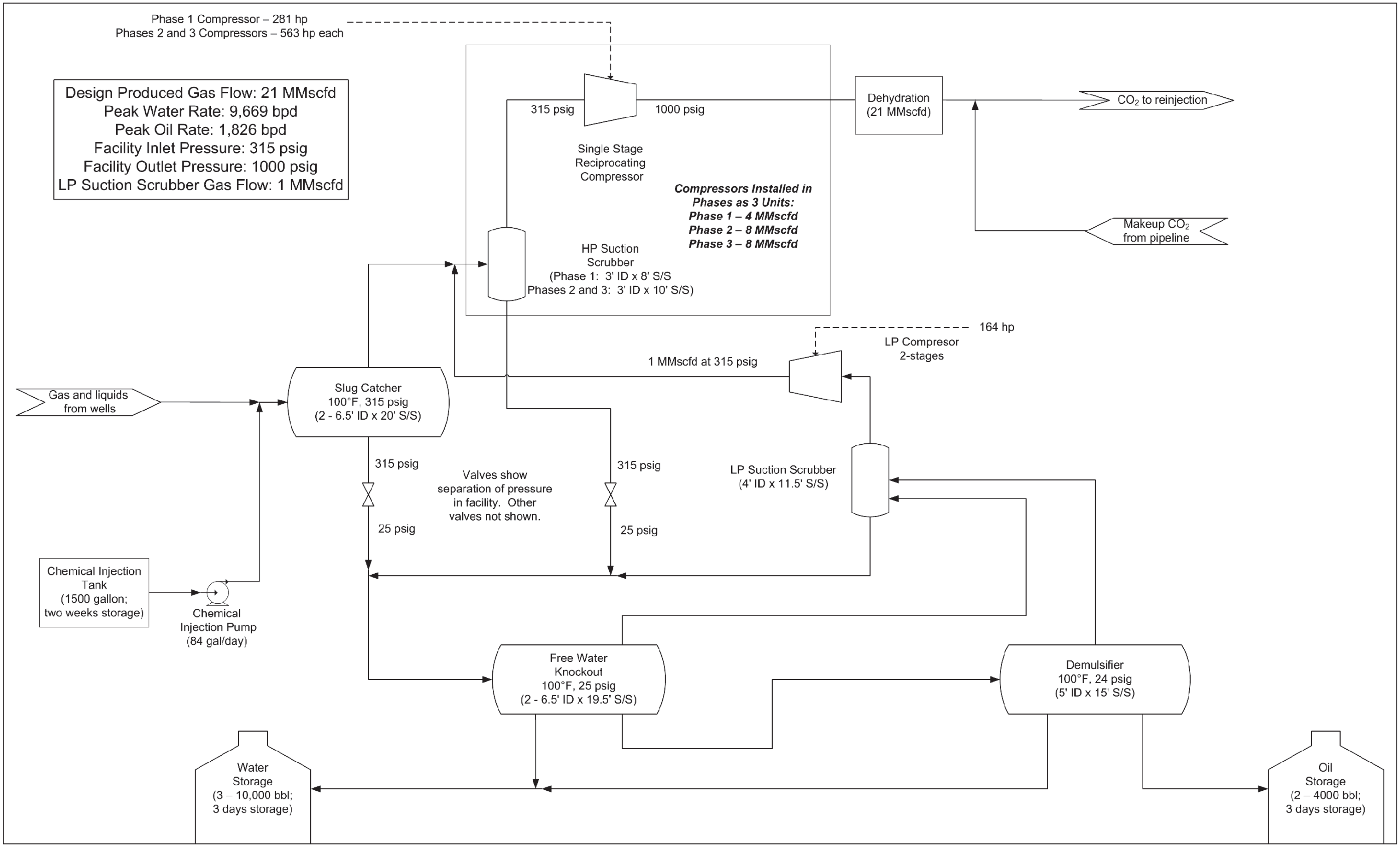

Figure A8 Process flow diagram for Case 8.

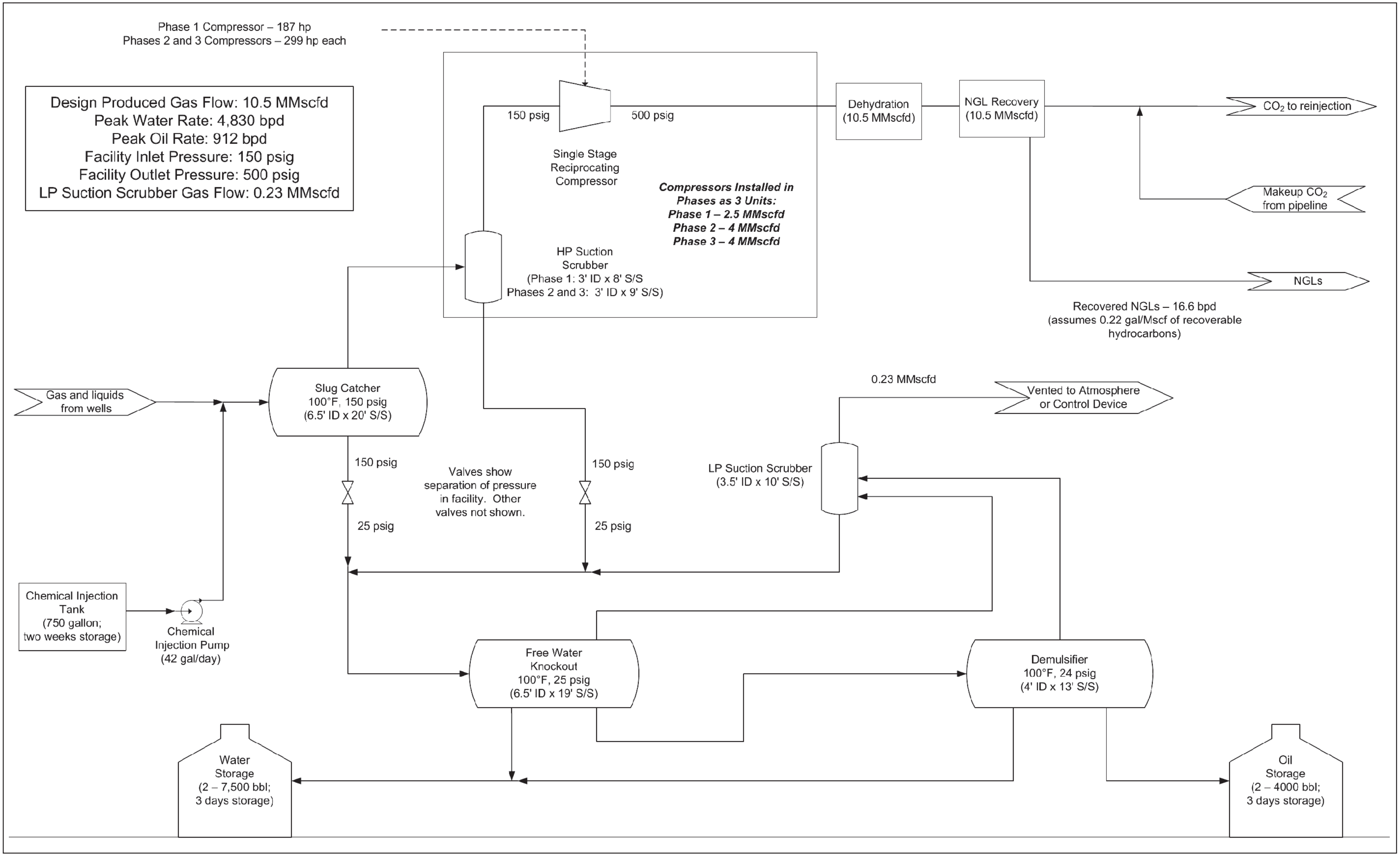

Figure A9 Process flow diagram for Case 9.

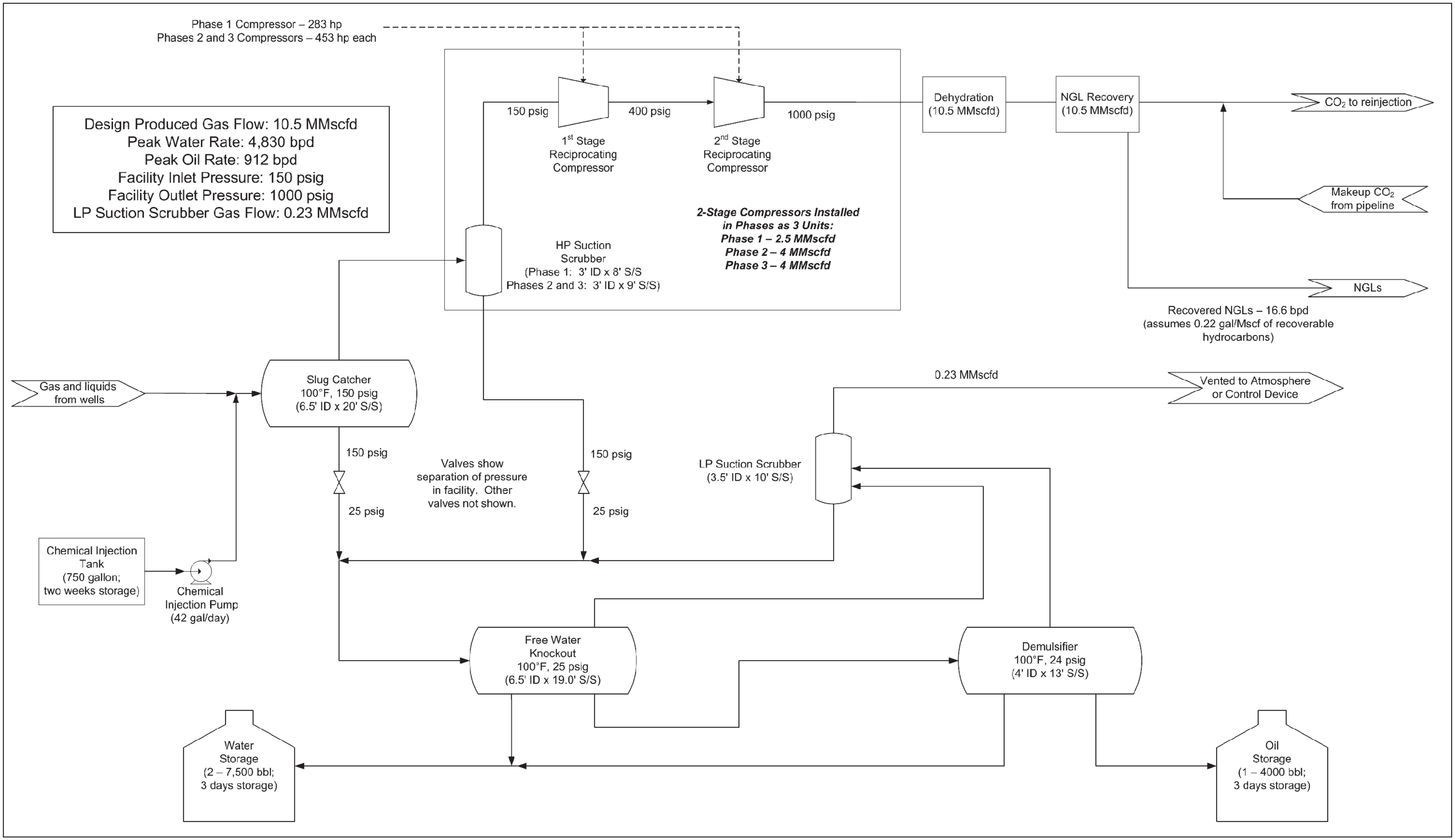

Figure A10 Process flow diagram for Case 10.

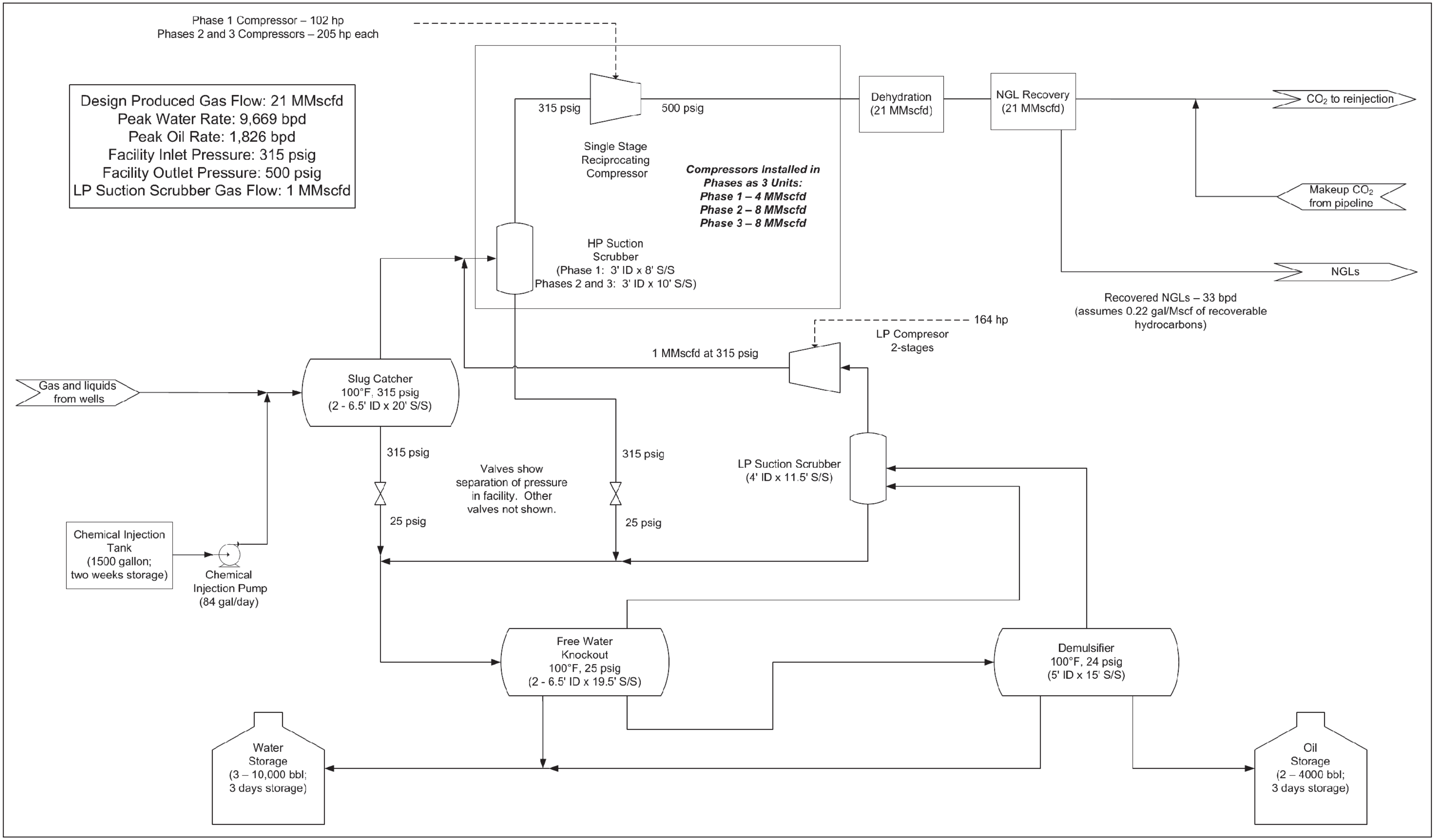

Figure A11 Process flow diagram for Case 11.

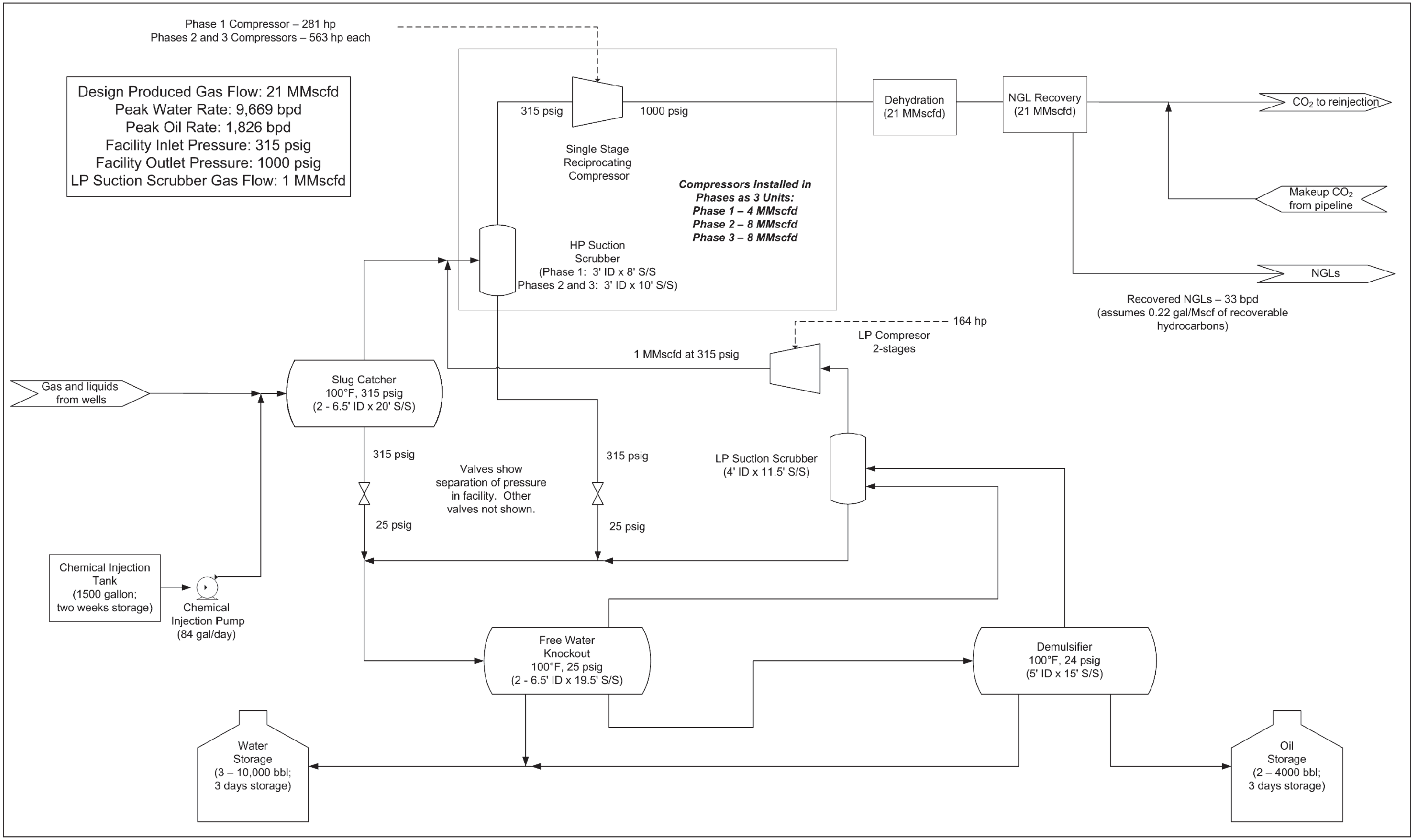

Figure A12 Process flow diagram for Case 12.

APPENDIX B

Equipment List and Purchased and Installed Costs for Cases 1–12

Table B1 Case 1—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 4 ft ID × 12.5 ft S/S length, cladded carbon steel	180	200	35,500	1	576.0	593.8	36,600	0.88	68,800
HP suction scrubber	Vertical separator vessel, 2.5 ft ID × 8 ft S/S length, cladded carbon steel	180	200	19,900	1	576.0	593.8	20,500	0.88	37,400
LP suction scrubber	Vertical separator vessel, 2.5 ft ID × 7.5 ft S/S length, cladded carbon steel	45	200	19,700	1	576.0	593.8	20,300	0.88	37,000
Free water knockout	Horizontal 3-phase separator vessel, 4 ft × 11.5 ft S/S length, cladded carbon steel	45	200	31,900	1	576.0	593.8	32,900	0.88	60,000
Demulsifier	Horizontal 3-phase separator vessel, 3.5 ft × 7 ft S/S length, cladded carbon steel	45	200	23,500	1	576.0	593.8	24,200	0.88	44,200
Chemical injection pump	4.2 gal/d to give 1,000 ppm by volume in oil	180	150	8,300	2	576.0	593.8	17,100	0.30	21,600
Chemical injection tank	60-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	300	1	404.0	593.8	400	0.88	600
Oil storage tank	210-bbl fiberglass tank (for 1-wk capacity at peak rate, 639 bbl)	Hydrostatic + 2.5 psig	150	5,700	3	593.8	593.8	17,100	0.88	32,100
Water storage tank	API storage tank 4,000-bbl capacity, carbon steel (for 1-wk capacity at peak rate, 3,381 bbl)	Hydrostatic + 2.5 psig	150	122,600	1	576.0	593.8	126,400	0.88	230,500
CO_2 compressor	76 hp from 150 to 500 psig; single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	439,200	1	576.0	593.8	452,800	0.49	654,400
Dehydration	After compression at 500 psig	600	150	114,100	1	593.8	593.8	114,100	0.72	196,300
Building		NA	NA	121,000	1	593.8	593.8	121,000	0.00	121,000
Total								983,400		1,503,900

[1]Production rates: 1,180 Sm^3/h (standard cubic meters per hour; 1.0 MMscfd [million standard cubic feet per day]) of CO_2, 14 m^3/d of oil (91 bopd [barrels of oil per day]), 77 m^3/d of water (483 bwpd [barrels of water per day]). Pressures: 1,034 kPag (kilopascal gauge; 150 psig [pounds per square inch gauge]) facility inlet, 3,448 kPag (500 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B2 Case 2—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 4 ft ID × 12.5 ft S/S length, cladded carbon steel	180	200	35,500	1	576.0	593.8	36,600	0.88	68,800
HP suction scrubber	Vertical separator vessel, 2.5 ft ID × 8 ft S/S length, cladded carbon steel	180	200	19,900	1	576.0	593.8	20,500	0.88	37,400
LP suction scrubber	Vertical separator vessel, 2.5 ft ID × 7.5 ft S/S length, cladded carbon steel	45	200	19,700	1	576.0	593.8	20,300	0.88	37,000
Free water knockout	Horizontal 3-phase separator vessel, 4 ft × 11.5 ft S/S length, cladded carbon steel	45	200	31,900	1	576.0	593.8	32,900	0.88	60,000
Demulsifier	Horizontal 3-phase separator vessel, 3.5 ft × 7 ft S/S length, cladded carbon steel	45	200	23,500	1	576.0	593.8	24,200	0.88	44,200
Chemical injection pump	4.2 gal/d to give 1,000 ppm by volume in oil	180	150	8,300	2	576.0	593.8	17,100	0.30	21,600
Chemical injection tank	60-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	300	1	404.0	593.8	400	0.88	600
Oil storage tank	210-bbl fiberglass tank (for 1-wk capacity at peak rate, 639 bbl)	Hydrostatic + 2.5 psig	150	5,700	3	593.8	593.8	17,100	0.88	32,100
Water storage tank	API storage tank 4,000-bbl capacity, carbon steel (for 1-wk capacity at peak rate, 3,381 bbl)	Hydrostatic + 2.5 psig	150	122,600	1	576.0	593.8	126,400	0.88	230,500
CO_2 compressor	114 hp from 150 to 1,000 psig; 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	515,700	1	576.0	593.8	531,600	0.49	768,400
Dehydration	Dehydration after 2nd stage of compression at 1,000 psig	1,100	150	91,300	1	593.8	593.8	91,300	0.72	157,000
Building		NA	NA	121,000	1	593.8	593.8	121,000	0.00	121,000
Total								1,039,400		1,578,600

[1]Production rates: 1,180 Sm^3/h (standard cubic meters per hour; 1.0 MMscfd [million standard cubic feet per day]) of CO_2, 14 m^3/d of oil (91 bopd [barrels of oil per day]), 77 m^3/d of water (483 bwpd [barrels of water per day]). Pressures: 1,034 kPag (kilopascal gauge; 150 psig [pounds per square inch gauge]) facility inlet, 6,895 kPag (1,000 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B3 Case 3—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 4 ft ID × 16 ft S/S length, cladded carbon steel	365	200	53,300	1	576.0	593.8	54,900	0.88	103,200
HP suction scrubber	Vertical separator vessel, 2.5 ft ID × 8 ft S/S length, cladded carbon steel	365	200	23,600	1	576.0	593.8	24,300	0.88	45,700
LP suction scrubber	Vertical separator vessel, 2.5 ft ID × 8 ft S/S length, cladded carbon steel	45	200	20,100	1	576.0	593.8	20,700	0.88	38,900
Free water knockout	Horizontal 3-phase separator vessel, 4.5 ft × 14 ft S/S length, cladded carbon steel	45	200	41,700	1	576.0	593.8	43,000	0.88	80,800
Demulsifier	Horizontal 3-phase separator vessel, 3.5 ft × 11 ft S/S length, cladded carbon steel	45	200	28,700	1	576.0	593.8	29,600	0.88	55,600
Chemical injection pump	8.4 gal/d to give 1,000 ppm by volume in oil	365	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	125-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	600	1	404.0	593.8	900	0.88	1,700
Oil storage tank	210-bbl fiberglass tank each (for 1-wk capacity at peak rate, 1,278 bbl)	Hydrostatic + 2.5 psig	150	5,700	6	593.8	593.8	34,200	0.88	64,300
Water storage tank	API storage tanks at 4,000-bbl capacity each, carbon steel (for 1-wk capacity at peak rate, 6,769 bbl)	Hydrostatic + 2.5 psig	150	122,600	2	576.0	593.8	252,800	0.88	475,300
CO_2 compressor	51 hp from 315 to 500 psig; single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	408,200	1	576.0	593.8	420,800	0.49	627,000
Dehydration	After compression at 500 psig	600	150	173,100	1	593.8	593.8	173,100	0.72	297,700
Building		NA	NA	142,000	1	593.8	593.8	142,000	0.00	142,000
Total								1,213,400		1,954,400

[1]Production rates: 2,478 Sm^3/h (standard cubic meters per hour; 2.1 MMscfd [million standard cubic feet per day]) of CO_2, 29 m^3/d of oil (183 bopd [barrels of oil per day]), 154 m^3/d of water (967 bwpd [barrels of water per day). Pressures: 2,172 kPag (kilopascal gauge; 315 psig [pounds per square inch gauge]) facility inlet, 3,448 kPag (500 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B4 Case 4—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 4 ft ID × 16 ft S/S length, cladded carbon steel	365	200	53,300	1	576.0	593.8	54,900	0.88	103,200
HP suction scrubber	Vertical separator vessel, 2.5 ft ID × 8 ft S/S length, cladded carbon steel	365	200	23,600	1	576.0	593.8	24,300	0.88	45,700
LP suction scrubber	Vertical separator vessel, 2.5 ft ID × 8 ft S/S length, cladded carbon steel	45	200	20,100	1	576.0	593.8	20,700	0.88	38,900
Free water knockout	Horizontal 3-phase separator vessel, 4.5 ft × 14 ft S/S length, cladded carbon steel	45	200	41,700	1	576.0	593.8	43,000	0.88	80,800
Demulsifier	Horizontal 3-phase separator vessel, 3.5 ft × 11 ft S/S length, cladded carbon steel	45	200	28,700	1	576.0	593.8	29,600	0.88	55,600
Chemical injection pump	8.4 gal/d to give 1,000 ppm by volume in oil	365	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	125-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	600	1	404.0	593.8	900	0.88	1,700
Oil storage tank	210-bbl fiberglass tank each (for 1-wk capacity at peak rate, 1,278 bbl)	Hydrostatic + 2.5 psig	150	5,700	6	593.8	593.8	34,200	0.88	64,300
Water storage tank	API storage tanks at 4,000-bbl capacity each, carbon steel (for 1-wk capacity at peak rate, 6,769 bbl)	Hydrostatic + 2.5 psig	150	122,600	2	576.0	593.8	252,800	0.88	475,300
CO_2 compressor	141 hp from 315 to 1,000 psig; single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	512,300	1	576.0	593.8	528,100	0.49	786,900
Dehydration	After compression at 1,000 psig	1,100	150	138,500	1	593.8	593.8	138,500	0.72	238,200
Building		NA	NA	142,000	1	593.8	593.8	142,000	0.00	142,000
Total								1,286,100		2,054,800

[1]Production rates: 2,478 Sm^3/h (standard cubic meters per hour; 2.1 MMscfd [million standard cubic feet per day]) of CO_2, 29 m^3/d of oil (183 bopd [barrels of oil per day]), 154 m^3/d of water (967 bwpd [barrels of water per day]). Pressures: 2,172 kPag (kilopascal gauge; 315 psig [pounds per squre inch gauge]) facility inlet, 6,895 kPag (1,000 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B5 Case 5—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	180	200	86,800	1	576.0	593.8	89,500	0.88	168,300
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	180	200	22,400	1	576.0	593.8	23,100	0.88	43,400
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 9 ft S/S length, cladded carbon steel	180	200	23,300	2	576.0	593.8	48,000	0.88	90,200
LP suction scrubber	Vertical separator vessel, 3.5 ft ID × 10 ft S/S length, cladded carbon steel	45	200	30,100	1	576.0	593.8	31,000	0.88	58,300
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19 ft S/S length, cladded carbon steel	45	200	68,300	1	576.0	593.8	70,400	0.88	132,400
Demulsifier	Horizontal 3-phase separator vessel, 4 ft × 13 ft S/S length, cladded carbon steel	45	200	37,500	1	576.0	593.8	38,700	0.88	72,800
Chemical injection pump	42 gal/d to give 1,000 ppm by volume in oil	180	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	750-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	1,800	1	404.0	593.8	2,600	0.88	4,900
Oil storage tank	API storage tanks at 4,000-bbl capacity each (for 3-d capacity at peak rate, 2,737 bbl)	Hydrostatic + 2.5 psig	150	122,600	1	576.0	593.8	126,400	0.88	237,600
Water storage tank	API storage tanks at 7,500-bbl capacity each, carbon steel (for 3-d capacity at peak rate, 14,491 bbl)	Hydrostatic + 2.5 psig	150	186,600	2	576.0	593.8	384,700	0.88	723,200
CO_2 compressor–1st phase	2.5 MMscfd from 150 to 500 psig (187 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	545,000	1	576.0	593.8	561,800	0.49	837,100
CO_2 compressor–2nd and 3rd phases	4 MMscfd from 150 to 500 psig (299 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	618,800	2	576.0	593.8	1,275,800	0.49	1,900,900
Dehydration	Treats 10.5 MMscfd after compression at 500 psig	600	150	454,600	1	593.8	593.8	454,600	0.72	781,900
Building		NA	NA	309,800	1	593.8	593.8	309,800	0.00	309,800
Total								3,433,500		5,383,000

[1]Production rates: 12,390 Sm^3/h (standard cubic meters per hour; 10.5 MMscfd [million standard cubic feet per day]) of CO_2, 145 m^3/d of oil (912 bopd [barrels of oil per day]), 768 m^3/d of water (4,830 bwpd [barrels of water per day]). Pressures: 1,034 kPag (kilopascal gauge;150 psig [pounds per inch gauge]) facility inlet, 3,448 kPag (500 psig) facility outlet. HP, high pressure; ID, inside diameter; S/S, seam to seam.

Table B6 Case 6—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	180	200	86,800	1	576.0	593.8	89,500	0.88	168,300
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	180	200	22,400	1	576.0	593.8	23,100	0.88	43,400
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 9 ft S/S length, cladded carbon steel	180	200	23,300	2	576.0	593.8	48,000	0.88	90,200
LP suction scrubber	Vertical separator vessel, 3.5 ft ID × 10 ft S/S length, cladded carbon steel	45	200	30,100	1	576.0	593.8	31,000	0.88	58,300
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19 ft S/S length, cladded carbon steel	45	200	68,300	1	576.0	593.8	70,400	0.88	132,400
Demulsifier	Horizontal 3-phase separator vessel, 4 ft × 13 ft S/S length, cladded carbon steel	45	200	37,500	1	576.0	593.8	38,700	0.88	72,800
Chemical injection pump	42 gal/d to give 1,000 ppm by volume in oil	180	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	750-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	1,800	1	404.0	593.8	2,600	0.88	4,900
Oil storage tank	API storage tanks at 4,000-bbl capacity each (for 3-d capacity at peak rate, 2,737 bbl)	Hydrostatic + 2.5 psig	150	122,600	1	576.0	593.8	126,400	0.88	237,600
Water storage tank	API storage tanks at 7,500-bbl capacity each, carbon steel (for 3-d capacity at peak rate, 14,491 bbl)	Hydrostatic + 2.5 psig	150	186,600	2	576.0	593.8	384,700	0.88	723,200
CO_2 compressor–1st phase	2.5 MMscfd from 150 to 1,000 psig (283 hp); 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	613,500	1	576.0	593.8	632,500	0.49	942,400
CO_2 compressor–2nd and 3rd phases	4 MMscfd from 150 to 1,000 psig (453 hp); 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	766,700	2	576.0	593.8	1,580,800	0.49	2,355,400
Dehydration	Treats 10.5 MMscfd after compression at 1,000 psig	1,100	150	363,700	1	593.8	593.8	363,700	0.72	625,600
Building		NA	NA	309,800	1	593.8	593.8	309,800	0.00	309,800
Total								3,718,300		5,786,500

[1]Production rates: 12,390 Sm^3/h (standard cubic meters per hour; 10.5 MMscfd [million standard cubic feet per day]) of CO_2, 145 m^3/d of oil (912 bopd [barrels of oil per day]), 768 m^3/d of water (4,830 bwpd [barrels of water per day]). Pressures: 1,034 kPag (kilopascal gauge; 150 psig [pounds per square inch gauge]) facility inlet, 6,895 kPag (1,000 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B7 Case 7—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	365	200	119,600	2	576.0	593.8	246,600	0.88	463,600
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	365	200	26,700	1	576.0	593.8	27,500	0.88	51,700
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 10 ft S/S length, cladded carbon steel	365	200	29,100	2	576.0	593.8	60,000	0.88	112,800
LP suction scrubber	Vertical separator vessel, 4 ft ID × 11.5 ft S/S length, cladded carbon steel	45	200	34,300	1	576.0	593.8	35,400	0.88	66,600
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19.5 ft S/S length, cladded carbon steel	45	200	69,000	2	576.0	593.8	142,300	0.88	267,500
Demulsifier	Horizontal 3-phase separator vessel, 5 ft × 15 ft S/S length, cladded carbon steel	45	200	45,700	1	576.0	593.8	47,100	0.88	88,500
Chemical injection pump	84 gal/d to give 1,000 ppm by volume in oil	365	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	1,500-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	3,600	1	404.0	593.8	5,300	0.88	10,000
Oil storage tank	API storage tank 4,000-bbl capacity each (for 3-d capacity at peak rate, 5,479 bbl)	Hydrostatic + 2.5 psig	150	122,600	2	576.0	593.8	252,800	0.88	475,300
Water storage tank	API storage tank 10,000-bbl capacity each, carbon steel (for 3-d capacity at peak rate, 29,008 bbl)	Hydrostatic + 2.5 psig	150	230,100	3	576.0	593.8	711,600	0.88	1,337,800
CO_2 compressor–1st phase	4 MMscfd from 315 to 500 psig (102 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	483,200	1	576.0	593.8	498,100	0.49	742,200
CO_2 compressor–2nd and 3rd phases	8 MMscfd from 315 to 500 psig (205 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	556,400	2	576.0	593.8	1,147,200	0.49	1,709,300
LP compressor	1 MMscfd from 25 to 315 psig (164 hp); two-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	583,700	1	576.0	593.8	601,700	0.49	896,500
Dehydration	Treats 21 MMscfd after compression at 500 psig	600	150	688,900	1	593.8	593.8	688,900	0.72	1,184,900
Building		NA	NA	520,000	1	593.8	593.8	520,000	0.00	520,000
Total								5,001,600		7,948,900

[1]Production rates: 24,780 Sm^3/h (standard cubic meters per hour; 21.0 MMscfd [million standard cubic feet per day]) of CO_2, 290 m^3/d of oil (1,826 bopd [barrels of oil per day]), 1,537 m^3/d of water (9,669 bwpd [barrels of water per day]). Pressures: 2,172 kPag (kilopascal gauge; 315 psig [pounds per square inch gauge]) facility inlet, 3,448 kPag (500 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B8 Case 8—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	365	200	119,600	2	576.0	593.8	246,600	0.88	463,600
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	365	200	26,700	1	576.0	593.8	27,500	0.88	51,700
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 10 ft S/S length, cladded carbon steel	365	200	29,100	2	576.0	593.8	60,000	0.88	112,800
LP suction scrubber	Vertical separator vessel, 4 ft ID × 11.5 ft S/S length, cladded carbon steel	45	200	34,300	1	576.0	593.8	35,400	0.88	66,600
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19.5 ft S/S length, cladded carbon steel	45	200	69,000	2	576.0	593.8	142,300	0.88	267,500
Demulsifier	Horizontal 3-phase separator vessel, 5 ft × 15 ft S/S length, cladded carbon steel	45	200	45,700	1	576.0	593.8	47,100	0.88	88,500
Chemical injection pump	84 gal/d to give 1,000 ppm by volume in oil	365	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	1,500-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	3,600	1	404.0	593.8	5,300	0.88	10,000
Oil storage tank	API storage tank 4,000-bbl capacity each (for 3-d capacity at peak rate, 5,479 bbl)	Hydrostatic + 2.5 psig	150	122,600	2	576.0	593.8	252,800	0.88	475,300
Water storage tank	API storage tank 10,000-bbl capacity each, carbon steel (for 3-d capacity at peak rate, 29,008 bbl)	Hydrostatic + 2.5 psig	150	230,100	3	576.0	593.8	711,600	0.88	1,337,800
CO_2 compressor–1st phase	4 MMscfd from 315 to 1,000 psig (281 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	580,400	1	576.0	593.8	598,300	0.49	891,500
CO_2 compressor–2nd and 3rd phases	8 MMscfd from 315 to 1,000 psig (563 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	799,900	2	576.0	593.8	1,649,200	0.49	2,457,300
LP compressor	1 MMscfd from 25 to 315 psig (164 hp); 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	583,700	1	576.0	593.8	601,700	0.49	896,500
Dehydration	Treats 21 MMscfd after compression at 1,000 psig	1,100	150	551,200	1	593.8	593.8	551,200	0.72	948,100
Building		NA	NA	520,000	1	593.8	593.8	520,000	0.00	520,000
Total								5,466,100		8,609,400

[1]Production rates: 24,780 Sm^3/h (standard cubic meters per hour; 21.0 MMscfd [million standard cubic feet per day]) of CO_2, 290 m^3/d of oil (1,826 bopd [barrels of oil per day]), 1,537 m^3/d of water (9,669 bwpd [barrels of water per day]). Pressures: 2,172 kPag (kilopascal gauge; 315 psig [pounds per square inch gauge]) facility inlet, 6,895 kPag (1,000 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B9 Case 9—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	180	200	86,800	1	576.0	593.8	89,500	0.88	168,300
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	180	200	22,400	1	576.0	593.8	23,100	0.88	43,400
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 9 ft S/S length, cladded carbon steel	180	200	23,300	2	576.0	593.8	48,000	0.88	90,200
LP suction scrubber	Vertical separator vessel, 3.5 ft ID × 10 ft S/S length, cladded carbon steel	45	200	30,100	1	576.0	593.8	31,000	0.88	58,300
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19 ft S/S length, cladded carbon steel	45	200	68,300	1	576.0	593.8	70,400	0.88	132,400
Demulsifier	Horizontal 3-phase separator vessel, 4 ft × 13 ft S/S length, cladded carbon steel	45	200	37,500	1	576.0	593.8	38,700	0.88	72,800
Chemical injection pump	42 gal/d to give 1,000 ppm by volume in oil	180	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	750-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	1,800	1	404.0	593.8	2,600	0.88	4,900
Oil storage tank	API storage tanks at 4,000-bbl capacity each (for 3-d capacity at peak rate, 2,737 bbl)	Hydrostatic + 2.5 psig	150	122,600	1	576.0	593.8	126,400	0.88	237,600
Water storage tank	API storage tanks at 7,500-bbl capacity each, carbon steel (for 3-d capacity at peak rate, 14,491 bbl)	Hydrostatic + 2.5 psig	150	186,600	2	576.0	593.8	384,700	0.88	723,200
CO_2 compressor–1st phase	2.5 MMscfd from 150 to 500 psig (187 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	545,000	1	576.0	593.8	561,800	0.49	837,100
CO_2 compressor–2nd and 3rd phases	4 MMscfd from 150 to 500 psig (299 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	618,800	2	576.0	593.8	1,275,800	0.49	1,900,900
Dehydration	Treats 10.5 MMscfd after compression at 500 psig	600	150	454,600	1	593.8	593.8	454,600	0.72	781,900
Building		NA	NA	309,800	1	593.8	593.8	309,800	0.00	309,800
Total								3,433,500		5,383,000

[1]Production rates: 12,390 Sm^3/h (standard cubic meters per hour; 10.5 MMscfd [million standard cubic feet per day]) of CO_2, 145 m^3/d of oil (912 bopd [barrels of oil per day]), 768 m^3/d of water (4,830 bwpd [barrels of water per day]). Pressures: 1,034 kPag (kilopascal gauge; 150 psig [pounds per square inch gauge]) facility inlet, 3,448 kPag (500 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B10 Case 10—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	180	200	86,800	1	576.0	593.8	89,500	0.88	168,300
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	180	200	22,400	1	576.0	593.8	23,100	0.88	43,400
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 9 ft S/S length, cladded carbon steel	180	200	23,300	2	576.0	593.8	48,000	0.88	90,200
LP suction scrubber	Vertical separator vessel, 3.5 ft ID × 10 ft S/S length, cladded carbon steel	45	200	30,100	1	576.0	593.8	31,000	0.88	58,300
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19 ft S/S length, cladded carbon steel	45	200	68,300	1	576.0	593.8	70,400	0.88	132,400
Demulsifier	Horizontal 3-phase separator vessel, 4 ft × 13 ft S/S length, cladded carbon steel	45	200	37,500	1	576.0	593.8	38,700	0.88	72,800
Chemical injection pump	42 gal/d to give 1,000 ppm by volume in oil	180	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	750-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	1,800	1	404.0	593.8	2,600	0.88	4,900
Oil storage tank	API storage tanks at 4,000-bbl capacity each (for 3-d capacity at peak rate, 2,737 bbl)	Hydrostatic + 2.5 psig	150	122,600	1	576.0	593.8	126,400	0.88	237,600
Water storage tank	API storage tanks at 7,500-bbl capacity each, carbon steel (for 3-d capacity at peak rate, 14,491 bbl)	Hydrostatic + 2.5 psig	150	186,600	2	576.0	593.8	384,700	0.88	723,200
CO_2 compressor–1st phase	2.5 MMscfd from 150 to 1,000 psig (283 hp); 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	613,500	1	576.0	593.8	632,500	0.49	942,400
CO_2 compressor–2nd and 3rd phases	4 MMscfd from 150 to 1,000 psig (453 hp); 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	766,700	2	576.0	593.8	1,580,800	0.49	2,355,400
Dehydration	Treats 10.5 MMscfd after compression at 1,000 psig	1,100	150	363,700	1	593.8	593.8	363,700	0.72	625,600
Building		NA	NA	309,800	1	593.8	593.8	309,800	0.00	309,800
Total								3,718,300		5,786,500

[1]Production rates: 12,390 Sm^3/h (standard cubic meters per hour; 10.5 MMscfd [million standard cubic feet per day]) of CO_2, 145 m^3/d of oil (912 bopd [barrels of oil per day]), 768 m^3/d of water (4,830 bwpd [barrels of water per day]). Pressures: 1,034 kPag (kilopascal gauge; 150 psig [pounds per square inch gauge]) facility inlet, 6,895 kPag (1,000 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B11 Case 11—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	365	200	119,600	2	576.0	593.8	246,600	0.88	463,600
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	365	200	26,700	1	576.0	593.8	27,500	0.88	51,700
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 10 ft S/S length, cladded carbon steel	365	200	29,100	2	576.0	593.8	60,000	0.88	112,800
LP suction scrubber	Vertical separator vessel, 4 ft ID × 11.5 ft S/S length, cladded carbon steel	45	200	34,300	1	576.0	593.8	35,400	0.88	66,600
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19.5 ft S/S length, cladded carbon steel	45	200	69,000	2	576.0	593.8	142,300	0.88	267,500
Demulsifier	Horizontal 3-phase separator vessel, 5 ft × 15 ft S/S length, cladded carbon steel	45	200	45,700	1	576.0	593.8	47,100	0.88	88,500
Chemical injection pump	84 gal/d to give 1,000 ppm by volume in oil	365	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	1,500-gal tank for 14 days of storage	Hydrostatic + 2.5 psig	150	$3,600	1	404.0	593.8	5,300	0.88	10,000
Oil storage tank	API storage tank 4,000-bbl capacity each (for 3-d capacity at peak rate, 5,479 bbl)	Hydrostatic + 2.5 psig	150	122,600	2	576.0	593.8	252,800	0.88	475,300
Water storage tank	API storage tank 10,000 bbl capacity each, carbon steel (for 3-d capacity at peak rate, 67,686 bbl)	Hydrostatic + 2.5 psig	150	230,100	3	576.0	593.8	711,600	0.88	1,337,800
CO_2 compressor–1st phase	4 MMscfd from 315 to 1,000 psig (281 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	580,400	1	576.0	593.8	598,300	0.49	891,500
CO_2 compressor–2nd and 3rd phases	8 MMscfd from 315 to 1,000 psig (563 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	799,900	2	576.0	593.8	1,649,200	0.49	2,457,300
LP compressor	1 MMscfd from 25 to 315 psig (164 hp); 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	583,700	1	576.0	593.8	601,700	0.49	896,500
Dehydration	Treats 21 MMscfd after compression at 1,000 psig	1,100	150	551,200	1	593.8	593.8	551,200	0.72	948,100
Building		NA	NA	520,000	1	593.8	593.8	520,000	0.00	520,000
Total								5,466,100		8,609,400

[1]Production rates: 24,780 Sm^3/h (standard cubic meters per hour; 21.0 MMscfd [million standard cubic feet per day]) of CO_2, 290 m^3/d of oil (1,826 bopd [barrels of oil per day]), 1,537 m^3/d of water (9,669 bwpd [barrels of water per day]). Pressures: 2,172 kPag (kilopascal gauge; 315 psig [pounds per square inch gauge]) facility inlet, 3,448 kPag (500 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

Table B12 Case 12—Major equipment list, and purchased and installed costs[1]

Name	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to May 2012, $	Number of units	Plant index equipment basis	May 2012 plant index	May 2012 total purchased equipment cost, $	Installation factor	Total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 6.5 ft ID × 20 ft S/S length, cladded carbon steel	365	200	119,600	2	576.0	593.8	246,600	0.88	463,600
HP suction scrubber–1st phase	Vertical separator vessel, 3 ft ID × 8 ft S/S length, cladded carbon steel	365	200	26,700	1	576.0	593.8	27,500	0.88	51,700
HP suction scrubber–2nd and 3rd phases	Vertical separator vessel, 3 ft ID × 10 ft S/S length, cladded carbon steel	365	200	29,100	2	576.0	593.8	60,000	0.88	112,800
LP suction scrubber	Vertical separator vessel, 4 ft ID × 11.5 ft S/S length, cladded carbon steel	45	200	34,300	1	576.0	593.8	35,400	0.88	66,600
Free water knockout	Horizontal 3-phase separator vessel, 6.5 ft × 19.5 ft S/S length, cladded carbon steel	45	200	69,000	2	576.0	593.8	142,300	0.88	267,500
Demulsifier	Horizontal 3-phase separator vessel, 5 ft × 15 ft S/S length, cladded carbon steel	45	200	45,700	1	576.0	593.8	47,100	0.88	88,500
Chemical injection pump	84 gal/d to give 1,000 ppm by volume in oil	365	150	8,300	2	576.0	593.8	17,100	0.30	22,200
Chemical injection tank	1,500-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	3,600	1	404.0	593.8	5,300	0.88	10,000
Oil storage tank	API storage tank 4,000-bbl capacity each (for 3-d capacity at peak rate, 5,479 bbl)	Hydrostatic + 2.5 psig	150	122,600	2	576.0	593.8	252,800	0.88	475,300
Water storage tank	API storage tank 10,000-bbl capacity each, carbon steel (for 3-d capacity at peak rate, 67,686 bbl)	Hydrostatic + 2.5 psig	150	230,100	3	576.0	593.8	711,600	0.88	1,337,800
CO_2 compressor–1st phase	4 MMscfd from 315 to 1,000 psig (281 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	580,400	1	576.0	593.8	598,300	0.49	891,500
CO_2 compressor–2nd and 3rd phases	8 MMscfd from 315 to 1,000 psig (563 hp); single-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	799,900	2	576.0	593.8	1,649,200	0.49	2,457,300
LP compressor	1 MMscfd from 25 to 315 psig (164 hp); 2-stage reciprocating compressor; assumes mix of cladded carbon steel and stainless steel components	NA	NA	583,700	1	576.0	593.8	601,700	0.49	896,500
Dehydration	Treats 21 MMscfd after compression at 1,000 psig	1,100	150	551,200	1	593.8	593.8	51,200	0.72	948,100
Building		NA	NA	520,000	1	593.8	593.8	$520,000	0.00	520,000
Total								5,466,100		8,609,400

[1]Production rates: 24,780 Sm^3/h (standard cubic meters per hour; 21.0 MMscfd [million standard cubic feet per day]) of CO_2, 290 m^3/d of oil (1,826 bopd [barrels of oil per day]), 1,537 m^3/d of water (9,669 bwpd [barrels of water per day]). Pressures: 2,172 kPag (kilopascal gauge; 315 psig [pounds per inch gauge]) facility inlet, 6,895 kPag (1,000 psig) facility outlet. HP, high pressure; LP, low pressure; ID, inside diameter; S/S, seam to seam.

APPENDIX C
NATURAL GAS LIQUID RECOVERY STUDY

The potential for natural gas liquid (NGL) recovery from the produced gas from enhanced oil recovery (EOR) CO_2 streams in the Illinois Basin (ILB) was evaluated. The NGL recovery study was first performed as a stand-alone task using a produced gas flow rate of 5,900 Sm^3/h (5 MMscfd [million standard cubic feet per day]). The results of the initial NGL study were then used to evaluate the economics for the cases outlined in Table 1. Natural gas liquids are considered C_{3+} components; methane and ethane separation from the produced CO_2 was not a goal of the evaluation. This section discusses the reasons for implementing or not implementing NGL recovery at a CO_2 EOR facility. An overview of the technologies that can be used to recover NGLs from high-CO_2 streams is also presented. Finally, the economics for NGL recovery with the EOR surface facility cases are presented and a sensitivity analysis is performed to define conditions in which NGL recovery would be favorable on EOR CO_2 streams in the ILB. This study is meant to be a high-level investigation into the potential feasibility of NGL recovery in the ILB. More detailed analysis and NGL technology selection may be warranted depending on the specific operating conditions and reservoir performance for actual EOR sites in this area.

Reasons for Natural Gas Liquid Recovery

Natural gas liquids are recovered on an EOR produced gas CO_2 stream for two main reasons. The first is to sell the recovered NGLs for profit. The second is to reduce the hydrocarbon content to a level at which it will not cause issues upon reinjection into the reservoir with recycled CO_2. The presence of methane in the CO_2 stream can raise the compressor power requirements and minimum miscibility pressure (MMP) of the crude oil. Noncondensable gases, such as nitrogen, in recycled CO_2 streams can also raise the MMP and may be removed from the CO_2 concurrently during NGL recovery (depending on the nitrogen concentration and the NGL recovery process selected). A general rule of thumb is that the hydrocarbon content needs to be at least 10 mol. % in the CO_2 stream to justify recovery, and even then it may be driven by the reinjection gas specifications as opposed to the economics of NGL recovery alone. The total hydrocarbons in the produced gas typically start at 95 mol. % when CO_2 injection is initiated but may drop to less than 10 mol. % soon after the start of CO_2 injection and to very low levels at the end of the field life. During initial start-up, the produced gas flow rate is very low, such that it may be vented, flared, or combined with a substantial amount of pipeline-quality CO_2 for reinjection so that the combined gas going into the injection wells will have a hydrocarbon content well below 10 mol. %.

It may be possible to recover NGLs for sale, but the combination of the assumed lean C_{3+} content and low produced gas flow rates for the EOR cases considered in this report may make it uneconomical. Natural gas liquids are recovered in some West Texas CO_2 EOR produced gas streams, but the produced gas in that area is typically considered very rich.

Typical Natural Gas Liquid Recovery Technologies for High CO_2 Content Streams

Many different technologies are available to recover NGLs. The technologies used depend on which components need to be separated (e.g., C_1/C_2 removal or only C_{3+} separation) as well as the necessary recovery efficiencies for the process. Background on different NGL recovery processes is provided below to facilitate discussion of the rationale for selecting the NGL recovery technology evaluated.

Refrigeration with Stabilizer

Mechanical refrigeration with a stabilizer is often a good NGL recovery option for CO_2 streams. Several of these plants in operation in West Texas economically achieve C_{3+} recovery from CO_2 gas streams. The process is relatively simple. The inlet gas is dehydrated to a water dew point near –23 °C (–10 °F). The dehydrated gas then flows through a gas–gas heat exchanger and feeds into a stabilizer column. The overhead gas stream from the stabilizer column is chilled with refrigerant to partially condense the gas and provide some liquid reflux, which goes back to the top of the stabilizer column. At the column bottom is the C_{3+} NGL product. The net overhead stream is used to cool refrigerant in a heat exchanger, and the net overhead stream is then compressed and reinjected. The produced gas may need to be compressed to operate at the appropriate pressure for mechanical refrigeration with the stabilizer option for NGL recovery (~1,379 kPag [200 psig] plus), but the CO_2 stream has no significant pressure drop in the NGL recovery process. Typically, only moderate NGL recoveries are possible, and the amount recovered depends on the ratio of CO_2 to hydrocarbons. Expected typical recoveries are in the range of 20% to 40% C_3, 60% to 80% C_{4s}, and 90% C_{5+}. A typical process flow diagram is shown in Figure C1, although the refrigeration process has many variations (GPSA 2004b).

Ryan/Holmes

Four large Ryan/Holmes plants are located in West Texas. They can be designed to operate with two to four columns depending on whether separate CO_2, NGL, and hydrocarbon gas product streams (methane and ethane) are produced. The main benefit of Ryan/Holmes over other CO_2-NGL recovery options is that very high NGL recovery, including of C_2, is possible. The lean oil in Ryan/Holmes is used to break the CO_2–C_2 azeotrope. An azeotrope is formed when a mixture of components boils to produce a vapor with the same composition as the liquid, which prevents the use of distillation alone to separate the components. The typical operating pressure of the Ryan/Holmes plants is approximately 2,413 kPag (350 psig). The capital and operating costs for this process are high; it is therefore unlikely to be used to process gas streams at production rates of 24,780 Sm^3/h (21 MMscfd) and less. Thus, further analysis of the Ryan/Holmes process was not part of this report. Figure C2 shows a flow diagram of the Ryan/Holmes process (GPSA 2004b).

Lean Oil Absorption

This process would technically be applicable for CO_2 gas streams. However, most lean oil plants have been shut down or replaced with more modern straight

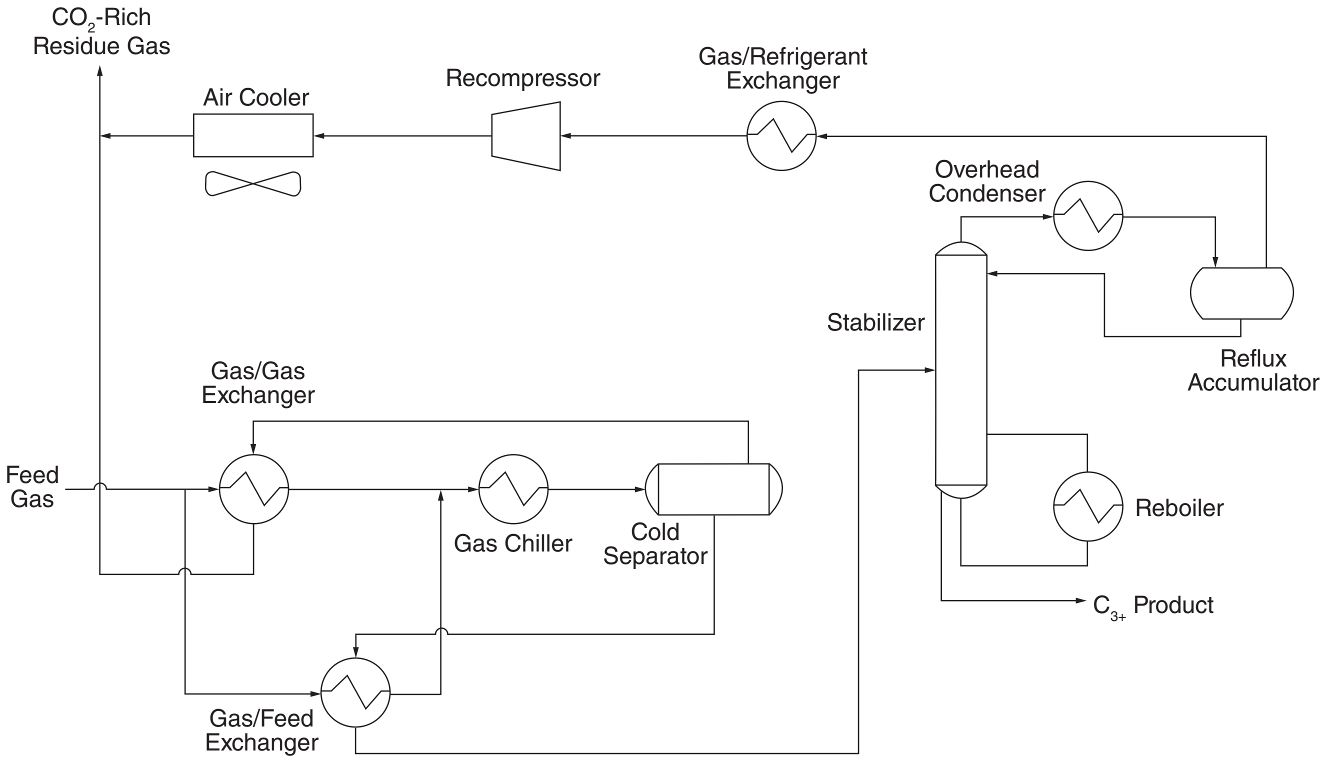

Figure C1 Example mechanical refrigeration system with stabilizer (GPSA, 2004b).

refrigeration plants. The lean oil process requires large processing equipment and equipment with high energy requirements. For these reasons, the lean oil absorption process would be less attractive than refrigeration with stabilization and was not evaluated further in this report.

Joule–Thomson Processes

These processes use the Joule–Thomson effect to reduce the gas temperature upon expansion from high pressure (often greater than 6,895 kPag [1,000 psig]). Such processes would not be a good fit for this application with a high CO_2 content in the feed stream because, on expansion to attain moderate C_{3+} recovery, CO_2 condensation would be very high. Stripping the CO_2 from the condensed liquids would result in high NGL losses, so the overall recovery would be low. The refrigeration–stabilization scheme avoids this problem by chilling the column overhead stream, causing the separation by liquefying the NGL C_{3+} product in the column, and leaving essentially all the CO_2 in the gas stream exiting the column rather than partially condensing the hydrocarbons and CO_2 and then stripping out the CO_2.

Membranes

Membranes have been used in conjunction with other processes for NGL recovery. The operating pressure needs to be high enough to provide the driving force for separation of the bulk CO_2 stream from the NGL-rich stream (on the order of 2,758 kPag [400 psig]). Here, the low-pressure, CO_2-rich permeate is compressed and reinjected, and the low CO_2/hydrocarbon-rich stream can be processed in an NGL facility. Recompression costs for the low-pressure, CO_2-rich stream can be significant. Because membranes can be installed at any capacity and are easily scalable, they may be applied for relatively low production rates (24,780 Sm^3/h [21 MMscfd] or less), but probably only if an NGL facility were located nearby to process the hydrocarbon-rich gas to recover the NGL product. A significant part of the membrane cost is pretreatment of the inlet gas stream (i.e., water and hydrocarbon dew point control, depending on the type of membrane used).

Amine or Other Solvent Processes

Some of the three-column Ryan/Holmes plants in West Texas have used absorption of CO_2 by amine solvents followed by recompression. The two products are fairly pure CO_2 and low-CO_2/hydrocarbon-rich gas, from which NGLs could then be recovered. As with membranes, the low-pressure CO_2 would need to be recompressed, and the cost of this recompression could be high. Other combination processes for NGL recovery from CO_2 streams likely exist, but identifying all the possible combinations is beyond the scope of this initial NGL evaluation.

Natural Gas Liquid Recovery Economics for Enhanced Oil Recovery Surface Facility Cases

Mechanical refrigeration with a stabilizer is a relatively simple, straightforward process commonly used for NGL recovery. Like some of the other technologies, the produced CO_2 gas stream needs to operate at pressures of approximately 1,379 kPag (200 psig or higher), which is why the process is located downstream from

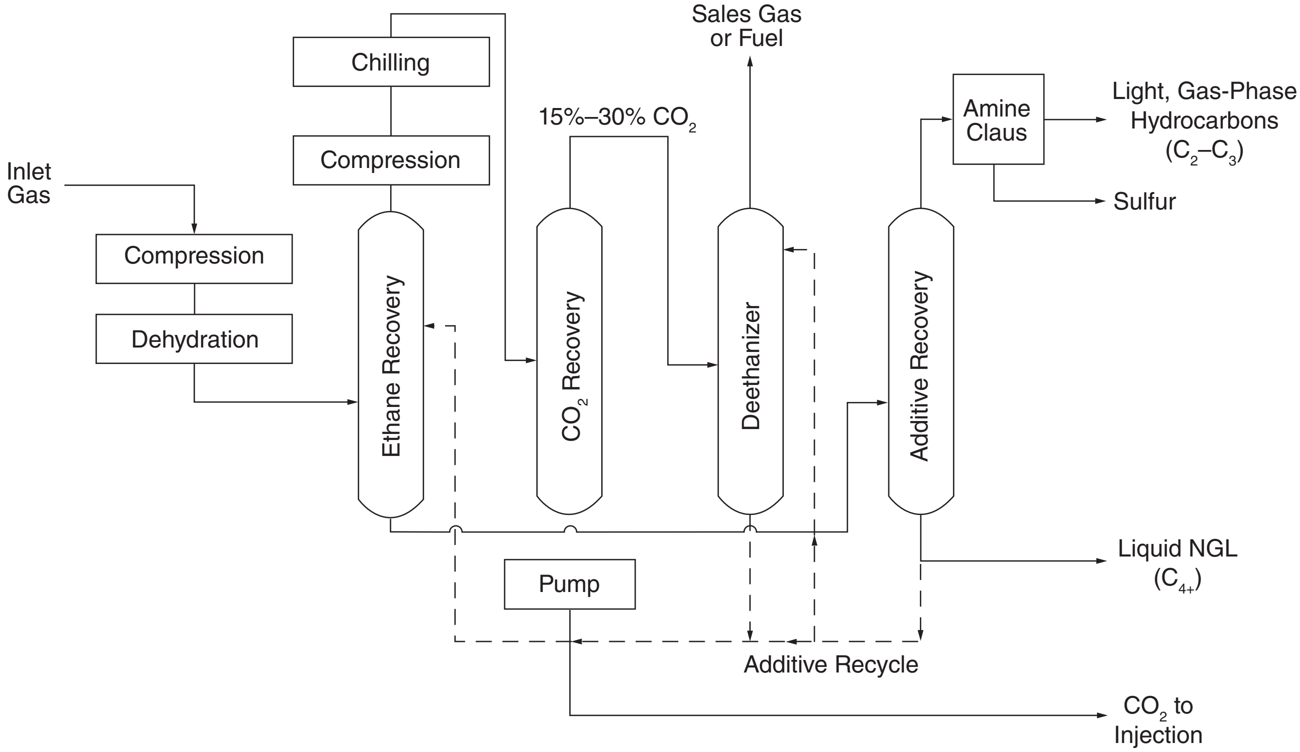

Figure C2 Typical Ryan/Holmes process flow diagram (GPSA 2004b).

compression for Cases 9–12. However, because the CO_2 stream has little to no pressure loss in the refrigeration process, the extra compression that would be needed with the Joule–Thomson, membrane, and amine processes is not required for mechanical refrigeration with a stabilizer. Thus, this refrigeration-based approach should give a reasonable approximation of a low-cost option for NGL recovery in this application and is therefore used as the basis for the economic evaluation. A gas production rate of 5,900 Sm^3/h (5 MMscfd) was used for the initial NGL feasibility study and has been carried through in this evaluation. The capital and operating costs for the initial 5,900 Sm^3/h (5 MMscfd) NGL recovery case did not include CO_2 compression and dehydration. However, these costs were included in Cases 9–12. The NGL results of the 5,900 Sm^3/h (5 MMscfd) study were then adjusted for comparison with Cases 9–12, as summarized in Table 1. The same gas composition over the lifetime of the facility was assumed for the 5,900 Sm^3/h (5 MMscfd) study, such that the recovered NGLs from the 5,900 Sm^3/h (5 MMscfd) case could be scaled up to the 12,390 and 24,780 Sm^3/h (10.5 and 21 MMscfd) cases in this report.

Mechanical refrigeration with a stabilizer on a gas stream with less than 0.13 L/m^3 (1.0 GPM [gallons of recoverable hydrocarbons in the gas per thousand standard cubic feet of gas], or less than 3.0 mol. %) of C_{3+} components is generally not effective because lower temperatures are required to recover NGLs from this stream, so a cryogenic refrigeration process would be required. However, it is unlikely to be economically practical in current economic and market conditions to construct and operate cryogenic NGL recovery plants on 5,900 Sm^3/h (5 MMscfd) gas streams with only 0.13 L/m^3 (1 GPM) of hydrocarbons in the feed stream. The minimum-size cryogenic NGL plant is typically approximately 29,500 Sm^3/h (25 MMscfd).

The gas composition assumed for this study is too lean for practical, economic NGL recovery by any process known to Trimeric at this time. Notwithstanding these limitations, an example case is evaluated below to provide representative economics for a hypothetical NGL recovery system on a lean produced gas. It was assumed that an NGL technology with a low treatment cost (similar to mechanical refrigeration with stabilization) could be used and that it would achieve a 52% recovery of total C_{3+} hydrocarbons. Also presented is a case in which 100% recovery is assumed, to illustrate the economic feasibility for NGL recovery at "best case" conditions for the lean produced gas used for this study in the ILB. Except where specifically mentioned, the results in the following paragraphs are based on the 52% NGL recovery case.

Peak NGL production of 2.4 m^3/d (15.3 bbl/d) was assumed to occur in Year 5, with approximately 3,540 Sm^3/h (3 MMscfd) of produced gas flow in Year 5 and 5,900 Sm^3/h (5 MMscfd) of peak gas production occurring in Year 20. The recovered NGL components were speciated into C_3, iC_4/nC_4, and C_{5+} by using data found in the literature (Vargas 2010). On the basis of this reference, the NGL was estimated to be composed of 45.8% propane, 29.5% butanes, and 24.6% C_{5+} on a per-mole basis. The value of the NGL products was also obtained from publicly available literature at \$46.90/bbl for C_3, \$63.60/bbl for C_{4s}, and \$54.80/bbl for

C_{5+}. Thus, at peak NGL production, the total annual value of the recovered NGLs would be $160,000 based on 52% NGL recovery (7.9 bbl/d).

Capital costs for the NGL refrigeration process were obtained from another reference for natural gas processing (Tannehill 2009). However, NGL recovery from CO_2 gas streams is much more difficult and more expensive than recovery from natural gas. For this reason, the costs from the literature source were scaled to match economic data for a large-scale NGL refrigeration application on CO_2 gas streams. The cost estimate for the refrigeration process includes the facilities to produce a single C_{3+} product with no fractionation and limited storage. The C_{3+} product was assumed to be trucked for off-site fractionation into C_3, C_{4s}, and C_{5+} components.[2] Glycol injection with regeneration of the glycol is included to prevent hydrate formation. Precompression or other treatment of the inlet gas is not included. The estimated fixed capital investment of the refrigeration plant with the capacity to handle peak NGL production is $3.6 MM. The basic operating expenses for the refrigeration plant include the electricity for the refrigerant propane compressors and the heat needed in the stabilizer reboiler and in the glycol regenerator. The cost of electricity was assumed to be $0.08/kWh and the fuel gas cost was assumed to be $7.5/MMBtu. The total operating cost for this process is approximately $58,000/yr at peak NGL production conditions. Additional operating labor beyond that required to support the CO_2 compression and dehydration facility was not included in the estimated operating costs. This evaluation approach gives a simple payback period of approximately 34 yr at peak NGL production conditions. In comparison, a 2-yr payback period for equipment and project selection is a common target for operators in the oil and gas industry. To have a 2-yr payback period with a CO_2 stream this lean in NGL content, the produced gas flow rate would need to be approximately 684,400 Sm^3/h (580 MMscfd), which is 115 times the 5,900 Sm^3/h (5 MMscfd) base case produced gas flow rate at peak NGL production. In the example case, NGL production peaks when the produced gas flow is 3,540 Sm^3/h (3 MMscfd; with 0.7 mol. % NGL), and the total produced gas flow is 5,900 Sm^3/h (5 MMscfd) at the end of a hypothetical 20-yr period of CO_2 injection. Even at 100% NGL recovery, the payback period would be 15 yr. This scenario represents the ideal situation, in which a low-cost NGL technology (with costs similar to mechanical refrigeration with stabilization) could be used to attain a moderate to high NGL recovery efficiency on a very lean CO_2 gas stream; in all likelihood, the capital and operating costs of a viable NGL recovery process would be considerably higher than refrigeration, making the economics even less favorable.

A third economic example of a CO_2 stream that is rich in NGLs is presented. The gas stream was assumed to contain 86.4 mol. % CO_2, 3.5 mol. % C_1, 2.5 mol. % C_2, 4.1 mol. % C_3, 2.3 mol. % C_{4s}, and 1.2 mol. % C_{5+}. The total NGL content of the gas (C_{3+}) is 7.6 mol. % (2.3 GPM [0.31 L/m^3]). A refrigeration with stabilizer unit could be used with this level of NGLs in the gas to yield approximately 52% overall C_{3+} recovery. At the same produced gas flow rate used previously (5,900 Sm^3/h [5 MMscfd]), this would be 25.8 m^3/d (162 bbl/d) of NGLs versus the 2.4 m^3/d (15.3 bbl/d) of NGLs at peak conditions for the leaner assumed base case. The NGL plant would cost more ($6.4 million), but the C_{3+} would provide a net revenue of $1.39 million/yr after accounting for operating expenses, giving a payback of approximately 4.6 yr. Other NGL contents were also evaluated, as shown in Table C1. This table shows the impact of the lean NGL content in the produced gas anticipated from ILB EOR facilities on project economics for NGL recovery.

Table C1 Example sensitivity study on natural gas liquid recovery economics[1]

Gas flow rate, Sm^3/h (MMscfd)	Recoverable hydrocarbons in gas, L/m^3 (gal/Mscf or GPM)	Payback period, yr	Gas flow rate for 2-yr payback period, Sm^3/h (MMscfd)
5,900 (5)	0.03 (0.22)	34.0	649,000 (550)
5,900 (5)	0.10 (1)	12.5	296,180 (251)
5,900 (5)	0.31 (2.3)	4.6	35,400 (30)
5,900 (5)	0.50 (4)	2.5	9,440 (8)
5,900 (5)	0.70 (5)	2.0	5,900 (5)

[1]Sm^3/h standard cubic meters per hour; MMscfd, million standard cubic feet per day; gal/Mscf, gallons per thousand standard cubic feet (GPM).

[2]The individual NGL components have a higher unit value than the blended NGLs; however, fractionation plants are typically installed only at larger natural gas processing facilities.