General Large-Scale Carbon Dioxide Enhanced Oil Recovery Facility Design for the Illinois Basin

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Front cover: *Typical vertical vessel high-pressure suction scrubber and compressor. Typical low-pressure suction scrubbers and compressors are similar in appearance. Photograph courtesy of Denbury Onshore.*

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Technical Report

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EXECUTIVE SUMMARY

The Midwest Geological Sequestration Consortium (MGSC) is leading a program to demonstrate the feasibility of carbon dioxide (CO_2) capture and storage, particularly in the Illinois Basin (ILB). One potential storage method uses CO_2 for enhanced oil recovery (EOR) by injecting it into 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 MGSC, 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:

- Determining the equipment that would be required for typical facilities;
- Identifying capacity breakpoints in the major equipment (i.e., 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 conceptual facility designs included the equipment required to separate produced liquids from the CO_2, storage and disposal of the produced liquids, and compression of the CO_2 to be reinjected. The current evaluation included CO_2 recycle rates ranging from 59,000 to 236,000 Sm^3/h (standard cubic meters per hour; 50 to 200 MMscfd [million standard cubic feet per day]) with facility inlet pressures of 1,034 and 2,172 kPag (kilopascal gauge; 150 and 315 psig [pounds per square inch gauge]) and a facility discharge pressure of 6,895 kPag (1,000 psig). An initial study performed in 2013 similarly evaluated EOR surface facilities with CO_2 flow rates ranging from 1,180 to 24,780 Sm^3/h (1 to 21 MMscfd) with the same facility inlet pressures of 1,034 and 2,172 kPag (150 and 315 psig) and with discharge pressures of 3,448 and 6,895 kPag (500 and 1,000 psig).

The feasibility of NGL recovery of propane and heavier (C_{3+}) components was also assessed in the 2013 evaluation for the smaller facilities. The 2013 evaluation concluded that the lean produced gas anticipated in the ILB would likely require costly cryogenic processing to achieve significant NGL recovery, and thus would be uneconomical. If the recycled CO_2 from actual operating EOR facilities in the ILB is eventually found to be richer in NGL components than originally expected, the economic feasibility of NGL recovery can be reevaluated. Lower crude oil prices observed in 2015 and early 2016 would also impede the implementation of NGL recovery.

The purchased equipment costs for the small-scale EOR facilities without NGL recovery were estimated to range from approximately \$1 million for the case with a 1,200 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 24,800 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 facility infrastructure in addition to the EOR equipment.

After the 2013 evaluation was completed, the MGSC requested that additional cases be evaluated for larger scale EOR applications without NGL recovery. In this study, EOR surface facility cases were evaluated with CO_2 recycle rates ranging from 59,000 to 236,000 Sm^3/h (50 to 200 MMscfd) at the same two facility inlet pressures as the smaller cases, but with only one discharge pressure of 6,395 kPag (1,000 psig). The estimated purchased equipment costs for the large-scale EOR facilities ranged from \$6.7 million for the case with a 59,000 Sm^3/h (50 MMscfd) CO_2 rate and 2,172 kPag (315 psig) inlet pressure up to \$27.2 million for the case with a 236,000 Sm^3/h (200 MMscfd) CO_2 rate and an inlet pressure of 1,034 kPag (150 psig). The estimated FCIs were approximately \$20 million and \$81.6 million for the same cases, respectively.

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). In this report, the design and costs are evaluated for *large-scale* surface facility 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 large-scale EOR CO_2 recycle rates considered in this work ranged from 59,000 to 236,000 Sm^3/h (standard cubic meters per hour; 50 to 200 MMscfd [million standard cubic feet per day]).

Previous work performed in 2013 evaluated smaller scale EOR facilities with CO_2 recycle rates from 1,200 to 24,800 Sm^3/h (1 to 21 MMscfd). In the 2013 evaluation, the natural gas liquid (NGL) recovery for propane and heavier (C_{3+}) compounds was also assessed, but was found to be uneconomical because the produced gas from the ILB is expected to be rather lean (0.03 L/m^3 or 0.22 GPM [gallons of recoverable hydrocarbon NGL per thousand standard cubic feet of gas; see Myers et al. 2017, *A Universal Methodology to Develop, Test, and Calibrate a Carbon Dioxide Enhanced Oil Recovery and Storage Capacity Estimate*]). More expensive NGL recovery processes, such as cryogenic technologies, would likely be required to recover substantial amounts of NGL from the lean gas, and the costs of such technologies would be prohibitive. Recycled CO_2 gas containing at least 0.66 to 0.92 L/m^3 (5 to 7 GPM) would likely be required to make the economics of NGL recovery potentially feasible. Details of the small-scale EOR evaluation and NGL study can be found in the final version of that report (Trimeric Corporation 2016).

The objective of this report is to provide information and calculation tools that could be used to help determine the feasibility of implementing *large-scale* 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 of the CO_2 delivery pipeline, injection wells, and production wells are not included in this evaluation, 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 covered in this report. The process configurations and costs provided in this report are intended as examples that are representative of typical large-scale 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 large-scale 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) 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 large-scale 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 facility for this large-scale evaluation.
- 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.
- Estimate the fixed capital investment (FCI) for these surface processing facilities.
- Provide information needed for any further economic analysis related to the surface processing facilities, including the following:
  - Unit operating costs
    - Electricity (kWh)
    - 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 (emulsion breakers)
    - Number of operators and labor costs
    - Maintenance costs (spare parts)
    - Consumable costs (compressor lubrication oil, filters)
  - Annual operating costs

The NGL recovery was not evaluated for large-scale EOR facilities in this report because during the 2013 EOR surface facility evaluation, it was found to be impractical for the lean produced gas expected from ILB oil fields, even at the higher CO_2 recycle rates considered for the large-scale EOR facilities.

Description of Cases

A list of cases was developed to cover the range of conditions (i.e., gas, oil, and water production rates, facility inlet pressure, and facility outlet pressure) anticipated for large-scale EOR facilities in the ILB. Table 1 shows the six cases selected for this evaluation. The 1,034 and 2,172 kPag (kilopascal gauge; 150 and 315 psig [pounds per square inch gauge]) facility inlet pressures were selected to show the impact of suction pressure on the compression costs required to achieve the same 6,895 kPag (1,000 psig) discharge pressure. The ISGS provided the facility outlet (injection) pressure of 6,895 kPag (1,000 psig) based on the anticipated miscible CO_2 flood surface and bottomhole pressure requirements. The temperature

Table 1 Case summary for the large-scale enhanced oil recovery surface facility study[1]

Case	Peak gas production rate, Sm^3/h (MMscfd)	Peak gas production rate, m^3/min (acfm)	Peak water production rate, m^3/d (bbl/d)	Peak oil production rate, m^3/d (bbl/d)	Facility inlet pressure, kPag (psig)	Facility outlet pressure, kPag (psig)	High-pressure compressor installation phases, Sm^3/h (MMscfd)
1	59,000 (50)	89 (3,140)	1,908 (12,000)	477 (3,000)	1,034 (150)	6,895 (1,000)	Single, 2 × 29,500 (25)
2	59,000 (50)	42 (1,470)	1,908 (12,000)	477 (3,000)	2,172 (315)	6,895 (1,000)	Single, 2 × 29,500 (25)
3	118,000 (100)	178 (6,280)	3,816 (24,000)	954 (6,000)	1,034 (150)	6,895 (1,000)	Multiple, 2 × 29,500 (25), 1 × 59,000 (50)
4	118,000 (100)	83 (2,930)	3,816 (24,000)	954 (6,000)	2,172 (315)	6,895 (1,000)	Multiple, 2 × 29,500 (25), 1 × 59,000 (50)
5	236,000 (200)	356 (12,570)	7,632 (48,000)	1,908 (12,000)	1,034 (150)	6,895 (1,000)	Multiple, 2 × 29,500 (25), 3 × 59,000 (50)
6	236,000 (200)	166 (5,860)	7,632 (48,000)	1,908 (12,000)	2,172 (315)	6,895 (1,000)	Multiple, 2 × 29,500 (25), 3 × 59,000 (50)

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

for 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 to cool the gas leaving the CO_2 compressors. Details on fluid temperatures are not addressed in this report.

The selected EOR CO_2 production (recycle) range was 59,000 to 236,000 Sm^3/h (50 to 200 MMscfd), the assumed minimum and maximum CO_2 production rates for large-scale facilities in the ILB. The gas was assumed to be mostly CO_2 but also to contain hydrocarbons, as shown in Table 2.

The first 59,000 Sm^3/h (50 MMscfd) of recycled CO_2 gas flow in each case will be processed with two 29,500 Sm^3/h (25 MMscfd) compressors operating in parallel. This setup provides additional operational flexibility at reduced throughput conditions as compared with installing one 59,000 Sm^3/h (50 MMscfd) compressor. Afterward, each additional increment of 59,000 Sm^3/h (50 MMscfd) will be processed with one additional 59,000 Sm^3/h (50 MMscfd) compressor. As the production rate increases over several years, additional compressors and other equipment will be added to accommodate increases in produced gas and oil rates. Water production rates are typically at their highest right after changing from waterflood to CO_2 flood operations, so it has been assumed in this report that an existing waterflood field in the ILB would already have adequate water processing, storage, and disposal equipment before beginning a CO_2 flood.

The central facility phased approach of installing additional, nearly identical sets of equipment, each with a 59,000 Sm^3/h (50 MMscfd) capacity per phase, as the CO_2 recycle and oil production rates increase over the life of the EOR flood is an approach that is often used in CO_2 EOR operations. Deployment of the central facility components in phases is sometimes related to development of the oil field in EOR flood phases. The central facility phased approach might result in a somewhat higher overall total cost at the end of facility build-out to full capacity as compared with installing fewer pieces of equipment with larger unit capacities at the beginning of the operation that are capable of processing the ultimate expected facility CO_2 recycle, produced oil, and produced water flow rates. However, the advantage of delaying much of the capital expenditures by several years is often a tradeoff that favors a phased approach. The phased approach also provides operation of equipment closer to design capacities (avoiding high turndown operations with lower efficiencies) and reduces the risk of purchasing equipment or equipment capacity that might not be needed if actual EOR flood operations differ from original projections. It is beyond the scope of this report to evaluate the pros and cons of the phased approach that would likely be driven by field-specific considerations in any case. Nonetheless, it is important to point out that the phased approach selected by Trimeric for this evaluation and the resulting costs that basically scale linearly with throughput capacity might not reflect the approach that would be taken by an EOR flood operator for large-scale CO_2 recycle facility design in all cases.

Oil and Water Production Rate Assumptions

The oil and water production rates provided in Table 1 for the large-scale EOR cases were based on an 80% water/20% oil ratio on a barrels-per-day basis with the total liquid flow rate scaled based on the CO_2 per recycle rate (MMscfd). The oil and water production rates are based on recent ILB CO_2 EOR reservoir simulation estimates performed by the ISGS. 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.

The peak water capacity was based on a water-to-gas ratio of 0.00134 m^3/m^3 (0.24 bbl/Mscf [oilfield barrels/thousand standard cubic feet]), and the peak oil capacity assumes an oil-to-gas ratio of 0.00034 m^3/m^3 (0.06 bbl/Mscf). These ratios are approximately 52% and 69% of the values used with the small-scale EOR cases, respectively. Thus, less water and oil were assumed to be produced for the large-scale EOR facilities than would have been estimated if the same ratios had been used as with the small-scale applications in the previous study. The estimated oil storage requirements and water disposal costs were 48% and 31% lower, respectively, than if they had been estimated using the same water-to-gas and oil-to-gas ratios as in the small-scale study.

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

Table 2 Assumed peak characteristics of the produced gas[1]

Component	Value
Carbon dioxide	97.8 mol. %
Methane + ethane	1.5 mol. %
NGLs	0.7 mol. %
NGLs	0.03 L/m^3 (0.22 GPM)

[1]NGLs, natural gas liquids, i.e., 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).

subsurface operations, which depend 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 storage, discharge, or both for the liquids.
   a. To capture or treat low-pressure gas, if necessary, from flash 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 six cases listed in Table 1 are provided in Appendix A, and the detailed equipment design and cost estimates are described in the following section. In Appendix B, the individual facility component costs are listed in tables so that the impact on the overall cost of the facility of removing or adding a particular component can be evaluated.

EQUIPMENT DESIGN AND COST ANALYSIS SUMMARY

This section describes the general approach used to size and select surface equipment for the large-scale EOR facilities. Included here are the equipment capital costs and the anticipated fluid processing rates for the plants. Important differences from the small-scale EOR evaluation conducted previously are also noted. The economic results from the large-scale EOR study are also presented.

Equipment Sizing

The surface equipment for the large-scale 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 at the inlet to the facility. The gas exits the top of the vessel and flows to the high-pressure compressor train, whereas the oil and water exit in a combined stream to downstream separation vessels. The slug catcher operates at a pressure slightly lower than the pressure of the wellhead (1,034 and 2,172 kPag [150 and 315 psig], depending on the inlet pressure for each case). The pressure drop in the wellhead choke and in the gathering lines brings the fluids to the central facility. The slug catcher is typically a horizontal vessel, sized to have a length-to-diameter ratio of about 3 and a liquid residence time of 7.5 min.
- *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 as a gas and be sent to the low-pressure suction scrubber. The free water knockout is typically a horizontal vessel. A horizontal free water knockout is shown in Figure 1. In many parts of the United States, heat from burning natural gas, sometimes transferred to the free water knock-out via use of an intermediate heat transfer fluid, is used 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. The free water knockout vessel was sized by using an approach in the literature for three-phase separators (Monnery and Svrcek1994).
- *Demulsifier.* In this vessel, the water and oil phases separate because chemicals are added in the upstream process to break any oil–water emulsions. A small amount of CO_2 gas 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. Depending on site operations, the actual pressure drop might be in the range of 34.5 to 68.9 kPa (5 to 10 psi), but these differences will not affect these early phase designs and cost esti mates. The demulsifier vessel typically has a horizontal orientation and 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 natural gas, fuel gas, electricity, or some form of waste heat input is available, but discussions with those experienced in oilfield operations in the ILB suggest that a chemical separation approach is used almost exclusively in ILB oil production facilities (Ken Hake of Baker Hughes, personal communication, July 15, 2015). The demulsifier was sized to have a length-to-diameter ratio of approximately 3 and a residence time of 30 min. A longer residence time is used in this vessel to obtain a high degree of separation of the oil and water phases.
- *High-pressure suction scrubber*. This vertical vessel is used to prevent liquids from entering the compressor cylinders and is typically made of carbon steel with an internal corrosion-resistant coating or stainless steel. 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 as well as any atomized drops or carryover, and (2) to remove any slugs of liquid from the slug catcher in upset conditions or if unexpectedly high fluid volumes come to the facility. The high-pressure suction scrubber will operate at a facility inlet pressure of either 1,034 or 2,172

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 and compressors are similar in appearance. Photograph courtesy of Denbury Onshore.

kPag (150 or 315 psig). The vendor quotes that were used to estimate the purchased costs for the large-scale compressors in this report included this suction scrubber. Therefore, sizing or estimating the costs for this equipment was not needed.

- *Low-pressure suction scrubber*. This vertical vessel is used to prevent any liquids from entering the low-pressure compressor cylinders. The vessel is typically made of carbon steel with an internal corrosion-resistant coating or stainless steel. 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 used to feed these gases to the suction of the high-pressure compression system. In all six cases in this large-scale EOR evaluation, the flash gas rates were high enough to justify the addition of a low-pressure compressor or compressors. The low-pressure suction scrubber operates at a low pressure of approximately 165 kPag (~24 psig). This pressure might be slightly lower depending on the actual operating conditions, but these differences will not affect the early-phase designs and cost estimates.

The material of construction for the slug catcher, free water knockout, and demulsifier was assumed to be coated carbon steel. The dimensions of these separators were based on the produced-gas rate and the oil and water capacities for Cases 1 and 2. Multiple separators of the same size as those in Cases 1 and 2 were then used to handle the higher flow conditions for Cases 3 through 6.

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 a concentration of 90 ppmv (parts per million by volume) of demulsifier in the oil–water mixture, as recommended by Ken Hake as a starting point for use in this evaluation. However, actual oil and water samples and laboratory testing will be used to determine the optimal additive type(s) and concentration(s) for a specific application. The final separation of oil and water will occur in the demulsifier vessel. The demulsifier chemical storage tank was sized to hold a 14-d supply of demulsifier chemical. The demulsifier injection pump was sized to transfer the appropriate amount of chemical for Cases 1 and 2. Multiple chemical injection pumps and demulsifier storage tanks were assumed for Cases 3 through 6.

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 the field. The oil would be stored in tanks until it could be piped off-site. For Cases 1 and 2, it was assumed that 1,431 m^3 (9,000 bbl) American Petroleum Institute-style steel tanks would be used to hold 3 d of oil production at the peak capacity rate. Multiple oil storage tanks were assumed for Cases 3 through 6. The oil storage requirement may be less, depending on the sales options available at the site.

Water Storage Tank

For the large-scale EOR facilities, water storage tanks were excluded from the study. This was because the fields were assumed to have been converted from an existing waterflood operation and would therefore already have existing water storage and disposal equipment. Water production generally decreases during CO_2 EOR, so fields with an existing waterflood may not need new water storage or handling facilities. The existing water-handling equipment was assumed to be adequate.

Air Coolers

Air-cooled heat exchangers are used to remove the heat of compression from the CO_2 stream after each stage of compression in both the high- and low-pressure trains although, as mentioned, CO_2 EOR facilities at this large scale are likely to incorporate heat integration to help cool the compressed CO_2 and transfer the heat to improve fluid separations. To simplify this early-stage evaluation, all heat of compression was assumed to be removed by air coolers. The air-cooled heat exchangers were not included in the vendor quotes that were used to estimate the compressor costs for the large-scale EOR cases. Thus, the exchanger duties were estimated from modeling using the WinSim Design II software. The air coolers were assumed to have stainless steel material in the tubes and in other areas in contact with the wet CO_2 gas.

CO_2 Compressor Trains

The high-pressure and low-pressure CO_2 compressor trains were first modeled with WinSim's Design II software using the Peng–Robinson equation of state to obtain an initial estimate of the horsepower requirements. The ISGS had expressed an interest in identifying the highest possible single-compressor-unit throughput capacities for the two inlet pressures. Trimeric worked with two equipment supplier contacts, Jason Sowels at Reagan Power and Compression and Dave Morse at Dresser-Rand (personal communication, July 2015), to determine the maximum feasible compressor sizes for the 1,034 to 6,895 kPag (150 to 1,000 psig) and the 2,172 to 6,895 kPag (315 to 1,000 psig) compression ratios. On the basis of input from these highly experienced contacts, 59,000 Sm^3/h (50 MMscfd) was judged to be the maximum expected throughput for the largest Dresser-Rand 7HOSS6 or similar Ariel KBZ6 frames for the higher compression ratio case with 1,034 kPag (150 psig) of suction pressure. Sowels and Morse identified single-unit options with throughputs of 59,000 Sm^3/h (50 MMscfd) and 88,500 Sm^3/h (75 MMscfd) for the lower compression ratio case with 2,172 kPag (315 psig) of suction pressure. Trimeric and the two contacts concluded that the 59,000 Sm^3/h (50 MMscfd) throughput was a more logical fit for both suction pressure cases because it fit in even increments with the recycle rates of 59,000 Sm^3/h (50 MMscfd), 118,000 Sm^3/h (100 MMscfd), and 236,000 Sm^3/h (200 MMscfd) in this evaluation.

Trimeric expected that the higher compression ratio case with 1,034 kPag (150 psig) of suction pressure would require two stages of compression, which was confirmed by the equipment suppliers. Trimeric expected that the lower compression ratio case with 2,172 kPag (315 psig) of suction pressure could be achieved in a single-stage compressor. However, the equipment suppliers also proposed a two-stage compressor for the lower compression ratio case. Trimeric reviewed this option with the contacts, who explained that at these high flow rates and these suction and discharge pressure conditions, a single-stage configuration on the largest frames, such as the Dresser-Rand 7HOSS6, would actually result in a lower unit capacity and a higher power requirement per standard cubic meter per hour (million standard cubic feet per day) of CO_2 throughput than if the equipment were configured for a two-stage operation.

Trimeric also discussed the low compression ratio application with two other industry contacts, Casey Saunier and Dirk Dailey, both with Pelstar Mechanical Services (personal communication, July 2015). Saunier and Dailey provided several reasons why they agreed that two stages of compression would be preferable in this application. They pointed out that any decrease in suction pressure, increase in suction temperature, increase in discharge pressure, or changes in the gas composition could lead to problems with a single-stage compressor in this application, including excessive cylinder discharge temperatures, high rod-load conditions, or exceeding the pressure relief valve set points. They offered that a suction pressure of at least 2,760 kPag (400 psig), suction temperatures in the range of 10 to 21 °C (50 to 70 °F), or some related combination of higher suction pressure and lower suction temperature would be needed to specify a single-stage compressor with adequate design margins for this application. Using two stages of compression for the high suction pressure case resulted in another difference from the 2013 evaluation for the smaller compressors. In those smaller facility cases, which could operate on compressor frames with a greater margin between the operating conditions and the maximum rod-load limits, Trimeric assumed that the higher suction pressure case with 2,172 kPag (315 psig) of suction pressure and 6,895 kPag (1,000 psig) of discharge pressure could be accommodated with single-stage compressors.

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 saturated and water could be present from condensation or carryover. Carbon steel is typically used on the discharge side of the compressor cylinders because the discharge is hot, near 149 °C (300 °F), and therefore well above the water dew point during normal operation.

Two 29,500 Sm^3/h (25 MMscfd) compressors would be installed in parallel to handle the 59,000 Sm^3/h (50 MMscfd) of CO_2 gas flow rate for Cases 1 and 2. Doing so would provide additional operational flexibility at reduced throughput conditions as compared with installing one 59,000 Sm^3/h (50 MMscfd) compressor. As discussed, a single 59,000 Sm^3/h (50 MMscfd) reciprocating compressor is the largest size recommended for this application. Therefore, additional compressors of this size would be installed to handle the additional gas flow rates shown in Table 1 for Cases 3–6. The additional 59,000 Sm^3/h (50 MMscfd) compressors were assumed to be installed over time as the gas production rate increased throughout the life of the field. Many companies that operate CO_2 EOR facilities elect to defer the relatively high capital cost of compression and related 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.

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. Compressor operating costs were based on the peak product throughput; however, the energy efficiency may be lower at times when the compressors are not fully loaded. Low-pressure compression trains would presumably be used to send flash gases from the free water knockout and demulsifier to the suction of the high-pressure compression train.

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 after compression. Without dehydration, the CO_2 leaving the compressor train could be saturated with water at some conditions. The CO_2 would cool as the gas flowed through aboveground and underground piping, increasing the potential for water to condense and cause increased corrosion. The injection pressure anticipated for ILB EOR facilities (6,895 kPag [1,000 psig]) is too low to take advantage of the increased water-holding capacity of CO_2 that occurs 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. However, triethylene glycol 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, triethylene glycol dehydration could be performed between the first and second stages of compression. 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, Major Equipment Lists and Purchased and Installed Costs for Cases 1–6, to show the cost impact of this unit operation and to facilitate the removal of these costs if dehydration is not required.

A single dehydration unit should be able to treat the gas flow in Cases 1 and 2 (59,000 Sm^3/h [50 MMscfd]) as well as in Cases 3 and 4 (118,000 Sm^3/h [100 MMscfd]). For the highest gas flow rate assumed in Cases 5 and 6 (236,000 Sm^3/h [200 MMscfd]), two 118,000 Sm^3/h (100 MMscfd) dehydration units were used in the process flow scheme. A single dehydration unit could possibly also be provided for the highest gas flow rate cases, but this difference would not have a significant impact on the cost estimates in this early-stage conceptual evaluation.

Buildings

Buildings to house compressors, controls, chemicals, and maintenance equipment were included in the EOR facility. The estimated size of the building(s) was determined based on past experience with other projects.

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 process equipment. The In-Plant Cost Estimator costs are from the first quarter of 2015. The purchased costs were adjusted to a January 2015 cost basis (the most recent index available at the time of this evaluation) using published plant cost indices (Chemical Engineering Plant Cost Index, Chemical Engineering Magazine 2015). The list below shows the source of the purchased equipment costs by equipment type:

- Separators (slug catcher, free water knockout, and demulsifier)—In-Plant Cost Estimator. The high- and low-pressure suction scrubbers were included in cost estimates for the CO_2 compressor trains, so the costs for these vessels were not estimated separately.
- Chemical injection pump—In-Plant Cost Estimator
- Chemical injection tank—in-house vendor data
- Oil storage tanks—In-Plant Cost Estimator
- Water storage tanks—not required because the water storage and disposal equipment were already assumed to exist from the waterflood operations before conversion to CO_2 flooding
- CO_2 compressor train interstage air coolers—scaled from a similar air cooler quote in 2013
- CO_2 compressor trains—In-Plant Cost Estimator and a vendor quote from 2014
- Dehydration—scaled from vendor quotes for other CO_2 projects obtained from 2008 to 2015
- Building—In-Plant Cost Estimator

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. The tables in Appendix B show the detailed equipment sizes and the estimated purchased and installed costs for the individual equipment components needed for the six cases. Table 3 provides a summary of the total purchased and installed costs for each case.

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

Differences in Small- and Large-Scale Enhanced Oil Recovery Studies

The small-scale EOR evaluation conducted previously has several important differences from the large-scale EOR cases evaluated in this report. These differences are summarized below for reference.

Oil and Water Production Ratios

The oil-to-gas and water-to-gas production ratios were higher for the small-scale EOR cases than for the larger CO_2 flow cases in this study. This means that the oil storage and water disposal requirements are less for the large-scale EOR cases than if we had used the same ratios from the previous study.

CO_2 Compressor Trains

High- and low-pressure suction scrubbers were included in the cost estimates for the large-scale compressor trains; however, the interstage air coolers were not. This is different from the small-scale EOR study, in which the coolers were included in the compressor quote from the compression vendor and the high- and low-pressure suction scrubbers were excluded. For this reason, different equipment sizing and cost estimates were required in the two studies to estimate the overall compressor costs, including suction scrubbers and coolers. In addition, the compressors used

Table 3 Total purchased and installed costs[1]

Case	CO_2 production, Sm^3/h (MMscfd)	Water production, m^3/d (bpd)	Oil production, m^3/d (bpd)	Facility inlet pressure, kPag (psig)	Total purchased cost, $	Total installed cost, $
1	59,000 (50)	1,908 (12,000)	477 (3,000)	1,034 (150)	7,458,000	11,538,000
2	59,000 (50)	1,908 (12,000)	477 (3,000)	2,172 (315)	6,673,000	10,374,000
3	118,000 (100)	3,816 (24,000)	954 (6,000)	1,034 (150)	13,843,000	21,343,000
4	118,000 (100)	3,816 (24,000)	954 (6,000)	2,172 (315)	12,301,000	19,058,000
5	236,000 (200)	7,632 (48,000)	1,908 (12,000)	1,034 (150)	27,214,000	42,114,000
6	236,000 (200)	7,632 (48,000)	1,908 (12,000)	2,172 (315)	24,157,000	37,588,000

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

to increase the pressure of the CO_2 from 2,172 to 6,895 kPag (315 to 1,000 psig) for the large-scale EOR cases required two stages of compression (per vendor input) instead of the single-stage compressors selected for the same suction and discharge pressure requirements in the small-scale EOR facility evaluation.

Water Storage Tanks

Capital costs for water storage tanks were not included with the large-scale EOR cases because they were already assumed to exist from waterflood operations. Carbon steel water tanks were included and sized to hold 3 d of capacity at the peak water rate for the small-scale EOR facility evaluation. However, operating costs for water disposal were included for both the small- and large-scale EOR facilities based on an assumed cost of $1/bbl of produced water. Trimeric assumed this value after discussing water disposal costs with one ILB oilfield operator and comparing the operator's input with water disposal cost data from other Trimeric projects. Operators often arrange for on-site disposal of the produced water or use it in the flood management of an EOR field to reduce costs for water disposal.

Natural Gas Liquid Recovery

Natural gas liquid recovery was not evaluated for the large-scale EOR cases because in the previous work, we concluded that NGL recovery was not economically justified, given the lean NGL content of the gas expected from the ILB (0.03 L/m³ [0.22 GPM]), even at the high CO_2 flow rates used in the large-scale facility evaluation. The NGL content in gases is typically characterized in terms of gallons of recoverable hydrocarbons in the gas per thousand standard cubic feet of gas (GPM).

Demulsifier Chemicals

In the small-scale EOR evaluation, a demulsifier concentration of 1,000 ppmv in only the oil phase was assumed based on past project experience. The resulting concentration in the total liquid volume of the oil and water phases would be approximately 200 ppmv, which is about 2.2 times the amount used with the large-scale EOR cases. The 90 ppmv concentration based on the total liquid volume (oil plus water) should be considered more up to date and accurate because it was recently obtained from a vendor specifically for the ILB large-scale facility evaluation (Ken Hake of Baker Hughes, personal communication, July 2015). A higher cost for the demulsifer chemicals was also used in this work ($24/gal) than in the small-scale EOR evaluation ($10/gal). The difference in concentration bases and costs resulted in an increase in annual demulsifer chemical costs of 8% in the large-scale evaluation. The chemical storage capacity for the large-scale EOR cases was approximately half that required if we had used the same concentration basis as in the small-scale EOR work. However, this expense is insignificant in terms of the overall costs for the EOR facilities.

Dehydration

Capital costs for dehydration in this report were based on more recent vendor quotes for units treating CO_2 streams in the larger flow range.

Operating Cost Information

Operating cost information for the six cases is shown in Table 4. The information is separated into two categories: variable costs (with the capacity utilization factor) and fixed costs. The operating cost information and bases are discussed in this section so that they can be combined with any field-wide operating costs developed by others.

As shown in Table 4, 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 facilities. The compression power includes the power required for both the high-pressure and the low-pressure compression trains. The annual electricity cost is estimated based on an assumed electricity cost of $0.09/kWh. The peak water rate is shown so that disposal costs for off-site disposal can be estimated ($1/bbl assumed). The peak oil rate is given to facilitate the estimation of transportation fees (not included). The total dehydration operating costs are included so that the operating expenses can be estimated for the entire EOR facility. The demulsifier chemical cost is $24/gal based on recent vendor input (Ken Hake of Baker Hughes, personal communication, July 2015).

The fixed costs include an estimate of the number of operators required to run the facility and 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 costs (typical factor). The fixed costs do not include the capacity utilization factor.

The total operating costs (variable and fixed items) ranged from $7.7 million for Case 2 with 59,000 Sm³/h (50 MMscfd) of CO_2 flow and an inlet pressure of 2,172 kPag (315 psig) to $35.3 million for Case 5 at 236,000 Sm³/h (200 MMscfd) of CO_2 flow and an inlet pressure of 1,034 kPag (150 psig). The cost for produced water disposal represents approximately 48% to 58% of the variable operating costs, and the annual electricity cost accounts for another 29% to 41%, depending on the inlet gas pressure (2,172 kPag [315 psig] or 1,034 kPag [150 psig], respectively). Approximately 32% to 76% of the fixed operating costs resulted from annual compressor maintenance, with the remaining amount pertaining to labor and overhead expenses.

Fixed Capital Investment

The purchased equipment costs for the EOR facility were multiplied by a factor of 3 to estimate the FCI cost. This factor accounts for the costs of items such as purchased equipment costs, purchased equipment installation, instrumentation and controls, piping, electrical systems, engineering and supervision, construction expenses, contractors' fees, and contingency. A multiplier of 3 times the purchased equipment costs is typically used to estimate the FCI for a mix of vendor-provided skid-mounted equipment,

Table 4 Operating cost summary[1]

Operating cost information	Unit	Case					
		1	2	3	4	5	6
Variable costs (includes capacity utilization factor)							
Capacity utilization factor	%	95	95	95	95	95	95
Electricity usage	kW	4,240	2,524	8,483	5,051	16,970	10,105
Motor efficiency	%	95	95	95	95	95	95
Annual electricity cost	$/yr	3,518,700	2,094,800	7,040,300	4,191,800	14,083,400	8,386,000
Chemical injection	$/gal	24	24	24	24	24	24
Chemical injection rate	gal/d	54	54	108	108	215	215
Annual chemical injection cost	$/yr	472,000	472,000	944,000	944,000	1,887,000	1,887,000
Produced water disposal	$/bbl	1	1	1	1	1	1
Produced water disposal rate	bpd	11,400	11,400	22,800	22,800	45,600	45,600
Annual produced water disposal cost	$/yr	4,161,000	4,161,000	8,322,000	8,322,000	16,644,000	16,644,000
Annual dehydration cost	$/yr	515,000	515,000	887,000	887,000	1,450,000	1,450,000
Oil transport capacity	bpd	2,850	2,850	5,700	5,700	11,400	11,400
Total variable operating costs	$/yr	8,666,700	7,242,800	17,193,300	14,344,800	34,064,400	28,367,000
Fixed costs							
Operating labor	Full-time equivalent	2	2	2	2	2	2
Cost of labor	$/h	29	29	29	29	29	29
Operating labor cost	$/yr	146,600	146,600	146,600	146,600	146,600	146,600
Supervisor labor	% of operating labor cost	20	20	20	20	20	20
Supervisor labor cost	$/yr	29,400	29,400	29,400	29,400	29,400	29,400
Compressor maintenance cost factor	$/(hp-yr)	40	40	40	40	40	40
Compressor horsepower	hp	5,984	3,562	11,973	7,128	23,951	14,260
Annual compressor maintenance cost	$/yr	239,400	142,500	479,000	285,200	958,100	570,400
Plant operating overhead	% of operating + supervisor cost	75	75	75	75	75	75
Plant operating overhead cost	$/yr	132,000	132,000	132,000	132,000	132,000	132,000
Total fixed operating costs	$/yr	547,400	450,500	787,000	593,200	1,266,100	878,400
Total operating costs	$/yr	9,214,100	7,693,300	17,980,300	14,938,000	35,330,500	29,245,400

[1]bbl, oilfield barrels; bpd, barrels per day.

on-site assembly (separators, tanks, etc.), and field 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. Trimeric considers these cost estimates a study estimate (factored estimate) that is based on the knowledge of major items of equipment and that has an expected accuracy of ±30% (Peters et al. 2003). Table 5 summarizes the estimated FCI for the surface equipment for all six cases. Observations regarding these cost data are described below.

The General Fixed Capital Investment Cost Relationship

As shown in Table 5, the FCI ranged from \$20 million for Case 2 with 59,000 Sm^3/h (50 MMscfd) of produced gas to as high as \$81.6 million for Case 5 with 236,000 Sm^3/h (200 MMscfd) of produced gas. Compression represents approximately 50% to 60% of the overall capital costs, followed by dehydration at approximately 20% and the separation of liquid phases and cooling of the gas phase totaling approximately 15%.

Figure 3 graphically represents the estimated FCI for the large-scale EOR facilities. As explained, the costs scale fairly linearly with the CO_2 recycle rate because of the modular approach assumed for the construction of these large-scale EOR recycle facilities. In addition, differences in the FCI are fairly minimal because the compressor vendors that Trimeric contacted regarding these cases recommended two-stage compressors for both suction pressure conditions. More details on this topic are provided in the CO_2 Compressor Trains discussion in the Equipment Design and Cost Analysis Summary section of this report.

Fixed Capital Investment Model Development

A model (see Equation 1) was developed to estimate the FCI based on the cost estimates from the previous small-scale EOR facility evaluation and the large-scale EOR cases summarized in this report.

A simple model was developed to estimate the FCI (first quarter of 2015) for the oil storage tanks based on the peak oil production rate and that for the rest of the surface equipment based on the peak CO_2 recycle rate as a function of inlet pressure ranging from 1,034 to 2,172 kPag (150 and 315 psig) with a discharge pressure of 6,895 kPag (1,000 psig). The FCI cost estimates derived from Equation 1 do not include any costs for water storage tanks for the reasons already noted:

$$\text{FCI} = (379{,}810 \times \text{CO}_2\ \text{Rate} + 2{,}851{,}322) \times (1.0955 - 0.0006 \times \text{Pressure}) + (207 \times \text{Oil Rate} + 131{,}211), \tag{1}$$

where CO_2 Rate is the peak CO_2 recycle rate (MMscfd), Pressure is the inlet gas pressure (psig), and Oil Rate is the peak oil production (barrels per day [bpd]).

Figure 4 shows a correlation graph for the estimated FCI cost and the modeled FCI cost for the data in the two studies. If the model correlated perfectly with the estimated costs, the data points would fall on the 45° line. As shown in the graph, the model correlates within 5% for the large-scale EOR cases and within approximately 28% for the small-scale cases. The model is valid for only the water-to-gas, oil-to-gas, and CO_2 compression ratios used in the two studies. Extrapolation to conditions that vary significantly from these could produce erroneous results.

Effect of Facility Inlet Pressure

Table 6 shows the effect of the facility inlet pressure on the high-pressure compressor purchased equipment costs. On the basis of cost estimates for the large-scale EOR cases, higher inlet pressures of 2,172 kPag (315 psig) result in approximately 25% lower purchased equipment costs for the compressors when compared with compressor costs when the CO_2 facility pressure is 1,034 kPag (150 psig).

Compressor size and cost are a function of the suction actual volumetric flow rate, and the motor power requirement (and cost) is a function of both the pressure ratio and mass flow rate. For cases with a similar mass flow rate (Cases 1 and 2, Cases 3 and 4, and Cases 5 and 6), the lower facility inlet pressure results in a higher pressure ratio and more work being required to achieve the same discharge pressure of 6,895 kPag (1,000 psig). The lower facility inlet pressure cases (Cases 1, 3, and 5) also have a higher actual volumetric flow. Both of these parameters (higher pressure ratios and higher actual volumetric flow rates) make the compressors more expensive for the cases with a lower facility inlet pressure (Cases 1, 3, and 5) than the compressors for the cases with a higher facility inlet pressure (Cases 2, 4, and 6).

MISCELLANEOUS COST ITEMS

The scope of Trimeric's work in this evaluation included estimates of the costs for potential environmental controls and the costs for flowlines to and from the EOR surface facilities.

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 about 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: the low-pressure suction scrubber and the oil storage tanks. Although crude oils from the ILB contain no hydrogen sulfide (H_2S), additional environmental controls may be necessary if H_2S is present in the produced fluids in other basins.

In this evaluation, the low-pressure gas flow rate is large enough to justify the installation of a low-pressure compressor in all cases. The flash gas generated in the low-pressure suction scrubber is compressed in a low-pressure compressor and combined with the inlet gas going to the high-pressure compressor train(s). Therefore, this potential emission source is eliminated. The costs for the low-pressure compressors are summarized in Table 7. These costs are also shown in the equipment cost tables for each case in Appendix B. The purchased equipment cost to recover the low-pressure flash gas constitutes approximately 9% to 13% of the total purchased equipment cost for each case.

Table 5 Summary of the total fixed capital investment[1]

Parameter	Unit	Case 1	2	3	4	5	6
Actual gas flow	acfm	3,142	1,466	6,283	2,931	12,570	5,862
Produced gas flow	MMscfd	50	50	100	100	200	200
Peak water capacity	bpd	12,000	12,000	24,000	24,000	48,000	48,000
Peak oil capacity	bpd	3,000	3,000	6,000	6,000	12,000	12,000
Inlet pressure	psig	150	315	150	315	150	315
Discharge pressure	psig	1,000	1,000	1,000	1,000	1,000	1,000
Installation phase(s)		Single	Single	Multiple	Multiple	Multiple	Multiple
Equipment cost information							
Purchased equipment costs, January 2015 basis							
Separators, coolers, and chemical injection	$	1,081,000	1,016,000	1,991,000	1,889,000	3,817,000	3,640,000
Oil tanks	$	219,000	219,000	438,000	438,000	876,000	876,000
Compression	$	4,066,000	3,346,000	8,131,000	6,691,000	16,262,000	13,382,000
Dehydration	$	1,612,000	1,612,000	2,444,000	2,444,000	4,888,000	4,888,000
Building	$	480,000	480,000	839,000	839,000	1,371,000	1,371,000
Total purchased equipment costs (PEC)	$	7,458,000	6,673,000	13,843,000	12,301,000	27,214,000	24,157,000
Total installed equipment cost	$	11,538,400	10,373,800	21,342,500	19,058,400	42,113,900	37,588,300
Capital cost information							
Factor for estimating the fixed capital investment (FCI) for the plant from the PEC		3	3	3	3	3	3
Total FCI	$	22,374,000	20,019,000	41,529,000	36,903,000	81,642,000	72,471,000

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

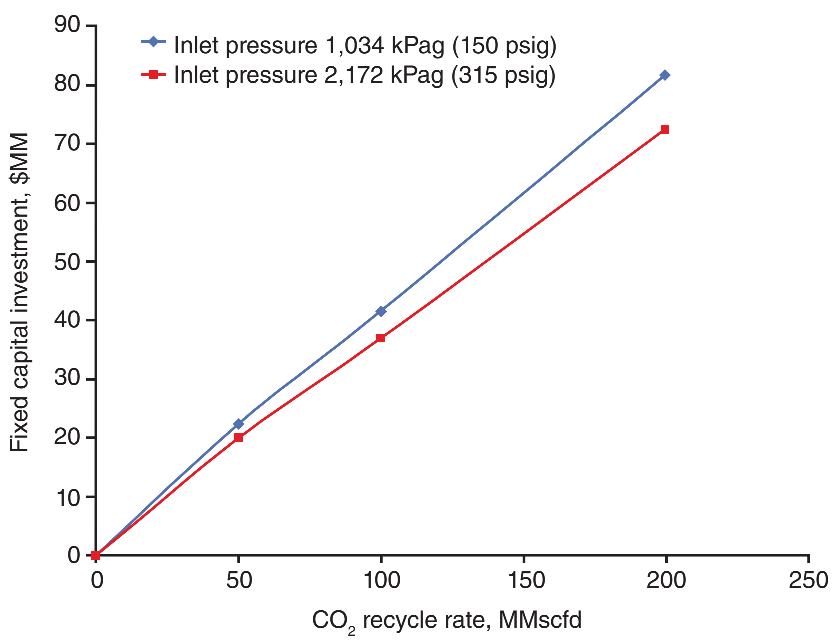

Figure 3 Fixed capital investment as a function of the CO_2 recycle rate and inlet pressure. MM, million; psig, pounds per square inch gauge.

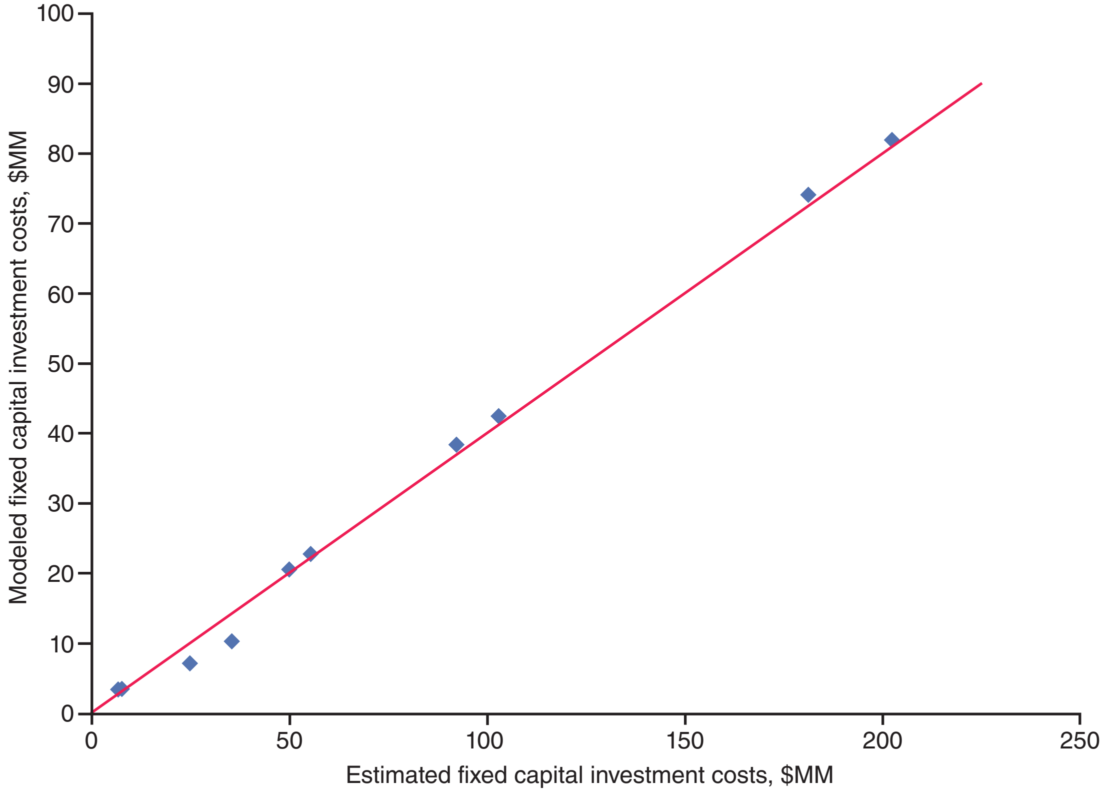

Figure 4 Fixed capital investment model comparison with estimated costs. MM, million.

Table 6 Effect of facility inlet pressure on the high-pressure compressor purchased cost[1]

		Case					
Parameter	Unit	1	2	3	4	5	6
Facility inlet pressure	psig	150	315	150	315	150	315
Facility outlet pressure	psig	1,000	1,000	1,000	1,000	1,000	1,000
Pressure differential	psi	850	685	850	685	850	685
Pressure ratio	dimensionless	6.7	3.2	6.7	3.2	6.7	3.2
Mass flow	lb/h	239,420	235,733	478,765	471,305	957,834	942,610
Standard volumetric flow	MMscfd	50	50	100	100	200	200
Actual volumetric flow	acfm	3,142	1,466	6,283	2,931	12,570	5,862
High-pressure CO_2 compressor power	hp	5,746	3,144	11,497	6,292	22,999	12,588
Number of 25 MMscfd compressors	—	2	2	2	2	2	2
Number of 50 MMscfd compressors	—	0	0	1	1	3	3
Number of compression stages per compressor	—	2	2	2	2	2	2
High-pressure CO_2 compressor and air cooler purchased cost	$MM	4.05	3.09	7.94	6.02	15.72	11.97

[1]psig, pounds per square inch gauge; psi, pounds per square inch; MMscfd, million standard cubic feet per day; acfm, actual cubic feet per minute; hp, horsepower; MM, million.

Table 7 Low-pressure compressor purchased equipment cost estimates[1]

Case	Vent gas flow rate, Sm^3/h (MMscfd)	Vent gas pressure, kPag (psig)	Recovery compressor discharge pressure,[2] kPag (psig)	Purchased equipment cost,[3] $
1	2,950 (2.5)	165 (24)	1,034 (150)	692,000
2	2,950 (2.5)	165 (24)	2,172 (315)	816,000
3	5,900 (5)	165 (24)	1,034 (150)	1,383,000
4	5,900 (5)	165 (24)	2,172 (315)	1,632,000
5	11,800 (10)	165 (24)	1,034 (150)	2,766,000
6	11,800 (10)	165 (24)	2,172 (315)	3,265,000

[1]Sm^3/h, standard cubic meters per hour; MMscfd, million standard cubic feet per day; kPag, kilopascal gauge; psig, pounds per square inch gauge.
[2]Assumes that the recovered vapor will be returned to the high-pressure compressor suction.
[3]All costs obtained using Aspen In-Plant Cost Estimator software.

The working losses[1] from the oil storage tanks could increase if the oil production rate increases with the change from waterflood to CO_2 flood. The characteristics of the vapors vented as working losses would depend on the composition of the produced oil, so it is difficult to specify which emission controls might be required for the storage tank vents. A flare (with or without an inlet air 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 called a vapor recovery unit, similar to the low-pressure compressor, could be used to send the recovered storage tank vapors to the suction of the low-pressure compressor, which ultimately goes to the suction of the high-pressure compressor, and then to reinjection.

Flowline Piping

The sizes for flowlines to carry the produced fluids to the EOR surface facility can be estimated by assuming a typical design velocity for two-phase flow of 9.1 m/s (30 ft/s) or less. Estimated material costs for selected piping diameters that may be applicable for the large-scale EOR facility cases are shown in Table 8. The actual piping diameter in a given application would depend on the total volume of fluid transported in that flowline. Separate flowline sizing calculations would also be required for flowlines to transport CO_2 from the central facility back to the injection wells. The estimated piping capital costs (NETL 2013) are shown, assuming carbon steel as the construction material for the piping. Stainless steel piping costs could range from approximately 1.5 to 4 times that of the carbon steel estimates. Carbon steel and stainless steel could be specified for flowlines going to and from the central facility. Carbon steel would likely be more common, given the anticipated difference in capital 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 work for this evaluation because such an analysis requires details or assumptions regarding the number of wells, distances between the wells, and intermediate satellite test facilities where production from multiple wells is measured and then aggregated and transported to the central facility, 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 correlation that can be used to estimate the FCI for central facilities as a function of CO_2 recycle rates ranging from 1,180 Sm^3/h (1 MMscfd) to 236,000 Sm^3/h (200 MMscfd) for suction pressures of 1,034 kPag (150 psig) or 2,172 kPag (315 psig) and a discharge pressure of 6,895 kPag (1,000 psig). The capital costs used as inputs to the correlation for the small recycle facilities (1,180 to 24,780 Sm^3/h [1 to 21 MMscfd]) were taken from an evaluation Trimeric performed for the ISGS in 2013 for smaller EOR central facilities. Costs for facilities with recycle rates ranging from 59,000 Sm^3/h (50 MMscfd) to 236,000 Sm^3/h (200 MMscfd) were developed in the current report. Together, these conditions represent ranges that might be expected for any early-phase CO_2 EOR floods in the ILB. Estimates of major operating costs for the larger EOR CO_2 recycle facilities are also provided.

The FCI for this type of facility is typically dominated by compressor costs. Smaller compressors are often installed in parallel at the beginning of an EOR flood operation rather than installing one larger compressor. This setup provides more operational flexibility as the CO_2 rate returning from the production wells begins to increase. Larger compressors are often installed in later years in a phased approach as the produced gas rate from the field continues to increase. This method reduces major capital expenditures for several years and reduces the risk of installing equipment or equipment capacity at the beginning of the flood that might not be needed if actual operations differ from original projections.

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

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

Table 8 Unit purchased material capital costs for flow-line piping[1]

Pipe diameter, mm (in.)	Carbon steel pipe cost, $/mi
50.8 (2)	115,000
101.6 (4)	126,000
152.4 (8)	162,000
304.8 (12)	220,000
406.4 (16)	298,000

[1]Piping material costs can fluctuate significantly. It is necessary to verify current pricing at the beginning of each project.

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.

ACKNOWLEDGMENTS

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APPENDIX A

Process Flow Diagrams for Cases 1–6

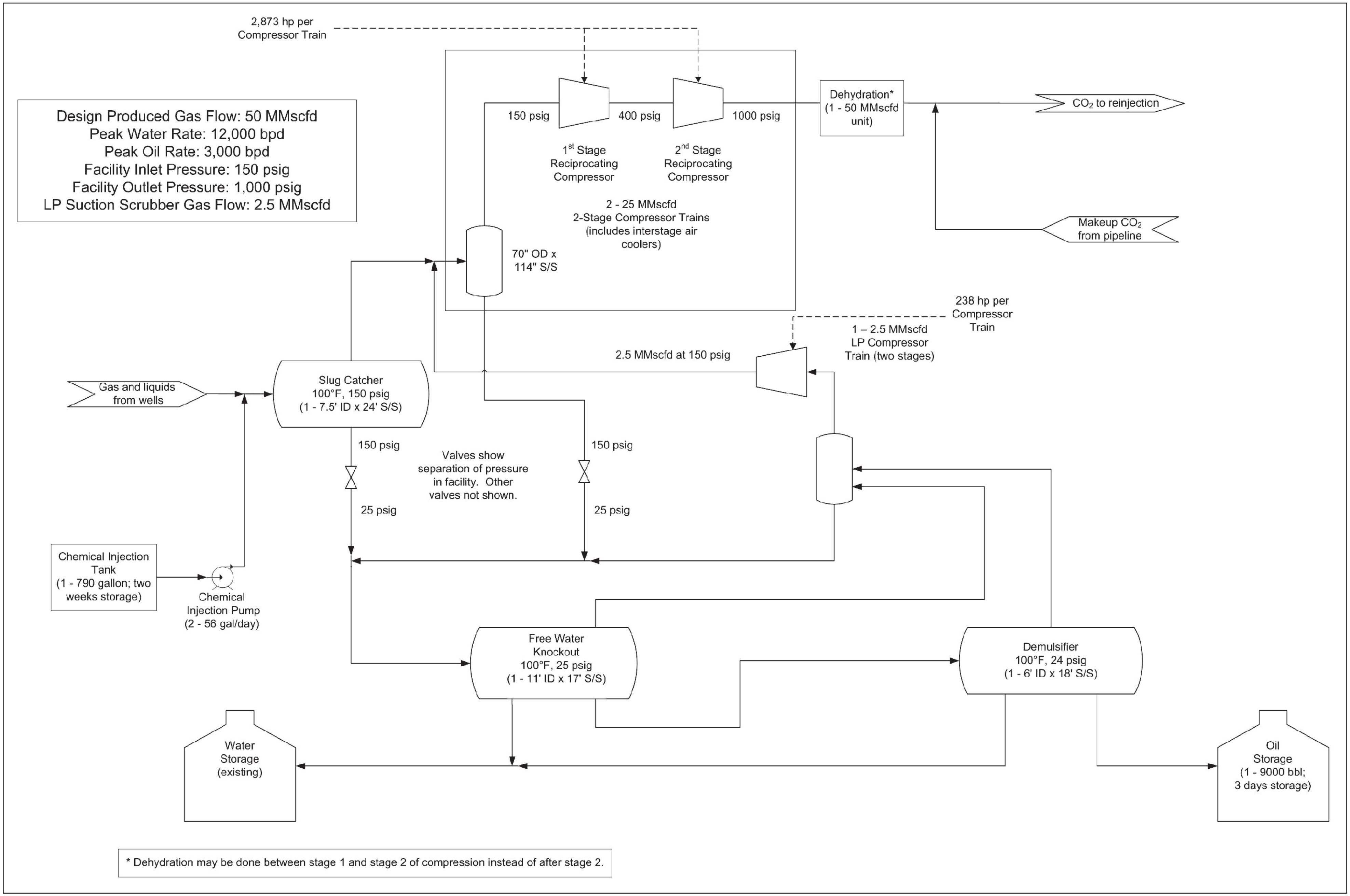

Figure A1 Process flow diagram for Case 1.

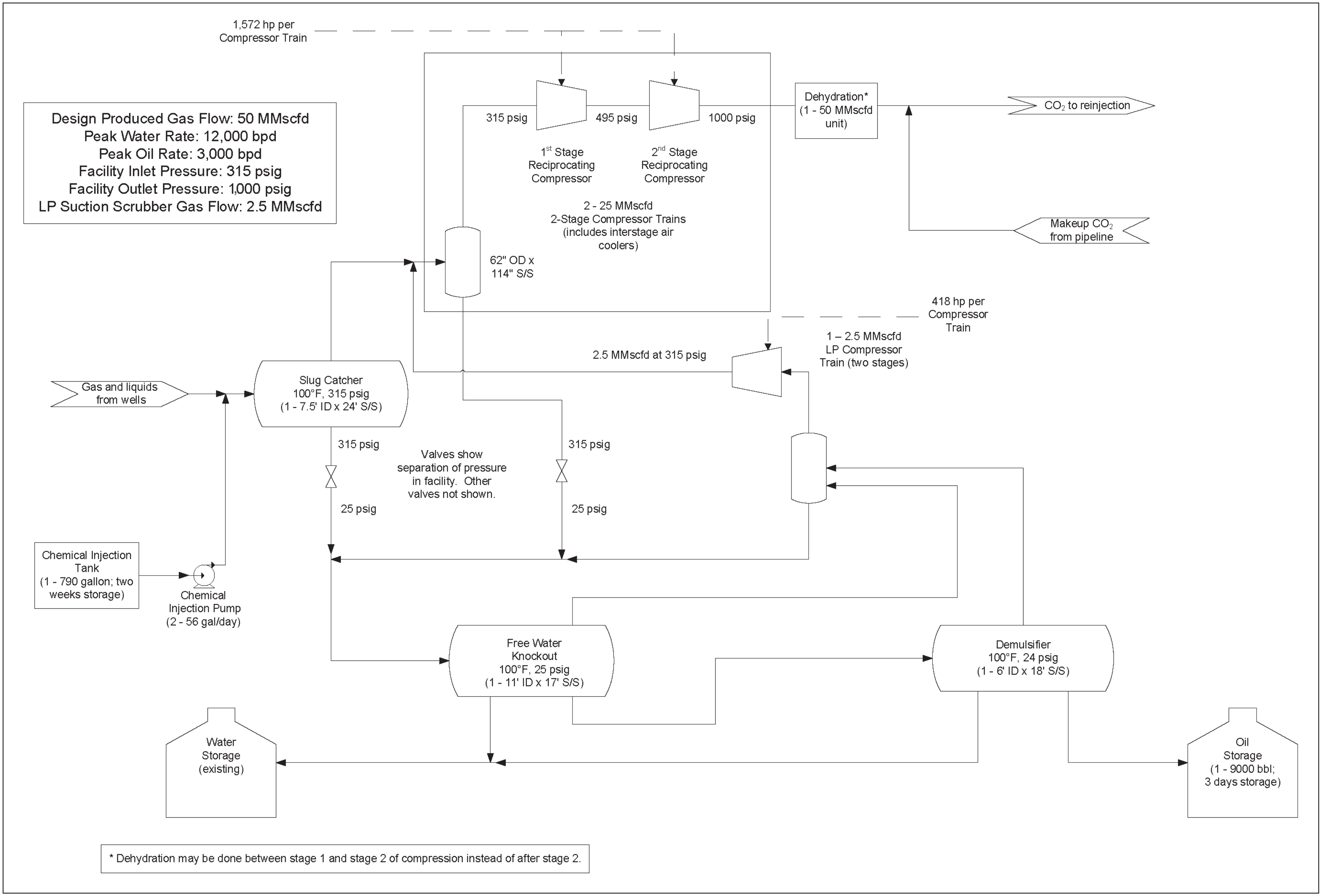

Figure A2 Process flow diagram for Case 2.

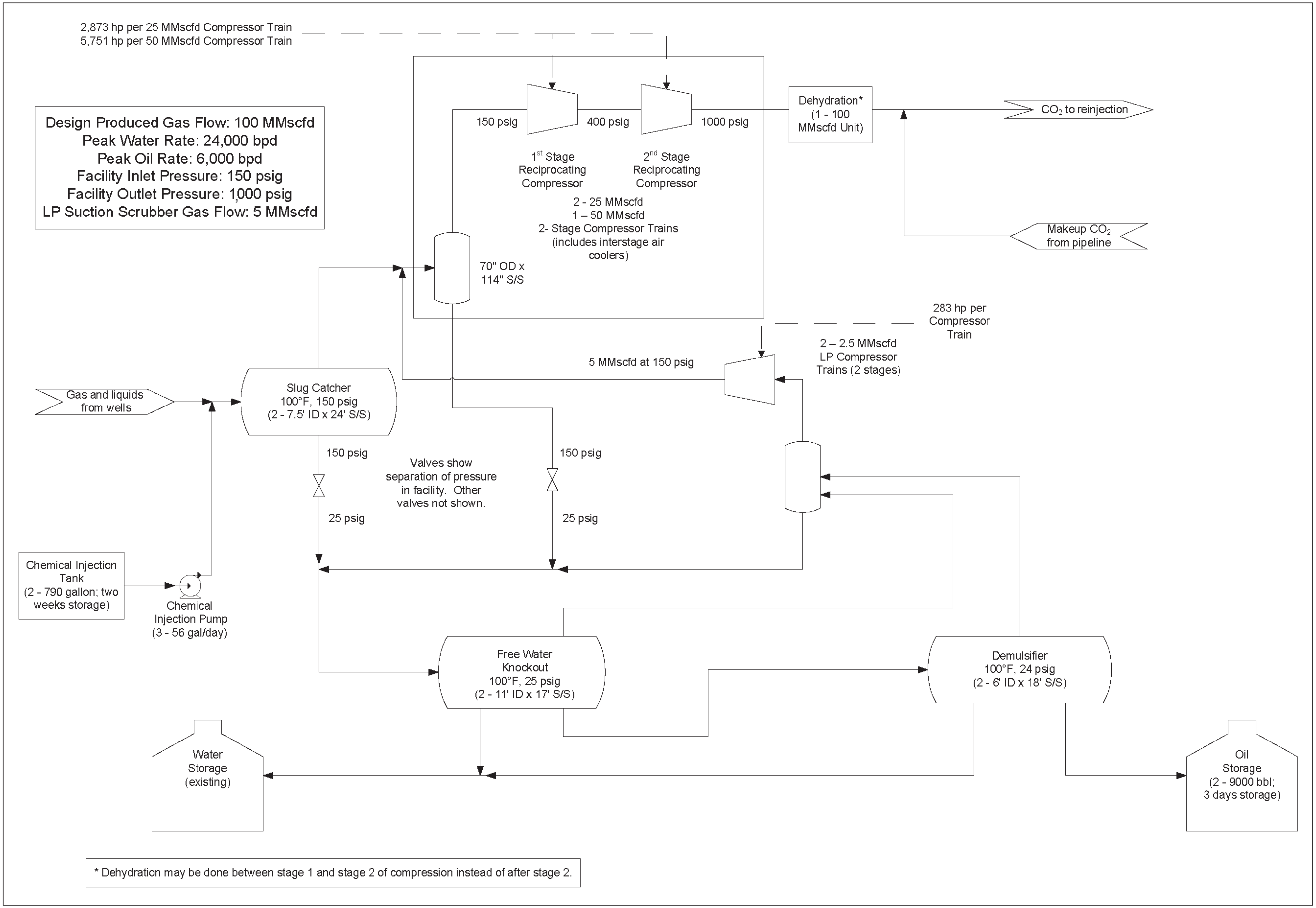

Figure A3 Process flow diagram for Case 3.

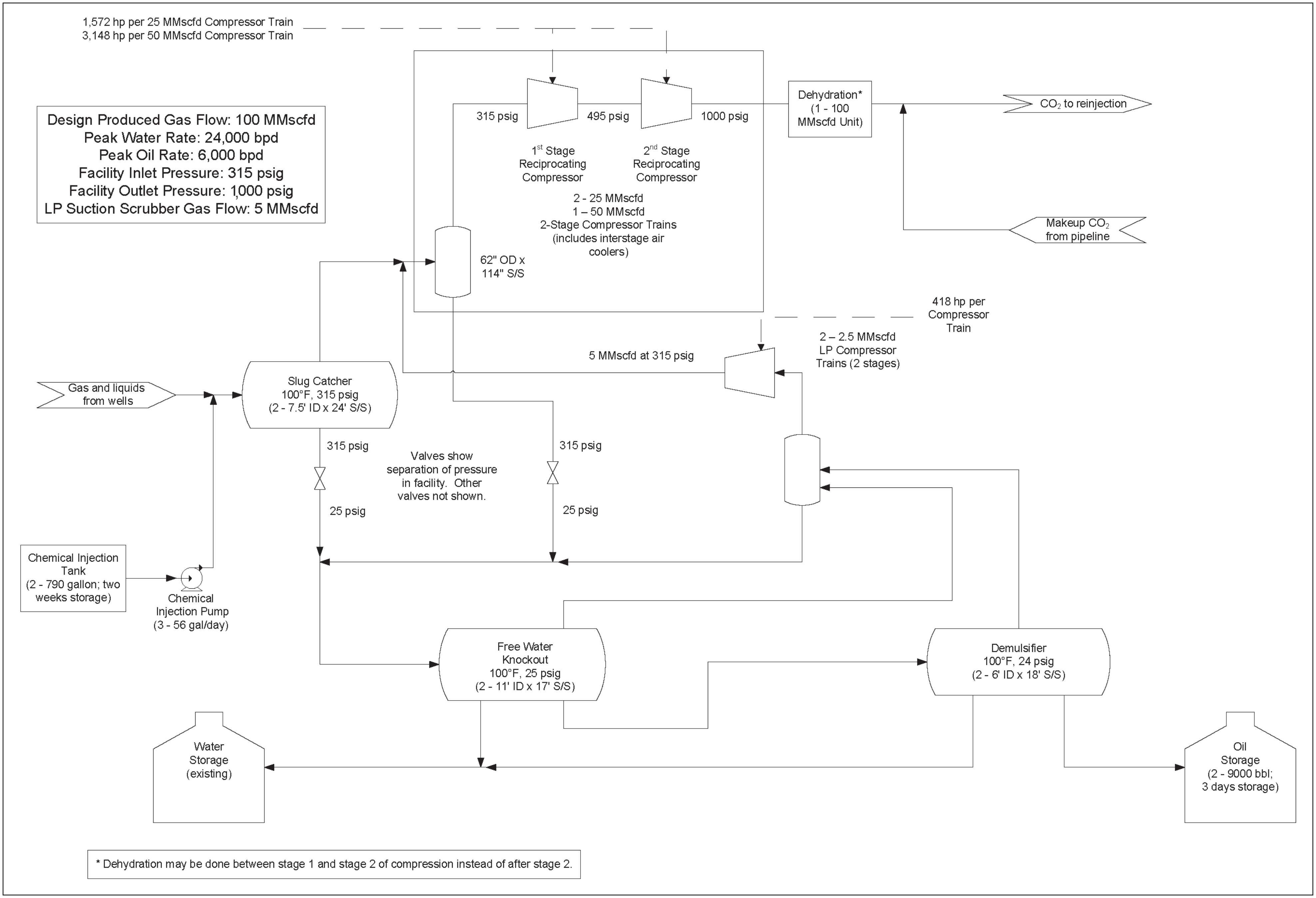

Figure A4 Process flow diagram for Case 4.

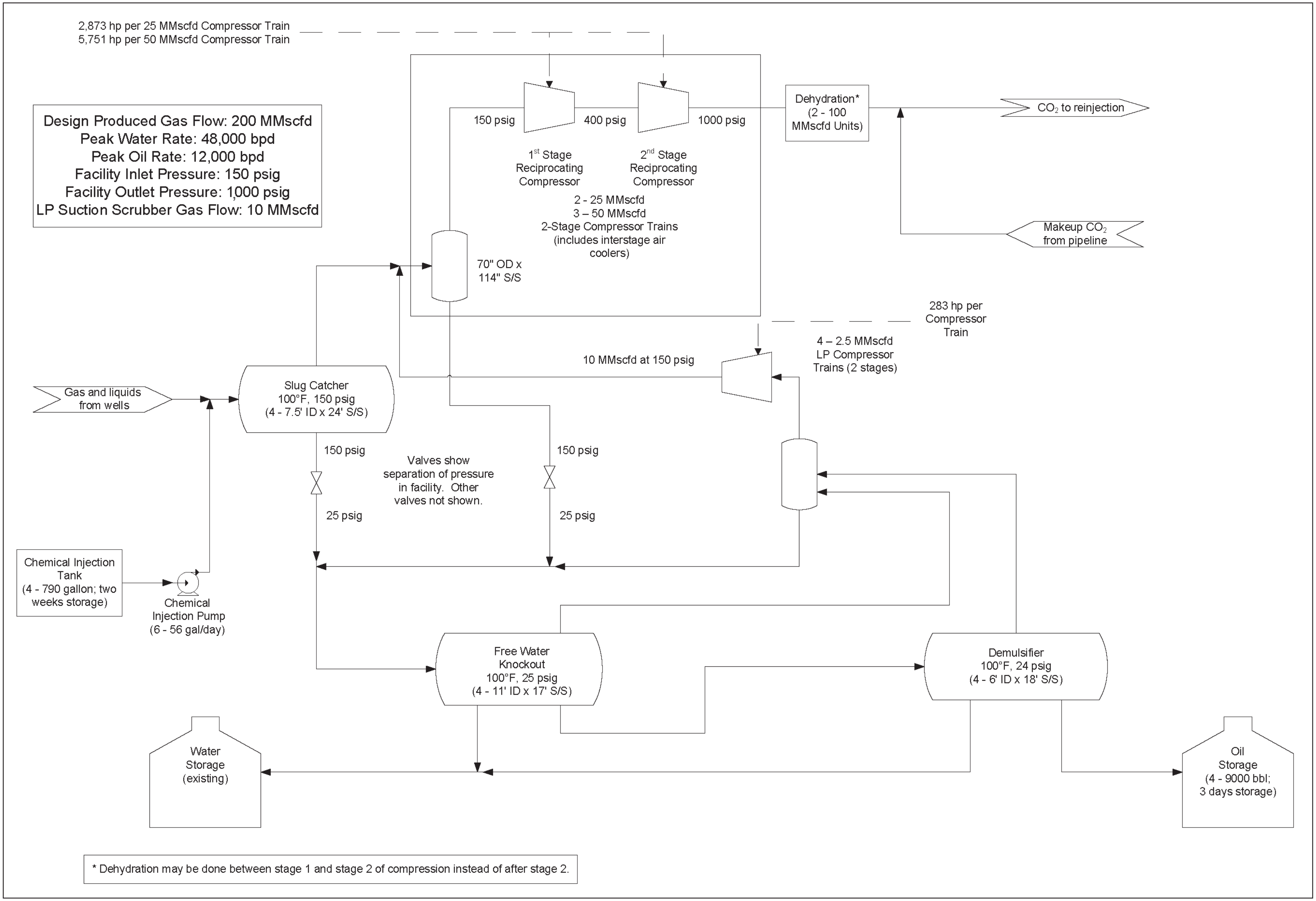

Figure A5 Process flow diagram for Case 5.

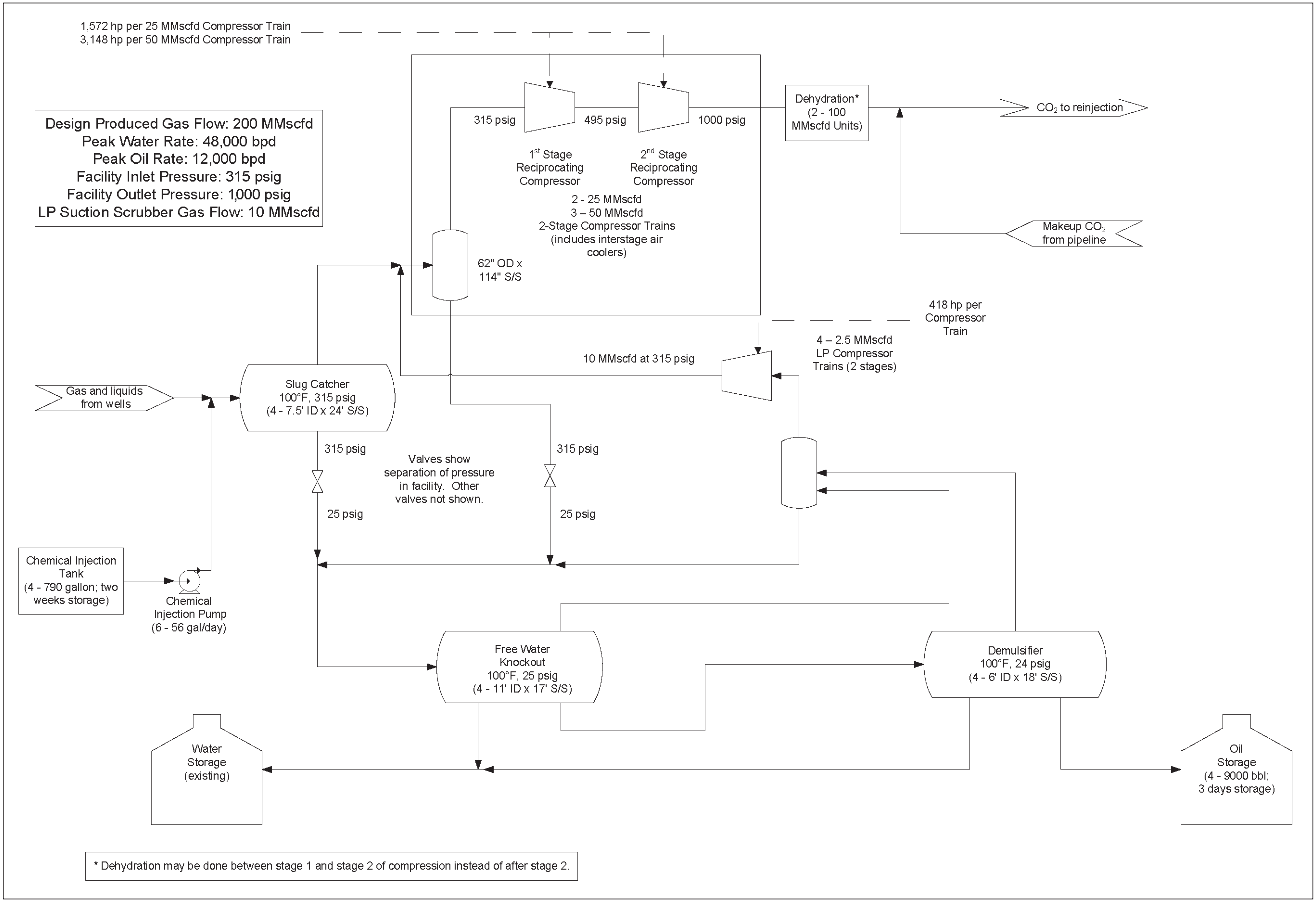

Figure A6 Process flow diagram for Case 6.

APPENDIX B

Equipment List and Purchased and Installed Costs for Cases 1–6

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 Jan 2015, $	No. of units	Plant index equipment basis	Jan 2015 plant index	Jan 2015 total purchased equipment cost, $	Installation factor	Jan 2015 total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 7.5 ft ID × 24 ft S/S length, coated carbon steel	250	200	129,400	1	572.8	573.1	129,500	0.88	243,500
HP air cooler–1st stage, 25 MMscfd	SS air cooler, 3.14 MMBtu/h	500	350	136,900	2	564.8	573.1	277,800	0.61	447,300
HP air cooler–2nd stage, 25 MMscfd	SS air cooler, 5.75 MMBtu/h	1,200	350	197,300	2	564.8	573.1	400,400	0.61	644,600
LP air cooler–1st stage	SS air cooler, 0.49 MMBtu/h	250	350	44,800	1	564.8	573.1	45,500	0.61	73,300
LP air cooler–2nd stage	SS air cooler, 0.50 MMBtu/h	250	350	45,300	1	564.8	573.1	46,000	0.61	74,100
Free water knockout	Horizontal 3-phase separator vessel, 11 ft × 17 ft S/S length, coated carbon steel	50	200	101,000	1	572.8	573.1	101,100	0.88	190,100
Demulsifier	Horizontal 3-phase separator vessel, 6 ft × 18 ft S/S length, coated carbon steel	50	200	57,500	1	572.8	573.1	57,500	0.88	108,100
Chemical injection pump	0.039 gpm to give 90 ppm by volume in oil/water mixture	50	150	5,700	2	572.8	573.1	11,400	0.30	14,800
Chemical injection tank	790-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	11,400	1	572.8	573.1	11,400	0.88	21,400
Oil storage tank	API storage tank 9,000 bbl capacity each (for 3-d capacity at peak rate)	Hydrostatic + 2.5 psig	150	218,800	1	572.8	573.1	218,900	0.88	411,500
HP compressor	25 MMscfd from 150 to 1,000 psig (2,873 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	1,706,400	2	579.7	573.1	3,373,900	0.49	5,027,100
LP compressor	2.5 MMscfd from 25 to 150 psig (238 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	691,200	1	572.8	573.1	691,600	0.49	1,030,500
Dehydration	Treats 50 MMscfd after compression at 1,000 psig	1,100	150	1,612,000	1	573.1	573.1	1,612,000	0.72	2,772,600
Building	3,844 ft^2	NA	NA	479,200	1	572.8	573.1	479,500	0.00	479,500
Total								7,456,500		11,538,400

[1]Production rates: 59,000 Sm^3/h (standard cubic meters per hour; 50 MMscfd [million standard cubic feet per day]) of CO_2, 476 m^3/d of oil (3,000 bopd [barrels of oil per day]), and 1,908 m^3/d of water (12,000 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; SS, stainless steel.

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 Jan 2015, $	No. of units	Plant index equipment basis	Jan 2015 plant index	Jan 2015 total purchased equipment cost, $	Installation factor	Jan 2015 total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 7.5 ft ID × 24 ft S/S length, coated carbon steel	450	200	176,100	1	572.8	573.1	176,200	0.88	331,300
HP air cooler–1st stage, 25 MMscfd	SS air cooler, 2.82 MMBtu/h	750	350	128,300	2	564.8	573.1	260,400	0.61	419,200
HP air cooler–2nd stage, 25 MMscfd	SS air cooler, 3.66 MMBtu/h	1,200	350	150,200	2	564.8	573.1	304,800	0.61	490,700
LP air cooler–1st stage	SS air cooler, 0.49 MMBtu/h	250	350	45,000	1	564.8	573.1	45,700	0.61	73,600
LP air cooler–2nd stage	SS air cooler, 0.52 MMBtu/h	450	350	46,500	1	564.8	573.1	47,200	0.61	76,000
Free water knockout	Horizontal 3-phase separator vessel, 11 ft × 17 ft S/S length, coated carbon steel	50	200	101,000	1	572.8	573.1	101,100	0.88	190,100
Demulsifier	Horizontal 3-phase separator vessel, 6 ft × 18 ft S/S length, coated carbon steel	50	200	57,500	1	572.8	573.1	57,500	0.88	108,100
Chemical injection pump	0.039 gpm to give 90 ppm by volume in oil	50	150	5,700	2	572.8	573.1	11,400	0.30	14,800
Chemical injection tank	790-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	11,400	1	572.8	573.1	11,400	0.88	21,400
Oil storage tank	API storage tank 9,000 bbl capacity each (for 3-d capacity at peak rate)	Hydrostatic + 2.5 psig	150	218,800	1	572.8	573.1	218,900	0.88	411,500
HP compressor	25 MMscfd from 315 to 1,000 psig (1,572 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	1,279,300	2	579.7	573.1	2,529,500	0.49	3,769,000
LP compressor	2.5 MMscfd from 25 to 315 psig (418 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	815,700	1	572.8	573.1	816,100	0.49	1,216,000
Dehydration	Treats 50 MMscfd after compression at 1,000 psig	1,100	150	1,612,000	1	573.1	573.1	1,612,000	0.72	2,772,600
Building	3,844 ft^2	NA	NA	479,200	1	572.8	573.1	479,500	0.00	479,500
Total						572.8	573.1	6,671,700		10,373,800

[1]Production rates: 59,000 Sm^3/h (standard cubic meters per hour; 50.0 MMscfd [million standard cubic feet per day]) of CO_2, 476 m^3/d of oil (3,000 bopd [barrels of oil per day]), and 1,908 m^3/d of water (12,000 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; SS, stainless steel.

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 Jan 2015, $	No. of units	Plant index equipment basis	Jan 2015 plant index	Jan 2015 total purchased equipment cost, $	Installation factor	Jan 2015 total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 7.5 ft ID × 24 ft S/S length, coated carbon steel	250	200	129,400	2	572.8	573.1	258,900	0.88	486,700
HP air cooler–1st stage, 25 MMscfd	SS air cooler, 3.14 MMBtu/h	500	350	136,900	2	564.8	573.1	277,800	0.61	447,300
HP air cooler–2nd stage, 25 MMscfd	SS air cooler, 5.75 MMBtu/h	1,200	350	197,300	2	564.8	573.1	400,400	0.61	644,600
HP air cooler–1st stage, 50 MMscfd	SS air cooler, 6.27 MMBtu/h	500	350	207,400	1	564.8	573.1	210,400	0.61	338,700
HP air cooler–2nd stage, 50 MMscfd	SS air cooler, 11.5 MMBtu/h	1,200	350	299,000	1	564.8	573.1	303,400	0.61	488,500
LP air cooler–1st stage	SS air cooler, 0.49 MMBtu/h	250	350	44,800	2	564.8	573.1	90,900	0.61	146,300
LP air cooler–2nd stage	SS air cooler, 0.50 MMBtu/h	250	350	45,300	2	564.8	573.1	91,900	0.61	148,000
Free water knockout	Horizontal 3-phase separator vessel, 11 ft × 17 ft S/S length, coated carbon steel	50	200	101,000	2	572.8	573.1	202,100	0.88	379,900
Demulsifier	Horizontal 3-phase separator vessel, 6 ft × 18 ft S/S length, coated carbon steel	50	200	57,500	2	572.8	573.1	115,100	0.88	216,400
Chemical injection pump	0.039 gpm to give 90 ppm by volume in oil	50	150	5,700	3	572.8	573.1	17,100	0.30	22,200
Chemical injection tank	790-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	11,400	2	572.8	573.1	22,800	0.88	42,900
Oil storage tank	API storage tank 9,000 bbl capacity each (for 3-d capacity at peak rate)	Hydrostatic + 2.5 psig	150	218,800	2	572.8	573.1	437,800	0.88	823,100
HP compressor, 50 MMscfd	50 MMscfd from 150 to 1,000 psig (5,751 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	3,412,900	1	579.7	573.1	3,374,000	0.49	5,027,300
HP compressor, 25 MMscfd	25 MMscfd from 150 to 1,000 psig (2,873 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	1,706,400	2	579.7	573.1	3,373,900	0.49	5,027,100
LP compressor	2.5 MMscfd from 25 to 150 psig (238 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	691,200	2	572.8	573.1	1,383,100	0.49	2,060,800
Dehydration	Treats 100 MMscfd after compression at 1,000 psig	1,100	150	2,444,000	1	573.1	573.1	2,444,000	0.72	4,203,700
Building	6,724 ft^2	NA	NA	838,600	1	572.8	573.1	839,000	0.00	839,000
Total								13,842,600		21,342,500

[1]Production rates: 118,000 Sm^3/h (standard cubic meters per hour; 100 MMscfd [million standard cubic feet per day]) of CO_2, 953 m^3/d of oil (6,000 bopd [barrels of oil per day]), and 3815 m^3/d of water (24,000 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; SS, stainless steel.

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

Item	Description	Design pressure rating, psig	Design temperature rating, °F	Unit equipment cost before indexing to Jan 2015, $	No. of units	Plant index equipment basis	Jan 2015 plant index	Jan 2015 total purchased equipment cost, $	Installation factor	Jan 2015 total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 7.5 ft ID × 24 ft S/S length, coated carbon steel	450	200	176,100	2	572.8	573.1	352,400	0.88	662,500
HP air cooler–1st stage, 25 MMscfd	SS air cooler, 2.82 MMBtu/h	750	350	128,300	2	564.8	573.1	260,400	0.61	419,200
HP air cooler–2nd stage, 25 MMscfd	SS air cooler, 3.66 MMBtu/h	1,200	350	150,200	2	564.8	573.1	304,800	0.61	490,700
HP air cooler–1st stage, 50 MMscfd	SS air cooler, 5.63 MMBtu/h	750	350	194,500	1	564.8	573.1	197,400	0.61	317,800
HP air cooler–2nd stage, 50 MMscfd	SS air cooler, 7.32 MMBtu/h	1,200	350	227,600	1	564.8	573.1	230,900	0.61	371,700
LP air cooler–1st stage	SS air cooler, 0.49 MMBtu/h	250	350	45,000	2	564.8	573.1	91,300	0.61	147,000
LP air cooler–2nd stage	SS air cooler, 0.52 MMBtu/h	450	350	46,500	2	564.8	573.1	94,400	0.61	152,000
Free water knockout	Horizontal 3-phase separator vessel, 11 ft × 17 ft S/S length, coated carbon steel	50	200	101,000	2	572.8	573.1	202,100	0.88	379,900
Demulsifier	Horizontal 3-phase separator vessel, 6 ft × 18 ft S/S length, coated carbon steel	50	200	57,500	2	572.8	573.1	115,100	0.88	216,400
Chemical injection pump	0.039 gpm to give 90 ppm by volume in oil	50	150	5,700	3	572.8	573.1	17,100	0.30	22,200
Chemical injection tank	790-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	11,400	2	572.8	573.1	22,800	0.88	42,900
Oil storage tank	API storage tank 9,000 bbl capacity each (for 3-d capacity at peak rate)	Hydrostatic + 2.5 psig	150	218,800	2	572.8	573.1	437,800	0.88	823,100
HP compressor, 50 MMscfd	50 MMscfd from 315 to 1,000 psig (3,148 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	2,558,600	1	579.7	573.1	2,529,500	0.49	3,769,000
HP compressor, 25 MMscfd	25 MMscfd from 315 to 1,000 psig (1,572 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	1,279,300	2	579.7	573.1	2,529,500	0.49	3,769,000
LP compressor	2.5 MMscfd from 25 to 315 psig (418 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	815,700	2	572.8	573.1	1,632,300	0.49	2,432,100
Dehydration	Treats 100 MMscfd after compression at 1,000 psig	1,100	150	2,444,100	1	573.1	573.1	2,444,100	0.72	4,203,900
Building	6,724 ft^2	NA	NA	838,600	1	572.8	573.1	839,000	0.00	839,000
Total						572.8	573.1	12,300,900		19,058,400

[1]Production rates: 118,000 Sm^3/h (standard cubic meters per hour; 100 MMscfd [million standard cubic feet per day]) of CO_2, 953 m^3/d of oil (6,000 bopd [barrels of oil per day]), and 3,815 m^3/d of water (24,000 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; SS, stainless steel.

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

Name	Description	Design pressure rating, psig	Design temperature rating, ºF	Unit equipment cost before indexing to Jan 2015, $	No. of units	Plant index equipment basis	Jan 2015 plant index	Jan 2015 total purchased equipment cost, $	Installation factor	Jan 2015 total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 7.5 ft ID × 24 ft S/S length, coated carbon steel	250	200	129,400	4	572.8	573.1	517,900	0.88	973,700
HP air cooler–1st stage, 25 MMscfd	SS air cooler, 3.14 MMBtu/h	500	350	136,900	2	564.8	573.1	277,800	0.61	447,300
HP air cooler–2nd stage, 25 MMscfd	SS air cooler, 5.75 MMBtu/h	1,200	350	197,300	2	564.8	573.1	400,400	0.61	644,600
HP air cooler–1st stage, 50 MMscfd	SS air cooler, 6.27 MMBtu/h	500	350	207,400	3	564.8	573.1	631,300	0.61	1,016,400
HP air cooler–2nd stage, 50 MMscfd	SS air cooler, 11.5 MMBtu/h	1,200	350	299,000	3	564.8	573.1	910,200	0.61	1,465,400
LP air cooler–1st stage	SS air cooler, 0.49 MMBtu/h	250	350	44,800	4	564.8	573.1	181,800	0.61	292,700
LP air cooler–2nd stage	SS air cooler, 0.50 MMBtu/h	250	350	45,300	4	564.8	573.1	183,900	0.61	296,100
Free water knockout	Horizontal 3-phase separator vessel, 11 ft × 17 ft S/S length, coated carbon steel	50	200	101,000	4	572.8	573.1	404,200	0.88	759,900
Demulsifier	Horizontal 3-phase separator vessel, 6 ft × 18 ft S/S length, coated carbon steel	50	200	57,500	4	572.8	573.1	230,100	0.88	432,600
Chemical injection pump	0.039 gpm to give 90 ppm by volume in oil	50	150	5,700	6	572.8	573.1	34,200	0.30	44,500
Chemical injection tank	790-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	11,400	4	572.8	573.1	45,600	0.88	85,700
Oil storage tank	API storage tank 9,000 bbl capacity each (for 3-d capacity at peak rate)	Hydrostatic + 2.5 psig	150	218,800	4	572.8	573.1	875,700	0.88	1,646,300
HP compressor, 50 MMscfd	50 MMscfd from 150 to 1,000 psig (5,751 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	3,412,900	3	579.7	573.1	10,122,100	0.49	15,081,900
HP compressor, 25 MMscfd	25 MMscfd from 150 to 1,000 psig (2,873 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	1,706,400	2	579.7	573.1	3,373,900	0.49	5,027,100
LP compressor	2.5 MMscfd from 25 to 150 psig (238 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	691,200	4	572.8	573.1	2,766,200	0.49	4,121,600
Dehydration	Treats 100 MMscfd after compression at 1,000 psig	1,100	150	2,444,000	2	573.1	573.1	4,888,000	0.72	8,407,400
Building	11,025 ft^2	NA	NA	1,370,000	1	572.8	573.1	1,370,700	0.00	1,370,700
Total								27,214,000		42,113,900

[1]Production rates: 236,000 Sm^3/h (standard cubic meters per hour; 200 MMscfd [million standard cubic feet per day]) of CO_2, 1,906 m^3/d of oil (12,000 bopd [barrels of oil per day]), and 7,630 m^3/d of water (48,000 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; SS, stainless steel.

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 Jan 2015, $	No. of units	Plant index equipment basis	Jan 2015 plant index	Jan 2015 total purchased equipment cost, $	Installation factor	Jan 2015 total installed cost, $
Slug catcher	Horizontal 2-phase separator vessel, 7.5 ft ID × 24 ft S/S length, coated carbon steel	450	200	176,100	4	572.8	573.1	704,800	0.88	1,325,000
HP air cooler–1st stage, 25 MMscfd	SS air cooler, 2.82 MMBtu/h	750	350	128,300	2	564.8	573.1	260,400	0.61	419,200
HP air cooler–2nd stage, 25 MMscfd	SS air cooler, 3.66 MMBtu/h	1,200	350	150,200	2	564.8	573.1	304,800	0.61	490,700
HP air cooler–1st stage, 50 MMscfd	SS air cooler, 5.63 MMBtu/h	750	350	194,500	3	564.8	573.1	592,100	0.61	953,300
HP air cooler–2nd stage, 50 MMscfd	SS air cooler, 7.32 MMBtu/h	1,200	350	227,600	3	564.8	573.1	692,800	0.61	1,115,400
LP air cooler–1st stage	SS air cooler, 0.49 MMBtu/h	250	350	45,000	4	564.8	573.1	182,600	0.61	294,000
LP air cooler–2nd stage	SS air cooler, 0.52 MMBtu/h	450	350	46,500	4	564.8	573.1	188,700	0.61	303,800
Free water knockout	Horizontal 3-phase separator vessel, 11 ft × 17 ft S/S length, coated carbon steel	50	200	101,000	4	572.8	573.1	404,200	0.88	759,900
Demulsifier	Horizontal 3-phase separator vessel, 6 ft × 18 ft S/S length, coated carbon steel	50	200	57,500	4	572.8	573.1	230,100	0.88	432,600
Chemical injection pump	0.039 gpm to give 90 ppm by volume in oil	50	150	5,700	6	572.8	573.1	34,200	0.30	44,500
Chemical injection tank	790-gal tank for 14 d of storage	Hydrostatic + 2.5 psig	150	11,400	4	572.8	573.1	45,600	0.88	85,700
Oil storage tank	API storage tank 9,000 bbl capacity each (for 3-d capacity at peak rate)	Hydrostatic + 2.5 psig	150	218,800	4	572.8	573.1	875,700	0.88	1,646,300
HP compressor, 50 MMscfd	50 MMscfd from 315 to 1,000 psig (3,148 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	2,558,600	3	579.7	573.1	7,588,400	0.49	11,306,700
HP compressor, 25 MMscfd	25 MMscfd from 315 to 1,000 psig (1,572 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	1,279,300	2	579.7	573.1	2,529,500	0.49	3,769,000
LP compressor	2.5 MMscfd from 25 to 315 psig (418 hp); 2-stage reciprocating compressor; assumes mix of coated carbon steel and SS components	NA	NA	815,700	4	572.8	573.1	3,264,500	0.49	4,864,100
Dehydration	Treats 100 MMscfd after compression at 1,000 psig	1,100	150	2,444,000	2	573.1	573.1	4,888,000	0.72	8,407,400
Building	11,025 ft^2	NA	NA	1,370,000	1	572.8	573.1	1,370,700	0.00	1,370,700
Total						572.8	573.1	24,157,100		37,588,300

[1]Production rates: 236,000 Sm^3/h (standard cubic meters per hour; 200 MMscfd [million standard cubic feet per day]) of CO_2, 1,906 m^3/d of oil (12,000 bopd [barrels of oil per day]), and 7,630 m^3/d of water (48,000 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; SS, stainless steel.