Illinois Basin – Decatur Project: Process Design and Operation of Carbon Dioxide Surface Facilities

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Front cover: *Pipeline (white) termination at the injection well (blue).*

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EXECUTIVE SUMMARY

Numerous organizations and researchers from government, academia, and industry have worked together as a team to carry out the Illinois Basin – Decatur Project (IBDP), a 1-million-tonne carbon dioxide (CO_2) storage demonstration project. The IBDP is led by the Midwest Geological Sequestration Consortium, one of seven U.S. Department of Energy Regional Carbon Sequestration Partnerships, and is managed by the Illinois State Geological Survey. The overall objective of this and other U.S. Department of Energy Regional Carbon Sequestration Partnership projects is to confirm that CO_2 injection and storage can be achieved safely, permanently, and economically.

This report addresses the process design and operation of the CO_2 surface facilities required to compress, dehydrate, and transport 1,102 ton/day (1,000 tonne/day) of CO_2 to the injection well over a 3-year injection period. Trimeric Corporation was responsible for the process engineering design of the compression and dehydration facility, and the company worked closely with the Archer Daniels Midland Company (ADM), which carried out the engineering, construction, and operation of the facility at the host site.

The initial challenge in designing the surface facilities was to establish a process design basis that specified the functional requirements for the surface facility equipment. The design basis addressed many special features that arose from the research nature of the project and site-specific details that would not normally have been encountered in a greenfield commercial project. For example, the shorter duration of the project led to placing greater emphasis on minimizing capital costs versus operating costs. The research objectives, combined with uncertainties in the required injection pressure, created the need for additional operational flexibility with respect to capacity turndown and the ability to deliver CO_2 over a wide range of surface injection pressures. Integrating the IBDP surface facilities into the ADM host site resulted in site-specific decisions, including the need for two separate equipment buildings to avoid covering underground piping and the need for an aboveground, insulated transmission pipeline.

The design basis provided the foundation for conducting a process design study that examined numerous configurations with regard to the purchased equipment and operating costs; availability and lead time of equipment; reliability, flexibility, and safety of the design; and process complexity. Ultimately, a process configuration was selected that consisted of a multistage centrifugal blower, followed by four stages of compression in two reciprocating compressors operating in parallel, with a triethylene glycol dehydration system configured between the third and fourth stages. A multistage centrifugal pump was also included after the reciprocating compressors to provide the capability of delivering the CO_2 at higher pressures if necessary to maintain the desired injection rate.

The preliminary process design was developed further into a detailed design together with a series of bid packages that were used to obtain firm, fixed-price bids from equipment suppliers for the major equipment. Concurrently, ADM began working on the civil, mechanical, electrical, and structural engineering needed to carry out the overall construction of the facilities. The design and construction phase of the project was completed in the fall of 2011.

The surface facilities were commissioned and started up in November 2011, and injection operations continued for 3 years until November 2014, when the injection goal was reached. During the 3-year injection period, 1,006,410.2 tons (913,000 tonnes) of net CO_2 was captured, after subtracting an estimated 95,901.1 tons (87,000 tonnes) of CO_2 emissions associated with generating the electrical energy required to operate the IBDP compression, dehydration, and transmission equipment. With a few minor exceptions, the CO_2 surface facilities met the operational requirements for the project. During the 3-year injection, the equipment was monitored closely and process data were collected and analyzed to characterize the performance of the system in terms of how well the equipment performed compared with the design requirements and to identify any lessons learned.

The total fixed capital investment for the compression, dehydration, and transmission facilities was \$20.3 million, the overall cost of injection was estimated at \$28.53/ton (\$31.45 per tonne) of CO_2, and the overall electrical requirements for compression were estimated at 101.6 kWh/ton (112 kWh/tonne). These results are not directly comparable to what might be expected for a large-scale, permanent commercial facility, primarily because of the small scale and short time frame over which the IBDP compression, dehydration, and transmission facility capital costs had to be recovered. The final section of the report (Full-Scale Power Plant Carbon Dioxide Compression and Dehydration) provides a detailed discussion of how the surface facilities would differ for a full-scale power plant application in terms of equipment selection, design, and overall costs.

The following significant lessons were learned on the project:

- Having a good estimate of the required surface injection pressure at the beginning of the design effort would be helpful.
- The time required to obtain permits, including the permit to inject, may be the most critical path on the project schedule because it affects the required equipment delivery times, the timing for the placement of equipment orders, and the timing of equipment warranties, which often expire 12 months after start-up or 18 months after shipment, whichever comes first.
- Economic conditions in general and particularly those related to the oil and gas industry may affect the cost and delivery time for the type of equipment described in this report.
- Realistic timelines need to be built into the construction and commissioning schedule for this type of project because of possible issues with equipment deliveries; weather; permitting; and the limited availability of key personnel, equipment, and site access needed for construction of this type of facility.

The following significant operational challenges and problems were encountered on this project:

- Acceptable cooling water return temperatures leaving the CO_2 process coolers need to be maintained over a wide range of seasonally affected cooling water supply temperatures. This issue has been addressed with a tempered water control loop at other facilities, but this solution was not selected by ADM.
- The equipment needs to be purged to remove residual moisture after equipment shutdowns.
- The type of cylinder lubrication oil needs to be carefully selected to optimize the oil injection rates.
- Our initial operational experience confirmed the prediction by the design team that insulation would be needed on aboveground piping to maintain a stable injection temperature and pressure regardless of ambient conditions and to make it easier to ensure compliance with permit conditions. The aboveground pipeline was insulated in the first few months of operation despite being a significant project expense.

INTRODUCTION

This report describes the process design and operation of the carbon dioxide (CO_2) compression and dehydration facilities associated with the Illinois Basin – Decatur Project (IBDP), a 1-million-tonne CO_2 storage demonstration project. The IBDP is led by the Midwest Geological Sequestration Consortium (MGSC), one of seven U.S. Department of Energy (U.S. DOE) Regional Carbon Sequestration Partnerships, and is managed by the Illinois State Geological Survey (ISGS) at the University of Illinois at Urbana-Champaign. In this report, we review the development of a process design basis to meet the overall CO_2 injection research objectives, and we evaluate the various process configurations that led to the final process design used by the Archer Daniels Midland Company (ADM) at the host site to complete the detailed engineering and construction for the project. The surface facilities were commissioned and started up in November 2011. Here, we present a detailed review of the operational performance of the surface facility equipment during the 3-year injection phase that followed. Actual performance is compared with the original design, and a detailed breakdown of the costs is presented. The total fixed capital investment for the compression, dehydration, and transmission facilities was $20.3 million. The overall capital and operating costs for compression, dehydration, and injection of the CO_2 were estimated at $28.53/ton ($31.45/tonne) injected, and electricity costs were estimated at 101.6 kWh/ton (112 kWh/tonne). These costs are reviewed and compared with what might be expected for a full-scale power plant application (approximately 10 times larger), which could be lower because of economies of scale and the longer project duration with a longer capital amortization schedule.

PROJECT BACKGROUND AND OBJECTIVES

The MGSC, one of seven Regional Carbon Sequestration Partnerships funded by the U.S. DOE, has completed injection at a large-scale CO_2 storage demonstration project called the IBDP, in which 1,102 tons (1,000 tonnes) of CO_2 was injected from November 2011 to November 2014. The host site was the Archer Daniels Midland Company (ADM) facility in Decatur, Illinois.

The ISGS is the lead organization in the MGSC, which provided overall management of the project; geologic expertise; and monitoring, verification, and accounting. Schlumberger Carbon Services, a division of Schlumberger, provided wellhead design and operations evaluation. Trimeric Corporation was responsible for the process engineering design of the compression and dehydration facility used to take atmospheric-pressure, water vapor-saturated CO_2 from the ADM ethanol fermentation process and deliver dehydrated, supercritical CO_2 to the injection well. In addition, Trimeric provided start-up and commissioning support, ongoing process engineering support, and data monitoring throughout the injection period. The Archer Daniels Midland Company, the site and permit owner, had overall responsibility for the design, engineering, construction, and operation of the surface facilities.

This report focuses on the surface facility equipment, which includes the CO_2 compression, dehydration, and transmission pipeline systems. Here, we provide an overview of the project design basis and requirements that led to the selected process design; a discussion of the facility construction and equipment installation, commissioning, and start-up; and lessons learned through the engineering design, construction, start-up, and injection operations. Also presented are a comparison of process data for the design versus actual equipment performance and maintenance requirements as well as process design considerations for scaling up the compression and dehydration equipment by a factor of 10. This larger scale represents a future 550-MW (net) power plant with 90% CO_2 capture. Discussion of the scale-up includes the difference in the process designs; equipment selection, construction, and installation; and a cost analysis.

SURFACE FACILITY DESIGN REQUIREMENTS

General Design Requirements

The general process design requirement was to compress and dehydrate a nominal 1,102 ton/day (1,000 tonne/day) of more than 99% pure CO_2 from ethanol plant fermenters and transport the compressed CO_2 to an injection wellhead located approximately 6,400 ft (1,950.7 m) from the compression facility. The compressed CO_2 was then injected into the Mt. Simon Sandstone.

Process Design Basis

The process design basis for the surface facilities defined the functional design requirements for the compression and dehydration equipment for CO_2 injection. The surface facilities were designed for 24-h operation, with no more than 30 days of downtime per year. Downtime included both scheduled and unscheduled downtime, as well as intentional shut-in days without CO_2 injection that were part of the overall test program. These requirements led to a minimum design rate of 1,096.8 ton/day (995 tonne/day) for the compression and dehydration equipment. The original CO_2 delivery requirements for the injection rate, pressure, and temperature are summarized in Table 1.

Each of the CO_2 delivery specifications in Table 1 is discussed below, along with its rationale:

- *Delivery flow*. As mentioned, the injection equipment was required to supply at least 1,096.8 ton/day (995 tonne/day).
- *Flow turndown*. The system was required to deliver CO_2 to the injection well at any specified flow rate of 275.6 to 1,096.8 ton/day (250 to 995 tonne/day). This range was requested by the ISGS to allow experimental flexibility over a wide range of injection rates.
- *Flow control*. The system was required to maintain CO_2 flow within ±10% of the targeted flow rate. For example, if a steady-flow test was being conducted at an injection rate of 551.2 ton/day (500 tonne/day), then the system had to be able to control the flow at 496.0 to 606.3 ton/day (450 to 550 tonne/day) during that test. This range was set based on the expected typical performance of compression systems operating at steady-flow conditions and the typical accuracy and performance of standard flow-control equipment.

Table 1 Original CO_2 delivery requirements

Delivery parameter	Project design requirement
Delivery flow	≥1,096.8 ton/day (≥995 tonne/day)
Flow turndown	Able to inject at rates ranging from 275.6 to 1,096.8 ton/day (250 to 995 tonne/day)
Flow control	Maintain within ±10% of target flow
Maximum wellhead inlet pressure	1,350–2,000 psig (93.1–137.9 barg)[1]
Minimum wellhead inlet pressure	During start-up: 0–1,057 psig (0–72.9 barg); during normal operations: 1,057 psig (72.9 barg)[1]
Pressure control	No specific requirement
Maximum temperature allowed at the injection wellhead	120 °F (48.9 °C)[1]
Minimum temperature allowed at the injection wellhead	88 °F (31.1 °C)[1]
Temperature control	NA

[1]The final underground injection control (UIC) permit (discussed in the Impact of Injection Permit Conditions on Carbon Dioxide Delivery Requirements section), which was issued after the design basis was finalized, specified a maximum injection pressure of 1,950 lb per square inch gauge (psig; 134.4 bar gauge [barg]) as measured at the wellhead, as well as minimum and maximum temperatures of 60 °F (15.6 °C) and 150 °F (65.6 °C), respectively, as measured near the wellhead. The minimum injection pressure requirement at the wellhead inlet was replaced by a requirement that the fluid should be in the supercritical state at the point of injection.

- *Maximum wellhead inlet pressure.* The system was required to deliver CO_2 at a maximum pressure of 1,350 to 2,000 lb per square inch gauge (psig; 93.1 to 137.9 bar gauge [barg]) at the wellhead. The lower limit of 1,350 psig (93.1 barg) was based on a compressor discharge pressure of 1,400 psig (96.5 barg) with an allowance for a pressure drop of up to 50 psi (3.4 bar) through the final discharge cooler and pipeline. (The compressor was specified to deliver CO_2 at up to 1,400 psig [96.5 barg]. If higher pressures were required to meet the desired injection rate, then a multistage centrifugal pump would be used to reach the final required injection pressure. This approach was taken because the exact pressure required to achieve the desired injection rate was not known at the time the compression equipment was ordered.)
- *Minimum wellhead inlet pressure.* During start-up, the system was required to accommodate the increase in injection wellhead pressure from 0 psig (0 barg) during initial start-up to the final operating pressure during normal operation. During normal operation, the original system requirement was to deliver CO_2 to the wellhead at pressures of at least 1,057 psig (72.9 barg), which was the minimum allowable wellhead pressure for injection as specified in the draft underground injection control (UIC) injection permit. (In the final UIC injection permit, the minimum injection pressure requirement at the wellhead was replaced by a requirement that the fluid should be in a supercritical state at the point of injection.)
- *Pressure control.* There was no design requirement to maintain control over pressure beyond meeting the minimum and maximum noted in Table 1. For example, if the wellhead pressure naturally settled out at 1,500 psig (103.4 barg) during injection, the system was not required to be capable of controlling the wellhead pressure to between 1,200 and 1,250 psig (82.7 and 86.2 barg). The pressure at the wellhead was generally determined by the reservoir characteristics and not by the surface equipment. It was anticipated, however, that the pressure would remain relatively steady during steady-state injection at a constant injection rate, and this proved to be the case during injection operations. The system operating procedures allowed operators to minimize rapid pressure changes during start-up and shutdown to less than 50 psig/min (3.4 barg/min). However, it was not possible to meet this requirement in all cases. For example, if a sudden loss of power occurred at the plant or an automatic compressor shut down, the wellhead pressure could drop rapidly and the pressure drop could exceed 50 psig/min (3.4 barg/min).
- *Maximum temperature allowed at the injection wellhead.* A temperature of 120 °F (48.9 °C) was selected because this is typical for the design of air-cooled heat exchangers during summer conditions. Actual temperatures were less than 120 °F (48.9 °C) because water-cooled exchangers were used to cool the CO_2. The final UIC permit raised this limit to 150 °F (65.6 °C).
- *Minimum temperature allowed at the injection wellhead.* A temperature of 88 °F (31.1 °C) was the original minimum temperature selected, based on the parameter limits in the draft UIC injection permit. During cold-weather start-ups and shutdowns, the CO_2 could be colder than 88 °F (31.1 °C) at the wellhead immediately after start-up. The final UIC permit lowered this limit to 60 °F (15.6 °C).
- *Temperature control.* There was no requirement to maintain the temperature within a specified range, other than between the minimum and maximum temperatures listed in Table 1. For example, there was no requirement to deliver CO_2 at temperatures between 100 and 110 °F (37.8 and 43.3 °C). However, the CO_2 temperature was expected to remain constant if the temperature of the cooling water supply was maintained at a constant

85 °F (29.4 °C). During injection operations, the CO_2 temperature at the wellhead was constant at approximately 98 °F (36.7 °C) after the transmission pipeline was insulated, even though the temperature of the cooling water supply varied over a wider range than envisioned during the process design phase of the project.

Impact of Injection Permit Conditions on Carbon Dioxide Delivery Requirements

The Illinois Environmental Protection Agency (EPA) required a Class I–nonhazardous UIC permit to be obtained for this project. (For the purposes of this report, all references are to the Class I UIC permit.)[1] Conditions in the final Class I UIC permit affected some of the delivery requirements and thus the process design of the surface facility. Table 2 includes the delivery requirements modified to comply with the final Illinois EPA Class I UIC permit requirements.

The specifications in Table 2 that differ from the original CO_2 delivery requirements in Table 1 are described below:

- *Maximum wellhead inlet pressure.* The final UIC permit lowered the maximum injection pressure at the wellhead to 1,950 psig (134.4 barg). The ISGS determined that this pressure would correspond to a bottom-hole pressure that was 75% of the fracture pressure gradient. This limit ensured that the injection operations would not cause any fracturing in the injection zone.
- *Minimum wellhead inlet pressure.* During start-up, the system was designed to accommodate an increase in the injection wellhead inlet pressure from 0 psig (0 barg) during initial start-up to the expected maximum operating surface pressure during normal operation. The draft UIC injection permit originally required the system to be capable of delivering CO_2 to the wellhead at a minimum pressure of 1,057 psig or 1,071 lb per square inch absolute (psia [72.9 barg]) during normal operations, which is the critical pressure of CO_2. The final UIC permit did not have a minimum pressure requirement at the wellhead. This condition was replaced by a requirement that the CO_2 should be in a supercritical state at the point of injection.
- *Maximum temperature as measured near the wellhead.* The final UIC permit had an upper limit of 150 °F (65.6 °C) for the CO_2 delivered to the inlet to the injection well; however, the process design specification remained at 120 °F (48.9 °C) because this is a common design outlet temperature for the air-cooled heat exchangers used in gas compression. The use of water-cooled exchangers resulted in lower surface injection temperatures year round, typically approximately 98 °F (36.7 °C).
- *Minimum temperature as measured near the wellhead.* The final UIC permit required injection of CO_2 in a supercritical state at the point of injection, which meant that the CO_2 had to be above its critical temperature of 88 °F (31.1 °C) and critical pressure of 1,057 psig or 1,071 psia (72.9 barg) when it reached the formation. It was noted in the process design phase of the project that transmission through an aboveground, uninsulated pipeline during cold weather, high winds, or heavy precipitation could result in a CO_2 temperature at the wellhead much colder than 88 °F (31.1 °C), even though the temperature of CO_2 would increase as it traveled down the injection well. The Archer Daniels Midland Company began insulating the pipeline when it was confirmed after a few weeks of initial operation that this was necessary, and insulation of the pipeline was completed approximately 5 months after injection operations began.
  Modeling of the downhole conditions of the CO_2 over the range of expected surface pressures and temperatures showed that the CO_2 pressure and temperature would increase because of hydrostatic compression of CO_2 in the injection well and that CO_2 delivered at 60 °F (15.6 °C) at the inlet of the wellhead would be heated to temperatures greater than 88 °F (31.1 °C) at the point of injection. This allowed us to monitor UIC permit compliance with surface pressure and temperature gauges, which are more accessible for calibration and maintenance than downhole instruments. The final UIC permit had a minimum temperature limit of 60 °F (15.6 °C) as measured near the wellhead.

Table 3 includes the purity specifications for the injection CO_2 stream in addition to these operational parameters. Each of the specifications in Table 3 and its rationale is discussed below:

- *Carbon dioxide.* The fermenter vent gas from the ethanol plant was generally high-purity CO_2 (>99 vol % on a dry basis). The CO_2 source for this project was downstream of an existing water-scrubbing system; thus, minimal (parts per million level) impurities remained in the CO_2 that was compressed and dehydrated before injection.
- *Oxygen.* Typical analysis of the source stream indicated that the normal oxygen content was less than 10 parts

Table 2 Underground injection control CO_2 delivery requirements

Operation parameter	Project design requirement
Injection rate	Range: 275.6–1,096.8 ton/day (250–995 tonne/day); permit maximum: 1,322.8 ton/day (1,200 tonne/day)
Maximum wellhead inlet pressure	1,350–1,950 psig (93.1–134.4 barg)[1]
Maximum temperature as measured near the wellhead	150 °F (65.6 °C)
Minimum temperature as measured near the wellhead	60 °F (15.6 °C)

[1]psig, pounds per square inch gauge; barg, bar gauge.

[1]The current UIC Class VI rule was not in place at the time the IBDP began. The project was initially permitted by Illinois EPA as a UIC Class I well and later converted to a Class VI well.

Table 3 Injection CO_2 purity specifications

Component	Purity specification[1]
Carbon dioxide	Minimum 99 vol %
Oxygen	100 ppmv maximum
Water	633 ppmv (30 lb/MMscf) maximum

[1]ppmv, parts per million by volume; lb/MMscf, pounds per million standard cubic feet.

per million by volume (ppmv). A measurement of oxygen significantly higher (10×) than this level would indicate an ingress of air from a possible loss of pressure control on a fermenter.

- *Water.* As stated above, the source point for the CO_2 was downstream of an existing water scrubber; thus, the CO_2 was considered water saturated. A typical CO_2 pipeline specification of 633 ppmv (30 lb/MMscf [pounds per million standard cubic feet]) was selected to prevent the formation of free water or hydrates in the transmission pipeline.

EQUIPMENT DESIGN FACTORS

Project- and location-specific factors affected the process design of the IBDP surface facilities.

Project Factors

Although the required equipment size was relatively large for what would be considered a demonstration facility, the research aspects of the project affected the process design. For example, the process equipment was highly instrumented to collect process data for future analyses. In addition, because the operation of the facility was not production critical, equipment sparing was minimal. Because of the turndown capability required, some simpler configurations (described in the Process Equipment Options Considered section) could not be considered.

Location Factors

Certain location-specific factors influenced the process design. One factor was the availability of cooling water to remove the compression heat. In many CO_2 compression applications, air-cooled heat exchangers (also known as fin-fan air coolers) are used for interstage cooling because no cooling water utility is available. Use of the cooling water utility resulted in a smaller footprint for the selected shell and tube heat exchangers than would the use of cooling water from the ADM host facility.

Another location-specific design factor was the location of the CO_2 source relative to the injection wellhead. The distance between these locations and the low pressure of the source CO_2 necessitated locating the compression and dehydration equipment near the CO_2 source, which allowed the use of a small-diameter transmission pipeline. In addition, because the equipment and transmission pipeline were to be located in an established facility, it was more economical to route the CO_2 transmission pipeline aboveground instead of underground.

A final location-specific factor was the unknown surface injection pressure required at the wellhead. During design, geologists and reservoir engineers modeled and estimated the wellhead pressure profile for the initial injection and throughout the 3-year injection period. The uncertainty associated with the surface injection pressure affected design considerations for the compressor discharge pressure.

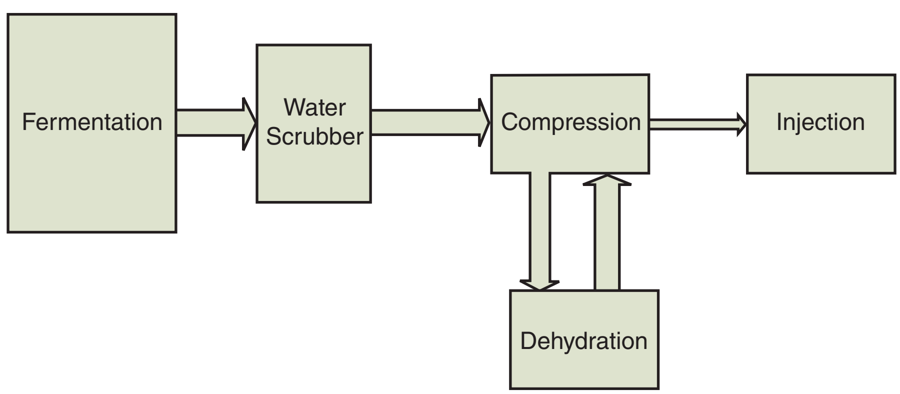

Figure 1 Block flow diagram.

Process Options Considered

A process study was conducted to identify the preferred equipment configuration for this project to compress the ethanol fermenter vent stream of CO_2 from approximately 1 psig (0.07 barg) to a maximum of 2,000 psig (137.9 barg) at the wellhead (Figure 1).

At the time the compressor configuration options were being analyzed, the final discharge pressure required was not known but was expected to be between 1,350 and 2,000 psig (93.1 and 137.9 barg), based on modeling work done by the ISGS and Schlumberger. Data from a simulation of the compression and dehydration equipment was used to estimate sizes for several configurations.

Process Equipment Options Considered

Several equipment options were initially considered for compressor configurations, process cooling, and dehydration. Each of these is described in the sections below.

Compression Configurations

Several configurations of compression equipment were considered to meet the design requirements of the project (Table 4).

The following is a final list of the combinations of compression equipment considered for CO_2 delivery:

- Blower, screw compressors, reciprocating compressors (Case 1)

Table 4 Compression equipment options considered

Case	Blower	Screw compressor	Reciprocating compressor	Chiller and condenser	Final-stage pump
1	X	X	X		
1a		X	X		
2	X	X	X		X
3	X	X	X	X	X
4	X		X		
4a			X		
5	X		X		X
6	X		X	X	X

- Blower, screw compressors, reciprocating compressors, pump (Case 2)
- Blower, reciprocating compressors (Case 4)
- Blower, reciprocating compressors, pump (Case 5)

A blower was considered for the initial compression of the CO_2 stream as a reliable, low-cost option for the initial compression stage to reduce the size and cost of a screw compressor or reciprocating compressor. For the remaining stages of compression, various combinations of screw and reciprocating compressors and multistage centrifugal pumps were considered. In addition, each of the configurations was evaluated without a blower to determine the impact of the blower on project costs.

The use of a centrifugal compressor was also considered. The centrifugal compressor vendors Trimeric contacted for budgetary quotes for this project chose not to provide them. One centrifugal compressor vendor did comment that this project was on the low end of flow rates at which centrifugal compressors are applicable. Centrifugal compressors also have longer equipment delivery times and more limited turndown capabilities than do reciprocating compressors. Because of these factors, centrifugal compressors were not considered further for this project. However, as explained in the Illinois Basin – Decatur Project Versus Full-Scale Power Plant Design Differences section, it is likely that multistage in-line or integrally geared centrifugal compressors would be favored over reciprocating compressors for a large-scale permanent installation.

Another option considered in the combinations described was the use of a multistage centrifugal pump. Once the density of CO_2 reaches approximately 35 lb/ft^3 (560 kg/m^3) or more, a multistage centrifugal pump is often used to raise the pressure of CO_2 further instead of using reciprocating or other types of compressors. Multistage centrifugal pumps in this application typically have lower capital and operating costs than do other types of compression equipment. A compression and liquefaction approach was also considered because it results in a liquid CO_2 stream that can be pumped for easy transportation. A site-specific analysis for the IBDP showed that liquefaction and pumping would not be more cost effective in terms of capital or operating costs relative to the other compression options considered.

Process Cooling

Process cooling for this project could have been provided either by air coolers (fin-fan coolers) or by shell and tube heat exchangers utilizing cooling tower water. A comparison of the two showed that shell and tube exchangers with a cooling water option would (1) provide tighter process temperature control within the surface facility equipment, (2) be more conducive to the planned indoor installation of the process equipment, and (3) have a smaller overall footprint, which was critical for this installation within the existing ADM facility. From a cost and performance standpoint, the air-cooled and water-cooled options were comparable; however, the water-cooled shell and tube heat exchangers had a slight advantage because lower interstage cooling and final compressor discharge temperatures could be achieved on a warm summer day. In many cases, air cooling would be the preferred option if cooling water were not available or where water resources were limited. Cooling water from an existing cooling water system was available for use on this project. Figure 2 shows a set of compressor aftercoolers, and Figure 3 shows the final cooler HE-204A/HE-204B that treated the combined discharge from both compressors.

Dehydration

The basis for setting the gas water content specification at 633 ppmv (30 lb/MMscf) was that it is typical of what might be specified for a commercial CO_2 pipeline transporting CO_2 for enhanced oil recovery or storage purposes. Dehydration prevents liquid water, solid CO_2–H_2O hydrates, or both from forming and helps prevent internal corrosion of the 6,400-ft (1,950.7-m) carbon steel pipeline. Both triethylene glycol (TEG) and solid desiccant (molecular sieve) dehydration systems were evaluated. A TEG dehydrator will normally dehydrate a CO_2 stream to a water content of approximately 148 ppmv (7 lb/MMscf). A molecular sieve or similar solid desiccant system typically dehydrates to a much lower water content, typically less than 1 ppmv. Triethylene glycol dehydration was selected because it provided an acceptable level of dehydration for the project requirements and a good safety margin below the requirements of the project design basis. The project requirements did not justify the higher capital and operating costs associated with molecular sieve dehydration.

PROCESS EQUIPMENT COMPARISON AND SELECTION

The following factors were used to select the final process equipment from the

Figure 2 Compressor aftercoolers.

Figure 3 Final cooler HE-204A/HE-204B.

options discussed in the Process Options Considered section:

- Budgetary purchased equipment cost (±20%) estimates
- Energy costs over a 3-year project life
- Overall system complexity
- Number of major process equipment skids
- Layout space requirements
- Equipment delivery time

In addition to the compression and dehydration options, this section includes the selection of transmission pipeline materials.

Compression Equipment

Providing detailed results from the comparison of all the compression options studied during the budgetary cost estimate stage of the project is beyond the scope of this report. In general, the sum of purchased equipment costs and 3 years of electrical energy costs varied over a fairly narrow range that was within the accuracy of the ±20% budgetary cost estimates provided by the suppliers of the compression equipment. Excluding the blower would have reduced the configuration complexity and layout space requirement; however, this option significantly increased the estimated purchased cost of the subsequent compression equipment by an average of 35% and increased the 3-year electrical cost by an average of 4%.

Because differences in the estimated purchased equipment cost were relatively minimal at the budgetary cost estimate stage, other factors, including system complexity and the number of major process skids, were used in combination with these differences in cost to select two options for which firm quotes were ultimately received. The first of these options included one blower, two screw compressors, one reciprocating compressor, and one multistage centrifugal pump. The second option included one blower, two reciprocating compressors, and one multistage centrifugal pump. The difference in purchased equipment costs plus 3 years of electrical energy costs for these two options was again minimal (<5%). Therefore, ADM chose the second option because the process was simpler and had one fewer major process skids. Thus, the compression equipment configuration, or compression train, selected for the project consisted of one blower, two reciprocating compressors, and one multistage centrifugal pump. Figure 4 shows a block flow diagram for the selected compression train. Initial, intermediate, and final pressures as well as the water content of the CO_2 are also shown.

The overall process flow diagram in Figure 5 provides additional details on

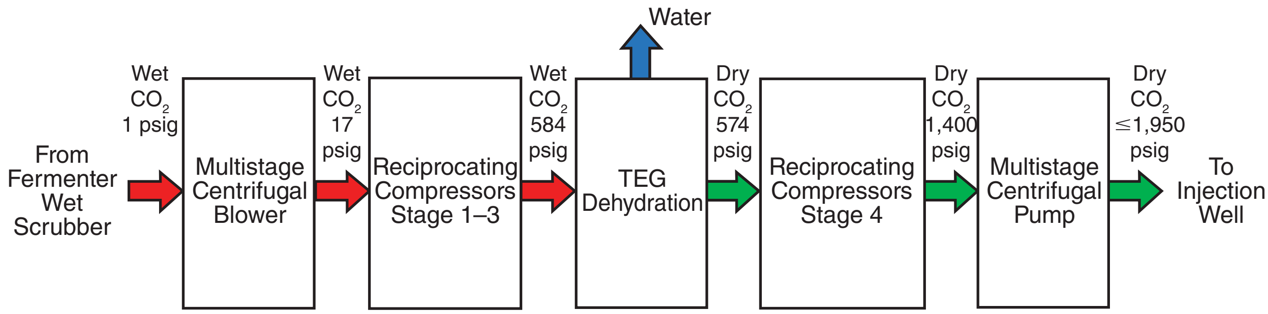

Figure 4 Block flow diagram of the compression train. Red arrows indicate wet CO_2, and green arrows indicate dry CO_2. psig, pounds per square inch gauge; TEG, triethylene glycol.

the type and quantity of equipment needed for the various components in the compression train, including the blower, reciprocating compressors, dehydration unit, and multistage centrifugal pump, and it shows how these subsystems are integrated. Important control parameters are also depicted.

Figure 6 shows the multistage centrifugal blower system. The blower achieved four stages of compression by a set of four impellers (wheels) mounted on a single shaft that rotated at a fixed speed of 3,550 rpm. The blower shaft was driven by a 4,160 V, 933 kW (1,250 hp) motor. Figure 7 shows one of the two reciprocating compressors that operated in parallel. Each compressor had four stages of compression and was driven by a 4,160 V, 2,425 kW motor (3,250 hp) that operated at a fixed speed of 715 rpm. Each compressor used two 26.5-in. (67.3-cm)-diameter cylinders for the first stage of compression, two 17.9-in. (45.4-cm) cylinders for the second stage of compression, one 12.5-in. (31.75-cm) cylinder for the third stage of compression, and one 7.2-in. (18.4-cm) cylinder for the fourth stage of compression. Finally, Figure 8 shows the multistage centrifugal pump that was capable of increasing the pressure of the dense-phase CO_2 to 1,950 psig (134.4 barg), as delivered to the wellhead. This pump had 26 stages and was driven by a 460 V, 149 kW (200 hp), 3,600 rpm motor. The pump motor was equipped with a variable-frequency drive, which allowed the pump to operate at reduced speeds, as required based on the flow rate and the head (pressure rise) developed across the pump.

Dehydration Equipment

The CO_2 vent stream leaving the post-fermentation water scrubber is saturated with water vapor. Dehydration of the CO_2 is needed to prevent corrosion and eliminate the potential for the formation of solid CO_2–H_2O hydrates. A significant portion of the water vapor is removed during each stage of compression, cooling, and phase separation. However, after the minimum saturation water content at the interstage cooling temperature and pressure for the CO_2 is reached, another dehydration technique is needed. Two common options considered for dehydration of the CO_2 were solvent absorption, such TEG, and adsorption using a solid desiccant, such as molecular sieves.

Solid desiccant systems, such as molecular sieve units, utilize fixed media beds to remove water from gas streams. When the media become saturated, they must be regenerated. Therefore, these systems have multiple vessels for dehydration as one dryer-bed vessel is typically undergoing regeneration. They also require a heater for the regeneration system. As mentioned previously, molecular sieve dehydration can achieve outlet concentrations lower than 1 ppmv. This is significantly lower than the CO_2 water content specification of 633 ppmv (30 lb/MMscf) for this project.

Liquid dehydration systems dehydrate gas streams by contacting the gas with an absorbing solution, such as TEG, in an absorber column. The rich (water-loaded) TEG solution is then regenerated by using a reboiler and a stripper column, and the lean TEG is returned to the absorber column.

A TEG dehydration unit was chosen for the IBDP dehydration equipment because the level of dehydration achieved by a molecular sieve unit was not required for this application and because TEG dehydration units are known to have lower capital and operating costs than do solid desiccant systems such as molecular sieves. Figure 9 shows the TEG contactor that absorbs the water vapor in the CO_2 into the TEG. Figure 10 shows the dehydration regeneration skid that evaporates the water from the TEG so that the TEG can be recycled to the contactor.

Transmission Pipeline

Several line sizes were evaluated for a pipeline to the injection well. These results are summarized in Table 5. The process conditions for the inlet to the transmission pipeline and for the piping components used in this evaluation were as follows:

- Temperature: 95 °F (35 °C)
- Inlet pressure: 1,415 psia (97.6 bar absolute [bara])
- Mass flow: 101,045 lb/h (45,833.2 kg/h)
- Density: 40 lb/ft^3 (640 kg/m^3)
- Viscosity: 0.0595 cP (5.95e^{-5} Pa·s)
- Pipe length: 3,200 ft (975.4 m)
- Fittings: 8–90° elbows
- Four full-bore ball valves
- Four run-through tees
- One swing check valve

The value of 3,200 ft (975.4 m) for the transmission pipeline length was the initial estimate selected for this evaluation because the actual pipeline length, which

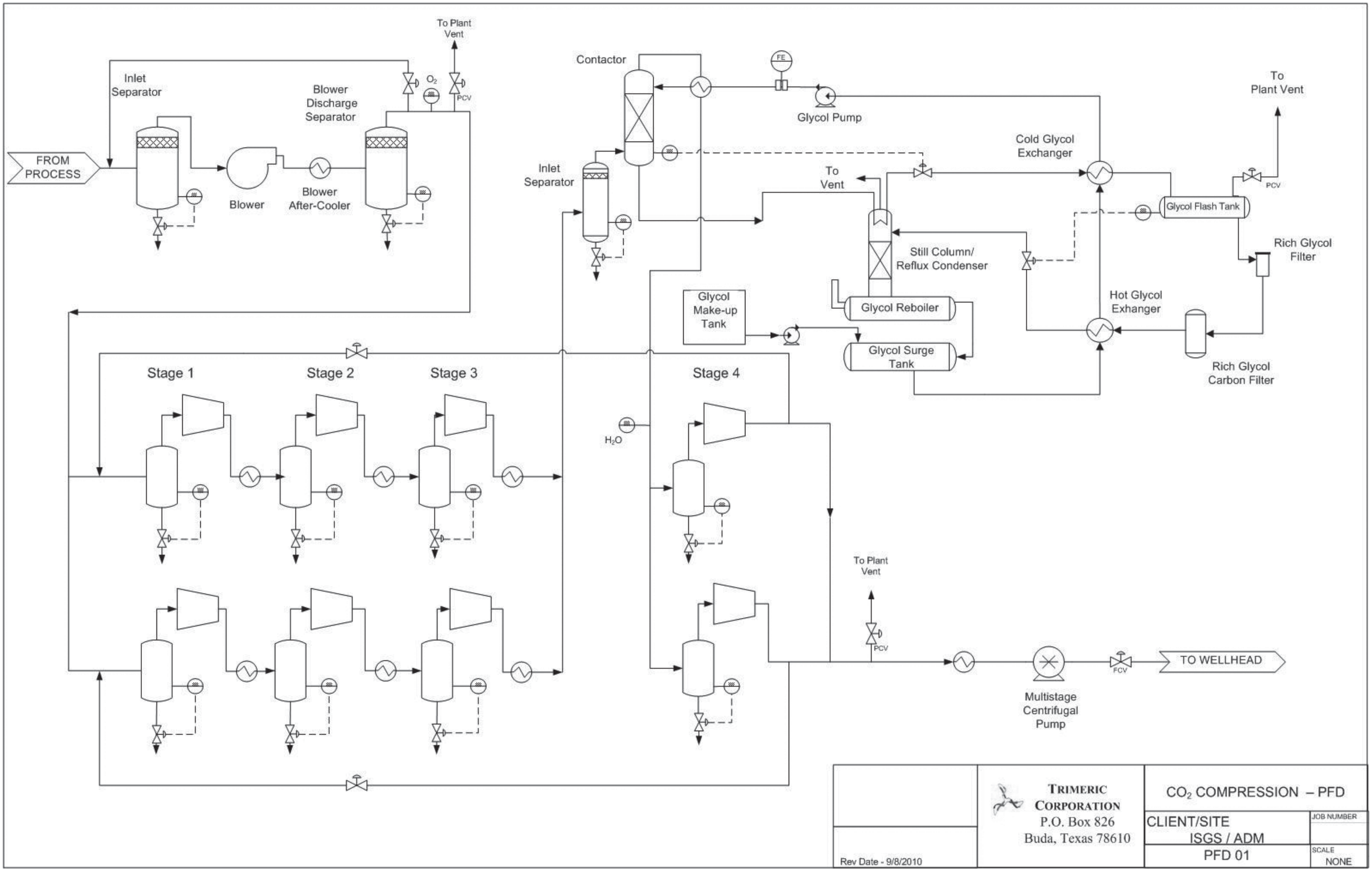

Figure 5 Overall process flow diagram.

Figure 6 Blower system.

Figure 7 Reciprocating compressor.

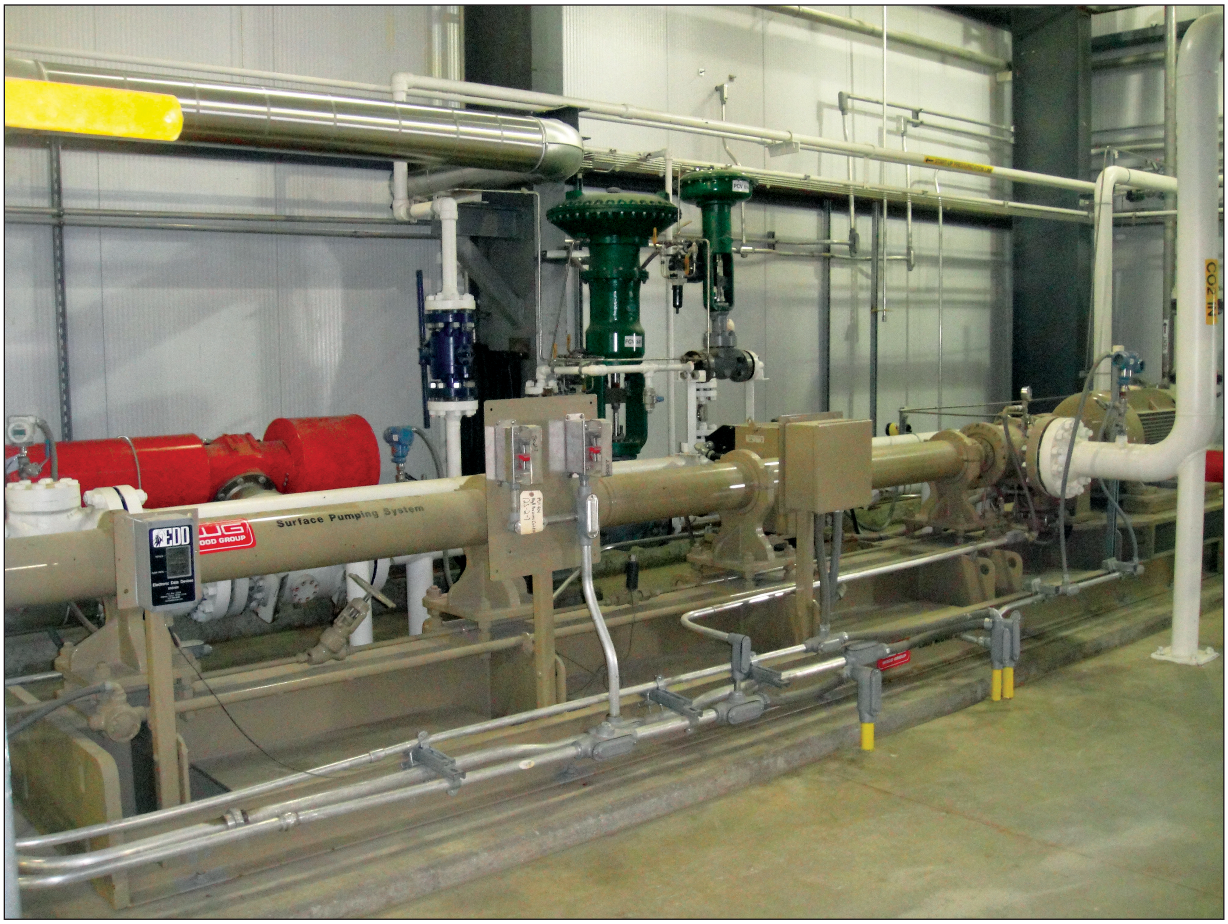

Figure 8 Multistage centrifugal pump.

Figure 9 Triethylene glycol contactor.

Figure 10 Dehydration regeneration skid.

Table 5 Pipeline sizing evaluation

Parameter	Unit	Pipe schedule 40	80	40	80	40
Nominal pipe size	in. (mm)	4 (101.6)	4 (101.6)	6 (152.4)	6 (152.4)	8 (203.2)
Outside diameter	in. (mm)	4.5 (114.3)	4.5 (114.3)	6.625 (168.3)	6.625 (168.3)	8.625 (219.1)
Inside diameter	in. (mm)	4.026 (102.26)	3.826 (97.18)	6.065 (154.05)	5.761 (146.33)	7.981 (202.72)
Fluid velocity	ft/s (cm/s)	7.80 (237.7)	8.76 (267.0)	3.50 (106.7)	3.87 (118.0)	2.01 (61.3)
Calculated pressure drop	psi (bar)	44.50 (3.07)	57.50 (3.96)	5.40 (0.37)	7.00 (0.48)	1.30 (0.09)

was twice that, at 6,400 ft (1,950.7 m), was not known at the time of the pipeline diameter and materials evaluation. The actual pipeline length was longer because of the need to fit within existing pipe rack structures and corridors to the greatest extent possible. Nevertheless, the results of this evaluation were still able to be used to make pipeline diameter and materials selections for the longer actual pipeline length.

The velocity in the 4-in. (101.6-mm) schedule 40 line was reasonable for CO_2 flow at the specified conditions; however, the associated pressure drop was unacceptable. The calculated pressure drop for the 8-in. (203.2-mm) piping indicated the line would have been significantly oversized.

In addition to the cost impact of the pressure drop on energy requirements, the various piping grades as well as the secondary impact on structural costs affected the purchased material cost because of the different weights of the pipeline materials. Table 6 includes a comparison of piping costs for 6-in. (152.4-mm) ASTM A-106 Grade B and 6-in. (152.4-mm) API 5L X52 pipe.

Trimeric performed an additional analysis to show the cost of the increased energy usage associated with using a 4-in. (101.6-mm) pipeline versus a 6-in. (152.4-mm) pipeline. The energy penalty used for the impact of the pressure drop on compressor operations was $5,628/year for the 4-in. (101.6-mm) pipeline and $804/year for the 6-in. (152.4-mm) pipeline. Figure 11 shows the purchased costs of pipeline materials and the costs of electrical energy for pipeline combinations of 4-in. (101.6-mm) and 6-in. (152.4-mm) ASTM A-106 Grade B and API X52 pipe.

Although purchased and electrical costs for 6-in. (152.4-mm) A-106 Grade B piping were less than those for 6-in. (152.4-mm) X52 piping, the selected pipeline size and material was 6-in.

Table 6 Transmission piping cost data[1]

Grade	Schedule	Material	Wall thickness, in. (mm)	Specified minimum yield, psig (barg)	Pipe cost, \$/ft (\$/m)[2]	Weight, lb/ft (kg/m)
ASTM A-106 Grade B	80	Carbon steel	0.432 (10.97)	35,000 (2,413.2)	28.83 (94.59)	28.57 (42.52)
API 5L X52	STD	Carbon steel	0.280 (7.11)	52,000 (3,585.3)	32.09 (105.28)	18.97 (28.23)

[1]ASTM, ASTM International (West Conshohocken, Pennsylvania); API, American Petroleum Institute (Washington, DC); psig, pounds per square inch gauge; barg, bar gauge.
[2]Pipe quotes from April 2009 (based on 3,200 ft [975.4 m]).

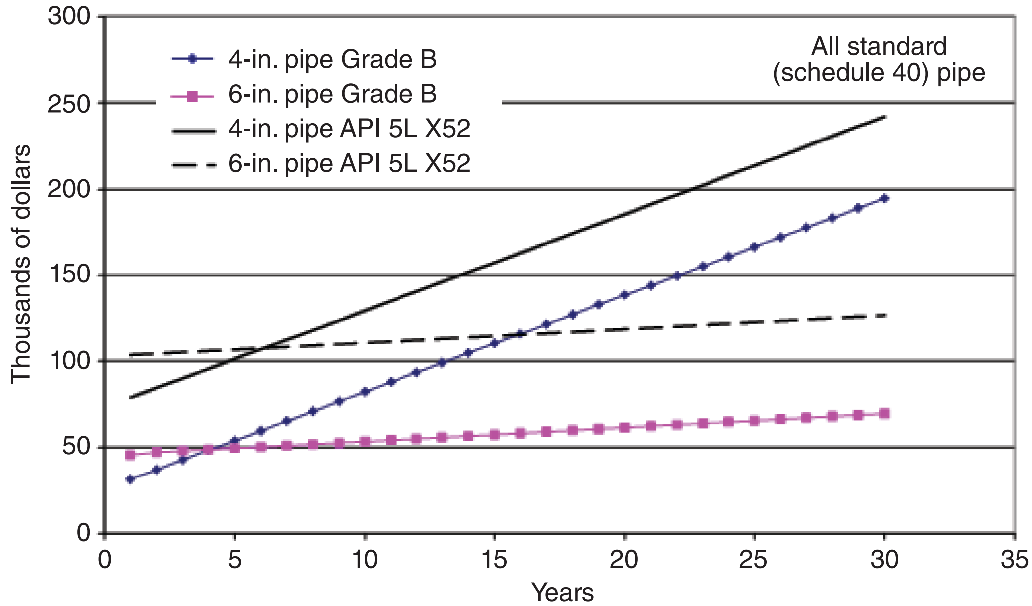

Figure 11 Cost analysis for a 4- versus 6-in. (101.6- vs. 152.4-mm) pipeline.

Figure 12 Pipeline (white) termination at the injection well (blue).

(152.4-mm) schedule 40 X52 because of other factors such as the higher anticipated structural cost to support the 3,200-ft (975.4-m) pipeline if 6-in. (152.4-mm) A-106 Grade B were chosen. Figure 12 shows the end of the pipeline that terminates at the injection well.

Equipment Specifications and Request for Quote Development

Trimeric developed equipment specification and bid request packages for the centrifugal blower, screw compressors, reciprocating compressors, and TEG dehydration unit and issued them to multiple suppliers as a request for quote (RFQ) on a fixed-price, firm-bid basis. The multistage centrifugal pump did not require a formal bid package, but Trimeric did provide the pump vendors with similar specifications for these quotes. The RFQ documents specified that major equipment was to be installed and shipped on process skids, including all interconnecting piping within the skids, which reduced the amount of field work required to install this equipment.

The formal bid requested documents for the blower, screw compressors, and reciprocating compressors included in the design parameters, material of construction requirements, piping, lubrication requirements, driver motor specifications, process control system requirements, and required codes and standards. The RFQ also defined the vendor-supplied drawing requirements, delivery and shipping schedule, equipment warranty period, recommended spare parts, availability and cost of start-up, operations and maintenance support, and on-site training requirements.

The bid request documentation for the TEG dehydration unit included the design basis for each component of the dehydration unit, a process flow diagram, the heat and material balance, the material of construction requirements, column internals, motor specifications, and requirements for the process control system. The RFQ also defined the vendor-supplied drawing requirements, delivery and shipping schedule, recommended spare parts, availability and cost of start-up, operations and maintenance support, and on-site training requirements.

Bid Evaluation, Ranking, and Comparison

After firm bids were received from vendors for the blower, screw compressors, and reciprocating compressors, Trimeric and ADM conducted a comparison of the bids. Each bid was first reviewed for completeness and any exceptions that the vendors had taken to the specifications in the RFQ. The purchased equipment cost and cost of electricity data were compiled for each configuration.

Used Equipment and Preengineered Compression Packages

In addition to obtaining quotes for new compression equipment, an effort was made to locate used equipment that might lower the project capital costs and shorten the project schedule. An effort was also made to locate preengineered off-the-shelf equipment that might shorten the project schedule.

One used compressor was located at an existing facility. The compressor was intact, but, as is often the case with used equipment, it needed substantial modifications and restoration to be used for compression at the IBDP. The estimated costs of modifications are shown in Table 7. The purchase cost for the used compressor was $700,000. After considering the additional cost to modify and refurbish this compressor, the total cost of the used compressor was within the range of budgetary quotes that Trimeric had received for new units. The search results for preengineered packages were limited because of the materials required for constructing the wet CO_2 stream; the preengineered equipment Trimeric identified was all carbon steel construction.

DETAILED PROCESS DESIGN SUPPORT

Piping and Instrument Diagrams and Detailed Engineering Drawing Reviews

Trimeric provided process engineering support in developing the initial piping and instrument diagrams (P&IDs) that incorporated the major equipment components and showed how they were interconnected from the inlet of CO_2 from the ADM wet scrubber source to the surface connection at the wellhead.

Trimeric reviewed the vendor-provided P&IDs for the blower, reciprocating compressors, and TEG dehydration unit. Markups were provided to clearly identify mechanical connections and process control signals between these units, the rest of the system, and the ADM host facility.

Table 7 Estimated used compressor costs

Item	Cost, $MM[1]
Compressor removal from existing building by a reputable demolition company	0.150
Refurbishment of compressor	0.500
Motor rewind for new application	0.125
New piping, intercoolers, and knockouts for CO_2 service	0.150
Additional cost of the foundation over a new, modern compressor	0.050
Miscellaneous	0.100
Total estimate, excluding purchase	1.075
Total estimate, including purchase	1.775

[1]$MM, million dollars.

Additional support included review of the detail-engineering firm's pipeline isometric drawings, development of utility P&IDs, sizing of low-pressure and high-pressure vent headers, and sizing of pressure-relief valve discharge lines to the vent headers.

Process Hazard Analysis and Process Hazard Analysis Revalidation

Several process hazard analysis (PHA) sessions were conducted for the IBDP surface facility equipment, using the hazard and operability study (HAZOP) technique to identify safety issues. Two separate HAZOPs were conducted before the detailed design. The blower, two reciprocating compressors, and the TEG dehydration unit were analyzed in the first session, whereas the multistage centrifugal pump, aboveground pipeline, and wellhead connections were analyzed in the second session. Recommendations from the HAZOP sessions were used to modify the process design and process control strategy, as necessary, to mitigate risks identified in the PHA sessions.

A final closeout PHA was also conducted to (1) review operational and process changes that occurred between the initial PHA sessions and commissioning of the facility; (2) evaluate any process changes that could have an impact on safety; and (3) review transmission pipeline procedures before the pipeline was placed into service.

Risk Management Evaluation

Given the scope and complexity of the overall project, a significant component of the overall project management effort involved a formal risk management program for the IBDP. This program, led by Schlumberger, is described in Hnottavange-Telleen (2014) in detail. The formal risk management program for IBDP was comprehensive in scope and included the following major areas of consideration:

- Air quality
- Site management
- Subsurface activities
- Surface facility engineering
- Community relations
- Ownership and the environment

The formal program involved a series of workshops and project team meetings, which led to the development of a list of 123 features, events, and processes, or FEPs, that were evaluated for their risks based on the estimated likelihood and severity of the consequences. These items, along with the specific actions required to mitigate the risk associated with each item, were assigned to various members of the project team and tracked by the risk management program leader as the project progressed.

Trimeric was assigned items that were associated with the design and operation of the surface facilities for compression, dehydration, and transport of the CO_2. Trimeric's risk mitigation items fell into four categories:

- Development of the "injection operations envelope"
- Preparation of an injection operations and shut-in plan
- Preparation of an equipment-sparing plan
- Preparation of an operational monitoring plan

Each of these activities is summarized briefly below.

Injection Operations Envelope

A general area of risk was the negative impacts resulting from excursions from the originally intended or "optimal" operating conditions. To mitigate these risks, the equipment was designed to operate over as wide a range of conditions as was practically achievable. The acceptable range of operating conditions was also described as an "operations envelope" and was defined in terms of process variables such as the injection flow rate, CO_2 purity, temperatures and pressures at various points in the process, and ambient conditions at the site. Analysis of the injection operations envelope involved systematically examining all the possible operating conditions with regard to risks associated with excursions from these conditions as they affected individual subsystems and operating parameters. This analysis was used as a resource during the subsequent design activities when detailed specifications for equipment were developed. In particular, examining the operations envelope led to a detailed understanding of what equipment would be required to adequately monitor and control the system during normal operations, what to expect during upset conditions, and how to control the system during upsets and normal or emergency shutdowns.

Injection Operations and Shut-in Plan

Trimeric developed a site-specific injection operations and shut-in plan that addressed several key elements identified during the risk management workshops:

- The scope of the plan included constant (normal) injection operations, as well as planned and unplanned interruptions in injections.
- The plan accounted for measures that were needed to adjust operations in response to weather stresses and expected variations in the injection rates.
- The plan specifically addressed such items as the potential reverse well flow of CO_2 and brine, reverse pipeline flow, stack venting of CO_2, brine separation and disposal, and procedures and documentation for immediately and automatically shutting in the injection well and suspending CO_2 injection for an indefinite period, if required.

Equipment-Sparing Plan

Trimeric developed an equipment-sparing plan that included a list of critical equipment, maintenance schedules, and equipment replacement lead times (procurement and installation). Both operational and monitoring equipment were included in the equipment-sparing plan, along with the loss, severity, hazard, and risk associated with the loss or malfunction of each item of critical equipment.

Operational Monitoring Plan

To enable a quick response to protect the health and safety of workers and the environment in the event of a hazardous condition, Trimeric helped develop a plan to continuously monitor live data streams from sensors measuring operational and key surface environmental parameters. This plan considered numerous technical items that were originally identified in the risk workshop meetings:

- Automated versus manual monitoring
- Visual inspections
- Implications of anticipated false-positive and false-negative readings
- Predefinition of sensor readings that trigger shutdown of any part of the injection system
- Information flow and communications issues
- Physical and stakeholder (human relations and communications) impacts of shutdown and notification scenarios
- Maintenance of system safety (preventing hazardous conditions and damaging impacts) in case of a malfunction of electronic or computer-controlled systems

Equipment Performance and Factory Automation Tests

Trimeric sent an engineer to two separate equipment fabrication facilities in Houston, Texas, to witness the blower performance test and to participate in a factory automation test (FAT) for the blower and reciprocating compressor process skids. Although the blower performance test used air instead of CO_2, the results did demonstrate that the unit functioned as designed based on the head (pressure rise) versus flow rate tests conducted at the manufacturing facility. The FAT was used to validate and refine the control methodologies for the blower and reciprocating compressor skids. This approach reduced the amount of time required for commissioning and start-up in the field, which resulted in an overall lower cost to the project. The dehydration unit and multistage centrifugal pump were considered standard units and did not warrant travel and labor costs for performance tests and FAT evaluations before shipment.

Process Control Narrative

An initial process control diagram was created to identify the major control loops and establish the process controls and communications among the blower, reciprocating compressor, TEG dehydration unit, and pump. A written control narrative was then developed and later expanded and revised following the FAT. The equipment vendors and ADM control engineers used these control documents for programmable logic controller (PLC) programming, and Trimeric used the documents to develop the start-up and operating procedures.

Instrumentation and Analyzers: Flow, Water, and Oxygen

Specific instruments and analyzers were required to maintain process control of the facility, collect research data, and ensure that the facility met the CO_2 delivery and UIC permit requirements. An orifice plate-type flowmeter was selected for CO_2 flow measurements. This flowmeter had real-time pressure and temperature measurements and used an on-board computer to calculate the CO_2 density at actual conditions to convert from the measured volumetric flow rate to a mass flow rate. This kind of orifice plate-type flowmeter is often used for custody transfer in CO_2 enhanced oil recovery applications. As an additional control measure and for research purposes, a second orifice flowmeter was also installed in-line. One flowmeter measured the combined flow leaving both reciprocating compressors upstream of the high-pressure vent valve, and the other measured the flow of CO_2 entering the transmission pipeline. Operational data collected by ADM and analyzed by Trimeric showed good agreement between the two flowmeters when the high-pressure vent valve was fully closed.

A moisture analyzer was selected to measure the water content of the CO_2 stream leaving the TEG dehydration unit and moving to the fourth stage of the reciprocating compressors. The water content analysis is very important not only to meet the CO_2 delivery requirements, but also because the selected material of construction from the fourth stage of compression through the pipeline up to the wellhead was carbon steel, which could corrode rapidly if CO_2 and liquid water were present at the same time. Because the CO_2 source was at a low pressure, we were concerned about possible air ingress from leaks in the fermentation unit or at the inlet of the IBDP compression facility. Therefore, an analyzer to measure oxygen content in the CO_2 was also included in the design.

CONSTRUCTION AND INSTALLATION

Impact of Facility Modification Versus Greenfield

The IBDP compression and dehydration equipment and transmission pipeline had to be integrated into a major industrial complex with minimal disruption to ongoing operations in the host facility. This task came with some obvious challenges when compared with installing this type of equipment in a greenfield location, but it also provided several advantages, which were difficult to anticipate but easy to appreciate in retrospect.

Figure 13 shows how the compression and dehydration facility and transmission pipeline were integrated into the existing ADM facility. As shown in Figure 13, the new equipment had to be carefully located within the overall facility, with consideration given to the available space and project requirements.

Different approaches were required to accommodate the installation of the compression and dehydration facility while still meeting the functional requirements for the new system. The low pressure of the source CO_2 (1 psig, or 0.069 barg) required the compression equipment to be located close to the ethanol fermentation unit, which was one project constraint. In addition, to avoid covering the underground facility piping, the compression and dehydration equipment had to be located in two buildings. The blower and the regeneration skid for the dehydration unit were installed in one building, and the two reciprocating compressors were installed in another. A greenfield installation would likely use only one building for all the equipment.

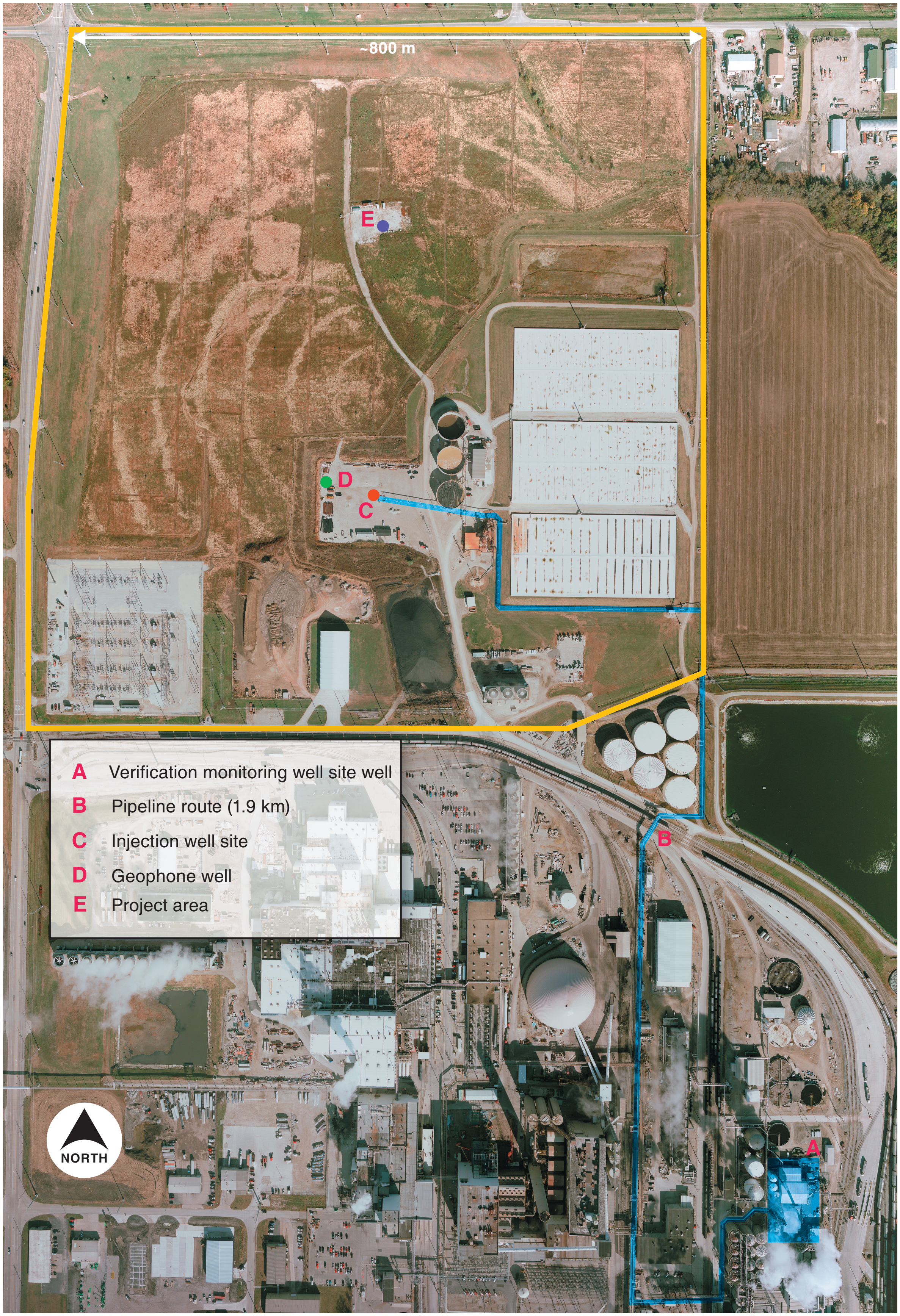

Figure 13 Aerial view of Illinois Basin – Decatur Project surface facilities within the existing Archer Daniels Midland Company facility.

The 6,400-ft (1,950.7-m)-long, 6-in. (152.4-mm)-diameter pipeline had to be installed aboveground in existing pipe racks. The CO_2 transmission pipeline would typically be installed underground in a greenfield installation. As discussed elsewhere in this report, the aboveground pipeline had to be insulated to maintain stable injection well inlet temperatures and pressures over a wide range of ambient conditions and to simplify operations that were required to stay above the minimum allowable injection temperature limit in the UIC injection permit. Insulation would likely have been unnecessary had the pipeline been installed underground.

Installing the IBDP compression and dehydration equipment and transmission pipeline within the ADM facility also had several advantages over a typical greenfield installation. First, a highly qualified ADM project team and supporting skilled-trade contractors were already working at the ADM facility and were available to support the installation of the IBDP compression and dehydration equipment and transmission pipeline. Other ADM staff for engineering, environmental, and electrical controls and instrumentation were available to the project on an ongoing and as-needed basis as dictated by the project requirements over the course of its execution. Readily accessible electricity, cooling water systems, and other key utilities were another key advantage.

Impact of Working Within an Existing Plant Versus Greenfield

All personnel involved in the installation, commissioning, start-up, and operation of the IBDP compression and dehydration equipment and transmission pipeline were required to comply with the protocols and policies of the ADM facility. This included meeting ADM requirements for training in contractor safety, participating in safety drills, and using the required personal protective equipment at all times while working in the facility. It was important for IBDP personnel to be aware of and comply with these requirements; thus, similar requirements would likely have to be developed for similar projects at a greenfield site. The IBDP personnel and project benefited from having these protocols and policies already in place, as evidenced by the outstanding safety record for the project during all phases of construction, installation, commissioning, start-up, and operation.

COMMISSIONING AND STARTUP, AND LESSONS LEARNED

Commissioning and Startup

Constructing facilities such as the IBDP compression and dehydration equipment and transmission pipeline is a complex, multiple-year operation involving a wide variety of skill sets from different disciplines, including construction management; safety, environmental, and permitting specialization; and mechanical, electrical, structural, civil, and process chemical engineering. After the facility was constructed and the equipment was installed, the commissioning and start-up process took approximately 12 weeks, from August through November 2011.

Archer Daniels Midland hired field-commissioning personnel from a major equipment supplier, Enerflex Energy Systems (Calgary, Canada), for on-site commissioning support. Enerflex support included checking for proper motor alignment and loop checking of all instrumentation to verify proper electrical continuity and voltage, proper instrument calibration and ranges, and a proper system response to alarm conditions, as indicated by these instruments and the PLC. Enerflex personnel were also responsible for ensuring that control loops on their skids were properly tuned to respond to changing conditions. Control loop tuning took longer than expected, primarily to balance the control loop responses to perform in a stable manner during normal operations and unusual conditions, such as those following the automatic shutdown of one of the compressors.

Electricians, instrument technicians, and control engineers from ADM had similar responsibilities for the off-skid equipment and for incorporating all the equipment controls and instrument displays into the ADM distributed control system (DCS). Archer Daniels Midland assigned a full-time engineer to the project during the commissioning phase. Trimeric engineers worked alongside ADM personnel to ensure that the overall construction was completed in accordance with the process design. Trimeric engineers rotated in on a weekly basis and compared the P&IDs with actual construction by physically inspecting every line and component in each major system in the facility, including the blower system, compressor systems, dehydration system, multistage centrifugal pump system, and transmission pipeline, to ensure that construction and commissioning were completed in accordance with the P&IDs, vendor specifications, project specifications, and other facility design requirements. Trimeric developed a construction punch list that ultimately included 96 items, and Trimeric personnel worked with ADM personnel and external contractors to ensure that each item on the punch list was properly addressed before start-up.

During commissioning, Trimeric also prepared a detailed “rounds and readings” sheet for operators to review and record system performance manually during each shift. These readings were taken in addition to automated parameters recorded in real time in the ADM DCS. In Trimeric’s experience, there is value in physically inspecting the equipment during each shift or on a daily basis, even when most of the parameters are automatically recorded in a DCS. This allows the personnel supporting the equipment to observe trends over time on a daily basis, particularly on items such as vibration; unusual noise; and gas, water, oil, and valve leaks that cannot always be measured by using a standard automated data collection system. Trimeric also developed written operating procedures for initial commissioning (“equipment lineup”), start-up, normal shutdown, and emergency shutdown, and Trimeric personnel trained ADM engineers and operators in the use of these procedures and other aspects of control and monitoring.

Lessons Learned from Commissioning and Startup

Trimeric’s experience with the IBDP commissioning and start-up and with similar commercial efforts suggests that a schedule for bringing this type of facility online needs to have realistic timelines built in. The limited availability of key personnel, equipment, and other resources can interrupt the flow of the commissioning and start-up effort. Weather can also have significant impacts, although this was not the case at the IBDP. As a result, a project like this benefits from having one or more leaders with the vision to set time-specific goals and the drive to motivate and lead others to achieve these goals.

On the IBDP, the time required to issue the UIC injection permit significantly increased the overall timeline for the project. This led to a delay of more than a year between installing the equipment and constructing the major facility and beginning the commissioning and start-up effort. Such a long gap in the project timeline has implications for equipment warranties. Most U.S. suppliers provide a warranty on this type of equipment for up to 12 months after start-up or for 18 months after shipment, whichever comes first. A delay in the commissioning and start-up effort that extends beyond 18 months after shipment can create difficulty in obtaining warranty support or coverage from a supplier. Trimeric was not aware of this being an issue on the IBDP, although it has been an issue on other similar commercial projects.

The long delay also made it more difficult for project participants to maintain the flow of project decision making. In addition, some small items, including a sound attenuator for a vent valve and some internal parts for the CO_2 flowmeters, were lost and had to be found or replaced. In cases where long delays are unavoidable, thorough documentation and record keeping can help reduce disruptions.

SYSTEM OPERATIONS

Operation Philosophy

A secondary objective at the IBDP was to operate at or above the nominal 1,102 ton/day (1,000 tonne/day) injection rate on the days the system was in operation, as opposed to running more days per year but more frequently at a reduced injection rate. Because the capital costs on this project were expensed over a 3-year period, as opposed to a more typical 10- to 30-year period on a commercial project, the U.S. DOE and project team made the decision not to purchase and install in-line spares for the major equipment, such as the blower, compressors, and multistage centrifugal pump. Two compressors were installed in parallel, but this decision was dictated by the capacity limits of the compressors rather than a sparing philosophy. To ensure that the system would meet the target injection rate of 1,102 ton/day (1,000 tonne/day), a 10% safety factor was added to the design. Thus, the design capacity for the system was 1,212.5 ton/day (1,100 tonne/day, or 21 million standard cubic feet per day [MMscfd]).

Some injection took place in November 2011, but most of that month was consumed by final commissioning, start-up, and shakedown of the compression and dehydration equipment. As discussed in detail elsewhere in the report, there were several reasons for the days that injection did not occur. These included plant outages during which the host facility was unable to supply CO_2, electricity, cooling water, or other utilities; scheduled maintenance for the compression and dehydration equipment; unscheduled maintenance for this equipment; and days when reduced injection rates or no injection occurred in support of the research aspects of the project, namely, when pressure and other formation responses were measured in response to step-rate changes in the injection rate and in response to shut-in conditions.

Operation Staffing

Typical days and weeks during the IBDP injection phase consisted of 24-hour-a-day continuous operation. The Archer Daniels Midland Company assigned a chemical engineer to oversee operations for the plant and to supervise the work of four operators who supported the IBDP compression and dehydration facility. One operator was on duty for each 8- to 12-hour shift, and the operators rotated so that continuous operator coverage was provided for the equipment. In Trimeric's experience, this represents an upper end of the amount of coverage that would typically be provided for this type of facility. By comparison, other similar plants that Trimeric has designed for commercial clients are staffed with a supervisor and one on-duty operator for 8 to 12 hours per day for 5 to 7 days per week, and an automated call-out system is used to notify off-site personnel if any issues occur during times when the plant is unattended. The costs associated with the ADM staffing plan were built into the project economics at the beginning of the project, and this level of staffing was very effective in ensuring that operations were properly controlled, monitored, and supported and that issues were promptly identified and corrected.

Quarterly Site Visits

The project closely monitored the operation of the equipment and collected related data during the 3-year injection period. During this time, Trimeric sent an engineer to the site approximately every 3 to 4 months (i.e., quarterly). The Trimeric quarterly site visits helped provide consistency as ADM underwent normal engineering and operator staffing changes during the 3-year injection period. Trimeric performed the following activities during these visits:

- Discussed operations and issues with ADM engineers and operators.
- Worked with ADM and equipment suppliers to improve operations and devise recommendations to correct operating issues.
- Compared the current facility configuration with the P&IDs and provided redlines (i.e., recommended safety limits) to ADM so that ADM could keep the P&IDs current.
- Used "rounds and readings" sheets to physically inspect all the compression and dehydration equipment and related instruments and record operating parameters.
- Collected CO_2 flowmeter data to verify flowmeter calculations, and used a standalone PC to confirm that the flowmeter calculations were correct.
- Tracked the consumption of cylinder lubrication oil for the reciprocating compressors and glycol in the dehydration unit.
- Collected hourly average operating values for several hundred parameters provided by ADM from their DCS for review and analysis in the Trimeric office. These data were the source of most of the values used in this report.

Quarterly Site Visit Trip Reports

A trip report with a standardized format was prepared and issued for each visit and was used as a guideline for subsequent visits to ensure consistency in the level of effort and in the items covered in each of the site visits. After each trip, these reports were reviewed and evaluated to determine the direction of the project. They were a valuable resource for documenting the history of system performance and for comparing operations from year to year.

Key Findings and Project Decisions Resulting from Quarterly Site Visits

Some key areas were addressed during the quarterly visits:

- *Managing cooling water temperatures.* The interstage CO_2 temperatures were controlled by modulating the amount of cooling water that flowed through each heat exchanger in the process. The maximum temperature in the original CO_2 outlet design was 95 °F (35 °C), based on the summer design condition of an 85 °F (29.4 °C) cooling water supply. Because of seasonal temperature changes, the cooling water entering the CO_2 compression facility during the winter months was much colder than in the summer design condition. As a result, the control valves modulating the water flow rate to each heat exchanger partially closed to limit the amount of cooling water flowing through the exchangers. The cooling water discharge temperature in this scenario rose to much higher temperatures than the design value of 110 °F (43.3 °C). At times, the cooling water return temperatures were greater than 160 °F (71.1 °C). Trimeric and ADM had concerns that at these temperatures, the stainless steel tube bundles in each exchanger might be subjected to stress corrosion cracking from the chlorides present in the cooling water. They were also quite concerned that the carbon steel of the cooling water return lines might corrode at these temperatures. Trimeric worked with ADM, the engineering firm managing the local cooling water treatment for ADM, and the compressor vendor (Enerflex) to mitigate this issue by lowering the original set point temperature of the process CO_2 from 95 to 80 °F (35 to 26.7 °C) and by lowering the recommended shutdown temperature from 80 to 65 °F (26.7 to 18.3 °C). (At 65 to 80 °F [18.3 to 26.7 °C], CO_2 remains above the pressure range at which liquid CO_2 or solid CO_2–H_2O hydrates can form within these operations.) After confirming that the rod load and other parameters were within the operating limits, we determined it may have been possible to operate with CO_2 process temperatures lower than 80 °F (26.7 °C), particularly upstream of the third stage of compression. However, from a process or operational standpoint, there was not a strong motivation to pursue CO_2 temperatures lower than 80 °F (26.7 °C).
- *Heat exchanger performance monitoring and maintenance.* Trimeric analyzed the operating data provided by ADM to monitor the performance of the heat exchangers in the compression and dehydration facility. A reduction in heat exchanger performance often indicates fouling of the heat exchanger. The cooling water provided to this facility was shared with several other units in the ADM facility, including units that cool streams containing solids. A tube leak in a heat exchanger in one of these other units during the first year of injection caused fouling in every heat exchanger in the compression and dehydration unit, as indicated by ADM operator readings and Trimeric analysis of the heat exchanger performance data. A similar event occurred in the second year of injection. Trimeric and ADM worked with the heat exchanger maintenance companies to develop a heat exchanger cleaning procedure that ADM could implement on its own (at a lower cost and with less downtime than by using outside service companies), which was successful in restoring the performance of the fouled heat exchangers to satisfactory levels.
- *Rod-load shutdown logic implementation.* The control system for the reciprocating compressors was originally programmed to shut down the reciprocating compressors on low or high suction pressure or on low or high discharge pressure. These shutdowns were intended to avoid any damage to the reciprocating compressor rods because of excessive force from the pistons on the piston rods (an issue known as rod load). The suction and discharge pressure shutdown set points were set at conservative levels, and during some start-up operations or plant upset conditions, the rod load caused automatic compressor shutdowns. Trimeric worked with ADM and the compressor vendor (Enerflex) to enter calculations into the control system to estimate the rod load based on the differential pressure across the cylinder(s) at each stage of compression and to shut down the machine based on high rod-load conditions. Low suction pressure and low discharge pressure shut-downs were disabled in the control system, which greatly reduced the number of nuisance trips based on pressure fluctuations during start-up and upset conditions. High suction pressure and high discharge pressure shutdowns remained active.
- *Operational problem tracking.* Trimeric and ADM worked closely together to identify equipment, systems, or both in the compression and dehydration facility that did not work as well as intended, including the effects of cylinder lubrication pump failures and failures of their rupture disk pressure relief devices on the reciprocating compressors and damage to separator water drain valves caused by debris. By tracking items that did not perform as well as designed, ADM, Trimeric, the equipment manufacturers, and others will be able to design around these issues in future plants. For example, the compressor manufacturer released a new kind of cylinder lubrication pump and a new pressure relief device for the cylinder lubrication system, and these subsequently worked much better at the IBDP facility and other locations.

Process Data—Actual Versus Design

Injection Rate

The process design basis for the compression and dehydration facility required the capacity to inject at least 1,096.8 ton/day (995 tonne/day) and as much as 1,212.5 ton/day (1,100 tonne/day) of CO_2 into the formation, with a turndown capacity as low as 275.6 ton/day (250 tonne/day). Turndown below 606.3 ton/day (550 tonne/day) was managed by venting some of the compressed CO_2 upstream of the HE-204A/HE-204B final

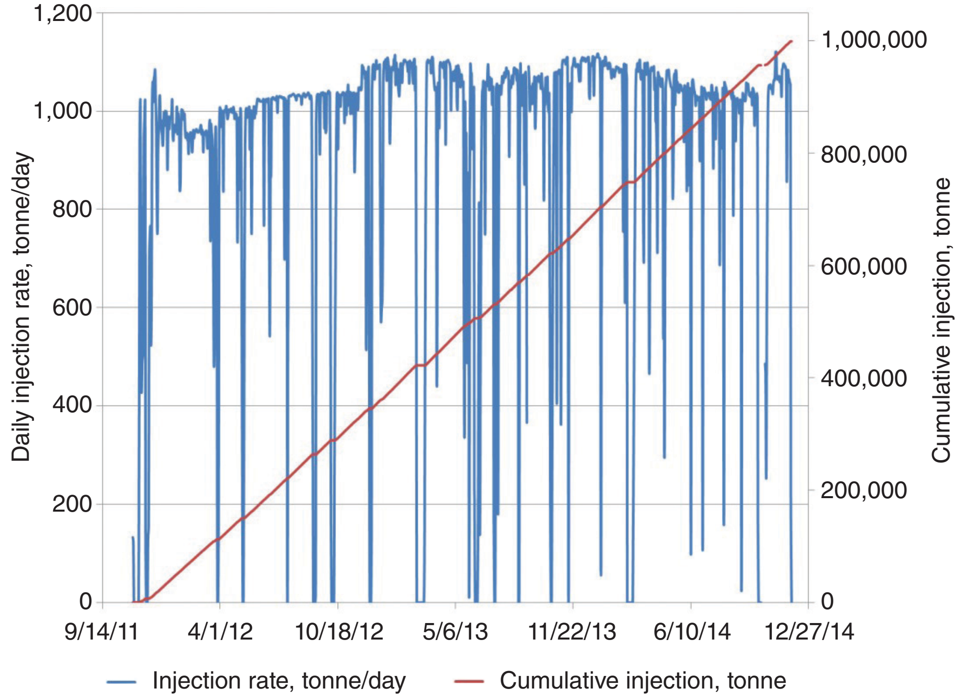

Figure 14 Injection rates over the 3-year injection period.

cooler through the vent header into the atmosphere. The injection flow rate was measured by orifice plate meter FIT-006. This meter was temperature and pressure compensated and utilized the Span–Wagner equation of state to calculate the density of the CO_2 as it passed through the meter. Figure 14 shows the daily and cumulative injection rates for the IBDP throughout injection.

The injection flow rate requirement of 1,096.8 ton/day (995 tonne/day) was met or exceeded by the facility for most of the injection period, except for a short time in early 2012. The injection rate was then reduced because ADM was developing the operating practices needed to comply with an administrative injection rate limit based on the size of the orifice plate in the injection flowmeter. This administrative limit was removed in December 2012, as evidenced by the clear uptick in the daily injection rate (see Figure 14), which confirmed that the IBDP system was capable of meeting its requirement to inject at rates up to 1,212.5 ton/day (1,100 tonne/day). However, ADM generally operated a little below this rate to ensure adequate margins for compliance with the revised permit limit.

Injection Pressure

The compression and dehydration facility was also designed to operate over a wide range of surface injection pressures, which were initially between 1,350 and 2,000 psig (93.1 and 137.9 barg). The equipment had to be ordered before drilling the injection well, so coring data and results from the water injection test that would have given a firmer estimate of the required surface injection pressure were not available at the time the equipment was designed and ordered. Initially, the expected range of surface injection pressure was estimated based on modeling work performed by other project team members, including personnel from the ISGS and Schlumberger. The actual injection pressure required was not known until IBDP injection operations began. Figure 15 shows the surface pressure at the wellhead as measured by pressure transmitter PIT-009.

As the facility came online and began to inject CO_2 into the formation, it became apparent that the injection pressure would not exceed approximately 1,400 psig (96.5 barg) during the initial phase of injection, even when injecting 1,212.5 ton/day (1,100 tonne/day). As a result, the multistage centrifugal injection pump was not commissioned at the beginning of injection. The low injection pressure in early 2012 corresponds to the lower injection flow rate associated with the administrative conditions in the permit, which can be seen in Figure 14. After the administrative restriction was removed, the surface injection pressure remained constant at approximately 1,350 psig (93.1 barg).

However, the surface injection pressure showed a slight upward trend (Figure 15) over the 3-year injection period. In the third year of injection, Trimeric worked with ADM and the compressor vendor (Enerflex) to raise the discharge pressure limits on the fourth (final) stage of the reciprocating compressors. Even then, ADM had to vent a small amount of CO_2 (nominally 55 ton/day [50 tonne/day] or less) in the final months of injection to stay within acceptable discharge pressure limits for the compressors. This result suggests that if injection operations had continued beyond 3 years, some actions may have been required to maintain the nominal rate of injection. Possible actions may have included adjusting the compressor capacity to reduce the throughput

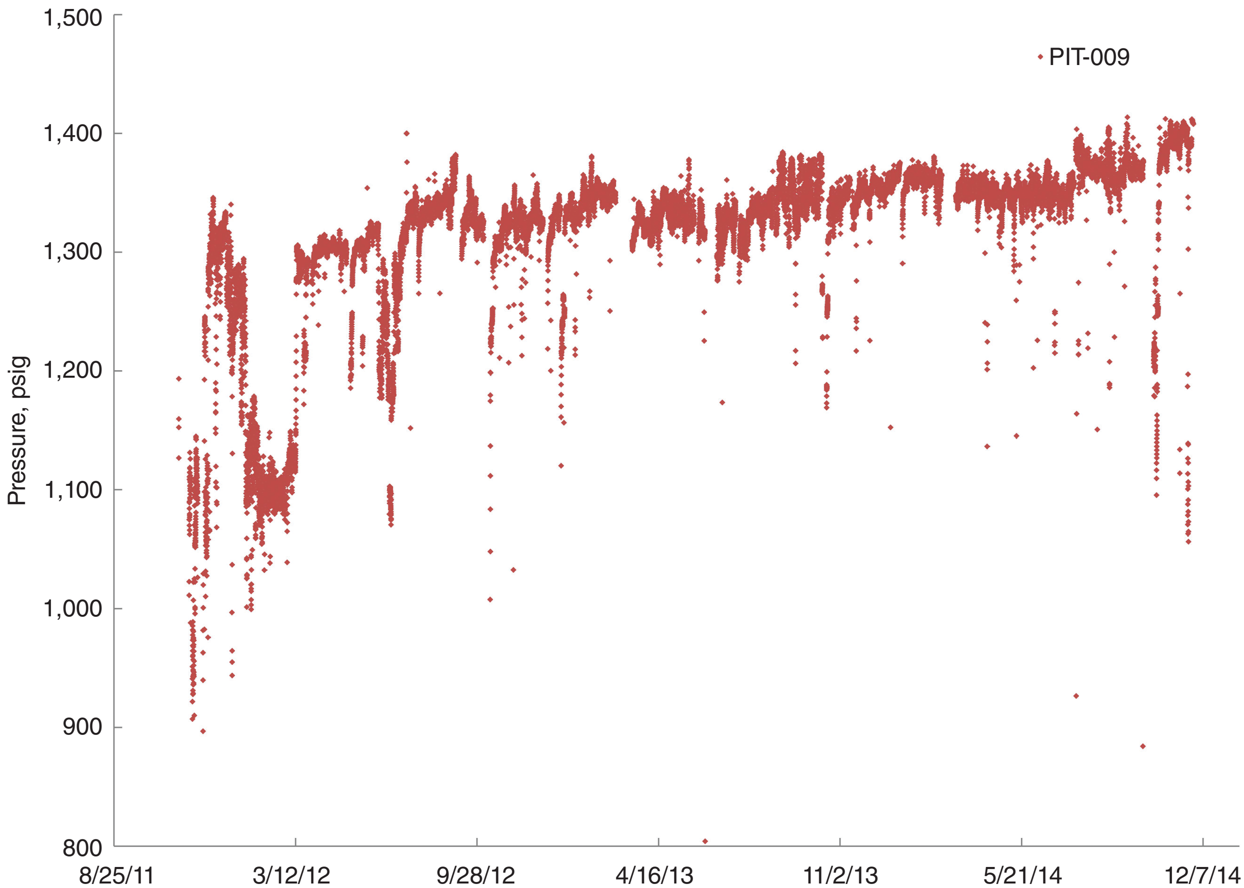

Figure 15 Injection well surface pressures throughout the injection period.

to achieve the same net injection rate without venting any CO_2 downstream of the compressors, cleaning the injection well, or using a multistage centrifugal pump.

The suction pressure for the facility was designed to be 15.0 psia (1.03 bara). Figure 16 shows the suction pressure at the blower inlet as measured by pressure transmitter PI-101A. The suction pressure of the blower was often below the design pressure of 15.0 psia (1.03 bara). In part, this was because the throttling valve used upstream of the blower suction scrubber worked as intended to prevent overloading the blower motor at times when the incoming gas was cooler, and therefore at a higher density, than in the summer design condition. Pressure at the inlet of the blower increased over the course of the injection period, primarily when ADM replaced its upstream wet scrubber packing with a different type of packing designed for a low-pressure drop and changed to a control method that provided a more stable inlet pressure to the IBDP surface facilities.

Carbon Dioxide Process Coolers

The IBDP compression and dehydration facility had a number of shell and tube heat exchangers that used cooling water to remove heat from the CO_2 as it was compressed in the blower and reciprocating compressors. Table 8 shows the actual performance of these exchangers compared with their design. This table compares the overall heat transfer rate, *Q*, and the approach temperature, which is the difference between the cold CO_2 outlet temperature and the cold cooling water inlet temperature. The calculated values are based on operating data during a period when the compression facility was operating at full rates and the temperature of the cooling water supply to the facility was close to the design maximum value of 85 °F (29.4 °C).

The project specifications required all heat exchangers to be designed with a safety factor of at least 10%. According to the data sheets from the heat exchanger manufacturer, the actual safety factors ranged from 14% to 25%. As shown in Table 8, all exchangers operated at a duty lower than design, which is in part because of the design safety factors. The approach temperatures were lower than the design, which indicates the heat exchanger performance was better than the design, except for HE-303 and HE-204A/HE-204B. Analysis of data from the trip report obtained prior to the data selected for comparison in Table 8 suggested that HE-303 restricted water flow, which could explain why the approach temperature was slightly higher than the design for this heat exchanger. The CO_2 discharge temperature leaving final cooler HE-204A/HE-204B was purposely maintained at 97 to 98 °F (36.1 to 36.7 °C), which could be achieved year round in this heat exchanger regardless of the

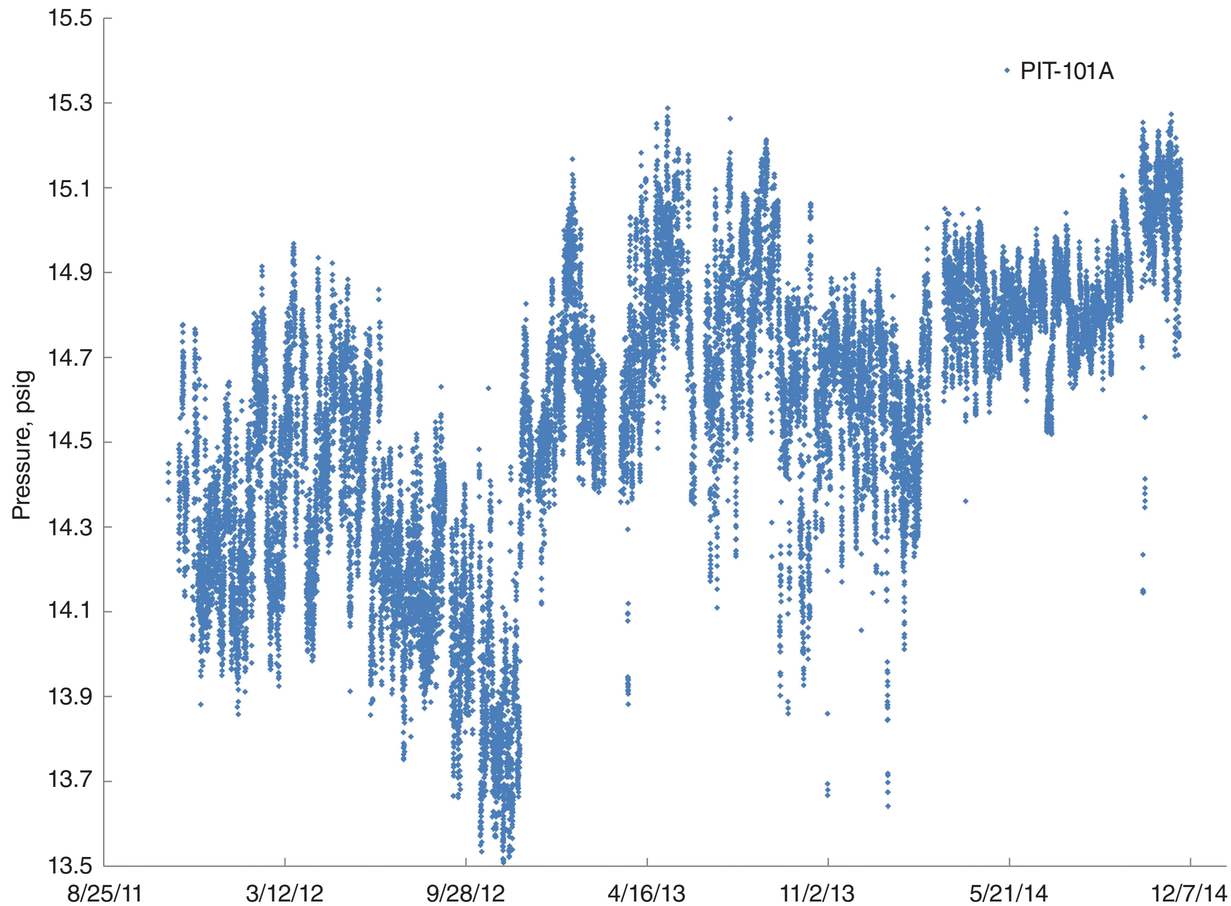

Figure 16 Blower suction pressures for the Illinois Basin – Decatur Project.

Table 8 Heat exchanger performance for the Illinois Basin – Decatur Project compression and dehydration facility

Exchanger	ID	*Q*, design, kW (MMBtu/h)	*Q*, calculated, kW (MMBtu/h)	Design approach temperature, °F (°C)	Actual approach temperature, °F (°C)
Blower aftercooler	HE-101	1,348 (4.6)	1,290 (4.4)	10 (–12.2)	8 (–13.3)
Compressor VC-201, Stage 1 aftercooler	HE-201	791 (2.7)	703 (2.4)	10 (–12.2)	6 (–14.4)
Compressor VC-201, Stage 2 aftercooler	HE-202	821 (2.8)	762 (2.6)	10 (–12.2)	6 (–14.4)
Compressor VC-201, Stage 3 aftercooler	HE-203	615 (2.1)	586 (2.0)	10 (–12.2)	9 (–12.8)
Compressor VC-301, Stage 1 aftercooler	HE-301	791 (2.7)	733 (2.5)	10 (–12.2)	8 (–13.3)
Compressor VC-301, Stage 2 aftercooler	HE-302	821 (2.8)	791 (2.7)	10 (–12.2)	8 (–13.3)
Compressor VC-301, Stage 3 aftercooler	HE-303	615 (2.1)	586 (2.0)	10 (–12.2)	11 (–11.7)
Final cooler	HE-204A/ HE-204B	3,224 (11.0)	3,107 (10.6)	10 (–12.2)	13 (–10.6)

temperature of the cooling water supply or fouling conditions. This explains the higher approach temperature for HE-204A/HE-204B.

Other than issues with fouling, the CO_2 process coolers performed as well as or better than the design requirements over the course of the 3-year injection period. As discussed elsewhere, cleaning methods were developed to remove fouling and restore heat exchanger performance when it became necessary. The ability of blower aftercooler HE-101 to perform as well as it did was particularly important to the performance of the overall IBDP compression and dehydration facility. However, this exchanger became fouled on several occasions, and cleaning was required to restore its performance. Proper performance of HE-101 provided a cooler, higher density feed to the reciprocating compressors, which allowed the compressors to work effectively; in addition, proper performance of HE-101 allowed a significant amount of free liquid water to be removed from the CO_2 gas before it left the blower system and moved to the reciprocating compressors. If HE-101 had not performed as well as it did, this aftercooler could have become a bottleneck for the facility.

Process Modifications and Improvements

Low suction pressure and low discharge pressure shutdowns of the reciprocating compressors were replaced with a rod-load-based shutdown limit that was calculated according to the differential pressure across each stage of compression. This modification reduced unnecessary equipment shutdowns, particularly during upset conditions and when intentionally taking some equipment offline. Similarly, the blower recycle valve was reprogrammed to allow the valve to open during high blower discharge pressure conditions. Opening this recycle valve in conjunction with the blower discharge vent valve during high blower discharge pressure conditions reduced the number of blower shutdowns caused by high discharge pressure on the blower as compared with the original design, which relied only on the blower vent valve to manage high blower discharge pressures.

The Archer Daniels Midland Company and Trimeric worked with the compressor vendor (Enerflex) to lower the CO_2 outlet temperature set point on several process coolers from the original design value of 95 °F (35 °C) to a revised design value of 80 °F (26.7 °C). This was primarily done to keep the return temperature of cooling water below an upper limit of 160 °F (71.1 °C), to reduce corrosion concerns in the heat exchangers and cooling water return lines. However, the lower outlet temperature also resulted in lower compression power requirements and reduced the water content in the CO_2 before dehydration.

Note that these kinds of process changes must be evaluated on a case-by-case basis to avoid operating in regions where issues with liquid CO_2 or solid CO_2–water hydrates could be encountered and to avoid any other undesirable consequences from the process change. For example, ADM found it best to keep the CO_2 outlet temperature set point leaving final cooler HE-204A/HE-204B at 98 °F (36.7 °C), even though a lower outlet temperature could be achieved when this exchanger was clean or when the cooling water supply temperature was lower than the summer design condition, or both. Because the transmission pipeline was insulated, keeping the temperature of CO_2 entering the transmission pipeline at 98 °F (36.7 °C) maintained the CO_2 delivered to the injection well at a constant temperature year round. This was done to minimize variations in both the temperature and pressure of CO_2 delivered to the injection well, by request of the larger project team. Because colder CO_2 has a higher density, a significant drop in temperature of the CO_2 delivered to the injection well could produce a lower surface injection pressure (all other things being equal); delivering colder, higher density CO_2 to the injection well would raise the hydrostatic pressure in the CO_2 column in the injection well.

The Archer Daniels Midland Company, Enerflex, and Trimeric worked with the compressor manufacturer (Ariel) to reduce the injection rates of cylinder lubrication by approximately 16%. This reduced the oil consumption and oil carryover into the CO_2 product stream. Additional information regarding the type of cylinder lubrication oil selected and its injection rates is discussed elsewhere in this report. The Archer Daniels Midland Company and Trimeric also worked together to reduce the reboiler temperature set point in the dehydration unit, which reduced fuel consumption as well as the potential for glycol degradation.

Power Consumption

Actual Versus Design Power Consumption

The largest operating cost in the IBDP compression and dehydration unit was the electricity used to drive the motors on the blower and on the two reciprocating compressors. Table 9 compares the actual power required and the original design power values provided by the equipment vendors for blower BL-101 to compress 1,212.5 ton/day (1,100 tonne/day) of CO_2 at the summer design condition. The same comparison was made for each of the reciprocating compressors to compress 606.3 ton/day (550 tonne/day) of CO_2. These values were taken from a representative period of operations when the IBDP compression and dehydration facility was operating near the design rate of 1,212.5 ton/day (1,100 tonne/day) of CO_2. Table 9 shows that the actual power required to operate the facility was in very good agreement with the design power requirements estimated by the equipment suppliers.

Total Power Required to Inject 1.1 Million Tons (1 Million Tonnes) of Carbon Dioxide

Trimeric used plant-operating data for the hourly average current in amperes for the blower and compressor motors and standard three-phase voltage calculations to estimate actual electrical power requirements in kilowatts. Using hourly average injection rates, Trimeric then estimated that the actual amount of electrical energy required for these three motors to compress and inject a nominal 1.1 million tons (1 million tonnes) of CO_2 at the IBDP was 101.6 gWh/ton (112 gWh/tonne) or 101.6 kWh/ton (112 kWh/tonne). The supplier design value was 99.8 kWh/ton (110 kWh/tonne) of CO_2 injected at the nominal injection rate under the summer design condition. Thus, quite good agreement was found

Table 9 Design versus actual power requirements for the blower and compressor motors

Machine	Design power hp	Design power kW	Actual operating power hp	Actual operating power kW	Difference, %
Blower BL-101	1,126	840	1,203	897	6.8
Compressor VC-201	2,800	2,088	2,726	2,033	–2.6
Compressor VC-301	2,800	2,088	2,804	2,091	0.1
Total power	6,726	5,016	6,733	5,021	0.1

between the calculated and supplier-predicted energy requirements (~2% difference).

This energy requirement did not include minor electrical costs such as dehydration unit pumps, lubrication oil pumps, or other small-horsepower motors. It also did not include any of the electricity required to operate the cooling tower pumps and fans that were part of the host facility cooling water system.

Dehydration System Performance

In this section, we review the performance of the TEG dehydration system by comparing selected process data recorded during the injection period with the dehydrator requirements in the original design. The primary performance indicators, such as the treated gas water content, process capacity, utility and glycol makeup requirements, and heat transfer performance, are reviewed. In addition to the analysis of process data, we briefly discuss the operability and availability of the unit, problems that were encountered, lessons learned, and implications for future projects.

Treated Gas Water Content

Perhaps the most important requirement for the dehydration system is to produce a dry CO_2 product that meets the specification for water content. Figure 17 shows that the treated CO_2 water content was generally within the specification of no more than 633 ppmv (30 lb/MMscf) of gas and an average value on the order of 169 ppmv (8 lb/MMscf) during normal operations.

The period of missing data in early 2012 is a result of the moisture analyzer being out of service after an upstream pressure regulator failed. While the analyzer was out of service, ADM monitored the reboiler operating temperature, glycol circulation rate, and other parameters to track the performance of the dehydration unit. The cause of the brief spike in August 2012 is unknown. The brief spike in October 2014 occurred during a start-up after a period of shutdown. In general, the treated CO_2 water content was approximately 148 to 169 ppmv (7 to 8 lb/MMscf), as expected.

Process Capacity

The design capacity for the dehydration unit was 1,212.5 ton/day (1,100 tonne/day, or 21 MMscfd). The amount of CO_2 treated averaged 1,116.6 ton/day (1,013 tonne/day). The average injection rate was lower than the design condition mainly because of the limit in an administrative permit that required ADM to operate at less than the full design condition until this limit was increased after approximately 1 year of injection. The CO_2 supply to the IBDP compression facility was limited at times because of operating conditions upstream of the IBDP compression, dehydration, and transmission equipment. All the injected CO_2 flowed through the dehydration unit; the injection rate data confirmed that the dehydration unit was able to operate at flows as high as 1,240 ton/day (1,124 tonne/day), which are slightly more than the actual design capacity of the unit.

The glycol circulation rate is a process variable closely related to the capacity of the dehydration system. Sufficient glycol must be circulated to absorb the required amount of water from the CO_2. The circulation rate also affects the operation of the heat exchangers and the fuel usage in the reboiler. The design circulation rate for lean glycol was 4.6 gal/min (gpm; 17.4 L/min). Initially, the glycol circulation pump provided more glycol circulation than necessary because a higher speed motor was installed than was called for in the design. This did not cause a problem other than slightly increasing the electricity usage for the pump and fuel usage in the reboiler that was needed to heat the additional glycol. The Archer Daniels Midland Company installed a variable-frequency drive on the glycol pump motor to reduce the speed of the pump until the correct glycol flow rate was achieved. Toward the end of the 3-year injection period, the maximum rate produced by the glycol circulation pump had declined to approximately 3.4 gpm (12.9 L/min). The wear on the gear pump internals is believed to have caused this decline, which normally occurs within 2 or 3 years with this type of pump. If the facility had been required to operate for a longer period, the pump heads would probably have needed to be replaced every few years or a more expensive type of pump selected.

Utilities and Glycol Makeup Requirements

Fuel for the reboiler was the main utility used by the dehydration system. The design heat duty for the reboiler was 101.4 kW (345,930 Btu/h), and the operating duty for the reboiler was 55.8 kW (190,271 Btu/h). The corresponding operating fuel rate would be 6.8 Mcfd (thousand cubic feet per day; 8 m^3/h) based on a 70% thermal efficiency in the reboiler. The actual fuel usage was not recorded, so a direct comparison was not possible between the actual fuel usage and what was expected. Electricity requirements to run the glycol circulation pump were minimal, less than approximately 4 kW.

Glycol makeup is needed to replace any glycol that is lost from the process, primarily through carryover into the treated

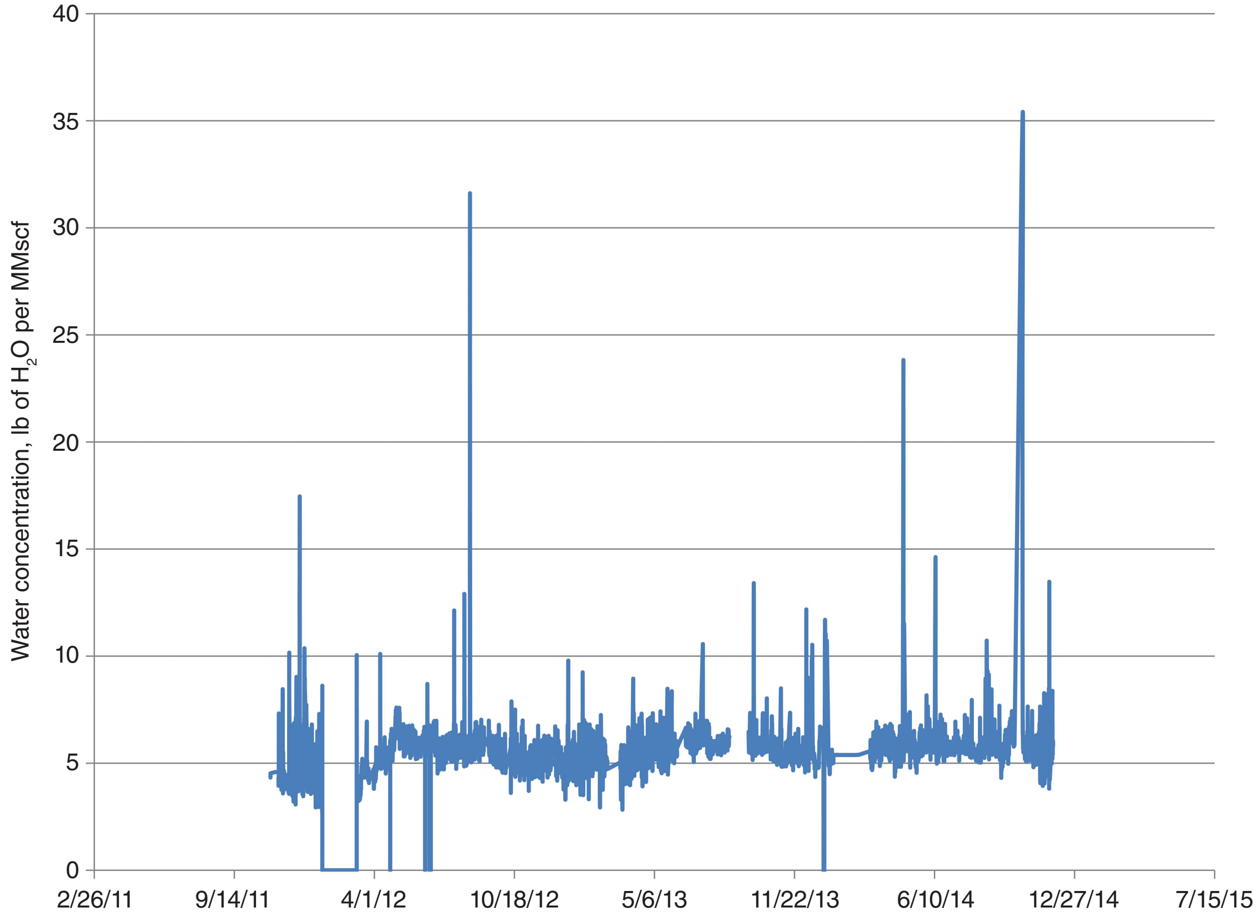

Figure 17 Dehydrator water concentration. MMscf, million standard cubic feet per day.

gas or exiting the system with the water vapor in the distillation (still) column vent. The expected losses for this type of system are generally in the range of 0.15 to 0.4 gal/MMscf (20 to 54 L/Msm^3) of gas treated, according to the Gas Processors Suppliers Association Engineering Data Book (GPSA 2004). In comparison, plant records indicate that approximately 0.16 gal/MMscf (21 L/Msm^3) of glycol was added to the system as makeup. This amount is on the low end of the range listed by the GPSA (2004), suggesting the glycol makeup rate for the IBDP dehydration system was acceptable.

Heat Transfer Performance

The dehydration unit included cold glycol and hot glycol heat exchangers (tag numbers HE-402 and HE-403, respectively) to recover heat from the hot lean glycol exiting the reboiler and to use this heat to preheat the colder, rich glycol from the absorber before it went to the reboiler. This heat recovery reduces the overall amount of fuel required for the reboiler and improves the energy efficiency of the process. If these heat exchangers become fouled, poor performance can result in increased fuel usage. If the problem is severe enough, the reboiler will be unable to provide enough heat to remove the water from the glycol and the dehydrator may no longer be able to dehydrate the CO_2 completely.

The performance of the heat exchangers can be assessed by examining the temperature of the glycol flowing into and out of the exchangers. Figure 18 shows some of the process temperature data for the lean glycol.

The designed maximum temperature for the lean glycol leaving the reboiler was 400 °F (204.4 °C). However, a temperature

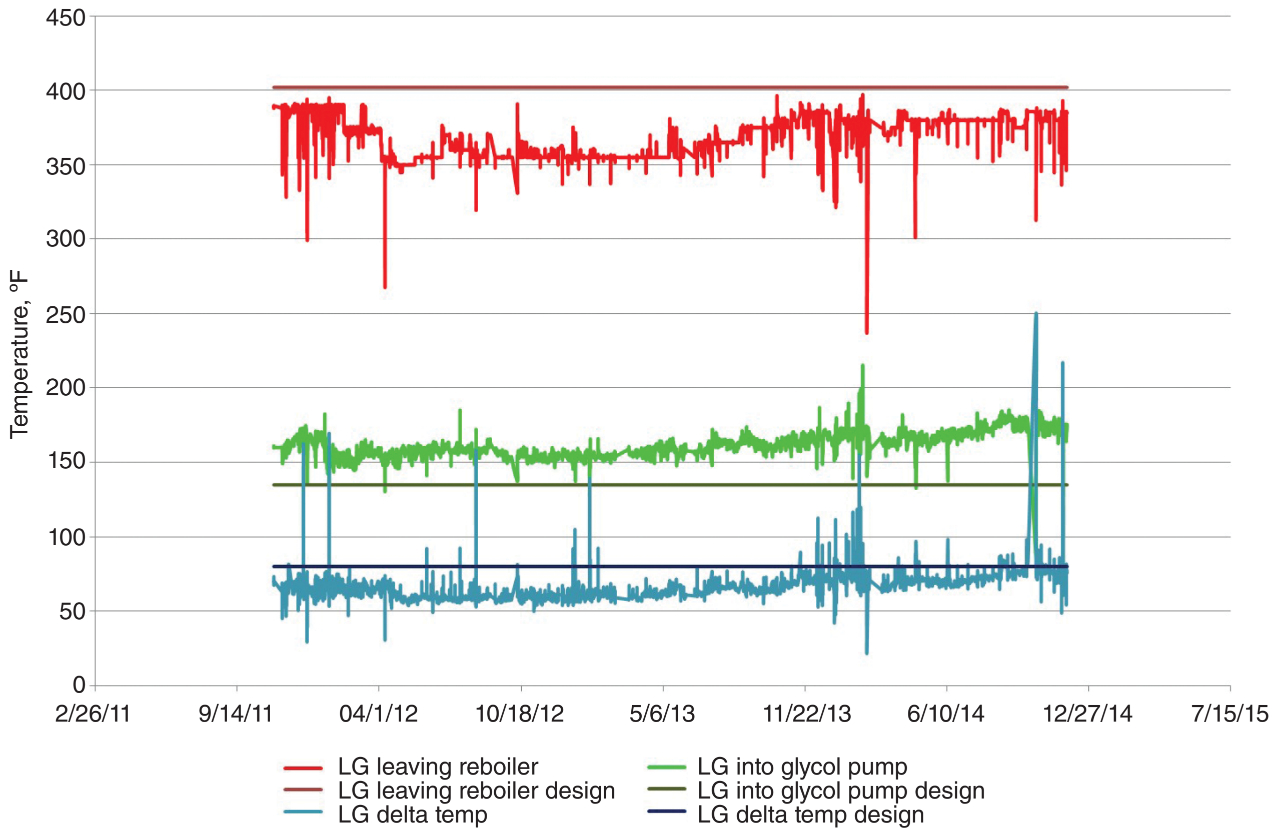

Figure 18 Actual and design temperatures of lean glycol (LG) in the dehydration unit.

of 400 °F (204.4 °C) is not required during normal operations, and the higher temperature results in higher fuel usage and potentially higher glycol consumption because of thermal degradation. The temperature in the reboiler was controlled at a set point determined by the operator, which ranged from approximately 360 °F (182.2 °C) to approximately 390 °F (198.9 °C) during injection. Figure 18 shows how this temperature varied during injection. Note that the reboiler was generally able to maintain temperatures near the process control set point, as indicated by the temperature of the lean glycol leaving the reboiler.

One indicator of heat exchanger performance is the difference between the temperature of the rich glycol entering the reboiler and the temperature of the reboiler, which is represented in Figure 18 by the trend for "LG delta temp." (This is also the difference between the lean glycol leaving the reboiler and the rich glycol leaving the hot glycol exchanger.) The design called for a difference in temperature of approximately 112 °F (44.4 °C) at maximum reboiler temperatures of 400 °F (204.4 °C). Under this condition, the rich glycol would be preheated to 320 °F (160 °C) by the 400 °F (204.4 °C) lean glycol exiting the reboiler. If the heat exchangers began to become fouled, the glycol would not be preheated as much and this temperature difference would increase. Figure 18 shows that the temperature difference was initially approximately 65 °F (18.3 °C), well within the design maximum of 80 °F (26.7 °C). However, this temperature difference began to climb slightly in the third year of operation, which could be an indicator of some fouling in the exchanger.

Another indicator of heat exchanger performance is the temperature of the lean glycol after it leaves the cold glycol heat exchanger (just before the glycol pumps). The design temperature at this point in the process was approximately 152 °F (66.7 °C). If the cold glycol or hot glycol heat exchanger is not performing properly, this temperature will increase as less heat is recovered from the hot lean glycol. Figure 18 shows that the temperature was near the design value for much of the injection period but that it began to climb noticeably during the last half of injection and was near 175 °F (79.4 °C) near the end of injection, which could possibly indicate fouling in the heat exchanger.

Figure 19 shows some of the process temperature data for the rich glycol. The design maximum temperature for the rich glycol entering the reboiler was 330 °F (165.6 °C); however, a temperature as high as 330 °F (165.6 °C) was not required during normal operations. The "RG into reboiler" trend in Figure 19 shows how this temperature varied during injection.

Another indicator of heat exchanger performance is the difference between the temperature of the rich glycol entering the cold glycol exchanger from the still column and the temperature of the lean glycol exiting the cold glycol exchanger.

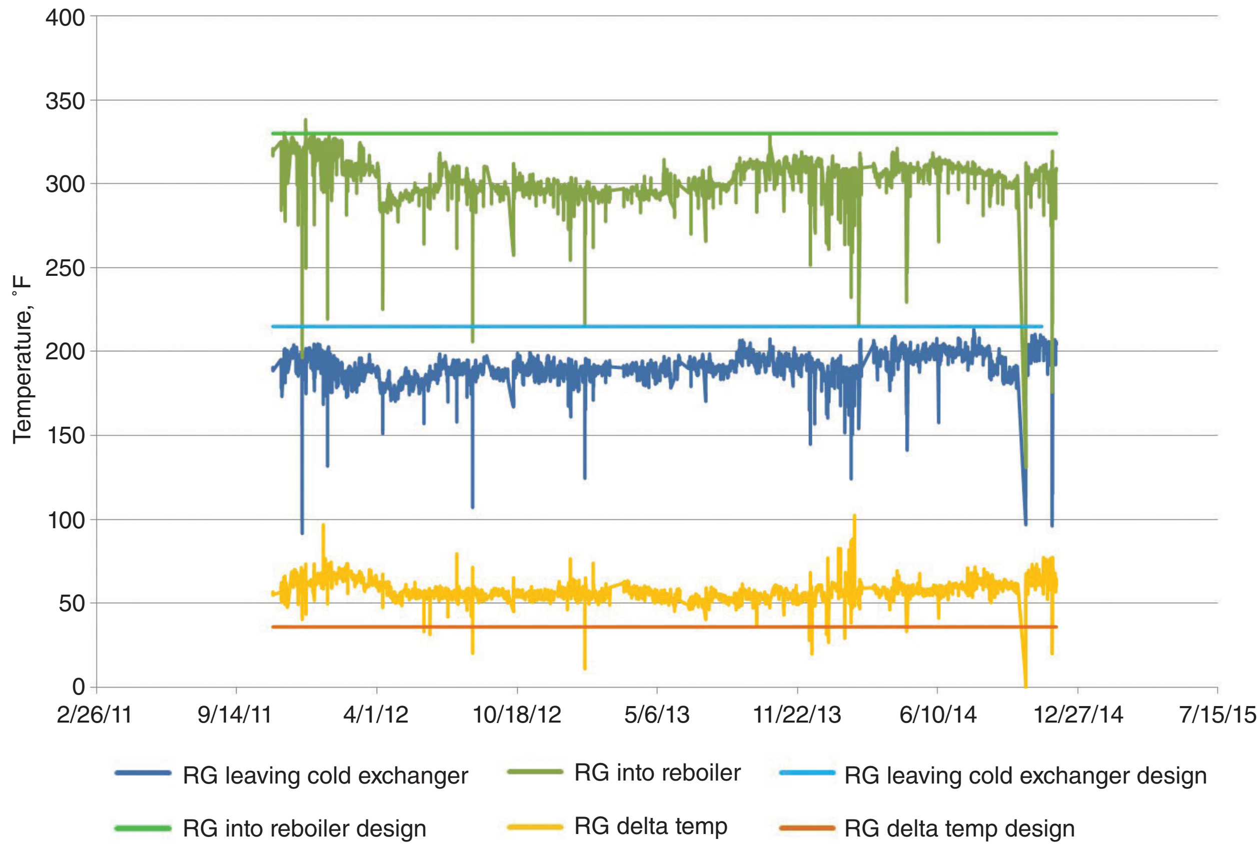

Figure 19 Actual and design temperatures of rich glycol (RG) in the dehydration unit.

The design called for a difference in temperature of approximately 36 °F (2.2 °C). Under this condition, the lean glycol would be cooled to 135 °F (57.2 °C) by the 99 °F (37.2 °C) rich glycol exiting the still column in the cold glycol exchanger. If the heat exchangers began to become fouled, the glycol would be warmer and the difference in temperature would increase. The trend for "RG delta temp" in Figure 19 shows that the difference in temperature was approximately 58 °F (14.4 °C). This higher than design temperature, especially toward the end of injection, could indicate that some fouling was beginning to occur.

Another indicator of heat exchanger performance is the temperature of the rich glycol exiting the cold glycol heat exchanger and entering flash tank TK-402. The design temperature at this point in the process was approximately 215 °F (101.7 °C), although this temperature is not required during normal operations. A temperature as high as 215 °F (101.7 °C) would not be expected unless the reboiler was set to run at its maximum design temperature of 400 °F (204.4 °C). The trend for "RG leaving cold exchanger" in Figure 19 shows that the temperature was near the design value for much of the injection.

Operability and Operating Problems Encountered

The dehydration unit operated continuously during the 3-year injection period, except for 5 days when the unit was taken offline because of issues with the reboiler burner. A problem with the burner flame quality in the dehydration reboiler led to an outage on February 18, 2014. The Archer Daniels Midland Company cleaned the burner fire tube, removed some buildup and residue, and restarted the unit. These issues did not recur after this maintenance.

Glycol Quality Monitoring

The glycol quality was monitored regularly during operations by visually inspecting the glycol for color and clarity. Activated carbon filters in the system were changed when the glycol became discolored. Glycol samples were analyzed periodically for basic quality indicators such as total glycol assay, water concentration, the presence of hydrocarbons, iron levels, pH, and chloride levels. A typical analytical result for the IBDP rich glycol is shown below:

- TEG (assay wt %): 94.87
- Water (wt %): 5.3
- Chloride (ppm): 50
- Iron (ppm): 1.689
- pH: 4.64

The TEG assay was primarily used to detect the ingress of large amounts of foreign material into the glycol circulating system, which could have been caused by carryover from an upstream unit or possibly an operator charging the system with the wrong material. The TEG assay should generally be above 90%. The water analysis was run to confirm that the system was properly regenerating the glycol by removing the water in the reboiler and that the absorber was effectively removing water from the incoming gas. The level of 5.3 wt % measured for this particular sample represents a normal amount of water in the rich glycol. The rich and lean glycol would both have had a much higher water content if the reboiler had not been properly removing the water from the glycol. Chlorides usually indicate some contamination from an upstream operation and are especially important if salt is known to be present upstream. With some alloys of stainless steel, high levels of chlorides can also create a concern for corrosion. The level of 50 ppm is not considered unusually high for this application. Iron levels above 10 ppm generally indicate corrosion occurring in the system, which is sometimes caused by introducing strong acids into the system; this is the reason for checking the pH. Acidic compounds can also be created under certain conditions when TEG degrades thermally, oxidizes, or both. Thermal degradation is usually caused by problems with the fire tube in the reboiler or excessive reboiler temperatures. Oxidative degradation can occur when oxygen levels in the gas become excessive, when the TEG is stored in the presence of oxygen or air for long periods, or when oxygen enters the reboiler under certain upset conditions. A pH of approximately 4.6 is typical of systems that are treating CO_2 because of the normal and expected presence of carbonic acid that forms when CO_2 is absorbed into the glycol–water mixture. Stronger acids, if present, would likely cause the pH to drop below 4 and would suggest increased potential for corrosion issues.

Implications for Future Projects

Although the dehydration unit ran for only 3 years, some signs of normal wear were appearing at the end of injection (e.g., a decline in heat transfer in the heat exchangers and pump rotor wear that limited the glycol circulation rate). Projects of longer duration would typically require some maintenance to address these problems, as is often the case with similar dehydration units running for longer periods in commercial applications.

Corrosion Monitoring

One of the requirements of the UIC injection permit for this project was to monitor the injected CO_2 for any corrosive properties. The Archer Daniels Midland Company designed a corrosion monitoring system that included a series of metal coupons of three different alloys of interest (L80, 316L, and 5LX52) that were exposed to the product CO_2 in a continuous slipstream just before the dehydrated CO_2 entered the transmission pipeline.

Corrosion coupon studies involve weighing specially prepared coupon samples of the alloy and then exposing them to the environment to be monitored. After the required exposure time (which ranged from 58 to 135 days in this project), the coupons are removed, cleaned, examined, photographed, and weighed again. Any signs of corrosion, such as pitting or cracking, are recorded. The average rate of corrosion is determined by noting the weight loss of the coupon and calculating the corresponding thickness of metal that would have to be lost to account for the weight loss. Figure 20 shows a typical photograph of one of the coupons for the 316L alloy.

The resulting corrosion rate is expressed in terms of mils thickness per year (MPY), where one mil is equal to 0.001 in. (0.0254 mm). Uniform average corrosion rates of less than 1 MPY are generally considered acceptable for many applications. As shown in Figure 21, the corrosion rates observed for all three alloys (shown left to right as L80, 316L, and 5LX52) were well below 0.4 MPY, which indicates a noncorrosive environment, as expected for dry CO_2. Reasons for the higher corrosion rates observed with the initial set of coupons were not identified.

Challenges and Problems with Process Operations

Several minor process improvements were discussed earlier in this section. The three most significant process operating challenges and problems were the need to maintain acceptable return temperatures for cooling water leaving the CO_2 process coolers, the need to purge the system to remove residual moisture after equipment shutdowns, and the need to carefully select the type of cylinder lubrication oil. Initial operating experience confirmed the design team's prediction that insulation would be needed on aboveground piping to maintain a stable injection temperature and pressure regardless of ambient conditions and to facilitate compliance with permit conditions. Even though it was a significant project expense, the aboveground pipeline was insulated in the first few months of operation.

Figure 20 Example corrosion coupon.

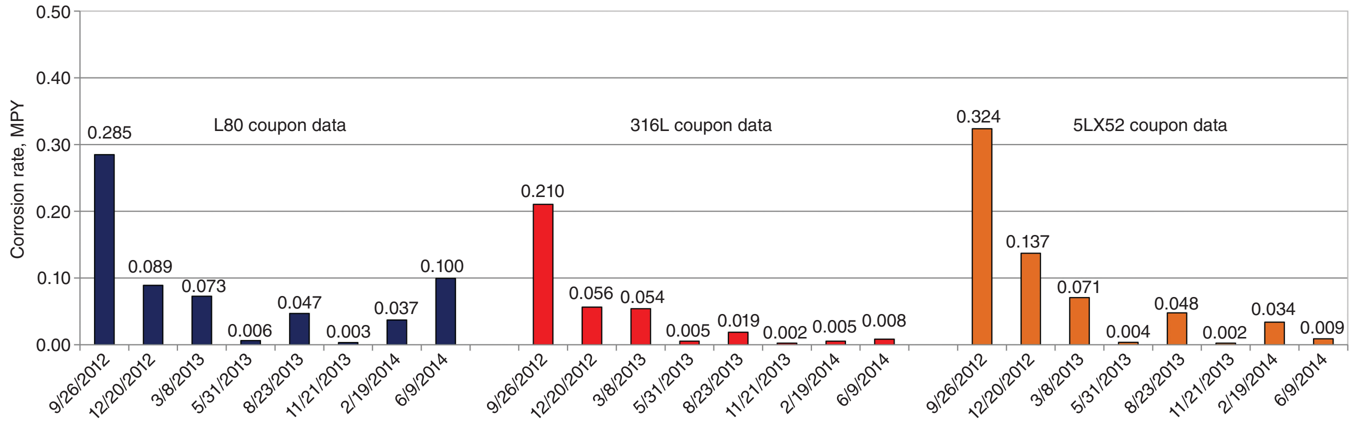

Figure 21 Corrosion coupon rates. MPY, mils thickness per year.

In the design phase, Trimeric recommended that ADM install a tempered water control loop that would feed some hot cooling water back to the cold cooling water supply to maintain a constant cooling water supply temperature to the IBDP CO_2 process coolers regardless of variations in ambient conditions over the seasons. The Archer Daniels Midland Company decided that implementing the tempered water control loop was not practical for the IBDP facility. The tempered water control loop had worked well on other similar commercial projects, and if one had been installed for the IBDP facility, higher year-round cooling water flow rates through the heat exchangers would likely have reduced the rate of fouling in the exchangers, the number of required heat exchanger back-flushes, and the requirements for heat exchanger cleaning. The tempered water control loop would also have reduced the amount of operator attention and intervention required to keep cooling water return temperatures from becoming too high, where corrosion of the heat exchanger tubes and cooling water return lines becomes more of a concern. As discussed previously, ADM selected the approach of lowering the set point temperature for CO_2 leaving the process coolers from 95 to 80 °F (35 to 26.7 °C) to increase the cooling water flow rate through these heat exchangers. Although ADM engineers and operators had to monitor and adjust cooling water flow rates to each of the heat exchangers frequently to keep the return temperatures of the cooling water within acceptable limits, lowering the CO_2 process temperature also reduced compression energy requirements and the amount of water that needed to be removed by the dehydration system. The tradeoffs of installing a tempered water control loop versus more frequent heat exchanger cleaning and a higher level of operator attention and intervention to maintain return temperatures of the cooling water would need to be considered in the design of a similar facility.

As part of risk management and in support of the UIC permit application process, Trimeric conducted process modeling to assess the injection operations envelope before the beginning of injection. This modeling predicted that some changes in ambient conditions, including rapid temperature drops, high winds, and heavy precipitation, would have a significant impact on the temperature of CO_2 entering the injection well. Colder CO_2 has a higher density, which will result in a lower surface injection pressure (all other things being equal) because of the higher hydrostatic head of the higher density CO_2 in the well. The lower surface injection pressure, in turn, changes the pressure profile over the entire vertical section of the well until the depth of the injection perforations is reached. Archer Daniels Midland preferred to commence injection operations without insulating the aboveground transmission pipeline to determine whether it was necessary to insulate the pipeline because this was a fairly expensive and labor-intensive effort. Early operations showed that insulating the pipeline was necessary to simplify the operations needed to comply with the minimum allowable injection temperature in the UIC permit and to avoid pressure variations in the injection well. Such variations were an initial concern because they could stress the bonding between the casing and the cement in the injection well over time. Insulation of the aboveground transmission pipeline was completed approximately 5 months into the first year of injection, which simplified the operations needed to meet the minimum requirement in the UIC permit for the injection temperature and stabilized the injection temperature and pressure at the surface and their profiles in the injection well.

During the design phase, the compressor packager and manufacturer recommended use of ISO 220 mineral oil-based lubricant for cylinder lubrication. During injection well maintenance about halfway through the second year of injection, a dry-to-the-touch, asphalt-like, dark substance was observed on well maintenance tools coming out of the injection well and on the inside of the injection well near the surface. Samples of the residue were analyzed and found to be 35% asphaltenes. A working group was formed that included personnel from Schlumberger, ISGS, ADM, Enerflex, Ariel, and Trimeric. The group determined that asphaltene deposits had been encountered downstream of other compressors when using mineral oil-based cylinder lubricants in some CO_2 compression applications. Some sites that had encountered this issue did periodic cleaning and others switched to synthetic oil, such as PAG ISO 100, for cylinder lubrication.

The group also discussed issues with mineral oil and synthetic oil forming gel-type emulsions if they were inadvertently mixed. Should mixing occur, the frame lubrication would also need to be changed to synthetic oil; therefore, careful measures were needed to keep the synthetic and mineral oils separate. The working group decided to continue injection with the lower cost mineral oil because no adverse impacts, such as an appreciable increase in the injection pressure, were observed. It was determined that the compressors were equipped to run with either mineral or synthetic oil, so retrofit costs would be less of a concern if it became necessary to change to synthetic oil at a later time. Ultimately, the entire 3 years of injection was completed with the mineral oil lubricant. Similar future projects should consider the tradeoffs between a lower cost mineral oil and a higher cost synthetic oil before selecting a cylinder lubricant.

The compressor manufacturer also determined that an approximately 28% reduction in mineral oil injection rates was possible because the original rates were set based on the maximum 1,000 rpm motor speed, whereas actual operations were at 715 rpm. The Archer Daniels Midland Company gradually reduced the mineral oil injection rates but opted to maintain an approximately 16% reduction when no further issues were encountered because of oil carryover. In addition, the cost savings would have been minimal for further reductions in the cylinder lubrication injection rates.

Net Amount of Carbon Dioxide Injected

The most recent emission factor from the U.S. DOE for CO_2 emissions for a pulverized coal-fired supercritical boiler power plant was 1,705 lb/MWh (net) or 773.4 kg/MWh (net) (U.S. DOE 2015). For the IBDP-calculated 101.6 kWh/ton (112 kWh/tonne) of CO_2 injected power requirement, this corresponds to CO_2 emissions of 191 lb (86.6 kg) from generating the power required to inject 1.1 tons (1 tonne) of CO_2. Because 1.1 tons (1 tonne) is 2,204.6 lb (1,000 kg), this calculation predicted that the tons of CO_2 avoided in emissions from CO_2 vent associated with ethanol production would be 1,826 lb/ton (913 kg/tonne) injected, which equals 91% of the CO_2 injected. In other words, for the 1.1 million tons (1 million tonnes) of CO_2 injected over the life of the project, 1,006,410.2 tons (913,000 tonnes) of net CO_2 was captured after estimating 95,901.1 tons (87,000 tonnes) of CO_2 emissions associated with generating the electrical energy required to operate the compression, dehydration, and transmission equipment.

MAINTENANCE—ACTUAL VERSUS DESIGN

Maintenance was an important consideration during the project. Reciprocating compressors are generally a higher maintenance item than centrifugal compression equipment, which includes multistage centrifugal blowers, pumps, and compressors. Reciprocating compressor maintenance did prove to be the dominant item requiring maintenance in the compression and dehydration facility.

Scheduled and Unscheduled Maintenance

The Archer Daniels Midland Company followed the scheduled maintenance programs outlined by the equipment suppliers, the details of which are beyond the scope of this report. One important item in the scheduled maintenance was the replacement of reciprocating compressor cylinder suction and discharge valves. These valves have a very tight tolerance, and they open and close hundreds of times per minute, so replacing them periodically is necessary and important for maintaining compressor efficiency. While performing the scheduled maintenance, a number of additional necessary maintenance items were identified, as discussed later in this section.

The blower motor failed near the end of the start-up and commissioning effort. The motor failure was attributed to windings arcing and phase-to-phase ground caused by poor insulation. The motor supplier replaced the motor as a warranty item. The only other significant maintenance issue with the blower was an undetected oil leak that led to motor bearing failures and 6 days of unscheduled downtime in the first year of injection. The Archer Daniels Midland Company moved the oil sight glass to the other side of the blower, where it was easier for the operators to see during their regular shift inspections.

In the first year of operation, the fourth-stage piston rods on both compressors were determined to be slightly bent. The compressor manufacturer provided replacements and changed the material of the rods from stainless steel to medium-carbon, low-alloy, cold-rolled steel that was iron nitrided to increase the hardness of the rod surface. In the second half of the 3-year injection period, two reciprocating compressor cylinders and two compressor rods had to be replaced. These items would not normally require replacement this early in the operating life of a reciprocating compressor.

The cylinders had pitting and corrosion, which was thought to have been caused by the accumulation of free liquid water in the system, most likely during shutdown conditions because the cylinders operate at temperatures well above the water dew point when the compressors are in operation. The compressor packager (Enerflex) suspected the bent rods in this instance might have been caused by liquid entering the cylinders during operations. The suction scrubbers were designed with a 10% safety factor and should have prevented any liquid from reaching the compressor cylinders during normal operations. It is possible that some liquid water accumulated in the suction pulsation bottles upstream of the cylinders when the compressors were shut down and that this water subsequently entered the cylinders when the compressors were restarted.

The reciprocating compressor systems were designed with an automated nitrogen purge system that was programmed to execute automatically after each compressor shutdown. However, the amount of nitrogen the purge process used caused operational problems in other parts of the overall facility. The nitrogen purge was run when the compressor was going to be down for a day or more, but not every time the compressors shut down. Trimeric recommended installing restrictive-flow orifice plates in the nitrogen supply lines so that the automated nitrogen purge could be used every time the compressors were shut down. This approach had worked on similar systems at other commercial facilities. The orifice plates were ordered but were not installed by the time the injection was complete. It is possible, but not certain, that the corrosion and free liquid issues and resulting cylinder and rod replacements could have been avoided if the automated nitrogen purge had been in service over the entire injection period.

As described in the Challenges and Problems with Process Operations section, a number of issues arose with the cylinder lubrication oil injection pumps and cylinder lubrication system rupture disks. Both types of failures caused automatic compressor shutdowns. In addition to that inconvenience, the rupture disk failures resulted in oil spilling onto the compressor that was difficult to clean up. The compressor manufacturer (Ariel) provided replacements based on a new design for the pumps and changed to a pressure relief valve instead of rupture disks. When the pressure relief valve opened, the oil would circulate from pump discharge to pump suction instead of spilling onto the compressor.

The dehydration system required little maintenance. The cartridge filters to remove particulates and the charcoal filters to remove hydrocarbons had to be replaced several times, which was part of scheduled maintenance for this system. The glycol pumps also needed to be replaced near the end of the third year of injection. These gear pumps are a normal wear item and replacing them would likely be necessary to restore system performance.

Carbon dioxide process coolers were shell and tube heat exchangers with wet CO_2 contained within stainless steel tubes and cooling water in the carbon steel shell. This design resulted in a lower cost because if the wet CO_2 had been on the shell side, stainless steel tubes and shells would have been required. However, it did make maintenance somewhat more difficult because it put the dirtier fluid (the cooling water) on the shell side. The tube side of a shell and tube heat exchanger is somewhat easier to clean because the heat exchanger head can be removed to clean the inside of the tubes. The shell side is less accessible for cleaning. When the cooling water was colder than the design, the lower cooling water velocity through the shell side of the heat exchangers likely also led to a higher fouling rate in the heat exchangers. The Archer Daniels Midland Company typically back-flushed the shell side of the heat exchangers every week or two. The back-flush allowed cooling water to flow in a reverse direction through the shell side of the heat exchanger, which helped to remove some debris. However, on two occasions, chemical cleaning of each of the CO_2 process coolers was required to restore their performance. Archer Daniels Midland developed a successful cleaning process that was easy and safe for their own personnel to perform. This reduced costs and downtime as compared with hiring a third party to clean the heat exchangers. Figure 22 shows how various heat exchanger cleaning methods were able to restore the heat exchanger performance.

Actual Days per Year of Injection Versus Design

The design basis called for injection to occur 335 days/year. The 30 days/year of downtime included scheduled and unscheduled maintenance, days when

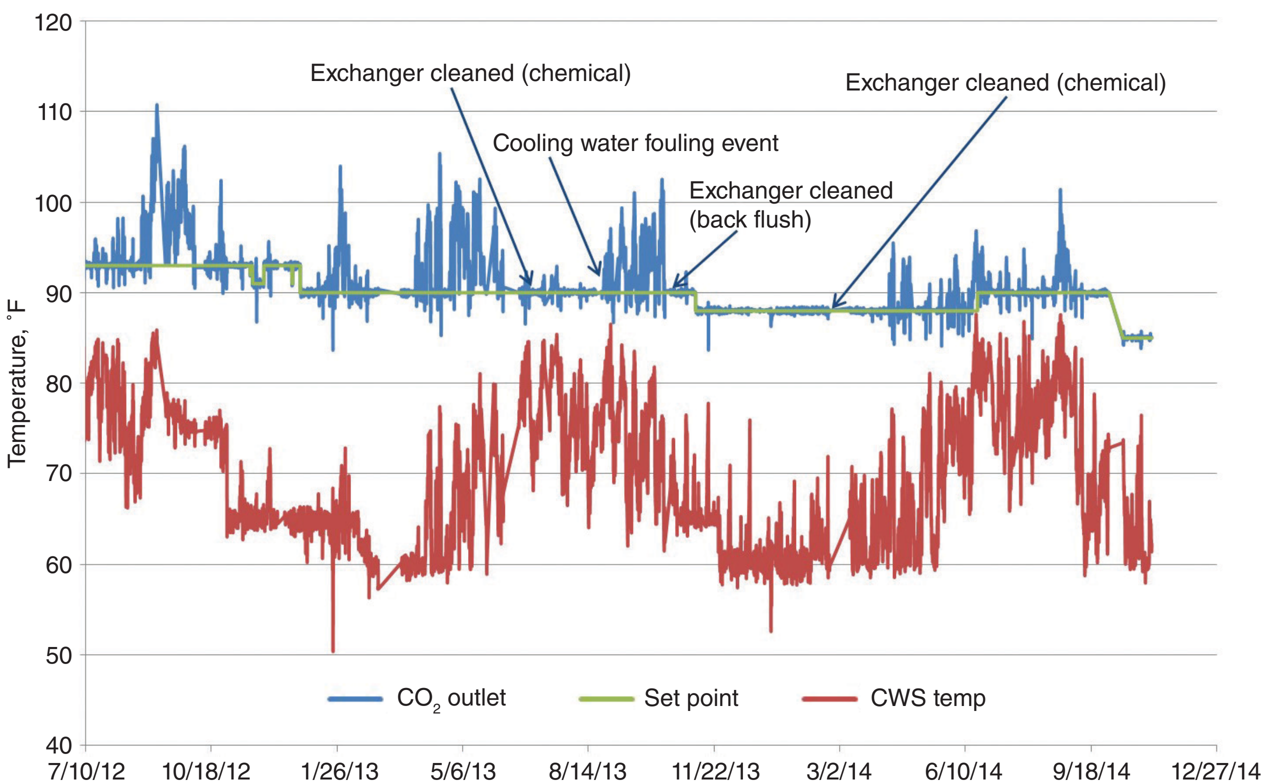

Figure 22 Blower aftercooler HE-101 performance tracking. CWS, cooling water supply.

injection was stopped for testing purposes, and days of ADM host site plant outages when CO_2, cooling water, electricity, or other items required for IBDP injection operations were not available. The first year of injection operations had 20 days without injection, the second had 41 days without injection, and the third had 28 days without injection. Over the 3-year injection period, the average was 30 days/year.

In the second year of injection, a relatively sudden reduction in the supply of CO_2 to the compression and dehydration facility caused by operational issues upstream of the IBDP facility resulted in lower than design suction pressure upstream of blower BL-101. This caused a large section of the CO_2 supply duct upstream of the IBDP facility to collapse. Injection was stopped for 7 days while ADM replaced the duct with sturdier piping material. The blower had a low-pressure shutdown, but the limit was originally set with the intention of protecting the blower and related equipment. The Archer Daniels Midland Company changed the set point for the low-pressure shutdown on the blower to ensure that equipment upstream of the compression and dehydration facility would also be protected. In the second year of injection, ADM experienced 17 days of plant outages, compared with 3 days in the first year and 10 days in the third year.

Table 10 Total capital costs for the Illinois Basin – Decatur Project surface facilities

Cost category	Costs, 2009–2011, US$	Actual, % of TDIC	Typical range, % of TDIC
Purchased equipment	6,145,000	30	15–40
Purchased equipment installation	1,883,000	9	6–14
Instrumentation and controls	958,000	5	2–12
Piping	5,108,000	25	4–17
Electrical systems	2,992,000	15	2–10
Buildings	720,000	4	2–18
Yard improvements	231,000	1	2–5
Total direct cost (TDC)	18,038,000	—	—
Engineering	1,873,000	9	4–20
Construction expenses	431,000	2	4–17
Total indirect cost (TIC)	2,304,000	—	—
Total direct and indirect cost (TDIC)	20,342,000	—	—

COST ANALYSIS

This section provides an analysis of the actual capital costs, the estimated operating costs, and the estimated total costs (capital and operating costs) per ton (tonne) of CO_2 injected at the IBDP. The costs presented here apply to the compression, dehydration, and transmission (pipeline) equipment, as well as instrumentation and controls for this equipment. Other project items, such as the injection well, monitoring wells, other monitoring and verification equipment, and instrumentation and controls related to that equipment, are not included in this cost analysis.

Capital Costs

Table 10 provides a detailed cost breakdown of the actual capital costs for the compression, dehydration, and transmission facilities for the IBDP. Costs are rounded to the nearest thousand-dollar value. The values for IBDP total direct and indirect capital costs shown in Table 10 are based on a careful analysis of the best available data. The total capital costs should be accurate because the project has been completed, and any uncertainty in the total amount would be due only to the complexity of tracking costs on this kind of project. Major equipment purchases occurred in 2009, but installation costs extended into 2010 and 2011 because of the lead time associated with the equipment and the overall project timeline, which was extended because of the time required to obtain the UIC permit for CO_2 injection.

The capital costs shown in Table 10 are sorted into commonly used categories for reporting facility construction costs in the chemical processing industry (Peters et al. 2003). Sorting into these categories was based on Trimeric's review of cost data provided by ADM, as well as Trimeric records of project engineering costs. The value of capital costs in any one of the categories listed in Table 10 may be less certain than the total capital costs because of the judgment required to interpret the descriptions recorded for each of the costs and to assign each cost to a particular category. Table 10 also shows typical ranges for these cost categories, which, as one might expect, could vary considerably from one project to another. The next few paragraphs provide comments on some differentiating aspects of the capital costs for the IBDP compression, dehydration, and transmission facility with respect to some of these cost categories.

In simple terms, because the system was used for 1,102,311 tons (1,000,000 tonnes) of injection, capital costs could be estimated at \$18.45/ton (\$20.34/tonne) of CO_2 injected. This capital cost estimate, when expressed on a per ton (per tonne) of CO_2 injected basis, may be higher than in a typical industrial project because the capital expense period and operational timeline would typically be much longer for a commercial project than the 3 years of injection operations for the IBDP.

Purchased Equipment Costs

As shown in Table 10, the purchased equipment costs were 30% of the total direct and indirect capital costs, which fell within the typical range of 15% to 40%. The purchased equipment costs category includes the major compression and dehydration equipment, including the blower, the two reciprocating compressors, the dehydration system, and the multistage centrifugal CO_2 pump. It also includes the motor and PLC systems for each of these items. The blower, reciprocating compressors, dehydration unit glycol regeneration system, and multistage centrifugal CO_2 pump were shipped to ADM on process skid assemblies that included the blower, compressors, pumps, motors, glycol reboiler, lubrication systems, control panels, separators and similar vessels, and some of the interconnecting piping required for these systems. The purchased equipment costs category also includes the costs for the system fabricators to install all these items on the process equipment skids. Installation time and costs were reduced by ordering skid-mounted equipment because much of the installation could be done in the fabrication shop rather than in the field. Additional items in the purchased equipment costs category include the off-skid portion of the lubrication system for the reciprocating compressors and the glycol storage tank and makeup pump for the dehydration unit. The costs of materials for the transmission pipeline were not included in the purchased equipment costs category. These were included in the piping costs category.

Purchased Equipment Installation Costs

As shown in Table 10, the purchased equipment installation costs were 9% of the total capital costs, which fell within the typical range of 6% to 14%. Literature values for the installation of purchased equipment are typically in the range of 25% to 55% of the purchased equipment cost (Peters et al. 2003). As shown in Table 10, purchased equipment installation costs were 31% of the purchased equipment costs.

Several factors affected the purchased equipment installation costs on this project. Thick foundations were required to support the reciprocating compressors, based on results of the soil survey at the site and the vibration analysis for the compressors. In addition to the foundations, significant structural work was required on this project to properly support and manage the stresses inherent in a system that used high-horsepower reciprocating compressors.

Charges for on-site equipment commissioning support from the skid fabrication companies were included in the purchased equipment installation costs category, whereas these costs might be accounted for as start-up expenses on other similar commercial projects. Finally, and more generally, installation of this relatively complex equipment within an existing, operating major industrial facility carried with it costs that would not be encountered for equipment installation at a greenfield site. However, as explained later in this section and elsewhere in this report, some benefits were also associated with installing the IBDP compression, dehydration, and transmission equipment within an existing industrial facility.

Instrumentation and Control Costs

Although the instrumentation and control costs shown in Table 10 were within the typical range, the level of instrumentation installed for the compression, dehydration, and transmission systems on the IBDP was higher than on a typical industrial project because of the unique research objectives of the IBDP. In addition to installing instruments in more locations than might be done on a typical industrial project, a greater percentage of the instruments were indicating transmitters rather than indicating instruments only. Adding the transmitter function to these instruments allowed ADM to incorporate more process readings into their automated DCS, which supported a higher level of data analysis than would be typical on a commercial (nonresearch) project. Costs included in the instrumentation and control costs category included control valves, instruments, analyzers, and labor costs specific to the installation, calibration, and testing of these items.

Piping Costs

As shown in Table 10, piping costs, including equipment, materials, and installation labor, were 25% of the total direct and indirect costs, which is higher than the typical range of 4% to 17%. The piping category included materials for the process, utility, and transmission pipelines. Types of piping included in the piping costs category included CO_2 process (including the transmission pipeline), glycol process, cooling water, instrument air, lubrication oil, fuel gas, vent headers, and drains. Basic manual valves, strainers, flanges, and similar items that were directly associated with these piping systems were also included in the piping costs category. Other materials and services required to install these piping systems were also included, such as pipe racks, supports and piers, sand (for sandblasting), X-ray inspection, paint, insulation, heat tracing, and expansion joints.

Piping costs for the IBDP compression, dehydration, and transmission facility made up a higher percentage of the total capital costs than might be encountered on a more typical project for specific reasons. Installation costs are higher for stainless steel piping, and a significant amount of stainless steel piping had to be installed in this project because of the corrosive nature of the water-saturated CO_2. Much of this piping had a relatively large diameter (16 to 24 in. [406.4 to 609.6 mm]) to minimize the pressure drop on the low-pressure part of the system (up to the inlet of the reciprocating compressors). Welding and related piping installation costs were comparatively high on this larger bore stainless steel piping. Furthermore, costs to install and insulate the 6,400-ft (1,950.7-m)-long, 6-in. (152.4-mm)-diameter carbon steel transmission pipeline were included in the piping costs category, whereas a more typical project might not require pipeline of this length to be installed and insulated within a network of pipe racks in an existing industrial facility.

Electrical System Costs

As shown in Table 10, electrical system costs, including equipment and materials, were 15% of the total capital costs, which is higher than the typical range of 2% to 10%. One significant reason for this difference is that a new 4,160 V, 7,500 kVA transformer had to be purchased and installed to support the IBDP compression and dehydration equipment. Costs for installing a new motor control center were also included in the electrical system costs category. In addition, this category included soft starters for the 4,160 V blower and compressor motors and variable-frequency drives for motors on the glycol pumps and on the multistage centrifugal CO_2 pump.

Engineering Costs

Although engineering costs were within the typical range, as shown in Table 10, some factors worth noting led to higher engineering costs when they were expressed as a percentage of the total capital costs. This was a first-of-its-kind research project with unique project constraints, including equipment performance requirements unique to this research project and a 3-year period for expensing capital costs. Typical capital expense periods on a similar industrial facility would be in the range of 10 to 30 years. The shorter period for expensing capital costs justified a higher level of engineering effort to minimize capital costs. These efforts are described elsewhere in this report. The previously stated challenges associated with constructing these new facilities within an operating major industrial facility also had an impact on engineering costs. Additionally, the IBDP compression and dehydration facility was relatively small compared with similar industrial-scale facilities. In a larger facility, engineering costs would not increase proportionally with an increase in facility size and cost, which would result in lower engineering costs when expressed as a percentage of the total capital costs. Costs included in the engineering costs category included those ADM paid to the equipment suppliers and local engineering firms, as well as the Trimeric process engineering costs paid by the University of Illinois at Urbana-Champaign on behalf of the MGSC.

Other Cost Categories

Other cost categories shown in Table 10 include direct cost categories for buildings and yard improvements and the indirect costs category of construction expenses. These categories were at or below the low end of the typical ranges shown in Table 10. This difference had to do primarily with the way costs were assigned to the other categories, as discussed. Because ADM provided very specific information regarding which construction costs were associated with the piping, electrical systems, and instrument and control categories, among others, Trimeric typically assigned construction costs to those other specific categories. Yard improvement costs might have been lower because the IBDP compression, dehydration, and transmission facility was installed within an existing facility that was originally designed for similar industrial purposes.

Operating Costs

This section provides an estimate of the operating costs for the IBDP compression, dehydration, and transmission facility. Wherever possible, actual values for key factors, such as the electricity used per ton (tonne) of CO_2 injected and the operator labor hours, were used to estimate these operating costs. However, as explained later in this section, average costs taken from public references for electricity costs in dollars per kilowatt-hour and for labor costs in dollars per hour are used to estimate electricity and labor operating costs because actual values are host-site confidential information. In addition, factors taken from the literature are used to estimate maintenance costs because actual maintenance cost information was not available during preparation of this report.

Estimate of Power Costs

The major operating cost for the IBDP compression, dehydration, and transmission facility was electricity. The major electricity requirements were for the 933 kW (1,250 hp) blower motor and for each of two 2,425 kW (3,250 hp) reciprocating compressor motors. These were the nameplate ratings for these motors, not the amount of power that they actually used during the injection operations. As discussed previously in this report, the average power requirement for the sum of these three motors was calculated at 4,721 kW (6,329 hp), and the supplier-predicted requirement at the 1,212.5 ton/day (1,100 tonne/day) summer design condition was 5,018 kW (6,726 hp). Power requirements for compression are lower in cooler weather. The average calculated power requirements were lower primarily because, on average, the plant operated at less than 100% of the design condition. The average injection rate was 1,116.6 ton/day (1,013 tonne/day), mainly because an administrative permit limit required ADM to operate at less than the full design condition until this limit was increased after about 1 year of injection. In addition, at times the CO_2 supply to the IBDP compression facility was limited because of operating conditions upstream of the IBDP compression, dehydration, and transmission equipment. When expressed on the basis of the energy required per ton (tonne) of CO_2 injected, the calculated average energy required was 101.6 kWh/ton (112 kWh/tonne), or 136.3 hph/ton (150.2 hph/tonne). The supplier-predicted energy required per ton (tonne) of CO_2 injected at the 1,212.5 ton/day (1,100 tonne/day) summer design condition was 99.8 kWh/ton (110 kWh/tonne), or 133.8 hph/ton (147.5 hph/tonne). Thus, agreement between the calculated energy requirement and supplier-predicted energy requirement was quite good (~2% difference).

Actual costs for electricity at the host site are confidential. Therefore, in this report Trimeric used the average October 2014 price for industrial customers in Illinois as reported by the U.S. Energy Information Administration, which is $0.0693/kWh. This assumed cost of electricity was multiplied by the calculated average energy requirement of 101.6 kWh/ton (112 kWh/tonne) of CO_2 injected to estimate the energy costs per ton (tonne) of CO_2 injected at $7.04/ton ($7.76/tonne) of CO_2 injected.

Estimate of Labor Operating Costs

The Archer Daniels Midland Company used a team of four operators supported by one supervisor to provide essentially 24-hour-a-day coverage during the project. Actual costs for labor at the host facility are confidential. Therefore, in this report, Trimeric used the average hourly wage for Plant and System Operators for Illinois according to the U.S. Bureau of Labor and Statistics. This value was $30.34/h for May 2014. Equation 1 was used to estimate the direct operating labor costs, based on a guideline in the American Institute of Chemical Engineers (AIChE) course titled "Practical Project Evaluation" (AIChE 2000):

$$\left(\frac{8\text{ hours}}{\text{shift}}\right)\times\left(\frac{21\text{ shifts}}{\text{week}}\right)\times\left(\frac{52\text{ weeks}}{\text{year}}\right)\times(\$1.20\text{ for overtime})\times\left(\frac{\$30.34}{\text{hour}}\right)+\left(\frac{40\text{ hours}}{\text{week}}\right)\times\left(\frac{3\text{ weeks vacation}}{\text{operator-year}}\right)\times(4\text{ operators})\times\left(\frac{\$30.34}{\text{hour}}\right)=\left(\frac{\$332,623}{\text{year}}\right). \quad (1)$$

The AIChE course used a multiplier of 1.32 to account for payroll overhead in addition to the direct operating labor, which resulted in an estimate for total operating labor costs of $439,063/year. Because approximately 367,436.7 tons (333,333 tonnes) was injected in each year of operation, operator labor costs including overhead were estimated at $1.20/ton ($1.32/tonne) of CO_2 injected.

Supervisor labor was estimated at 20% of the direct operating labor, which was $66,525/year, based on the same AIChE reference. Because approximately 367,436.7 tons (333,333 tonnes) was injected in each year of operation, supervisor labor was estimated at $0.18/ton ($0.20/tonne) of CO_2 injected.

Maintenance Costs

Actual maintenance cost data were not available for the preparation of this report. Typical costs for maintenance and repairs on an annual basis ranged from 2% to 6% of the fixed capital investment for a simple chemical process (Peters et al. 2003). Trimeric selected the 2% value to estimate maintenance costs for the IBDP compression, dehydration, and transmission facility. The lower end of the range was chosen because most of the maintenance costs were associated with the two reciprocating compressors, and the cost of the reciprocating compressors was a significant part of, but not the only contributor to, the fixed capital investment for the overall facility. Using the value for total fixed capital investment from Table 10 multiplied by 2% gave an estimated annual maintenance cost of $406,840. Because approximately 367,436.7 tons (333,333) tonnes was injected in each year of operation, maintenance costs were estimated at $1.11/ton ($1.22/tonne) of CO_2 injected.

Other Operating Costs

Other minor operating costs in addition to electricity, operating labor, and maintenance were associated with the operation of the IBDP compression, dehydration, and transmission facility. These included cooling water; fuel gas and glycol for the dehydration unit; lubrication oil for the blower and reciprocating compressors; and filters, gaskets, and other minor consumable items. Trimeric estimated these other operating costs on an annual basis at 1% of the fixed capital investment. Multiplying the total fixed capital investment from Table 10 by 1% gave an estimate for these other annual operating costs of $203,420. Because approximately 367,436.7 tons (333,333 tonnes) was injected in each year of operation, these other operating costs were estimated at $0.55/ton ($0.61/tonne) of CO_2 injected.

Plant overhead, which includes items such as office buildings; the personnel department; the safety department; warehouses; shop, maintenance, and laboratory facilities; a locker room, uniforms, and laundry; roads; wastewater treatment; solid waste disposal; and the like, was not included for the IBDP compression, dehydration, and transmission facility because this facility was installed in a larger host facility that already provided these functions. A typical estimate for plant overhead would be 60% of the total operating labor, supervisor labor, and maintenance costs (Peters et al. 2003). Trimeric did not include plant overhead because these functions were provided by the host facility. If these costs had been included, they would have been estimated at $669,509 on an annual basis, or $1.82/ton ($2.01/tonne) of CO_2 injected because approximately 367,436.7 tons (333,333 tonnes) was injected each year of operation.

Summary of Costs per Ton (Tonne) of Carbon Dioxide Injected

Table 11 provides a summary of the capital and operating costs per ton (tonne) of CO_2 injected, which were estimated by using the values derived earlier in this section. The capital costs shown in Table 11 should be very accurate because the project is completed and any uncertainty in the capital costs would be only because of the complexity of tracking costs on this kind of project. However, the operating costs shown in Table 11 are estimates based on typical assumed values for the cost of electricity, labor, maintenance, and other operating costs. As explained, these are estimates because actual costs for electricity and labor are confidential and the actual maintenance and other operating cost data were not available.

Table 11 Estimated costs per ton (tonne) of CO_2 injected

Cost category	Cost, $/ton ($/tonne) of CO_2 injected
Capital costs	18.45 (20.34)
Electrical power	7.04 (7.76)
Total operating labor	1.20 (1.32)
Supervisor labor	0.18 (0.20)
Maintenance costs	1.11 (1.22)
Other operating costs	0.55 (0.61)
Total cost	28.53 (31.45)

FULL-SCALE POWER PLANT CARBON DIOXIDE COMPRESSION AND DEHYDRATION

This section discusses scale-up from the MGSC Phase III Project/IBDP by a factor of 10, which is close to the amount of CO_2 compression and dehydration that would be associated with 90% CO_2 capture from a 550-MW (net) power plant. For this discussion, Trimeric has referenced the latest publication from U.S. DOE's National Energy Technology Laboratory regarding projected equipment and costs for CO_2 injection associated with 90% capture from a 550-MW (net) power plant, titled "Cost and Performance Baseline for Fossil Energy Plants, Revision 3" (U.S. DOE 2015).

Design Differences in the Illinois Basin – Decatur Project Versus a Full-Scale Power Plant

For installation of the carbon capture compression and dehydration unit in a full-scale power plant, consideration would need to be given to the type and size of compression equipment and to the final use or storage of the CO_2. For this discussion, the end use was generally assumed to be the same as in the U.S. DOE report and as in the IBDP, injection in a deep saline formation. For a large-scale, long-term injection operation, additional consideration needs to be given to reliability, turndown capability, number of compression trains needed to meet the design requirement, and equipment sparing.

Compressors

Because the CO_2 flow rate would be nominally 10 times that of the IBDP facility, it would likely not be economically feasible to use a similar type of reciprocating compressor as those used in the IBDP facility and scale up to the required number of reciprocating compressors. Furthermore, the space requirement would be larger and the process controls would be more complex as compared with the IBDP or with a lesser number of other compressor types that are better suited for the full-scale power plant application.

The current commercially available compressors likely to be considered for the full-scale power plant application are in-line (barrel) and integrally geared (I-G) multistage centrifugal compressors. The U.S. DOE report describes the compressor used in their most recent evaluation as an eight-stage centrifugal compressor with cooling after each stage of compression. One advantage of an I-G centrifugal compressor over a conventional in-line (barrel) centrifugal compressor is that cooling using conventional heat exchange equipment is possible after every stage of an I-G compressor, whereas this is often not the case with in-line centrifugal compressors. In addition to improving efficiency, this feature of the I-G compressor allows more choices for the dehydration unit operating pressure, which can lead to a lower cost, more efficient dehydration system. Centrifugal compressors are usually driven by electric motors, steam or gas turbines, or turboexpanders. The driver selection would be based on what is available and most cost effective at a given site. Integration between the power plant and the capture system (including CO_2 compression) is a key consideration in reducing the energy requirements for CO_2 capture and use or storage, and the type of driver for the compressors is part of the integration that must be considered in the design phase.

Carbon Dioxide Process Cooling

At the IBDP facility, extra cooling water was available from the ADM host site for compressor cooling after each stage of compression. This might not always be the case in full-scale power plant applications. Another option that might be considered for the large-scale power plant application if cooling water is available would be the use of wet surface air coolers (WSAC) for interstage cooling. The WSAC design essentially combines the evaporative cooling effect from a cooling tower and the wet contact cooling of a traditional shell and tube heat exchanger locally in a cluster of several conventional heat exchanger tube bundles. Here, both evaporative and direct water contact cooling take place for heat exchange, and the process CO_2 is inside the tube bundles. The WSAC might allow the use of lower quality water than would a cooling tower and could typically achieve cooler process (CO_2) outlet temperatures than would conventional cooling tower and shell and tube heat exchanger systems, which would lead to increased efficiency in the compression and dehydration processes.

If cooling water were not available, the design would likely use air-cooled (fin-fan) heat exchangers. One consideration with air coolers for the full-scale power plant application is that they have a larger footprint than do shell and tube heat exchangers; thus, air coolers require more space for their installation. This often leads to air coolers being installed farther away from the compressors. In CO_2 applications, stainless steel piping is often used on CO_2 return lines from the coolers to the next stage of compression because free liquid water might be present in these CO_2 streams. If the air coolers were positioned far away from the compressors, stainless steel piping might also be needed for the lines leaving each compression stage and going to the air coolers if there were a chance that the hot CO_2 leaving the compressors could cool below its water dew point before reaching the air coolers. The need for more stainless steel piping would increase the project costs.

Dehydration

Some common industrial gas dehydration methods are solid desiccant, solvent absorption, and gas pressure increase accompanied by cooling. The choice is usually driven by the required level of dehydration needed based on downstream use or storage objectives as well as the transportation distance between the compression facility and the point of use and storage. Dehydration requirements might be significantly different for an application that would use nearby injection wells on privately owned land compared with an application that would require several miles of CO_2 transportation by pipelines.

The level of dehydration achievable by using solid desiccants, such as molecular sieve systems, far exceeds what is needed to meet typical CO_2 pipeline water content specifications, and for the same reason as in the IBDP, a power plant would likely not choose this dehydration option. Instead, a solvent absorption option such as TEG, the dehydration technology used in the IBDP facility, would be a more likely selection for the full-scale power plant application. DexPro is an emerging process that uses partial CO_2 stream recycling to generate cooling to dehydrate the bulk CO_2 stream. This would also be a dehydration option worthy of consideration for the full-scale power plant application. The use of multistage compressors with cooling after each stage of compression would also be a good fit for DexPro and other processes that use gas compression and cooling to achieve dehydration.

Reliability, Sparing, and Turndown

Equipment reliability requirements for the compression and dehydration facility at a power plant would likely be greater than those for the IBDP. Because of the research nature of the IBDP, more time was allotted for outages than would likely be acceptable in a full-scale power plant application. The use of centrifugal compressors, as discussed in the Illinois Basin – Decatur Project Versus Full-Scale Power Plant Design Differences section, would be expected to result in higher compression equipment reliability than that achieved in the IBDP because reciprocating compressors typically require more maintenance than do centrifugal compressors.

If power plant operations would be affected by the availability of the CO_2

capture system, an even higher level of equipment reliability might lead to more sparing of slightly more process equipment, instrumentation, and analyzers, but it is not likely that equipment with a high capital cost, such as compressors, would be spared. The most recent U.S. DOE report (U.S. DOE 2015) cited above did not include spare compressors. It used two 50% capacity units operating in parallel.

Turndown capabilities must be considered because a variable load on the power plant can result in changes in the flow rate of CO_2 captured, thus requiring some level of turndown and associated controls for the CO_2 compression and dehydration equipment. For centrifugal compressors, care must be taken to avoid unstable conditions, such as compressor surge and stonewall (also known as choking). In compressor surge, the flow rate in the compressor reaches a minimum point at the same time the discharge pressure of the compressor increases beyond the maximum acceptable limit for the current flow rate, which results in reverse flow through the compressor. In stonewall or choking, the pressure downstream of the compressor is much lower than the design condition, which results in flow through the compressor in the normal direction but at an unacceptably high flow rate and velocity. Compressor surge or stonewall (choking) can result in excessive vibration and, in a few severe cases, serious damage to the compressor. Modern compressor designs use features such as recycle valves and backpressure control valves to prevent these upset conditions, but doing so imposes some limitations on turndown, efficiency at turndown conditions, or both.

Impact of Outdoor Installation

The IBDP compression equipment and some of the dehydration equipment were located inside fully enclosed buildings specifically built to house this equipment, mainly because of the harsh winter conditions that could occur at the project site. Geographic factors and the larger scale compressor and dehydration equipment needed for a full-scale power plant application might result in the need to install the equipment outdoors or in partially enclosed buildings. Because many industries have extensive experience with outdoor installation of compression and dehydration equipment, outdoor installation of CO_2 compression and dehydration systems for full-scale power plant CO_2 capture applications should be possible in suitable geographic locations. Additional insulation, heat-tracing, weatherproof motors, analyzer housing, and similar considerations might be necessary if more of the CO_2 compression and dehydration equipment were installed outdoors or in partially enclosed buildings.

Process Piping

Regarding process piping selection for the full-scale power plant application, several site-specific considerations would need to be made, including the location of the CO_2 source relative to the compression equipment and the compression equipment location relative to interstage cooling, dehydration, and transmission piping or point of use or storage. Carbon dioxide leaving the capture source is likely to be saturated with water vapor, but this depends on the capture process. When CO_2 is saturated or contains enough water vapor that condensation of liquid water is possible, a corrosion-resistant piping system is often selected. In cases such as hot compressor discharge, when the CO_2 is well above its water dew point, carbon steel piping is often used, but non-steady-state conditions, such as scheduled and unscheduled shutdowns and idle periods as well as ambient temperature effects, must also be considered.

The use of stainless steel piping is one option for instances in which free liquid water and CO_2 might be present at the same time (which could be highly corrosive to carbon steel), but other options might be less expensive. Carbon steel piping with an internal corrosion resistant liner or coating is a commonly used alternative to stainless steel in wet CO_2 applications. However, care must be taken to install the liner or coating properly before it is placed into service so that it covers all the carbon steel and prevents it from being damaged by operations, maintenance, or modifications after it has been placed into service. Any exposed carbon steel, particularly around welds and flanges, can become a point of rapid, localized corrosion. This can lead to unscheduled downtime and expensive repairs that can offset some or all the cost savings relative to the use of stainless steel piping. Fiberglass-based piping is also used in lower pressure wet CO_2 applications. Pressure ratings of the piping materials at normal and maximum design temperatures must be considered when evaluating the piping materials. Once the CO_2 has been sufficiently dehydrated to ensure that free liquid water cannot form under any normal or off-design operating, shutdown, or idle conditions, then regular carbon steel piping is the common choice.

In addition to internal corrosion, other piping aspects need to be considered, including operability considerations and external corrosion. Finally, a cost analysis should be done for each piping segment, from the CO_2 source to compression, to cooling, back to compression, to dehydration, and to the point of transmission or final use or storage. As noted earlier, other piping cost components, including weight and associated structural requirements, need to be considered along with the cost of the piping itself.

Transmission Pipeline

For the design of the transmission pipeline, consideration must be given not only to the pipeline diameter, elevation changes, and pressure drop, but also to any applicable regulations or codes that affect design factors or that address fluid categorization, location classification, or pipeline routing. In many cases, the U.S. Department of Transportation Pipeline and Hazardous Materials Safety Administration regulations are applicable to CO_2 pipelines. When selecting the pipeline diameter, it is important to consider any potential future CO_2 flow rate requirements as well as the current design requirements.

Carbon steel is the most common and most economical choice for materials of construction for dry (dehydrated) CO_2 pipelines. Attention should be given to proper operation of the pipeline when using carbon steel. For example, during CO_2 pipeline blowdown, very low temperatures could be reached if liquid CO_2 were allowed to form in the pipeline before blowdown. Most CO_2 pipelines operate in excess of the CO_2 critical pressure, which is 1,057 psig (72.9 barg) for pure CO_2, but shutdown and other off-design conditions need to be considered with respect to the potential for liquid CO_2 to form and the generation of temperatures lower than the minimum allowable temperature for carbon steel. Although internal corrosion issues might not be a concern with CO_2 once it has been dehydrated, consideration should be given to external pipeline coatings and cathodic protection to minimize the potential for external corrosion.

An important caution for carbon steel CO_2 transmission lines is ensuring that any water from pipeline hydrostatic pressure testing (hydrotesting) is sufficiently purged before putting the pipeline into CO_2 service. Measuring the dew point at various locations in the pipeline during commissioning should be considered.

Number of Injection Wells

Considering that the CO_2 captured from a full-scale power plant would be injected into an underground formation for many years, a number of injection wells might be required. Projects of this type typically include geologic studies to select appropriate formations for long-term CO_2 storage and to estimate the number of injection wells that might ultimately be required to support long-term operations. The management of different and changing wellhead pressures and the compressed CO_2 piping network might require additional considerations, although such studies are beyond the scope of this report.

Cost Comparison of the Illinois Basin – Decatur Project with a Full-Scale Power Plant

The cost information from the IBDP provided a basis for scale-up and comparison with U.S. DOE cost estimates for full-scale power plant CO_2 capture costs. These costs are available in a report titled *Cost and Performance Baseline for Fossil Energy Plants, Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity, Revision 3, July 2015* (U.S. DOE 2015).

Purchased Equipment Cost Comparison

An exact comparison could not be made when using the IBDP compression and dehydration unit purchased equipment costs to scale up for a full-scale power plant application because the IBDP facility used reciprocating compressors as the base type of compression equipment, whereas the full-scale power plant application would likely use centrifugal compressors. Furthermore, a factor of 10 is often considered the upper limit for reasonable scale-up of purchased equipment costs when the equipment or facility capacities differ. Trimeric does have in-house budgetary cost data indicating that costs for large-scale CO_2 reciprocating and centrifugal compressors would be comparable. Thus, having clearly stated the caveats regarding different compressor types and the large difference in capacities, here we present some economic comparisons between the IBDP and a full-scale power plant application.

The purchased equipment costs for the IBDP compression and dehydration facility were $6,145,000 (June 2009 U.S. dollars). Adjusting to June 2011 dollars and using a 0.69 scaling exponent for scale-up from 1,212.5 to 12,676.5 ton/day (1,100 to 11,500 tonne/day, or 21 to 219 MMscfd) of CO_2 for compression equipment, the estimated equipment costs for the full-scale power plant application based on scaling up IBDP costs was $35,851,000. This amount is 28.6% lower than the U.S. DOE (2015) reported equipment cost of $50,211,000 for the 550 MW (net) power plant. This comparison was prepared by calculating the ratio of the full-scale power plant capacity to the IBDP capacity, raising this ratio by an exponent of 0.69, and multiplying this result by the IBDP equipment cost to estimate the full-scale power plant equipment cost. The 0.69 exponent is the typical scaling exponent for reciprocating compressors, which were the main components of the IBDP facility costs (Peters et al. 2003). The Chemical Engineering Plant Cost Index (Chemical Engineering 2009–2011) published in *Chemical Engineering* magazine was used to scale up equipment costs from the applicable time period on the IBDP to the June 2011 time period used in the U.S. DOE (2015) report. Equation 2 shows the overall calculations used to estimate the full-scale purchased equipment cost based on the IBDP estimates:

$$\text{IBDP full-scale purchased equipment cost, \$MM} = \$6.145\text{MM} \times \left(\frac{\text{CEPCI June 2011, 588.9}}{\text{CEPCI June 2009, 508.9}}\right) \times \left(\frac{\text{219 MMscfd}}{\text{21 MMscfd}}\right)^{0.69}. \tag{2}$$

Total Facility Cost Comparison

The fixed capital investment for the IBDP compression and dehydration facility was $20,342,000 (2009–2011 U.S. dollars). Adjusting to June 2011 dollars and using a 0.69 scaling exponent for scale-up from 1,212.5 to 12,676.5 ton/day (1,100 to 11,500 tonne/day, or 21 to 219 MMscfd) of CO_2 for the compression equipment, the estimated total installed equipment cost for the power plant was $108,538,000. This amount is 10.3% higher than the estimated installed equipment cost of $98,381,000 in the U.S. DOE (2015) report. The comparison was prepared by calculating the ratio of the full-scale power plant capacity to the IBDP capacity, raising this ratio by an exponent of 0.69, and multiplying this result by the IBDP total facility cost to estimate the full-scale power plant facility cost. The 0.69 exponent is the typical scaling exponent for reciprocating compressors, which were the main components of the IBDP facility costs (Peters et al. 2003). The Chemical Engineering Plant Cost Index (Chemical Engineering 2009–2011) was used to scale up equipment costs from the applicable time period on the IBDP to the June 2011 time period used in the U.S. DOE (2015) report. Equation 3 shows the overall equation used to estimate the full-scale fixed capital investment based on the IBDP estimates:

$$\text{IBDP full-scale fixed capital investment, \$MM} = \$20.34\text{MM} \times \left(\frac{\text{CEPCI June 2011, 588.9}}{\text{CEPCI June 2010, 556.4}}\right) \times \left(\frac{\text{219 MMscfd}}{\text{21 MMscfd}}\right)^{0.69}. \tag{3}$$

Installation Factors

The Lang factor is often used to estimate the fixed capital investment for a facility when only an estimate of the purchased equipment costs is available. The fixed capital investment includes the delivered equipment cost, site improvements, foundations, buildings, piping, electrical system, controls, engineering and construction costs, and overhead costs, including insurance, taxes, contingency fees, and contractor fees. Because both the fixed capital investment and the purchased equipment costs are known for the IBDP compression and dehydration facility, the Lang factor can be calculated for this facility. According to the cost data in Table 10, the calculated Lang factor for the IBDP compression and dehydration facility is 3.31. This value is lower than the common Lang factor of 4.74 used in order-of-magnitude cost estimates for fluid process plants; however, order-of-magnitude estimates are typically ±30% (Peters et al. 2003). When full-scale power plant CO_2 compression and dehydration facility costs in the U.S. DOE (2015) report were used, the calculated Lang factor was even lower, 1.96. This result may have been due to the higher purchased equipment costs relative to other project costs (economies of scale) and the U.S. DOE basis of a greenfield installation. Installation costs could be higher if a CO_2 compression and dehydration facility needed to be retrofitted within an existing power plant or other existing facility. As discussed, both challenges and benefits were associated with retrofitting the IBDP compression and dehydration facility within the existing ADM facility.

Energy Comparison

A larger scale CO_2 compression facility with a longer period of operation would likely be designed to have lower energy requirements. The higher capital costs for more efficient compressors would be justified for an injection rate 10 times larger than that for the IBDP and an assumed operational period 10 times longer, which is 30 years in the U.S. DOE (2015) report. Power requirements in the U.S. DOE (2015) report are 67.1 kWh/ton (74 kWh/tonne), which is 34% lower than the actual calculated IBDP power requirement of 101.6 kWh/ton (112 kWh/tonne). The compression system modeled in the U.S. DOE (2015) report is more efficient, at approximately 90% polytropic efficiency compared with approximately 80% polytropic efficiency for the IBDP facility. The U.S. DOE (2015) report includes eight stages of compression, whereas five were used in the IBDP. However, for a more comparable compression ratio, the compression facility in the U.S. DOE (2015) report used seven stages rather than five to achieve an outlet pressure comparable to the IBDP surface injection pressures. The eighth stage of compression for the compressor system in the U.S. DOE (2015) report reached a higher pressure than was required during IBDP injection operations. Using a greater number of stages to achieve a given compression requirement is typically more efficient, particularly when cooling occurs after each stage of compression, as was the case both in the U.S. DOE (2015) report and for the IBDP facility.

CONCLUDING REMARKS

In summary, this project was a success. Major accomplishments include the injection of a nominal 1.1 million tons (1 million tonnes) of CO_2 in a 3-year period, with an estimated 1,006,410.2 tons (913,000 tonnes) of net CO_2 captured. (See Table 11 for a breakdown on estimated cost per tonne of CO_2 injected.) The CO_2 surface facilities were designed, constructed, and operated as intended to compress, dehydrate, and transport 1,102 ton/day (1,000 tonne/day) of CO_2 to the injection well. Equipment and process data were collected and monitored to characterize and optimize performance and to predict scheduling of the maintenance activities needed to sustain system performance. Design and commissioning challenges for the IBDP surface facilities involved requiring operational flexibility with respect to capacity turndown and the ability to deliver CO_2 over a wide range of surface injection pressures, integrating the IBDP surface facilities into the existing ADM host site, and allowing time for the injection permit to be issued. Significant operating challenges and problems were related to the need to purge the system to remove residual moisture after equipment shutdowns and the need for careful selection of the type of cylinder lubrication oil, as well as the need to optimize the oil injection rate. Our initial operating experience confirmed the prediction by the design team that insulation would be needed on aboveground piping to maintain a stable injection temperature and pressure over the full range of ambient conditions encountered and to facilitate compliance with permit conditions. The aboveground pipeline was insulated within the first few months of operation even though insulation was a significant project expense.

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