DESIGN CONSIDERATIONS FOR CO_2 RECOVERY AND SEQUESTRATION FROM A GAS PROCESSING FACILITY

Laurance Reid Gas Conditioning Conference
February 26-29, 2024 – Norman, Oklahoma USA

Brad Piggott, P.E.
Trimeric Corporation
PO Box 826
Buda, Texas USA 78610
+1 512-295-8118
brad.piggott@trimeric.com

Darshan Sachde, Ph.D.
Trimeric Corporation
PO Box 826
Buda, Texas USA 78610
+1 512-295-8118
darshan.sachde@trimeric.com

Ethan Pulfrey
Trimeric Corporation
PO Box 826
Buda, Texas USA 78610
+1 512-295-8118
ethan.pulfrey@trimeric.com

Joe Lundeen, P.E.
Trimeric Corporation
PO Box 826
Buda, Texas USA 78610
+1 512-295-8118
joe.lundeen@trimeric.com

Carrie Beitler
Trimeric Corporation
PO Box 826
Buda, Texas USA 78610
+1 512-295-8118
carrie.beitler@trimeric.com

ABSTRACT

Natural gas production and processing often includes acid gas removal units (AGRU) that separate CO_2 and H_2S from natural gas to meet transport or downstream processing specifications. This highly concentrated CO_2 stream represents a potential opportunity for deployment of carbon capture, utilization, and storage (CCUS) projects. As government incentives make CCUS more economically feasible, natural gas processors are more often considering CO_2 recovery options from natural gas processing. Methods to gather the CO_2 from multiple sources into “hubs” are also being considered to further improve overall project economics.

This paper focuses on the process of CO_2 recovery from a representative gas processing facility in the southern United States to meet specifications for pipeline transport and sequestration. Example CO_2 injection specifications are presented, although the specifications for injection can vary based on the requirements for a particular site or transport system. Many different technologies can be used depending on the source gas characteristics and injection conditions, and a summary of the options considered for the representative facility are briefly discussed. In this paper, the selected technology to meet the necessary specifications involves use of reciprocating compressors and triethylene glycol dehydration. Reciprocating compressors are commonly used in the natural gas industry and have also become widely accepted for compressing CO_2 as well. TEG dehydration is also common to both natural gas and CO_2 service. Important operating and design factors for this equipment specific to CO_2 service will be presented including limiting operating conditions, material selection and compatibility in CO_2 service, temperature management and phase behavior, off-design operating conditions (e.g., start-up and shutdown), and capacity control.

DESIGN CONSIDERATIONS FOR CO_2 RECOVERY AND SEQUESTRATION FROM A GAS PROCESSING FACILITY

Brad Piggott, P.E., Trimeric Corporation, Buda, TX
Darshan Sachde, Ph.D., Trimeric Corporation, Buda, TX
Ethan Pulfrey, Trimeric Corporation, Buda, TX
Joe Lundeen, P.E., Trimeric Corporation, Buda, TX
Carrie Beitler, Trimeric Corporation, Buda, TX

Introduction

Natural gas processing facilities typically produce CO_2 as a byproduct or waste stream from different unit operations, at varying purities. In these facilities, an acid gas removal unit (AGRU) is often included and vents CO_2 at high purity > 99 vol% (dry basis), making this vent stream an ideal candidate for CO_2 capture. This high purity CO_2 vent stream can be compressed to required pipeline conditions, and either sold to a CO_2 pipeline company for sequestration, utilized in an enhanced oil recovery (EOR) process, or sequestered on site if the local geology allows.

Other sources of CO_2 at a gas treatment plant are typically lower purity, such as exhaust from fired heaters (7-14 vol% CO_2) or gas turbine / engine drivers (~ 4 vol% CO_2). These sources are also low pressure (i.e., near atmospheric pressure flue gas), meaning the partial pressure of CO_2 in the stream is low. These low purity sources require substantial additional processing, such as an amine unit, to separate CO_2 from other bulk gas components to make it suitable for transport and storage. This additional processing increases the amount of purchased equipment required, as well as the operating cost of the CO_2 capture facility. With the current tax credit being $85 per metric ton of CO_2 sequestered [1], and the comparatively low quantity of CO_2 from these sources, the higher cost of capture may outweigh the economic benefit of additional CO_2 being captured from these low purity streams.

This paper presents an example case for CO_2 capture and compression at a representative natural gas processing facility in the southern United States where the recovered CO_2 is transferred to a third-party pipeline for eventual sequestration. The representative natural gas processing facility treats 400 MMscfd of sour gas (~ 0.01 to 0.02 mole% H_2S) with the use of two amine AGRUs to bring the natural gas to pipeline specifications. The example CO_2 capture process downstream of the AGRU consists of the following steps:

1. The separated acid gas exits the AGRU regenerator vents and goes to an H_2S removal unit. In addition to reducing the H_2S content to meet the CO_2 transport specification, the H_2S removal unit also reduces the pressure of the product CO_2 stream to near atmospheric. This will have practical impacts on downstream equipment selection and processing. Any H_2S removal that occurs downstream of the AGRU should not significantly reduce the purity of the CO_2 (e.g., does not result in dilution of the already separated CO_2 stream with other gas species).

   Note that H_2S removal is not always required and depends on the H_2S content of the feed natural gas, specifications for H_2S in the downstream pipeline and injection, and any associated costs of leaving H_2S in the CO_2 product stream (e.g., metallurgy considerations). In addition, H_2S removal can occur at different points in the natural gas processing path using a variety of different technologies (e.g., scavengers upstream of the amine AGRU or even at the gas production sites; regenerable processes such as liquid redox, solid or liquid scavengers, other processes downstream to treat the acid gas). However, H_2S removal is not the focus of this paper and readers are referred to other literature for details on this topic e.g., [2].

2. The low sulfur acid gas, or wet CO_2 product stream (referred to simply as CO_2 moving forward) exiting the H_2S removal unit is compressed to an intermediate pressure, dehydrated, and then the dry CO_2 is compressed up to injection conditions.

3. The CO_2 flows to a third-party pipeline company for sequestration.

This paper discusses the source CO_2 conditions and required outlet CO_2 conditions, typical technologies used for compression, dehydration, pipeline transport and sequestration, and the impact of CO_2 service on reciprocating compressors and TEG dehydration units.

CO_2 Processing Inlet and Sequestration Injection Conditions

Table 1 shows the source CO_2 and delivery conditions for the CO_2 compression facility at the example natural gas processing plant in this paper as well as typical injection conditions for sequestration.

Table 1 – CO_2 Source and Delivery Conditions

	Example Facility		**Literature Range in Transport/Injection Conditions for Sequestration**
Parameter	**CO_2 Source (after H_2S removal)**	**CO_2 Delivery Requirement for this Example**	
Pressure, psig	0.1	2,200[2]	1,300 to 5,000 [3]
Temperature, F	115	120[3]	< 90 – 120[3] [4]
Composition			
Water	9.87 mole%	630 ppmv / 30 lb/MMscf	20 to 650 ppmw [5]
Methane	0.20 mole%	0.01 to 4 mole% [5]	0.01 to 4 mole% [5]
Ethane	<10 ppmv	0 to 1 mole% [5]	0 to 1 mole% [5]
Propane	<1 ppmv	0 to 1 mole% [5]	0 to 1 mole% [5]
Nitrogen	<1 ppmv	< 4 mol%	0.01 to 7 mole% [5]
Oxygen	Negligible	< 10 ppmw	0.001 to 4 mole% [5]
Carbon Dioxide	89.93 mol%	>95 mol%	90 to 99.8 mole% [5]
Hydrogen Sulfide	< 20 ppmv[1]	< 20 ppmv	0.002 to 1.3 mole% [5]
Glycol	N/A	N/A	< 0.3 gal/MMSCF [4]

Note 1: H_2S removal technologies can achieve a range of treated gas H_2S specifications, including lower levels than the 20 ppmw required in the delivery spec (e.g., liquid redox can achieve < 5 ppmv in some cases).
Note 2: Maximum delivery pressure may be lower in practice, but 2,200 psig is typical design pressure for ANSI 900 pipeline [4].
Note 3: CO_2 delivery temperature requirements may need to be discussed directly with pipeline operators as it can impact pipeline design (e.g. MAOP, underground coatings). In this example case, the gas processing plant only has air cooling available, so the pipeline temperature specification is tailored to accommodate the use of air cooling.

Depending on the natural gas processed at a facility, the amount of hydrocarbons in the source CO_2 can vary, and may also contain low levels of BTEX. The hydrocarbon content of the amine acid gas, and ultimately the CO_2 gas, impacts density and other properties that may affect the entire CO_2 capture, transport and injection system design and operation. For example, the presence of hydrocarbons in the amine unit acid gas can impact the water holding capacity of the acid gas stream/CO_2 gas and change the phase behavior, including the critical pressure of the CO_2 stream. Therefore, it is important to understand the potential effects that variations in the acid gas hydrocarbon content can have on the CO_2 system performance and try to avoid operating the AGRU at conditions that may cause high hydrocarbon content to occur. For example, it may be possible to operate the rich amine unit flash tank at lower pressures to evolve more hydrocarbons from the amine solution in the flash gas and lowering the hydrocarbon content of the amine unit acid gas. Other operational changes to the amine unit could potentially be considered.

The operating pressure of the acid gas exiting an AGRU may range from ~5 to 15 psig. Some AGRU regenerators run at elevated pressures (i.e., upper end of the operating pressure range), which can reduce the overall compression requirements and eliminate the requirement for a blower upstream of the reciprocating compressor. Operating at higher AGRU regenerator pressure comes at the cost of thermal energy, as the amine in the AGRU regenerator must be heated to a higher temperature, and more importantly, may lead to increased amine degradation which can limit the acceptable upper operating pressure/temperature for the regenerator. In this example case, H_2S is removed from the acid gas after exiting the amine AGRU, resulting in a near atmospheric pressure CO_2 stream that may necessitate a blower upstream of the reciprocating compressor to improve cost and efficiency. As will be shown, it is important to note that the capture facility becomes the primary pressure control on the AGRU regenerators.

The range of potential delivery pressures is driven by characteristics of different geological formations where the potential injection well is located, as well as different potential transport systems. If the specific injection well has not been characterized at the beginning of a project, the required wellhead pressure estimates will be just that, estimates, with significant differences possible. In some situations, equipment must be ordered before a refined estimate of the required compression discharge pressure can be obtained. There is still some potential for the discharge pressure to not be fully known even up to commissioning and operation. Table 2 shows examples of actual discharge pressures versus the expected discharge pressure at an early stage in project.

Table 2 – Projected Versus Actual Discharge Pressures for CO_2 Projects

Project	Actual Discharge Pressure at Commissioning (psig)	Projected Discharge Pressure at Equipment Order (psig)
Local Injection	1,300	1,800
Local Injection	700	1,500
Pipeline Network w/Multiple Sources	Not Running Yet	1,405 – 2,114

Overview of CO_2 Pipeline Transport and Sequestration Technology

There are many different types of compression and dehydration technologies that can be used for pipeline transport and sequestration of CO_2. This section presents an overview of the commonly used equipment configurations and the reasons that reciprocating compressors and TEG dehydration were selected for the example application in this paper.

Use of an Upstream Blower

If compression is required from low pressure (e.g., near atmospheric), a blower may be used at the beginning of the process to reduce the size and improve the efficiency of the downstream compressor depending on the type of compressor used. Due to CO_2 behaving as an ideal gas at low pressure (compressibility factor > 0.99), a blower doubling the pressure from ~14.7 psia to 30 psia will nearly cut the volumetric flow in half. With a positive displacement compressor, such as a reciprocating or screw compressor, a multistage centrifugal blower added upstream would increase the mass throughput capacity [6] for a given compressor size or reduce the required size of the cylinders. For example, reciprocating compressors may require multiple large bore cylinders per stage of compression for low suction pressure applications and only be able to compress a small mass or molar amount of gas. On the other hand, a blower may not be required for low-pressure applications with centrifugal type compressors because they can typically handle lower suction pressures without a large cost impact. For integrally geared centrifugal compressors, a blower can be worth the added expense if a pinion gear can be saved. An additional pinion gear is used for each odd stage added (3rd stage, 5th stage, 7th stage).

Compressor Options

The compressor technologies that were considered for this example application include reciprocating, screw, and centrifugal compressors. Details of the mechanical differences in these types of compressors can be found in the literature [6]. Standard oil-flooded screw compressors can compress CO_2 from approximately 10-20 psig to 350-400 psig. Special configurations with a discharge pressure of up to 700 psig or higher may be possible [7]. At these high pressures, oil-losses and contamination of the CO_2 may become an issue. Since this discharge pressure is below the critical pressure for CO_2 (1,070 psia), the CO_2 would need to be cooled to a low temperature with a refrigerant system to liquefy the CO_2 so that it could be pumped to 2,200 psig. This approach may be useful when liquefaction is already required (e.g., for truck or ship transport). Alternately, screw compressors could be used in combination with a different compression technology to achieve the required injection pressure. For example, a blower-screw compressor-reciprocating compressor combination could be used or the blower could potentially be replaced with just a screw compressor as well. Because it was anticipated that there would be insufficient cost savings to justify the added complexity of using screw compressors with liquefaction or in a hybrid compression system, the use of screw compressors was not considered further in the example design/application discussed in this paper.

Therefore, the primary consideration for this project was between a single stage compression train with a centrifugal machine and multiple parallel trains of reciprocating compression. Table 3 shows a high-level comparison of the advantages and disadvantages for these two compression options.

Reciprocating compressors were selected for the example application given the relatively low CO_2 feed gas flow (30 MMscfd) and flexibility in handling variations in feed gas conditions. Reciprocating compressors are a generally adaptable technology in that additional stages can be added to go to higher pressures, interstage coolers can be air or water cooled, and the machines can be easily retrofitted for new conditions. Reciprocating compressors are often used in natural gas processing so this type of compressor would be familiar to operators in this setting as well. Centrifugal compressors tend to become more economic relative to other compressor options when the gas flow rates exceed approximately 50 MMscfd. Centrifugal machines are also sensitive to changes in feed gas conditions and are less common in the natural gas industry. Moreover, a centrifugal machine is limited in its ability to handle varying differential pressure due to its characteristic flow versus head operating curve. As such, reciprocating compressors are much more forgiving in this example application where the discharge pressure is a significant unknown, there could be changes to the feed gas temperature and composition, etc.

Table 3 – Comparison of Reciprocating and Centrifugal Compressors **[6]**

	Reciprocating Compressors		**Centrifugal Compressors**	
Parameter	**Advantages**	**Disadvantages**	**Advantages**	**Disadvantages**
Capacity Constraints	• High flexibility • Capable of handling smaller volumes	• High-capacity applications may require multiple trains/increased complexity	• Largest volume per train • High efficiency	• Economics and efficiency deteriorate at lower volume flows
Pressure Ratio Constraints	• 4:1 max per stage • Multiple stages for high pressure		• Multiple stages can be used	• Typically, <2-4 per stage
Maintenance / Reliability		• Higher maintenance • Regular maintenance • Overhauls (~every 3 yrs)	• Lower maintenance • Higher reliability (~90% availability) • 5 years run time w/o service	
Turndown / Operation	• Reasonable turndown (50% before recycle needed) • Tolerates changes in inlet gas composition and feed/discharge pressure • Can retrofit machine			• Limited turndown (75-80% before recycle needed) • Limited flexibility (e.g., dP) • Hard to repurpose • Sensitive to changes in inlet gas conditions
Familiarity	• Commonly used in natural gas processing and			• Not as common in natural gas processing

	Reciprocating Compressors		Centrifugal Compressors	
Parameter	**Advantages**	**Disadvantages**	**Advantages**	**Disadvantages**
	enhanced oil recovery (EOR)			
Equipment Variation	• Interstage air or water coolers used	• Upstream blower used	• Upstream blower may not be needed	
Capital Cost	• Typically, lower capital • Shorter lead time up to 52 weeks)			• Typically, higher capital • Longer lead time (> 70 weeks) • Expensive spare parts
Size		• Large footprint with multiple trains • Significant equipment between stages	• More compact	
Other		• Entrained lubricant • Torsional, harmonic, pulsation studies required	• No vibration and little harmonic concern • No oil in contact with process gas	

Compressor Drivers

Several types of compressor drivers can be used including engines, turbines, and electrical motor drives (EMDs). In CO_2 capture, fired drivers are typically only used if electricity is not available at the site due to associated emissions. For centrifugal compressors, EMDs or turbines are typically used. For reciprocating compressors, EMDs or engines are more common.

In general, gas-fired turbines and engines will both emit CO_2 in their exhaust that may be difficult to recover. Turbines and engines also require more preventive maintenance than EMDs and are permitted as emission sources. Turbines are an economic option for speed control with inline centrifugal machines, but speed control may not be practical for integrally geared centrifugal machines. Likewise, engines also allow for low-cost speed control, but the speed control on reciprocating compressors may be limited because of pulsation and/or torsional concerns. Engines are typically more expensive than EMDs, potentially about up to 30% depending on the conditions. In the representative gas processing facility, there is sufficient electrical infrastructure for EMDs to be used.

CO_2 Pump

A CO_2 pump can be used as the last stage of compression if the density of the supercritical CO_2 at the suction of the pump is greater than 30 lb/ft^3; in this example a higher density criteria of 35 lb/ft^3 was used for a conservative margin when specifying the pump. The density will be dependent on the operating pressure, type of interstage cooling media used (water vs air), and presence of light ends (e.g., N_2, light hydrocarbons) in the gas. Higher densities can be achieved at lower pressures when cooling water exchangers are utilized since lower process temperatures are generally possible.

Utilizing CO_2 pumps as the last stage of compression can have several benefits including:

- Potential for reduced overall capital cost.
- Improved management of fluctuations in injection pressure with use of a VFD so that the upstream compressor equipment can operate at the desired operating pressure.
- Added flexibility in the design if the exact delivery conditions are not known.

Some disadvantages with CO_2 pumps as the last stage of compression include:

- Additional rotating equipment that can cause unintended outages.
- Additional leakage point to atmosphere through the pump seal.
- Additional process control instruments to keep the pump within the design operating envelope.

Dehydration

Triethylene glycol dehydration is commonly used to meet CO_2 pipeline transport and sequestration specifications since the gas only needs to be dehydrated to 30 lb/MMscf. The glycol dehydration unit is often located at an intermediate pressure (roughly 300 to 800 psig) in the compressor train. This is because, as shown in Figure 1, at a fixed temperature the water content of the source CO_2 generally decreases with increasing pressure until it reaches a minimum. Then, as the pressure increases, the water holding capacity of the CO_2 increases again. The minimum water content varies with the pressure, temperature, and composition of the source CO_2. This is unlike what is seen in a natural gas stream where the water content continually decreases as pressure increases.

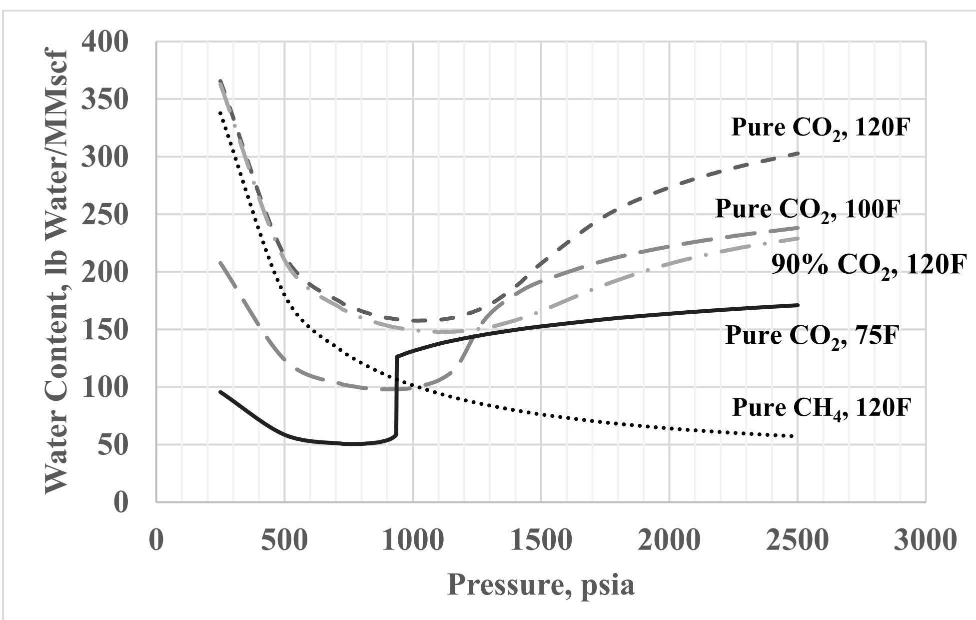

Figure 1 – Saturated Water Content for CO_2 and Methane as a Function of Pressure, Temperature and Composition

Incorporating the dehydration unit near the minimum water content pressure results in a smaller dehydration system with lower operating expenses. However, the pressure at which dehydration occurs is also minimized to prevent significant TEG losses into the CO_2 caused by evaporation from increased solubility of TEG in CO_2 at higher pressure. A literature source indicates that TEG is not suitable above pressures of approximately 950 psia because of higher solubility of TEG into the CO_2 at elevated pressures (i.e., higher TEG losses) [8]. Selecting an optimal pressure for dehydration unit operation will be application specific and should consider the trade-off in the size/operating cost of the dehydration unit, design pressure of the dehydration unit (TEG contactor tower), glycol losses, materials of construction, and other factors. If dehydration were required at elevated CO_2 pressures, then glycerol could be used instead of TEG to dehydrate the gas. TEG is 10 to 200 times more soluble than glycerol in CO_2 over a pressure range of 1,200 to 2,000 psig. [9] At these pressures, TEG would have excessive vapor losses and makeup requirements, making TEG dehydration uneconomic.

The TEG dehydration unit can be modified to include a Stahl column with stripping gas to lower the dry gas water content further if needed; some operators will design for as low as 7 lb/MMSCF.

One consideration of utilizing a TEG dehydration unit is air emissions. The dehydration unit may have a BTEX removal unit to remove these species from the dehydration still off-gas because the glycol absorbs some BTEX (but not all) from the CO_2. CO_2 also gets absorbed into the TEG that will flash off from the solution either because of pressure reduction or heat added for regeneration. Vent gas from the flash tank can be recycled to the compressors, while CO_2 from the regenerator is often vented to the atmosphere as allowed by environmental requirements. The amount of CO_2 absorbed in the TEG can be considerably higher than for a similar dehydration application of natural gas; a literature source reports methane absorption to be <10% of the CO_2 absorption at the same temperature and pressure [8].

TEG dehydration is usually less expensive than other technologies (e.g., mole sieve) and has a low pressure drop. TEG carryover to downstream compression stages and injection site can still occur so the unit should be equipped with an appropriate separation device on the treated gas. TEG units are also commonly used in natural gas operations so they would be familiar to operators in this setting. For these reasons, TEG dehydration was selected for the example application.

Mole sieve dehydration could be used if deep water removal (1-10 ppmv) is required. Mole sieve dehydration uses a solid adsorbent to remove water from the CO_2. The water adsorbs onto the solid. The process usually involves at least two beds so that one is operating while the other is regenerated with hot gas or a pressure swing. Since deep water removal is not required for this example application, the added cost of the mole sieve dehydration system is not warranted. Other solid sorbent technologies (e.g., silica gel) can also be used, but have similar trade-offs as the mole sieve unit.

Other refrigeration-type dehydration processes can be used depending on the conditions of the CO_2 stream. External refrigeration requires a separate refrigerant system and can be expensive. Given the potential for low cooling temperatures, care must be taken so that hydrates and/or liquid CO_2/water ice do not form. Internal refrigeration utilizing a valve to reduce the pressure and cool the CO_2 via the Joule-Thomson effect is another option. Internal refrigeration processes can be competitive with TEG dehydration in some cases. The process is relatively simple with no additional rotating parts (except possibly a methanol injection pump) and has a small footprint. There are also no emissions generated from the system. However, this process does require a large pressure drop for the Joule-Thomson effect and requires subsequent re-compression of a recycle CO_2 stream. Some advanced refrigeration-type dehydration processes can also recover hydrocarbon liquids if present in significant quantity in the source CO_2.

Selected Configuration

Figure 2 shows the chosen process flow configuration for the CO_2 recovery system from the representative natural gas processing facility. The system includes two 15 MMscfd blowers in parallel to increase the pressure of the CO_2 from the H_2S removal unit from about 0 psig to 15 psig. Three 10 MMscfd reciprocating compressor trains with electric motor drives are used to compress the CO_2 from 15 psig to about 1,700 psig. A TEG dehydration unit is used between the 4th and 5th stages at about 745 psig. A single TEG unit will dry the gas from the three compression trains. TEG dehydration was selected since deep water removal was not required. Finally, two-15 MMscfd CO_2 pumps are used in parallel to increase the pressure from 1,700 psig to 2,200 psig as required by the pipeline company.

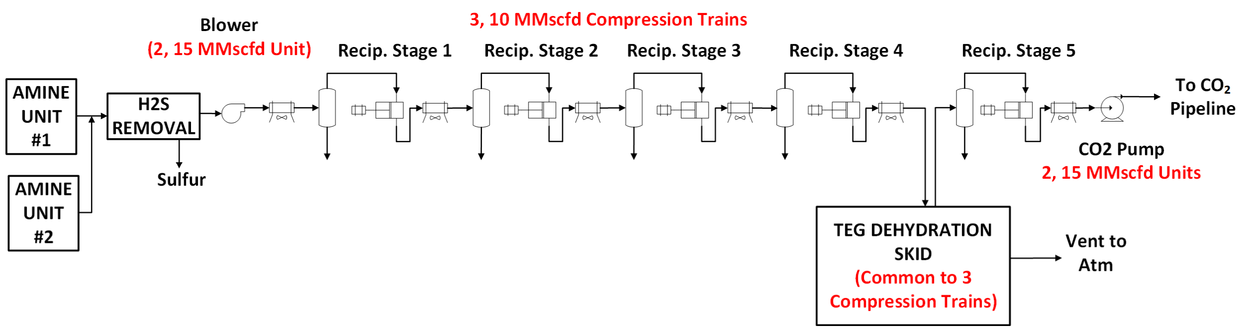

Figure 2 – Flow Scheme for Example CO_2 Recovery Application

Impact of CO_2 Service on Reciprocating Compressors and TEG Dehydration

Compression Train Design Guidelines

Reciprocating compressor technology in the oil and gas production area evolves over time along with every other technology. Many of the reciprocating compressors built for new service in the gas treating industry are “medium speed” reciprocating compressors that are designed to ISO standard 13631, which provides requirements and recommendations for packaged, skid-mounted reciprocating compressors. In the gas treating industry, medium speed reciprocating compressors are a common choice for compressor and are usually chosen over more traditional “slow speed” reciprocating compressors designed to API-618. Numerous manufacturers design medium speed compressors; some typical features of this type of reciprocating compressor include:

- Crankshaft rotating speeds at or above 600 RPM [10].
- Compressor power limited to 10,000 HP or less.
- Compressor and possibly the compressor driver are mounted on a skid along with some or possibly all interstage separators.

- Short piston stroke lengths, usually between 5-7 inches in length. For this paper, the piston stroke length is set at 7 inches.
- Up to six cylinders per compressor.
- Cylinder bore diameters less than 30 inches. For this paper, the maximum cylinder bore is set at 26.5 inches.

Medium speed compressors operate at a range of RPMs and a given frame will be rated for a certain speed, perhaps greater than 1,000 RPM. Most compressor operators do not operate their compressors at the maximum speed to maximize valve and piston ring/rider band life. Additionally, high molecular weight gases like CO_2 will limit the allowed piston speed in order for the compressor valves to operate correctly [11]. As a result, there is a tradeoff between compressor size (capital cost) and service life between required maintenance periods.
Reciprocating compressor wear part life can be reduced by several factors, some of which are not in the designer's control, but some guidelines to consider are:

- Keeping cylinder discharge temperatures low; this paper considers a maximum discharge temperature of 300 °F. Depending on the criticality of the compressor service, some operators may allow higher discharge temperatures while others may have a lower maximum discharge temperature limitation.
- Limiting the rotational speed of the compressor and the stroke length of the compressor. The rotational speed of the compressor sets the number of times the valves are actuated, while the stroke length in combination with the rotating speed of the compressor sets the piston speed.
    - For this paper, the piston speed limit is less than 850 ft/min. Depending on the criticality of the compressor service, some operators may allow a higher piston speed while others may have a lower allowed piston speed.
    - Rotational speed is set by the rated speed of the electric motor. For this paper, the rotational speed is set at a maximum of 713 RPM.

Most CO_2 capture processes, whether they be alkanolamine solvents, physical absorbents, membranes, solid adsorbents, etc. separate CO_2 from a bulk gas stream with a low concentration of CO_2 and then release the CO_2 at a much higher concentration but typically at low pressure. On the other hand, most CO_2 destinations require the CO_2 to be above the critical pressure of 1,070 psia. The CO_2 compressor ideally needs to be able to compress the CO_2 from the low feed pressure of the AGRU up to the required delivery pressure.

In the example case considered here, the feed CO_2 leaving the H_2S removal unit is available at 0.1 psig (shown in Table 1). The required delivery pressure for the CO_2 capture facility is 2,200 psig so the compressor must reach that delivery pressure while still operating within the bounds established above. Figure 3 shows a potential compression process on a Pressure-Enthalpy diagram with interstage cooling that is necessary to reach the 2,200 psig delivery pressure, and the 300 °F and 120 °F isotherms shown as dashed lines. 120 °F is the summer outlet temperature of the interstage coolers and the aftercooler.

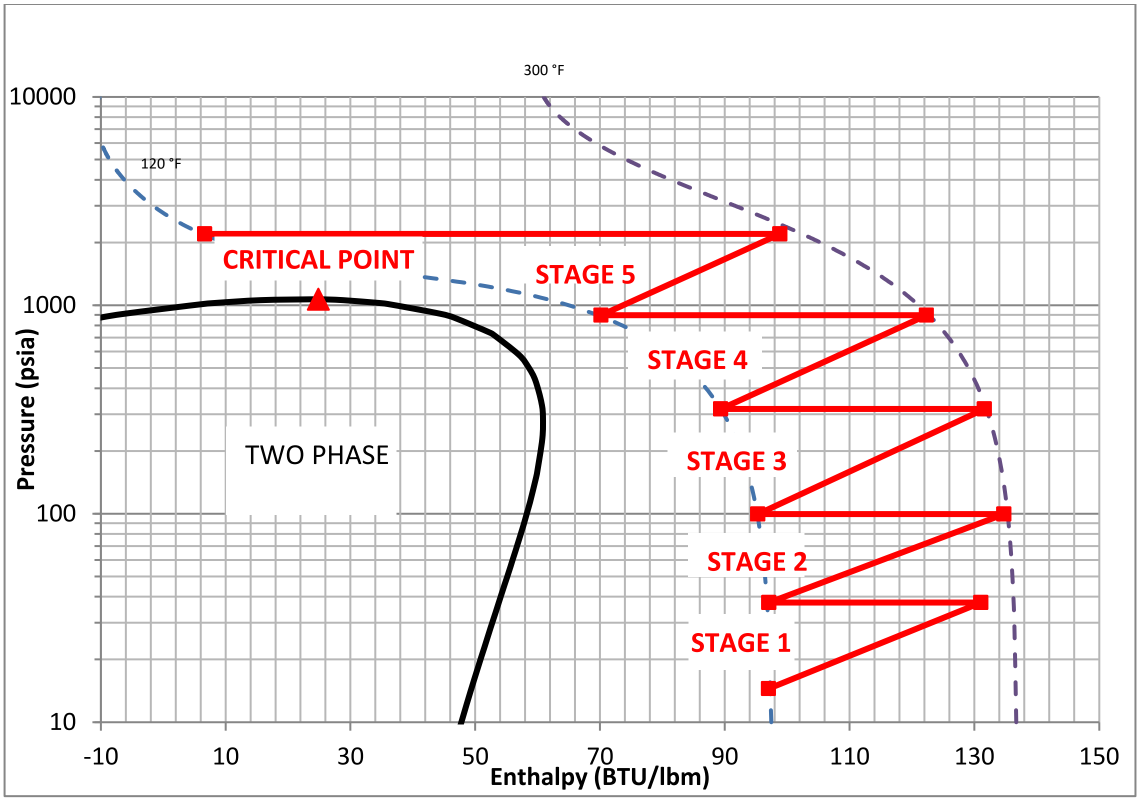

Figure 3 –Pressure-Enthalpy Diagram for CO_2 Capture Unit

The figure shows that at least five-stages of compression will be necessary to compress the CO_2 from the source conditions up to the required discharge pressure. In a six-cylinder reciprocating compressor for this service there is one "spare" cylinder that is not already assigned to a stage of compression. The capacity of a reciprocating compressor is largely determined by the volume of gas compressed in the first stage; thus, it is logical to assign the spare cylinder to the 1st stage of compression and essentially double the volume of gas that can be compressed in the 1st stage of compression (and subsequent stages). At the conditions shown in Table 1 the capacity of a single compressor is 6.1 MMSCFD so 5 compressors are required to process all the CO_2. The low capacity of a single compressor is due to the low density of 0.098 lb/ft^3 of the feed gas. If the gas density increases (at increased suction pressure), the capacity of the compressor increases substantially, as shown in Figure 4.

The installation of a multistage centrifugal blower upstream of the reciprocating compressor can increase suction pressure from 14.5 psia to as high as 34.5 psia. A centrifugal blower is a dynamic style of compressor and can efficiently compress high volumes of gas to a high enough pressure that the density of the CO_2 more than doubles to 0.23 lb/ft^3 and the capacity of a single downstream reciprocating compressor also more than doubles. In this case, one or more blowers must be installed as part of the project, but the number of compressors required to process all the CO_2 reduces to 3. The revised pressure-enthalpy diagram is shown in Figure 5. Note that in this case the fluid is cooled to a temperature warmer than 120 °F in the next to last stage of compression to keep the gas suitably distant from the critical point where physical properties are not stable.

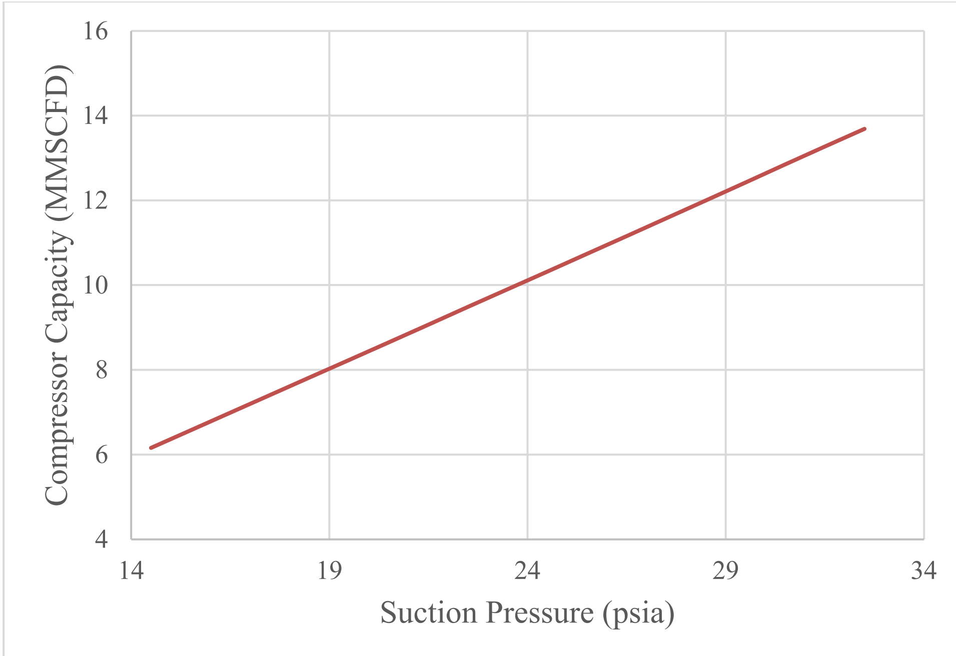

Figure 4 – Compressor Capacity as a Function of Suction Pressure

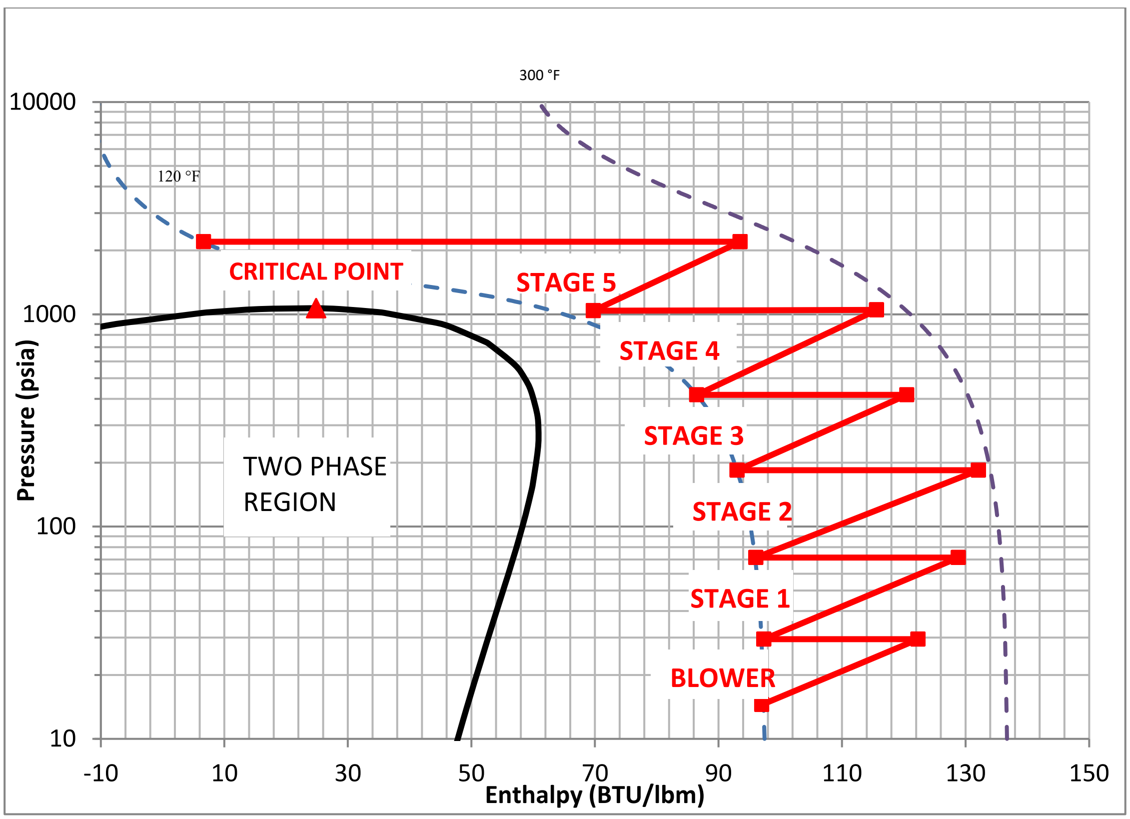

Figure 5 – Pressure-Enthalpy Diagram for CO_2 Capture Unit with Blower

There are several benefits to using centrifugal blowers as the first stage of compression when processing CO_2 from atmospheric pressure up to 2,200 psig.

1. A lower purchased equipment cost. A reciprocating compressor package for this service will cost around $4 million per machine, while each blower package will cost around $1.3 million (costs reflect approximate market conditions at the time of this paper). Table 4 shows the purchased equipment cost for a design that includes centrifugal blowers and the purchased equipment cost for a design that does not include centrifugal blowers.

Table 4 – Purchased Equipment Cost for CO_2 Compression Trains (All Costs in Million USD)

Compression Train Arrangement	No. of Centrifugal Blowers	Cost per Centrifugal Blower Package	No. of Reciprocating Compressors	Cost per Reciprocating Compressor	Purchased Equipment Cost
No Blowers	0	$1.3	5	$4	$20
Blowers	2	$1.3	3	$4	$14.6

2. Less compression ratio required for the reciprocating compressor. Figure 3 shows that the reciprocating compressors' discharge temperatures from each stage approach 300 °F. In fact, when selecting a compressor, it may be difficult to stay below 300 °F on each stage due to the fact that the compression ratio on a given stage of compression is a function of the upstream and downstream cylinders, which are only available in a finite number of sizes. Figure 5 shows that the reciprocating compressors' discharge temperatures from each stage are all below 300 °F, which should increase the lifespan of the compressor valves.
3. A lower balance of plant installation cost. By employing fewer reciprocating compressors, the balance of plant costs for piping, foundations, structural steel, etc. should all be lower. The blower is a compact centrifugal machine with only two separators and an aftercooler associated with it, so it has a much smaller footprint and associated installation cost than a multiple stage reciprocating compressor.

Temperature Management of Interstage Coolers and Aftercooler of Reciprocating Compressor

In many compression applications, the colder the gas can be entering the compressor cylinder, the better; the compressor can process more gas since the gas is denser and the discharge temperature at each stage may also be lower than it would be in a warmer condition. In CO_2 compression applications, there are situations where the CO_2 gas can be too cold. In wet CO_2 service, hydrate formation is possible when the CO_2 temperature is still well above the freezing point of water; as high as 55 °F at pressures above 500 psig, so care must be taken when ambient temperatures drop in the winter [12]. Some mitigation strategies for avoiding hydrate formation include:

- Air cooler louvers and air cooler fan speed control. Louver controls should be automated.
- Hot air recycle or air heaters in the air cooler box.
- If using cooling water, a hot water return to the basin or a basin heater may be required.

The temperature of the CO_2 also becomes important in the higher-pressure stages of the reciprocating compressor even if the gas has been dehydrated and the hydrate formation potential is mitigated due to potential condensing of the CO_2 itself. Table 5 shows the dewpoint temperature for CO_2 at pressures near or above typical dehydration pressures.

Table 5 – CO_2 Liquid Formation Temperatures as a Function of Pressure

Pressure (psia)	Temperature (°F)
600	43.9
700	55.1
800	65.1
900	74.2
1000	82.4

The critical point for CO_2 is at 1,070 psia and 87 °F. At pressures near or above 1,070 psia, the density and compressibility of CO_2 will vary widely at different temperatures. If the inlet pressure to a stage of compression is near the critical point, it may be necessary to control the inlet temperature of that stage suitably above the critical temperature to ensure stable fluid properties. One compressor manufacturer recommends staying 40-50 °F above the critical temperature, so reasonably tight control of the gas temperature in winter conditions may be required [11].

To avoid operating a compression stage near the critical point and to reduce compression ratio (and therefore discharge temperature) on the individual compression stages, another option to consider is the addition of a dense-phase pump on the discharge of the reciprocating compressor. There may be additional benefits to adding this piece of rotating equipment:

- Dense-phase pumps will have a shorter lead time than reciprocating compressors. The lead time for a reciprocating compressor may exceed 52 weeks and if the ultimate discharge pressure is uncertain, it may be desirable to get the compressor operating conditions defined so it can be ordered to meet the project schedule.
- It may be possible to reduce the number of stages in the compressor by further cooling the gas to each stage suction and thus allowing a higher compression ratio. In the example case, this is not possible due to the use of air coolers, but if cooling water was employed it might be possible to reduce the compressor stages to 4, which should reduce the capital cost of the project or allow additional gas compression capacity for each compressor.
- By reducing the compression ratio across the reciprocating compressor, discharge temperatures through the compressor should reduce, which will increase compressor valve life. The last stage of compression will occur below 800 psig, which is within the range for TEG dehydration and suitably below the critical pressure so control of the CO_2 temperature may be less stringent in winter conditions. TEG dehydration could be done at the lower stage with an inlet pressure of around 400 psig, but that would require more water removal in the TEG unit and thus a larger TEG unit. There would be some offset materials costs by using less stainless steel in the unit with the lower pressure dehydration step, but that is not considered here for simplicity.

In the example case, it is necessary to compress the CO_2 to 1,700 psig to reach the required density of 35 lb/ft^3. Figure 6 shows the pressure-enthalpy diagram for an equipment configuration that includes two blowers, three reciprocating compressors, and two dense-phase pumps.

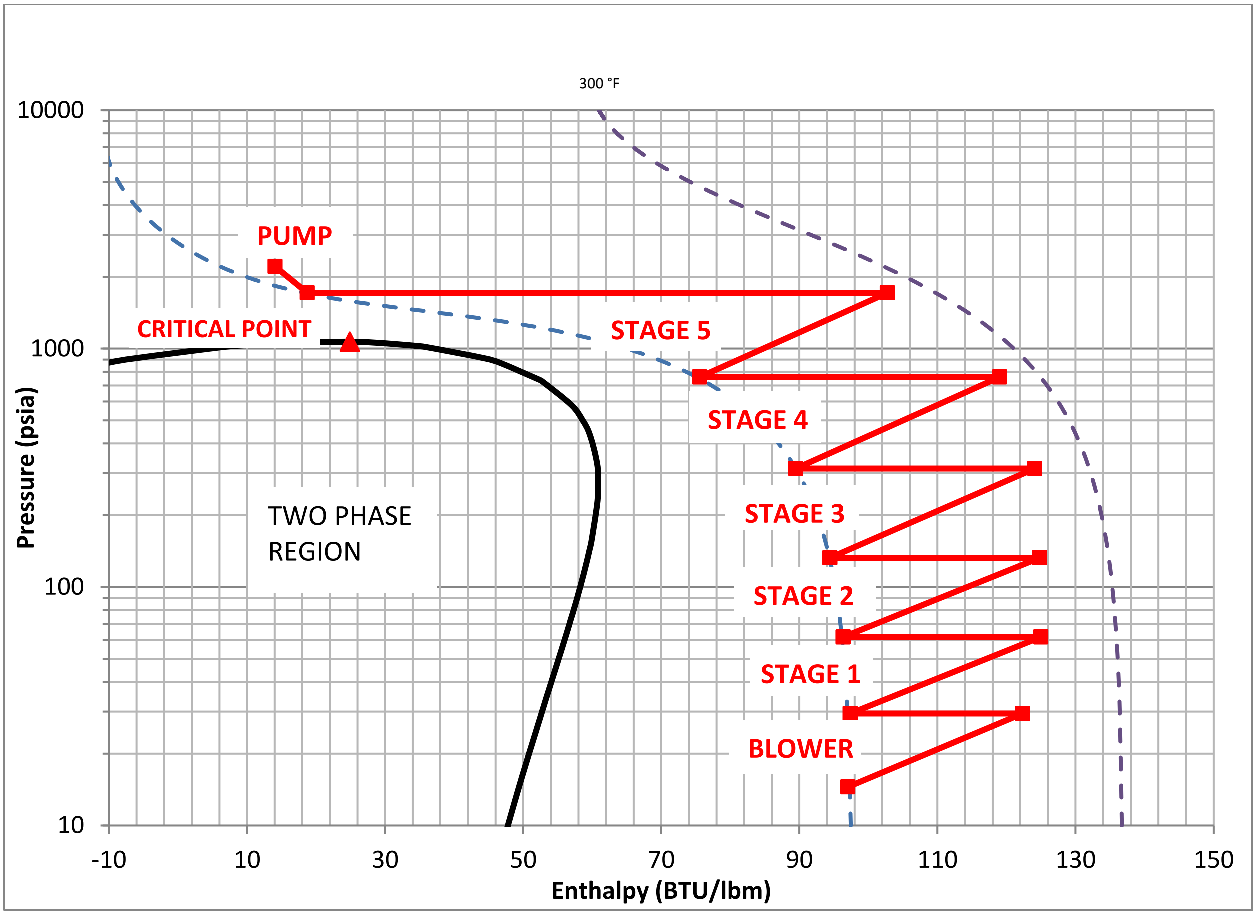

Figure 6 – Pressure-Enthalpy Diagram for Blower-Compressor-Pump CO_2 Capture Unit

Capacity Control for Compression Train and Start Up / Shut Down Concerns

Controlling the capacity of the CO_2 capture unit is an important design consideration for several reasons:

1. CO_2 capture is by nature horsepower intensive; electricity costs (or other driver costs) usually make up approximately 80% of the operating cost for the CO_2 capture unit. Minimizing these costs is a priority of the design effort.
2. The CO_2 capture unit is not the central unit operation for the natural gas treatment facility and should be designed to respond to changes in the natural gas treatment facility, namely swings in CO_2 feed rates from the AGRU. In many natural gas treatment plants, the actual CO_2 fraction in the incoming gas is lower than the specification so when designing the CO_2 capture unit, it is important to consider what the design feed rate of CO_2 could be and compare that to what the design CO_2 feed rate actually is in day-to-day operation. This difference may lead to some over-sizing of the CO_2 capture unit that then needs to be efficiently managed by the capacity control system.
3. The CO_2 capture unit will typically become the pressure control device for the regenerator, whether the AGRU is an amine, membrane, physical solvent, etc. instead of a simple and

reliable back pressure control valve typically employed. As a result, accurate and reliable capacity control for the CO_2 capture unit is critical to the operation of the AGRU.

For the CO_2 capture unit to be able to process all CO_2 in all conditions and for the AGRU to be able to continue to operate when the CO_2 capture unit is offline, the following control devices may be considered:

1. Control valves to vent CO_2 when part or all the CO_2 capture unit goes offline. In some locations, venting of CO_2 may not be allowed and instead may be routed through a pollution abatement device, such as a thermal oxidizer. In this instance, venting may need to be kept to a minimum.
2. Control valves to recycle part or all the CO_2 to balance the CO_2 capture unit when the upstream AGRU is not producing as much CO_2 as the design.
3. Variable frequency drives (VFDs) to control the speed of rotating equipment motors to modulate the amount of CO_2 processed by the CO_2 capture unit.
4. Automated or manual unloader devices on the reciprocating compressors. These can be on the head end of one or more compressor cylinder or located over the suction valves of each compressor cylinder.

CO_2 vent valves may be provided between major rotating equipment, but care must be taken to avoid venting cold, high-pressure CO_2 that could form dry ice in piping or exceed the minimum design metal temperature of the vent header. At a minimum, an automated, full-flow vent control valve should be located at the interface point between the AGRU and the inlet to the CO_2 capture unit so that the AGRU can continue to operate and vent CO_2 to atmosphere or the appropriate location if the CO_2 capture unit is offline.

The centrifugal blowers are dynamic machines, so some form of surge control is necessary to protect the machine. This can be as simple as a full flow recycle control valve or a more complicated surge control system. In some instances, a control valve on the inlet of the blower can be used to reduce the blower capacity more efficiently, but this may introduce a vacuum condition on the inlet of the blower and the potential for oxygen ingress. VFDs may be employed as well, but the speed reduction possible with a VFD on these centrifugal blowers may be very small and not justify the expense of a medium voltage VFD (each blower in this example is appx. 945 BHP). However, they have been used successfully in the past. Figure 7 shows the potential control scheme around the AGRU and CO_2 capture interface along with the centrifugal blower controls.

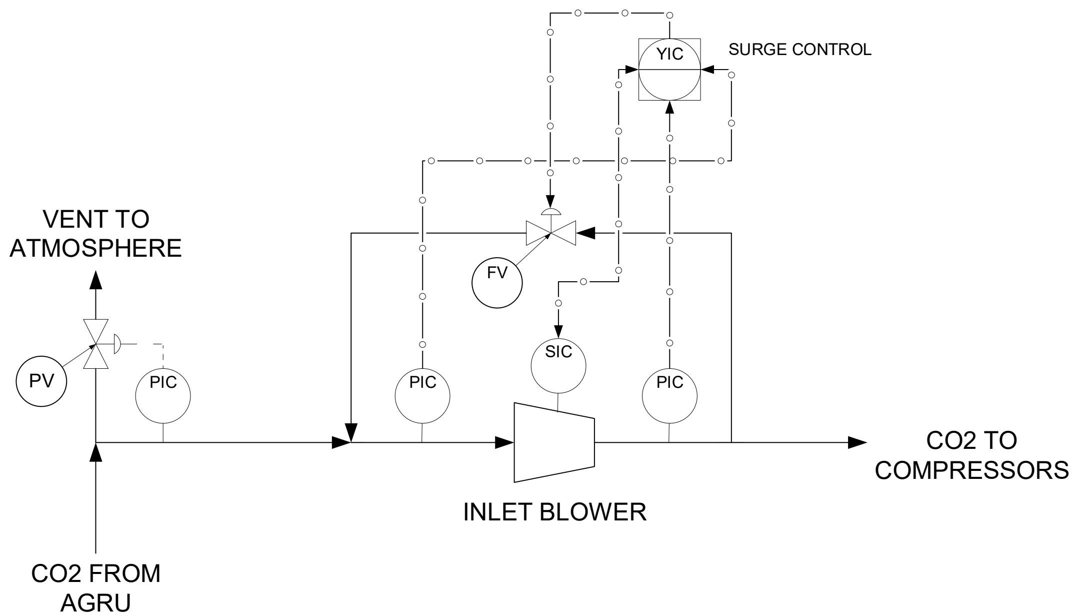

Figure 7 – Typical Pressure Control Scheme for AGRU Interface and Blower

The CO_2 compressors themselves can be controlled in several ways, as noted above. While one or more recycle valves around compressors are very common in natural gas or other gases, the isenthalpic expansion of CO_2 from the cool stage 5 discharge conditions to almost any other pressure will result in a 2-phase mixture of either liquid and gas CO_2 or even solid and gas CO_2 if the final pressure of the expansion is below the triple point of CO_2. To address this concern, the following techniques have been used:

1. Recycle hot, high-pressure CO_2 from the discharge of the compressor upstream of the compressor aftercooler. The expanded CO_2 is likely too warm for continuous recycle (the compressor will shut down on high suction or discharge temperature after a brief period), so recycle coolers can be used. Avoiding wet CO_2 ingress to this recycle cooler is an important design consideration if the tubes are not made of stainless steel to avoid potential corrosion. To limit the size and cost of this recycle system, a station recycle valve set up capable of recycling a portion (the capacity of one compressor for instance) can be installed and if recycle needs exceed the capacity of the recycle system, one or more compressors can be shut down. In this case, the compressors should be able to operate indefinitely recycling some or all the CO_2.
2. Recycle a mixture of hot and cold high-pressure CO_2 from the discharge of the compressor and the discharge of the aftercooler. This requires additional control valves and some temperature control but has been used successfully in some projects. In this case, the compressors should be able to operate indefinitely recycling some or all the CO_2.
3. Do not install continuous modulating recycle valves and only have on/off recycle valves for startup and shut down of individual CO_2 compressors. In this case, there is no normal recycling of gas around the compressors. This option must be carefully considered against the potential for unintended shutdowns of the CO_2 compressors due to pressure imbalances and the resultant frequent venting of the CO_2 if CO_2 rates from the AGRU drop below the minimum rate for compressor operation. However, if the AGRU's operating history shows constant and predictable rates of CO_2, manual variable volume clearance pockets (VVCPs) can be set so the compressor processes the appropriate amount of CO_2. There will also be

a "natural" capacity control that takes place according to the compressor capacity as a function of suction pressure as noted above in Figure 4. As the upstream blower throttles or recycles to maintain pressure in the AGRU, the suction pressure to the compressors will decrease, which reduces how much CO_2 the compressor processes. The suction pressure to the compressors will decrease until the amount of CO_2 the compressors process equals the CO_2 processed by the blower or the compressors shut down, usually on low suction pressure related to rod load issues in the compressor. The extent of this turndown capability can be determined during compressor selection and can be prioritized if this control scheme is utilized but in no cases will it allow infinite turndown like recycle valves.

Other control devices for reciprocating compressors include VFDs and pneumatic unloaders. VFDs may be an attractive option if power costs are high, but additional acoustic pulsation studies will be required to determine what speeds are possible with the compressor design. Failure to consider all possible operating conditions at different speeds may result in substantial shaking forces in and around the compressor skids, leading to unplanned downtime and high maintenance costs.

Pneumatic unloaders can be installed on head-end pockets of the compressor cylinders or over suction valves on the compressor cylinder. This can allow for partial or complete "deactivation" of the head end of the cylinder, which reduces the compressor capacity in discrete steps down to as low as 50% of the design flow. Like VFDs, how the compressor operates at each of these discrete load steps needs to be considered since the compression ratios across each stage will change and subsequent rod loads and discharge temperatures needs to be considered.

There are many options around capacity control of the reciprocating compressors and some options may be more attractive or familiar to different facilities than other options. For example, some rural power grids may require the large CO_2 compressors to utilize a VFD to start the compressor to avoid large current draws on the power grid. In this instance, it may be attractive to plan to use the VFD as a normal capacity control device since it needs to be purchased and installed in any event. In some cases, multiple compressors can be started with the same VFD and then transferred to operate at the normal fixed speed. In this case, a single compressor speed could be modulated to control capacity while the other compressors operate at a constant speed. This may save money on the cost of medium voltage VFDs.

The dense-phase pump is a small piece of rotating equipment that will have substantial reductions in speed possible if a VFD is used. In many cases, a low voltage (480 VAC) motor is selected for these pumps and the VFD that can be used to control the motor speed will be a low cost. Controlling the suction pressure to the dense-phase pump by modulating the speed of the pump is a typical solution and when combined with the "natural" capacity control method for the reciprocating compressors results in a complete facility pressure control scheme shown in Figure 8. The additional vent valve between the compressors and the dense-phase pump allows for controlled start up and shut down while the back pressure control valve on the dense-phase pump allows the pump to stay within operating bounds when the pipeline pressure is low.

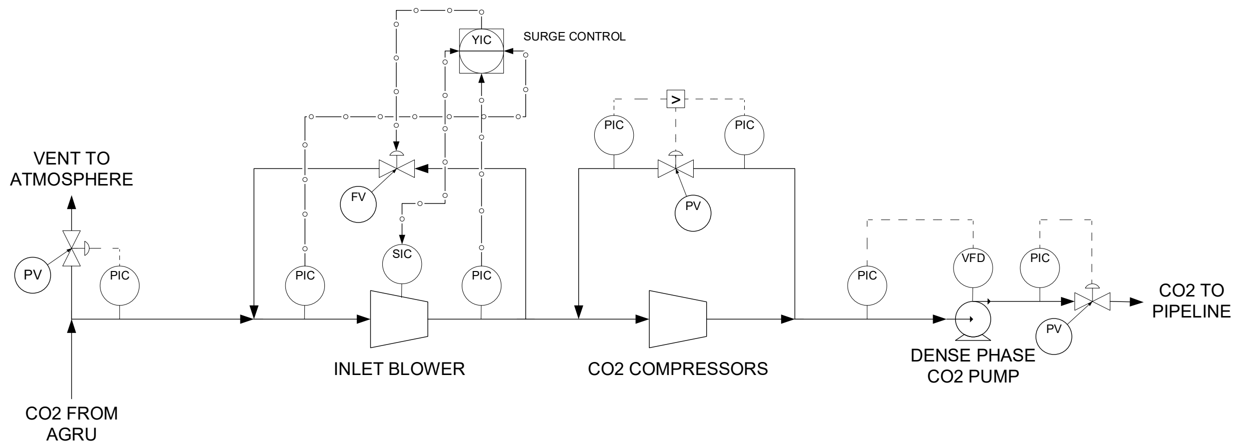

Figure 8 – Overall CO_2 Capture Unit Pressure Control Scheme

Materials of Construction for Compression Train and TEG Dehydration Unit

CO_2 in the presence of liquid water will dissolve into the water to form carbonic acid, which is corrosive to carbon steel. As the partial pressure of CO_2 increases over the liquid water phase, the amount of CO_2 dissolved in the water increases, which accelerates corrosion of carbon steel even further. To minimize the chances for corrosion in the CO_2 capture facility, areas of the plant where a liquid water phase could reasonably be expected to form during normal operation are made of either stainless steel or coated carbon steel. Some designs consider piping and vessels that operate more than 100 °F above the dewpoint as suitably hot to use carbon steel even though the CO_2 is not dehydrated. In practice, this means that the following areas of the compression train can be constructed of carbon steel:

- Discharge pulsation dampeners and discharge piping from each stage of the reciprocating compressors.
- All equipment and piping downstream of the dehydration unit.

The reciprocating compressor cylinders are not constructed of alloy materials but some upgrades to the compressor are recommended, such as:

- Stainless steel piston rods
- Stainless steel compressor valves

The remainder of the compression train should be constructed of stainless steel or coated carbon steel. When the compression train shuts down for maintenance or due to a process upset, the compressor should be blown down and purged before the compressor cylinders cool off to ambient temperature to prevent any chance of liquid water formation on the compressor cylinders or in other carbon steel vessels or piping. Corrosion in the compressor cylinder body, particularly in the actual cylinder bore will quickly render the compressor inoperable and potentially require a complete cylinder replacement.

The centrifugal blower package should be constructed of all stainless steel or coated. In many instances, the piping, inlet/outlet separator, and aftercooler are constructed of stainless steel while the blower impellers are coated with a baked phenolic material.

Similar concerns exist within the TEG dehydration unit; the presence of CO_2 and water in the rich TEG streams lead to selecting stainless steel as the material of construction for parts of the TEG dehydration unit. Figure 9 shows a typical TEG dehydration unit flow diagram. Equipment and piping denoted with dashed lines can be constructed out of carbon steel while those depicted with solid lines represent equipment and lines that should be constructed out of stainless steel.

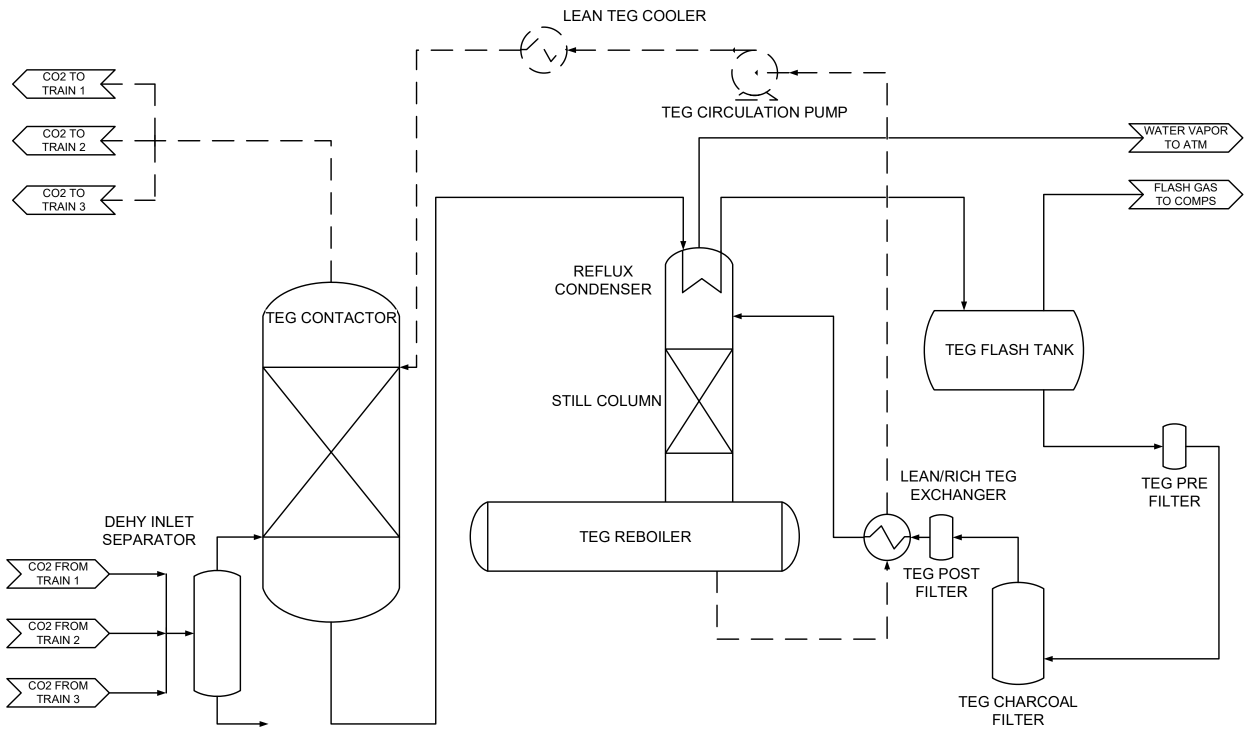

Figure 9 – Material Selection Diagram for TEG Dehydration Unit in CO_2 Service

Conclusions

CO_2 capture units are still not common operating systems in many natural gas treating facilities. Increasing government incentives and the public's perception of the overall benefits of carbon capture and sequestration projects are making it more likely that CO_2 capture units will begin to be installed in new gas treating facilities and existing gas treating facilities will be retrofitted to capture at least some of their CO_2 emissions.

The current incentive regime around capturing CO_2 can make high purity CO_2 emission sources, such as the AGRU vent stream attractive to capture and sequester. Other low purity CO_2 emission sources require additional processing and may still be too expensive to capture economically.

Most acid gas removal technologies release CO_2 at a low pressure and, in general, the CO_2 must be compressed up to the dense phase to be injected for sequestration or to flow in common CO_2 pipelines. A blower-reciprocating compressor-dense phase pump arrangement can be an optimal arrangement for the CO_2 capture unit by combining the different technologies to achieve the high compression ratio and relatively high gas volume processing requirements.

References

[1] "Congressional Research Service," 25 August 2023. [Online]. Available: https://crsreports.congress.gov/product/pdf/IF/IF11455. [Accessed 10 October 2023].

[2] Echt, B., D. Leppin, D. Mamrosh, D. Mirdadian, D. Seeger, B. Warren, "Fundamentals of Low-Tonnage Sulfur Removal and Recovery," in *Laurance Reid Gas Conditioning Conference*, Norman, Oklahoma, February 26 - March 1, 2017.

[3] Klewicki, J. Kenneth, B. Kobelski, A. Karimjee, and C.H. Sham, "Acid Gas Injection in the United States," in *5th Annual Conference on Carbon Capture and Sequestration - DOE/NETL*, May 8-11, 2006, 2006.

[4] Energy Equipment and Infrastructure Alliance, "Meeting the Dual Challenge: A Roadmap to At-Scale Deployment of Carbon Capture, Use, and Storage - Key Excerpts from National Petroleum Council Report," [Online]. Available: https://www.eeia.org/post/CCUS-Pipeline-Transport-Meeting-the-Dual-Challenge.pdf.

[5] DOE-NETL, "Quality Guidelines for Energy System Studies CO2 Impurity Design Parameters," DOE/NETL-341/011212, January 2012.

[6] Piggott, B., D. Mamrosh, A. Ryan, R. McKaskle, and D. Myers, "Overview of Natural Gas Compression Options in Upstream and Midstream Operations," in *Laurance Reid Gas Conditioning Conference*, Norman, OK, 22-25 February 2021.

[7] "High Pressure Screw Compressor Unit GH-Series," Mayekawa, [Online]. Available: http://www.mayekawa.eu/en/products/mycom/special-products/high-pressure-screw-compressor-unit/gh-series. [Accessed 12 October 2023].

[8] Carmody, P., "Design Tips for Dehydrating CO2 with Glycol," in *Laurance Reid Gas Conditioning Conference*, Norman, OK, 2022.

[9] Fisher, Kevin; J. Lundeen, C. Beitler, C. Wallace, R. Holt, D. Purvis, "Use of Glycerol (Glycerin) to dehydrate CO2 above the Critical Pressure at Denbury's Mississippi Facilities," in *10th Annual CO2 Flooding Conference*, Midland, TX, 2004.

[10] Gas Processors Suppliers Association, Engineering Data Book, Tulsa, OK: Gas Processors Association, 2012.

[11] Ariel Corporation, "Application Manual," Ariel Corporation, [Online]. Available: https://www.arielcorp.com/support/application-manual/applications/carbon_dioxide_mixture.html. [Accessed 26 October 2023].

[12] Carroll, J., D. Palfreyman, J. Johnson, "Fundamentals of Acid Gas Injection," in *Laurance Reid Gas Conditioning Conference*, Norman, Oklahoma, February 27 - March 2, 2005.