Plains CO_2 Reduction (PCOR) Partnership
Energy & Environmental Research Center (EERC)

OPPORTUNITIES AND CHALLENGES ASSOCIATED WITH CO_2 COMPRESSION AND TRANSPORTATION DURING CCS ACTIVITIES

Plains CO_2 Reduction Partnership Phase III
Task 6 – Deliverable D85

Prepared for:

Andrea M. Dunn
National Energy Technology Laboratory
U.S. Department of Energy
626 Cochrans Mill Road
PO Box 10940
Pittsburgh, PA 15236-0940

Cooperative Agreement No. DE-FC26-05NT42592

Prepared by:

Melanie D. Jensen
Charles D. Gorecki
Edward N. Steadman
John A. Harju

Energy & Environmental Research Center
University of North Dakota
15 North 23rd Street, Stop 9018
Grand Forks, ND 58202-9018

May 2015
Approved

2015-EERC-06-08

EERC DISCLAIMER

ACKNOWLEDGMENT

This material is based upon work supported by DOE National Energy Technology Laboratory under Award No. DE-FC26-05NT42592.

DOE DISCLAIMER

NDIC DISCLAIMER

Plains CO_2 Reduction (PCOR) Partnership
Energy & Environmental Research Center (EERC)

OPPORTUNITIES AND CHALLENGES ASSOCIATED WITH CO_2 COMPRESSION AND TRANSPORTATION DURING CCS ACTIVITIES

ABSTRACT

Carbon capture and storage (CCS) holds the potential to reduce carbon dioxide (CO_2) emissions from large stationary sources. The majority of CCS research to date has focused on the capture, injection, and subsequent monitoring of the CO_2 in the geologic formation, but efficient incorporation of compression into an integrated system may offer opportunities to reduce the cost of CCS, which could help to advance widespread implementation of the concept. The CO_2 is transported as a supercritical fluid in pipelines during CCS activities. Because the CO_2 stream exiting all CO_2 capture technologies is in the gas phase, compression is required prior to pipeline transport. The choice of compression approach is based upon the power demands and investment cost. A liquefaction approach has not been proven to be more efficient or cost-effective than traditional gas compression techniques, although the shock wave-based Dresser-Rand SuperCompressor shows promise, especially for postcombustion capture.

Compression plays an important role in overall CO_2 capture plant efficiency. Selection of an appropriate compression approach for the quantity of CO_2, desired pipeline pressure, and type of capture technology is crucial. The best plant efficiency and capture economics will be achieved by integrating the capture technology, dehydration step, compression approach, and integration of the compressor waste heat into the overall capture plant. Effective optimization will require that these steps be determined iteratively.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

NOMENCLATURE AND ABBREVIATIONS LIST

π	pi, approximately equal to 3.14159
ε	roughness height of the inner surface of the pipe
ΔD	relative difference
ΔP	change in pressure
°C	degrees Celsius
°F	degrees Fahrenheit
ASU	air separation unit
BHP	brake horsepower
CCS	carbon capture and storage
CF	capacity factor
cm	centimeter
CO_2	carbon dioxide
D	diameter
D_{cur}	diameter initial guess during iterative calculation
D_{new}	new value for the diameter during iterative calculation
DOE	U.S. Department of Energy
EOR	enhanced oil recovery
f_D	Darcy or Moody friction factor
FE	Fossil Energy
f_f	Fanning friction factor
ft	feet
g	acceleration due to gravity
h	hour
HP	high pressure
h_{in}	elevation at the inlet of a pipe segment above a reference elevation
hp	horsepower
H_2S	hydrogen sulfide
h_{out}	elevation at the outlet of a pipe seqment above a reference elevation
ID	inner diameter
IEAGHG	IEA Greenhouse Gas R&D Programme
in.	inch
K	Kelvin
kg	kilogram
kJ	kilojoule
km	kilometer
kPa	kilopascal
L	length
m	meter
M	molecular weight
m^3	cubic meters
mi	mile

Continued. . .

NOMENCLATURE AND ABBREVIATIONS LIST (continued)

MIT	Massachusetts Institute of Technology
MMcfh	million cubic feet per hour
MMscfd	million standard cubic feet per day (at oil and gas standard conditions of 60°F and 1 atmosphere)
mol	mole
MPa	megapascal
Mt	million tonnes
MW	megawatt
N_2	nitrogen
NA	not applicable
NETL	National Energy Technology Laboratory
Pa	pascal
Pa-s	pascal-second
P_{ave}	average pressure
ρ_{CO_2}	density of CO_2
P_{in}	pressure at the inlet of a pipe segment
P_{out}	pressure at the outlet of a pipe segment
psi	pounds force per square inch
psia	pounds force per square inch absolute
q_{max}	maximum mass flow rate
q_{av}	average annual mass flow rate
R	universal gas constant
Re	Reynolds number
s	second
SRI	Southwest Research Institute
T_{ave}	average temperature
TBD	to be determined
T_c	critical temperature
TEG	triethylene glycol
μ	viscosity
yr	year
Z_{ave}	compressibility factor for CO_2

Plains CO_2 Reduction (PCOR) Partnership
Energy & Environmental Research Center (EERC)

OPPORTUNITIES AND CHALLENGES ASSOCIATED WITH CO_2 COMPRESSION AND TRANSPORTATION DURING CCS ACTIVITIES

EXECUTIVE SUMMARY

Carbon capture and storage (CCS) holds the potential to reduce carbon dioxide (CO_2) emissions from large stationary sources, such as power plants and industrial facilities, thereby helping to achieve national and international CO_2 reduction goals. Although the majority of the research on CCS to date has focused on the capture, injection, and subsequent monitoring of the CO_2, efficient incorporation of compression into an integrated system may offer opportunities to reduce the cost of CCS, which could help to advance widespread implementation of the concept. This report provides basic information about CO_2 transport and compression and discusses some of the opportunities offered by thoughtful integration of them into a total CCS system.

CO_2 can be transported as a gas, a liquid, or a solid, although commercial-scale transport of CO_2 is usually accomplished as either a gas or liquid in tanks, pipelines, or ships. As a gas, CO_2 occupies less volume if it is compressed, so when commercial quantities are transported by pipeline, the CO_2 is compressed, generally to a high pressure. The volume occupied by the CO_2 can be further reduced by compressing the CO_2 to its supercritical state (over 7.4 MPa, or 1080 psi) or liquefying it.

During enhanced oil recovery (EOR) using CO_2 or CCS activities, the CO_2 is transported by pipeline at pressures exceeding 7.4 MPa (1080 psi). This approach is based on the quantity of CO_2 that must be transported, the diameter of the pipeline required for transport of that quantity, the cost of the compressors needed to achieve the transport pressure, the cost of any pressure booster stations required along the pipeline route, and the pressure requirements at the injection site.

Pipeline diameter is calculated as a function of allowable pressure drop per unit length, frictional resistance, CO_2 density, and CO_2 mass flow rate. A rigorous, iterative approach is used for more accurate calculations, although correlations between pipeline diameter and CO_2 flow rates can be used for estimates. The rigorous calculations show that supercritical CO_2 can be transported in a smaller and therefore less expensive pipeline than if the CO_2 remains in the gas phase. This approach also requires fewer recompression stations.

The CO_2 stream exiting all CO_2 capture technologies will be in the gas phase; therefore, compression is required prior to pipeline transport. Three approaches can be taken to compress CO_2 for transport in a pipeline: 1) a near-adiabatic method in which heat is neither gained nor lost

by the system; 2) a second approach in which the gas-phase CO_2 is compressed in stages and cooled until the conditions are above the critical point, at which time the CO_2 is cooled to form a supercritical fluid that is pumped to the final pressure; and 3) a third method that utilizes some of the compression stages, then cools the CO_2 to form a liquid, which is pumped to the desired final pressure. The choice of compression approach for a given capture system is based upon the power demands and investment cost.

The underlying premise of the liquefaction approach is that significantly less power is required to raise pressure by liquid pumps and that the pumps are considerably less expensive than gas compressors. However, it is crucial that the refrigeration process be carefully assessed when determining the system power requirements. Two power loads must be considered for the liquefaction option: the refrigeration compressor and the cryogenic pump. Some studies have found that the liquefaction approach does not result in a more efficient or lower-cost system.

Compression is an important piece of the overall CO_2 capture plant efficiency. Selection of an appropriate compression technology for the quantity of CO_2, desired pipeline pressure, and type of capture technology is crucial. For example, centrifugal compression appears to be the most appropriate for all three capture platforms (pre-, oxy-, and postcombustion). The shock wave compression offered by the Dresser-Rand SuperCompressor is well-suited to postcombustion but not to oxycombustion. Placement of the dehydration step within the compressor train affects integration of the heat produced during compression as well as compressor design. Optimization of compression within a plant requires integration of the heat of compression so as to maximize plant efficiency. The Dresser-Rand SuperCompressor, for example, offers the opportunity for significant waste heat recovery. The best plant efficiency and capture economics will be achieved by integrating the capture technology, dehydration, compression approach, and heat integration of the compressor waste heat into the overall plant. Effective optimization will require that heat integration, dehydration design, and compressor selection be determined iteratively.

Further studies of the effects of various dehydration schemes on compression could be of value when determining the best approaches to efficiently and cost-effectively integrate the entire CO_2 capture system into a power plant or industrial facility. Additional studies of the integration of the SuperCompressor into a capture facility are also recommended as the SuperCompressor is sufficiently different from other compressor technologies as to require a fresh examination of how heat integration and dehydration could be most effectively applied.

Plains CO_2 Reduction (PCOR) Partnership
Energy & Environmental Research Center (EERC)

OPPORTUNITIES AND CHALLENGES ASSOCIATED WITH CO_2 COMPRESSION AND TRANSPORTATION DURING CCS ACTIVITIES

INTRODUCTION

Carbon capture and storage (CCS) holds the potential to reduce carbon dioxide (CO_2) emissions from large stationary sources, such as power plants and industrial facilities, thereby helping to achieve national and international CO_2 reduction goals. CCS is essentially a four-step process: capture from a large stationary facility, compression, transport (most likely via pipeline), and injection of the CO_2 into a secure geologic formation for permanent storage. Technologies exist for all of the CCS steps, but they have only recently been integrated into a single large-scale CCS project at the Boundary Dam power plant in Canada. Although, the majority of the research on CCS to date has focused on the capture, injection, and subsequent monitoring of the CO_2, efficient incorporation of compression into an integrated system may offer opportunities to reduce the cost of CCS, which could help to advance widespread implementation of the concept. This report provides basic information about CO_2 transport and compression and discusses some of the opportunities offered by thoughtful integration of them into a total CCS system.

APPROACHES TO CO_2 TRANSPORT

CO_2 can be transported as a gas, a liquid, or a solid, although commercial-scale transport of CO_2 is usually accomplished as either a gas or liquid in tanks, pipelines, or ships (Doctor and others, 2005). As a gas, CO_2 occupies less volume if it is compressed, so when commercial quantities are transported by pipeline, the CO_2 is compressed, generally to a high pressure (Doctor and others, 2005). The volume occupied by the CO_2 can be further reduced by compressing the CO_2 to its supercritical state (over 7.4 MPa, or 1080 psi) or liquefying it.

CO_2 is compressed to enable more efficient transport within a pipeline. Depending on the pipe diameter, mass, CO_2 flow rate, and pipe roughness factor, there is typically a frictional loss of pressure of about 4–50 kPa/km (1.0–11.8 psi/mi) (Wong, 2005). Generally, larger-diameter pipelines have lower frictional losses (Wong, 2005). Maintenance of the CO_2 in the dense phase for the length of the pipeline requires that the pressure at the pipeline inlet be sufficiently high as to overcome all of the losses along the pipeline length while still maintaining a pressure of at least 7.46 MPa and a temperature of at least 31°C (1080 psi and 88°F), the critical point at which CO_2 becomes a supercritical fluid. Alternatively, booster stations can be installed along the pipeline route every 100–150 km (62–93 mi) to make up the pressure losses. Industry preference is to operate the compressor at the pipeline inlet so that the CO_2 stream is at a pressure of at least 10.3 MPa (1494 psi) to ensure that the CO_2 remains supercritical throughout the length of the pipeline (Wong, 2005).

During enhanced oil recovery (EOR) using CO_2 and other CCS-related activities, the CO_2 is transported by pipeline at pressures exceeding 7.4 MPa (1080 psi). This approach is based on the quantity of CO_2 that must be transported, the diameter of the pipeline required for transport of that quantity, the cost of the compressors needed to achieve the transport pressure, the cost of any pressure booster stations required along the pipeline route, and the pressure requirements at the injection site.

Pipeline diameter is calculated as a function of allowable pressure drop per unit length, frictional resistance, CO_2 density, and CO_2 mass flow rate. A rigorous, iterative approach is used for more accurate calculations, although correlations between pipeline diameter and CO_2 flow rates can be used for estimates. Table 1 shows this type of estimation, as made by the Massachusetts Institute of Technology (MIT), for CO_2 at 25°C and 2292 psi (Carbon Capture and Sequestration Technologies Program, 2009).

Table 1. Estimated CO_2 Pipeline Design Capacity

	CO_2 Flow Rate			
	Lower Bound		Upper Bound	
Pipeline Diameter, in.	Mt[1]/yr	MMscfd	Mt/yr	MMscfd
4			0.19	10
6	0.19	10	0.54	28
8	0.54	28	1.13	59
12	1.13	59	3.25	169
16	3.25	169	6.86	357
20	6.86	357	12.26	639
24	12.26	639	19.69	1025
30	19.69	1025	35.16	1831
36	35.16	1831	56.46	2945

[1] Million tonnes.

The FE/NETL (Fossil Energy/National Energy Technology Laboratory) CO_2 Transport Cost Model (2014) provides three rigorous equations that can be used to calculate the minimum inside diameter of the pipeline (Morgan and others, 2014). These equations are defined in the text that follows. Two of the equations are very similar. McCollum and Ogden (2006), Heddle and others (2003), and Massachusetts Institute of Technology (2009) provided the following equation for the inner diameter:

$$ID = \left\{ \frac{32 \times f_f \times q_{max}^2}{\pi^2 \times \rho_{CO_2} \times \left(\frac{\Delta P}{L}\right)} \right\}^{0.2} \qquad \text{[Eq.1]}$$

where:

ID = inner diameter of pipe (m)

q_{max} = maximum mass flow rate of CO_2 in pipe (kg/s)

f_f = Fanning friction factor (dimensionless)

ρ_{CO_2} = density of CO_2 (kg/m^3)

ΔP = change in pressure along pipe segment (Pa)

L = length of pipe segment (m)

To solve Equation 1, the user of the model specifies the maximum mass flow rate, the maximum allowable pressure drop in a pipe segment, and the length of the pipe segment. The maximum mass flow rate depends on the capacity factor for the pipeline.

$$q_{max} = \frac{q_{av}}{CF} \qquad \text{[Eq. 2]}$$

where:

q_{av} = annual average mass flow rate of CO_2 in pipe (kg/s)

CF = capacity factor of the pipeline (dimensionless), assumed to be 0.80 for this analysis

The pressure drop is the pressure lost because of friction plus the pressure lost or gained by an increase or decrease in elevation along the pipe segment.

$$\Delta P = (P_{in} - P_{out}) - (h_{out} - h_{in}) \times \rho_{CO_2} \times g \qquad \text{[Eq. 3]}$$

where:

P_{in} = pressure at the inlet of the pipe segment (Pa)

P_{out} = pressure at the outlet of the pipe segment (Pa)

h_{in} = elevation of the inlet of the pipe segment above a reference elevation (m)

h_{out} = elevation of the outlet of the pipe segment above a reference elevation (m)

g = acceleration due to gravity (= 9.80665 m/s^2)

The Fanning friction factor is a dimensionless quantity that is defined as one quarter of the Darcy or Moody friction factor. Although the Darcy friction factor must be determined empirically, there are a number of correlation equations for determination of the Darcy friction factor as a function of the Reynolds number, the inside diameter of the pipe, and the roughness of the inner surface of the pipe. The U.S. Department of Energy (DOE) pipeline model uses the Colebrook equation to estimate the Darcy friction factor:

$$\frac{1}{\sqrt{f_D}} = -2 \times \log_{10}\left(\frac{\left(\frac{\varepsilon}{D}\right)}{3.7} + \frac{2.51}{Re\sqrt{f_D}}\right) \quad \text{[Eq. 4]}$$

where:

ε = roughness height of the inner surface of the pipe (m)

Re = Reynolds number (dimensionless)

f_D = Darcy or Moody friction factor (dimensionless)

The Reynolds number is a dimensionless quantity defined by the following equation for flow in a circular pipe:

$$Re = \frac{4 \times q_{max}}{\pi \times \mu \times D} \quad \text{[Eq. 5]}$$

where:

μ = viscosity of CO_2 in the pipe (Pa-s)

Equations 1, 3, and 4 are interdependent: Equation 1 (for diameter D) depends on the Fanning friction factor (f_f), which depends on diameter D and the Reynolds number (Re). The Reynolds number also depends on diameter D (see Equation 5). Therefore, to determine the pipe diameter, an iterative procedure is required. The following procedure is used by the DOE pipeline model:

- Step 1: Provide an initial guess for the diameter: D_{cur}.
- Step 2: Calculate the Reynolds number using D_{cur} in Equation 5.
- Step 3: Calculate f_D using Equation 4. Equation 4 is an implicit equation and is solved using the Newton–Raphson method.
- Step 4: Calculate a new value for the diameter, D_{new}, using Equation 1.
- Step 5: Calculate the relative difference between the two estimates for the diameter as follows:

$$\Delta D = \text{abs}\left(\frac{D_{new} - D_{cur}}{D_{new}}\right) \qquad \text{[Eq. 6]}$$

- Step 6: The two values are considered to have converged if the relative difference (ΔD) is less than 10−6. At that point, D_{new} is considered to be the minimum inner diameter needed for the pipeline. If the relative difference ΔD is greater than or equal to 10−6, then D_{cur} is set equal to D_{new}, and the procedure returns to Step 2.

McCoy and Rubin (2008) utilized a similar procedure, although they began with an energy balance on the pipe segment and developed Equation 7 for the inner diameter of the pipe. McCoy and Rubin (2008) indicated that their derivation was adapted from that provided in Mohitpour and others (2003).

$$D = \left\{\frac{-64 \times Z_{ave}^2 \times R^2 \times T_{ave}^2 \times f_f \times q_{max}^2 \times L}{\pi^2 \times \left(M \times Z_{ave} \times R \times T_{ave} \times \left[P_{out}^2 - P_{in}^2\right] + 2g \times P_{ave}^2 \times M^2 \times [h_{out} - h_{in}]\right)}\right\}^{0.2} \qquad \text{[Eq. 7]}$$

where:

R = universal gas constant (8.314 m^3-Pa/K-mol)
M = molecular weight of CO_2 (44.01×10^{-3} kg/mol)
Z_{ave} = compressibility factor for CO_2 (dimensionless)
T_{ave} = average temperature of CO_2 in the pipeline (K), assumed to be the ground temperature (about 285 K or 12°C or 53.3°F)
P_{ave} = average pressure of CO_2 in the pipe (Pa)

When using the McCoy and Rubin approach, Equation 7 replaces Equation 1 in the above procedure for calculating the minimum inner diameter for a pipe.

In the above equations, the average pressure and temperature in the pipeline are used to calculate the density and compressibility factor using the Peng–Robinson equation of state.

Equations 1 and 7 yield estimates that are within 1% of each other when there is no elevation difference (i.e., $h_{in} = h_{out}$). When there is an elevation difference, Equation 7 should be used because it explicitly includes the influence of elevation on the potential energy of the fluid in the pipe.

When the density is low, as is the case when CO_2 is in the gas phase, the diameter of the pipeline is larger to transport the same quantity of CO_2. To provide an example of this, the FE/NETL CO_2 Transport Cost Model (2014) was used to estimate the nominal pipe diameter and the number of booster pumps that would be necessary for a 373-km (232-mi) pipeline with an elevation change of 572 m (1877 ft) that transports 13.9 Mt/yr. If the CO_2 were to be transported as a gas at a pipeline inlet pressure of 5.52 MPa (800 psi) and an exit pressure of 5.52 MPa (800 psi), the nominal pipeline inside diameter would be 91 cm (36 in.) and 15 booster pumps would be required. If the same metrics are applied to a supercritical CO_2 stream with a pipeline inlet pressure of 15.1 MPa (2200 psi) and a pipeline outlet pressure of 12.4 MPa (1800 psi), a pipeline having a nominal inside diameter of 61 cm (24 in.) would be required. This pipeline would only need three booster pumps. Pipeline capital costs for these two cases could differ by as much as $421,000/km ($670,000/mi). Clearly, supercritical CO_2 can be transported in a smaller and therefore less expensive pipeline than if the CO_2 remains in the gas phase.

APPROACHES TO CO_2 COMPRESSION

The CO_2 stream exiting all CO_2 capture technologies will be in the gas phase. Generally, CO_2 from a capture process will be at a pressure between 0.1 and 2.4 MPa (14.5 and 350 psia) and at a temperature ranging from 20° to 40°C (68° to 104°F) (Jensen and others, 2011). Compression outlet pressure generally ranges from about 10.0 MPa (1450 psia), which ensures that a CO_2 stream can be maintained in its supercritical state, to 18.7 MPa (2700 psia), the pressure at which the CO_2 leaves the Great Plains Synfuels Plant. (This pressure was chosen so as to deliver the CO_2 at the pressure needed at the Weyburn oil field.) A typical pressure for transporting CO_2 by pipeline in the United States is 2000 psia.

Three approaches can be taken to compress CO_2 for pipeline transport (shown on a pressure–enthalpy diagram pictured in Figure 1):

- Path C, in which heat is neither gained nor lost by the system. This is also called a near-adiabatic pathway. The gas is compressed in separate steps or stages and is cooled between the stages to remove heat that is generated during the compression. This is how CO_2 usually is compressed.

- Path B, where the gas-phase CO_2 is compressed in stages and cooled, as in the near-adiabatic approach. This continues until the conditions are above the critical point (at the top of the dome, where CO_2 reaches the supercritical phase). The CO_2 is then cooled to a more dense supercritical fluid and is pumped to the final pressure.

- Path A, which utilizes some of the compression stages, then cools the CO_2 to form a liquid, i.e., the pathway crosses the two-phase dome. The liquid is then pumped to the desired final pressure.

Choice of compression approach for a given situation is based upon the power demands and investment cost (International Energy Agency Greenhouse Gas R&D Programme, 2011). The primary caveat is to ensure that a compressor or pump is not working at conditions that place the CO_2 under the dome (i.e., in the two-phase regime) in Figure 1. Pumps and compressors cavitate under two-phase conditions and can be damaged if operated in this regime.

Near-Adiabatic Compression

Figure 2 summarizes the approximate ranges of pressures and inlet flow rates that are handled by various types of CO_2 compressors and pumps. Descriptions of the various types of compressors shown in Figure 2 can be found in the report entitled “Opportunities and Challenges

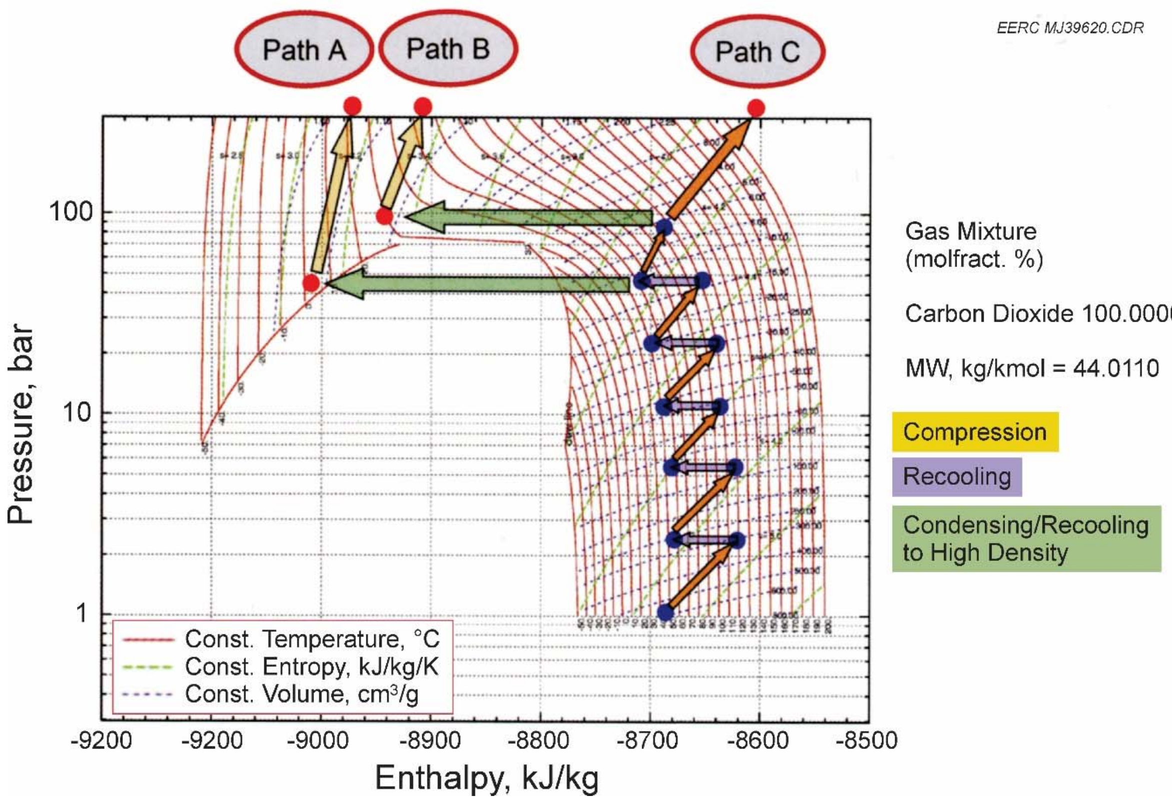

Figure 1. Three compression pathways toward a target pressure of 200 bar (20 MPa, 2900 psi) (taken from Winter, 2009).

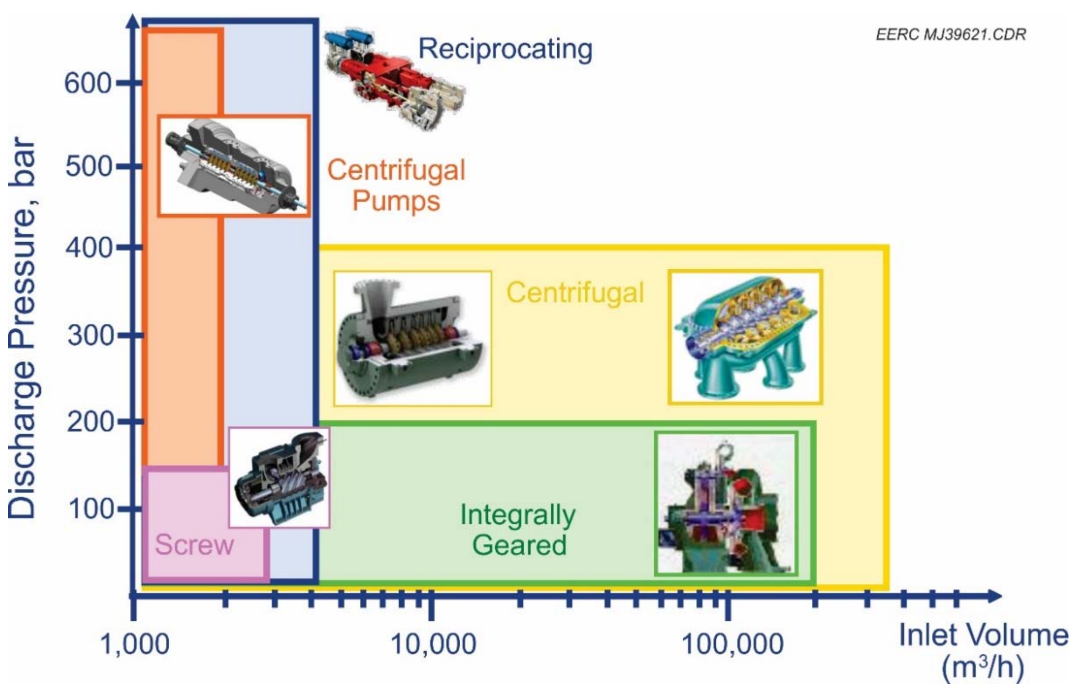

Figure 2. Types of compressors and pumps and the approximate ranges of inlet volumetric flow rates and pressures at which they are used (taken from Wadas, 2010). It should be noted that 500 bar = 50 MPa = 7252 psi and 100,000 m^3/h = 3.53 MMcfh.

Associated with CO_2 Compression and Transportation During CCS Activities" (Jensen and others, 2011). The final selection of compressor type for a particular application is made while considering a number of factors such as safety aspects with CO_2 (and possibly H_2S, depending on the composition of the CO_2 stream), especially with respect to seal type and composition; maintenance access; machine complexity; intercooler type (i.e., water or air); and overall power consumption (Weatherwax and others, 2012).

One type of compressor that is not shown on Figure 2 is an advanced compression concept for which development is nearing completion. The Dresser-Rand SuperCompressor is a high-efficiency gas compressor originally developed by Ramgen Power Systems that utilizes the same shock compression technology that is used by supersonic aircraft inlet systems. It features a rotating disk that operates at the high peripheral speeds necessary to achieve supersonic effect in a stationary environment. The disk is designed so that gas flow mimics the effect of the centerbody and channels of a conventional ramjet inlet. When gas enters the annular space between the supersonically spinning disk and the outer edge of its casing, a "ramming" effect is created, generating shock waves and gas compression analogous to the ramjet inlets on supersonic aerospace vehicles. This compression process is very efficient because the compressor has few aerodynamic leading edges and minimal drag. Additional information about the Dresser-Rand SuperCompressor (formerly called the Rampressor) can be found in the report entitled "Preliminary Design of Advanced Compressor Technology" (Jensen and others, 2009).

The SuperCompressor has been shown during testing to be capable of a single-stage pressure ratio of 8.9:1 (Baldwin and Williams, 2009). Such a high pressure ratio results in the production of considerable heat during each stage. Heat recovery from this type of compressor could be of significant value when fully integrated into a CO_2 capture and transport system (Baldwin, 2009).

The SuperCompressor offers the opportunity for significant waste heat recovery (Dresser-Rand, 2008). During its development at Ramgen (when the SuperCompressor was known as the Rampressor), a two-stage 100:1 pressure ratio Rampressor was compared to conventional integrally geared and in-line compressor configurations using modeled data. The results are summarized in Table 2.

Compression with Cooling

In the situation of Path B on Figure 1, the choice of a compressor or a pump for compression of CO_2 becomes a question of density rather than phase because of the special characteristics of supercritical CO_2 (Jockenhövel and others, 2009). An intercooled gear-type compressor can be used to compress the CO_2 to a supercritical state, followed by cooling of the CO_2 stream to change its density into the liquid range. At this point, a pump or high-density compressor can be used to increase the pressure to the desired condition (Jockenhövel and others, 2009). By achieving supercritical conditions prior to cooling, a two-phase condition (i.e., the area under the dome of Figure 1) is avoided.

Table 2. Comparison of the SuperCompressor[a] to Conventional CO_2 Compressors[b]

Parameter	SuperCompressor	Integrally Geared Turbo Compressor	Inline Process Turbo Compressor
lb/h	150,000	150,000	150,000
icfm	21,411	21,411	21,411
Stages	2	8	12
Intercoolers	1	7	2
Casings	1	1	3
kW	7333	7382	8312
hp	9830	9899	11,147
bhp/100	45.9	46.2	52.1
Isothermal Efficiency	65.8%	64.0%	56.9%
Approximate Average Stage/Casing Discharge Temperature, °F	470	210	380
Maximum Thermal Recovery Temperature, °F	250	250	250
kW Equivalent of Heat	5263	554	4172
% of Heat That Is Recoverable	71.8%	7.5%	50.2%
Shaft Power kW – Heat Recovery kW	2070	6828	4141

[a] Comparison was performed by Ramgen using modeled numbers for what was then called the Rampressor (now known as the SuperCompressor).
[b] Taken from Jensen and others, 2009.

Liquefaction

The final approach is shown in Figure 1 as Path A, in which a CO_2 stream is compressed part of the way to the desired pressure, then cooled through the two-phase region to reach a liquid, and finally pumped to the desired pressure using high-pressure pumps. A liquefaction process studied by the Southwest Research Institute (SRI), DOE NETL, Dresser-Rand, and BP utilized a refrigeration system to cool a CO_2 stream compressed to a pressure of roughly 1.72 MPa (250 psia) to −29°C (−20°F). The liquid CO_2 was then pumped using a cryogenic pump to a pressure of 15.27 MPa (2215 psia) (Moore and others, 2009).

Another liquefaction approach (and variation on Path A) found in the literature is one in which the CO_2 is cooled by an ammonia absorption refrigeration process and then compressed to the desired pressure (Duan and others, 2013). This approach makes use of low-quality heat to drive the refrigeration process and the authors say that the process can lower energy consumption over traditional compression methods when abundant low-quality heat is available.

Comparison of CO_2 Compression and Liquefaction

The underlying premise of the liquefaction approach is that significantly less power is required to raise pressure by liquid pumps and that the pumps are considerably less expensive than gas compressors. However, it is crucial that the refrigeration process be carefully assessed when determining the system power requirements (Baldwin and Williams, 2009; Moore and others, 2009). Two power loads must be considered for the liquefaction option, namely the refrigeration compressor and the cryogenic pump (Baldwin and Williams, 2009). An economizer can be employed to offset some of the refrigeration load and provide initial CO_2 cooling by the cryogenic pump discharge (Baldwin and Williams, 2009).

SRI performed thermodynamic analysis to indicate the power requirements of various compression technology options. This analysis used conventional Dresser-Rand ten-stage centrifugal compression with air cooling between stages as the base case. SRI estimated power requirement reductions for several different approaches, with improvements ranging from 7.44% to as much as 36.17%. Two options ("high ratio compression with 90% efficiency" with either limited or no interstage cooling) actually exhibited power requirement increases of 6.36% and 47.06%. The most efficient approach was calculated to be isothermal compression at 70°F and 80% efficiency, which reduced the power requirements by 36.17%. Compression using a centrifugal compressor with air cooling to 1.72 MPa (250 psia), followed by refrigeration to −32°C (−25°F), and liquid cryogenic pumping to 15.27 MPa (2215 psia) was calculated to require 34.86% less power than the base case. The results of SRI's thermodynamic analysis are presented in Table 3.

According to Baldwin and Williams (2009), the use of a shock wave CO_2 compressor (i.e., the SuperCompressor) to compress CO_2 from 1.7 MPa (220 psia) to 15.27 MPa (2215 psia) was 9.5% more efficient than liquefying the CO_2 and pumping it to pressure. These findings can be seen in Table 4. Similar analyses were completed for a matrix of liquefaction pressures ranging from 1.5 to 6.2 MPa (220 to 900 psia). These results, presented in Table 5, show that the compression auxiliary power benefit switches from liquefaction to gas compression at about 3.4 MPa (500 psi).

In a pipeline, liquid CO_2 would have to be kept at conditions that would allow it to maintain a liquid state, that is, at a temperature below 31°C (87.8°F). Typically, liquid CO_2 is maintained at −20°C (−4°F) and 2 MPa (300 psi) when transported by truck or rail tanker (Metz and others, 2005). Kept at a pressure higher than the critical pressure of 7.4 MPa (1080 psi) and the critical temperature of 31°C (88°F), CO_2 will remain supercritical. Therefore, the addition of booster stations to the pipeline can maintain the state of the supercritical CO_2 through elevation differences due to topography changes along the pipeline route, friction loss caused by the pipeline material, and temperature changes. The CO_2 that is transported by pipeline in the United States is transported in the supercritical phase.

Table 3. Thermodynamic Comparison of Compression and Liquefaction Options[a]

Compression Technology	Power Requirements, BHP[b]	Difference from Base Case, %	Cooling Technology
Conventional Dresser-Rand Centrifugal Ten-Stage Compression, base case	23,251	0.00	Air-cool streams between separate stages
Conventional Dresser-Rand Centrifugal Ten Stage Compression with Additional Cooling	21,522	−7.44	Air-cool streams between separate stages using ASU[c] cool N_2 stream
Isothermal Compression at 70°F and 80% Efficiency	14,480	−36.17	T_c[d] = 70°F inlet temperature throughout
Semi-Isothermal Compression at 70°F, pressure ratio = ~1.55	17,025, with required cooling power TBD[e]	−26.78	T_c = 70°F between each stage
Semi-Isothermal Compression at 100°F, pressure ratio = ~1.55	17,979, with required cooling power TBD	−22.67	T_c = 100° between each stage
High-Ratio Compression at 90% Efficiency, no interstage cooling	34,192	47.06	Air cool at 2215 psia only
High-Ratio Compression at 90% Efficiency, intercooling on final compression stage	24,730	6.36	Air cool at 220 and 2215 psia
Centrifugal Compression to 250 psia, liquid cryopump from 250 to 2215 psia	16,198 (includes 7814 BHP for refrigeration)	−30.33	Air cool up to 250 psia, refrigeration to reduce CO_2 to −25°F to liquefy
Centrifugal Compression to 250 psia with Semi-Isothermal Cooling at 100°F, Liquid Cryopump from 250 to 2215 psia	15,145 (includes 7814 BHP for refrigeration)	−34.86	Air cool up to 250 psia between centrifugal stages, refrigeration to reduce CO_2 to −25°F to liquefy

[a] From Moore and others, 2009.
[b] Brake horsepower.
[c] Air separation unit.
[d] Critical temperature.
[e] To be determined.

Table 4. Comparison of Gas Compression Using the SuperCompressor and Liquefaction[a–c]

	Gas Compression, hp	Liquefaction Option Without Economizer, hp	Liquefaction Option with Economizer, hp
HP Compressor	15,904	NA[d]	NA
Refrigeration Compressor	NA	18,772	18,772
Economizer Credit	NA	NA	−2999
Cryogenic Pump	NA	1809	1809
Total	15,904	20,581	17,582

[a] Baldwin and Williams, 2009.
[b] 1.5 to 6.2 MPa (220 to 900 psia).
[c] Values do not include the low-pressure compressor auxiliary power requirement that is common to both options.
[d] Not applicable.

Table 5. Comparison of the SuperCompressor and Liquefaction for a Matrix of Liquefaction Pressures[a,b]

Liquefaction Pressure, MPa (psi)	Compressor with Economizer, hp	Cryogenic Pump, hp	Total Liquefaction System Auxiliary Load, hp	Gas Compression, hp	Power Savings, hp
1.5 (220)	18,099	1810	19,909	17,314	2595
1.7 (250)	15,773	1809	17,582	15,904	1678
2.1 (300)	13,146	1802	14,948	13,848	1100
2.8 (400)	9521	1779	11,300	10,822	478
3.5 (500)	7056	1746	8802	8711	91
4.1 (600)	5279	1706	6985	7125	−140
4.8 (700)	4015	1659	5674	5853	−179
5.5 (800)	2941	1607	4548	4803	−255
6.2 (900)	2247	1550	3797	3899	−102

[a] Baldwin and Williams, 2009.
[b] Values do not include the low-pressure compressor auxiliary power requirement that is common to both options.

INTEGRATION OF COMPRESSION INTO A CO_2 CAPTURE SYSTEM

Considerable research has been devoted to reducing the costs associated with various approaches for capturing CO_2, but much less attention has been paid to compression. CO_2 compression plays an important role in the total capital requirement of and energy penalty associated with a capture technology. Different capture technologies produce CO_2 streams that are at different pressure and temperature conditions, affecting the compression requirements. Different types of compressors produce different quantities of heat that can be removed between stages with the potential for use in the capture process. Most compression incorporates CO_2 stream dehydration, either through condensation of water in the compression intercoolers, in a separate dehydration step, or both. It is clear that thoughtful, optimized integration of compression and dehydration into a capture system could produce cost and energy savings and improve the efficiency of a power plant or industrial process.

The first step in integrating compression into a CO_2 capture facility is to define the process requirements for compressing the CO_2. One of the most important parameters to be considered is the final water content specification since the pressure at which drying is required will define the location of the dehydration system within the compression train. Pipeline transport of CO_2 in the United States generally follows the Kinder Morgan specification, which is a maximum of 600 ppm water by weight.

The wet CO_2 stream from a postcombustion capture process or an oxycombustion process must first be cooled to condense and separate the water (Romeo and others, 2009). The compression process is divided into several stages, generally with some type of cooling between the stages as this reduces the work required. A triethylene glycol (TEG) dehydrator is often used to remove water once a pressure of 3 MPa (435 psi) has been reached (Romeo and others, 2009). If the CO_2 is leaving an oxycombustion process, however, it is likely that a TEG dehydration system would not be used as the glycol degrades in the presence of oxygen (International Energy Agency Greenhouse Gas R&D Programme, 2011). Molecular sieves or silica gel could also be used to remove water from the CO_2 stream. A water concentration of 60 ppm can be reached using any of these systems (Romeo and others, 2009). After dehydration, the CO_2 stream can be further compressed to reach the desired pipeline pressure.

Intercooling between stages reduces the power requirement for compression and therefore the compressor size (Romeo and others, 2009). In general, the heat is rejected to low-temperature cooling equipment in order to reduce the compression penalty. This strategy can benefit operation, especially in cold locations. However, in locations with higher temperatures, larger heat exchangers would be required to cool the gas (Romeo and others, 2009).

Foster Wheeler Italiana studied compression in CCS systems for the International Energy Agency Greenhouse Gas R&D Programme (IEAGHG). Their study targeted the basic compression requirements of pre-, oxy-, and postcombustion capture processes (International Energy Agency Greenhouse Gas R&D Programme, 2011). Integral to their study was the determination of how the compression could be better integrated with the capture system to provide a more energy- and cost-efficient process. This report (International Energy Agency Greenhouse Gas R&D Programme report, 2011) is lengthy, and the reader is encouraged to review it for more detail. Some of the findings included the following:

- The specification for final water content of the compressed CO_2 was found to be an important parameter as this affects the selection of a drying step that is in addition to compressor after-cooling and water knockout. Mole sieve dryers require a CO_2 recycle stream as well as a heat source with which to regenerate the adsorption bed. Oxycombustion processes require a very dry stream because of the required cryogenic processing conditions. Glycol cannot be used in this instance mainly because of the presence of oxygen, which causes glycol degradation. The pressure at which drying takes place is fixed by the parameters of the oxycombustion CO_2 cleanup process, meaning that its place within the compression train is also fixed. There is more flexibility with respect to the pressure at which the drying step takes place for pre- and postcombustion capture processes.

- In precombustion capture, increasing the number of solvent flash stages in the acid gas removal unit improved the overall plant economics. The number and operating conditions of each flash stage must be determined while also considering the characteristics of the chosen compressor so as not to introduce complications in the design of the compressor train. This can be minimized if additional solvent flash stages are introduced at a pressure close to the compressor stage discharge conditions.

- Increasing the number of compression stages exhibited both capital and operating expense improvements. For some compressor types, such as centrifugal compressors, the single-stage compression ratio cannot be reduced acceptably.

- Liquefaction of CO_2, as opposed to compression, may be economically attractive in cooler climates. In warmer climates, the pipeline would need to be designed for transport of CO_2 at temperatures below the critical temperature of 31°C or for a fluid whose physical properties are likely to change quickly as it heats up along the pipeline route.

- Early liquefaction appears promising for application to precombustion capture because the large amount of low-temperature waste heat from the cooling unit can be recovered in an absorption refrigeration system.

- Centrifugal (as opposed to reciprocating) compressors are considered to be the most appropriate for large-scale CCS applications because of their greater reliability, higher efficiency, and the fact that they are easier to maintain.

- If the stripper operating pressure in a postcombustion capture system can be increased, less compression would be required. However, the economics of operating the stripper this way are not necessarily attractive and would likely outweigh the benefits to the compression step.

- Shock wave compression (e.g., the Dresser-Rand SuperCompressor), while not yet commercial, offers potential for better economics and performance. The SuperCompressor concept is well suited to postcombustion but was found in the IEAGHG study not to be as effective for precombustion and not at all appropriate to current oxycombustion plant designs. The technology offers slightly higher overall power plant efficiency with greater simplicity in the compression step and potentially lower capital cost.

CONCLUSIONS AND RECOMMENDATIONS

Compression plays an important role in the overall CO_2 capture plant efficiency. Selection of an appropriate compression technology for the quantity of CO_2, desired pipeline pressure, and type of capture technology is crucial. For example, centrifugal compression appears to be the most appropriate for all three capture platforms (pre-, oxy-, and postcombustion). The shock wave compression offered by the Dresser-Rand SuperCompressor is well-suited to postcombustion but not to oxycombustion. Placement of the dehydration step within the compressor train affects integration of the heat produced during compression as well as compressor design. Optimization of compression within a plant requires integration of the heat of compression so as to maximize plant efficiency. The Dresser-Rand SuperCompressor, for example, offers the opportunity for significant waste heat recovery. The best plant efficiency and capture economics will be achieved by integrating the capture technology, dehydration, compression approach, and integration of the compressor waste heat into the overall plant. Effective optimization will require that heat integration, dehydration design, and compressor selection be determined iteratively.

Further studies of the effects of various dehydration schemes on compression could be of value when determining the best approaches to efficiently and cost-effectively integrate the entire CO_2 capture system into a power plant or industrial facility. Additional studies of the integration of the SuperCompressor into a capture facility are also recommended as the SuperCompressor is sufficiently different from other compressor technologies as to require a fresh examination of how dehydration and integration of the considerable quantity of usable heat generated could be most effectively applied.

REFERENCES

Baldwin, P., 2009, Low-cost, high-efficiency CO_2 compressors: Carbon Capture Journal, issue 11, 19–21.

Baldwin, P., and Williams, J., 2009, Capturing CO_2: gas compression vs. liquefaction: Power, v. 153, no. 6, 68–71.

Carbon Capture and Sequestration Technologies Program, 2009, Carbon management GIS: CO_2 pipeline transport cost estimation, Massachusetts Institute of Technology, Report for U.S. Department of Energy National Energy Technology Laboratory under contract DE-FC26-02NT41622.

Doctor, R., Palmer, A., Coleman, D., Davison, J., Hendriks, C., Kaarstad, O., Ozaki, M., and Austell, M., 2005, Chapter 4 – transport of CO_2, *in* Metz, B., Davidson, O., de Coninck, H., Loos, M., and Meyer, L., eds., Intergovernmental Panel on Climate Change special report on carbon dioxide capture and storage: Cambridge University Press, p. 181–192.

Dresser-Rand, 2008, Dresser-Rand invests in the development of game-changing compressor technology focused on green energy: investor.dresser-rand.com/releasedetail.cfm?releaseid=346744 (accessed May 2015).

Duan, L., Chen X., and Yang Y., 2013, Study on a novel process for CO_2 compression and liquefaction integrated with the refrigeration process: International Journal of Energy Research, v. 37, 1453–1464.

Fossil Energy/National Energy Technology Laboratory CO_2 Transport Cost Model, 2014, available at www.netl.doe.gov/research/energy-analysis/analytical-tools-and-data/co2-transport (accessed May 2015).

Heddle, G., Herzog, H., and Klett, M., 2003, The economics of CO_2 storage: Laboratory for Energy and the Environment, Massachusetts Institute of Technology, publication number LFEE 2003-003 RP.

International Energy Agency Greenhouse Gas R&D Programme, 2011, Rotating equipment for carbon dioxide capture and storage: Cheltenham, Gloucestershire, UK, IEAGHG Report 2011/07, 334 p.

Jensen, M., Steadman, E., Harju, J., and Belshaw, K., 2009, Preliminary design of advanced compression technology, Energy & Environmental Research Center, Report for U.S. Department of Energy National Energy Technology Laboratory under cooperative agreement DE-FC26-05NT42592.

Jensen, M., Cowan, R., Pei, P., Steadman, E., and Harju, J., 2011, Opportunities and challenges associated with CO_2 compression and transportation during CCS activities, Energy & Environmental Research Center, Report for U.S. Department of Energy National Energy Technology Laboratory under contract DE-FC26-05NT42592.

Jockenhövel, T., Schneider, R., Sandell, M., and Schlüter, L., 2009, Optimal power plant integration of post-combustion CO_2 capture: presented at POWER-GEN Europe 2009, Cologne, Germany, May 26–29, 2009.

Massachusetts Institute of Technology, 2009, Carbon management GIS: CO_2 pipeline transport cost estimation: Carbon Capture and Sequestration Technologies Program, Massachusetts Institute of Technology.

McCoy, S., and Rubin, E., 2008, An engineering-economic model of pipeline transport of CO_2 with application to carbon capture and storage: International Journal of Greenhouse Gas Control, v. 2, 219–229.

McCollum, D., and Ogden, J., 2006, Techno-economic models for carbon dioxide compression, transport, and storage and correlations for estimating carbon dioxide density and viscosity: Institute of Transportation Studies, University of California at Davis, publication UCD-ITS-RR-06-14.

Metz, B., Davidson, O., de Coninck, H., Loos, M., and Meyer, L., eds., 2005, IPCC special report on carbon dioxide capture and storage: New York, Cambridge University Press, 431 p.

Mohitpour, M., Golshan, H., and Murray, A., 2003, Pipeline design and construction: New York, New York, ASME Press.

Moore, J., Blieske, M., Delgado, H., Lerche, A., Alsup, C., Pacheco, J., Bough, M., and Byard, D., 2009, Novel concepts for the compression of large volumes of CO_2 phase II: Presented at the Annual NETL CO_2 Capture Technology for Existing Plants R&D Meeting, Pittsburgh, Pennsylvania, March 24–26, 2009.

Morgan, D., Grant, T., Simpson, J., Myles, P., Poe, A., and Valenstein, J., 2014, FE/NETL CO_2 transport cost model: description and user's manual: U.S. Department of Energy National Energy Technology Laboratory report DOE/NETL-2014/1660.

Romeo, L., Bolea, I., Lara, Y., and Escosa J., 2009, Optimization of intercooling compression in CO_2 capture systems: Applied Thermal Engineering, v. 29, 1744–1751.

Wadas, B., 2010, CO_2 capture and compression technologies, *in* Proceedings of the CO_2 Summit: Technology and Opportunity, Vail, Colorado, June 6–9, 2010: http://dc.engconfintl.org/cgi/viewcontent.cgi?article=1035&context=co2_summit (accessed February 2015).

Weatherwax, M., Patel, V., Musardo, A., Giovani, G., Pelella, M., and Cipriani, S., 2012, *in* Proceedings of the Forty-First Turbomachinery Symposium, Houston, Texas, September 24–27, 2012.

Winter, T., 2009, The right solution for CO_2 compression–integrally geared compressors from Siemens: Carbon Capture Journal, issue 11, p. 22, 22–24.

Wong, S., 2005, CO_2 compression and transportation to storage reservoir *in* Building capacity for CO_2 capture and storage in the APEC region—A training manual for policy makers and practitioners: module 4, p. 4–1–4–16.