

Analysis of CO₂ Separation from Flue Gas, Pipeline Transportation, and Sequestration in Coal

Eric P. Robertson

September 2007



The INL is a U.S. Department of Energy National Laboratory
operated by Battelle Energy Alliance

Analysis of CO₂ Separation from Flue Gas, Pipeline Transportation, and Sequestration in Coal

Eric P. Robertson

September 2007

**Idaho National Laboratory
Idaho Falls, Idaho 83415**

<http://www.inl.gov>

**Prepared for the
U.S. Department of Energy
Assistant Secretary for Fossil Energy
Under DOE Idaho Operations Office
Contract DE-AC07-05ID14517**

DISCLAIMER

This information was prepared as an account of work sponsored by an agency of the U.S. Government. Neither the U.S. Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. References herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise, does not necessarily constitute or imply its endorsement, recommendation, or favoring by the U.S. Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the U.S. Government or any agency thereof.

1 Forward

This report was written to satisfy a milestone of the Enhanced Coal Bed Methane Recovery and CO₂ Sequestration task of the Big Sky Carbon Sequestration project.^a The task was directed by Eric Robertson at the Idaho National Laboratory (INL). The report also satisfies a fiscal year 2006 Performance Evaluation Management Plan (PEMP) milestone for INL. The U.S. Department of Energy Idaho Operations Office approved the milestone as follows:

1.1 Milestone Statement

This task will begin to assess the costs associated with separating the CO₂ from flue gas and then injecting it into a coal seam. The technical challenges and costs associated with CO₂ separation from flue gas and transportation of the separated CO₂ from the point source to an appropriate sequestration target will be analyzed. An interim status report will be prepared that details the data generated and analysis performed in FY-06, as described as follows:

- 1) The report will include the selection of a specific coal-fired power plant for the application of CO₂ separation technology. An appropriate CO₂ separation technology will be identified from existing commercial technologies.
- 2) The report will also include a process design for the chosen technology tailored to the selected power plant that will be used to obtain accurate costs of separating the CO₂ from the flue gas. In addition, an analysis of the costs for compression and transportation of the CO₂ from the point-source to an appropriate coal bed sequestration site will be included in the report.

^a <http://www.bigskyco2.org/>

2 Introduction

Future regulations of carbon dioxide (CO₂) and other greenhouse gases emissions may tip the scale of currently unprofitable sequestration technologies by providing incentives for compliance. The reduction of greenhouse gases with the goal of addressing global warming concerns may generate economic benefits to some areas of the United States that contain extensive sequestration sinks. Geologic sequestration options in Wyoming include many large, partially depleted oil and gas fields; vast coal beds; large saline aquifers; and oil shales. All these sinks lie close to significant point-source for CO₂ and a growing regional CO₂ pipelining infrastructure.

Technical and economic data on potential CO₂ separation technology and CO₂ pipeline transportation have been collected and capital and operating costs have been determined. The first goal of this interim report is to compare these technical and economic results in order to select an appropriate technology for separating CO₂ from a typical coal-fired power plant in the U.S. intermountain area. The second goal is to determine anticipated costs associated with transporting the separated CO₂ from the emission source to a disposal location.

This project has worked closely with the University of Wyoming as part of the 2005-2006 Senior Design program under the direction of professors Gordon Harris and David Bell from the Chemical Engineering Department of the College of Engineering. The directive of the Design program was to identify three commercially available CO₂ separation technologies and apply each of them to a specific coal-fired power plant in Wyoming. Facilities for each of the technologies was designed and sized to fit the same location in order to fairly compare them. A technical and economic evaluation was done using Aspen Plus, a process simulator marketed by AspenTech. A large portion of this interim report was taken from the final reports of the Senior Design groups.

Separation technologies examined include chemical absorption, adsorption, and membrane separation processes. The Wyodak pulverized coal power plant located in Campbell County was selected as the point source emitter to provide the design specifications used in all three processes. A comparative economic analysis was performed to determine the break-even cost of CO₂ for the various separation designs.

Transportation of CO₂ is an important focus for determining the cost and economic feasibility for sequestration. Transportation costs of CO₂ often depend on matching CO₂ sources and storage locations, generating considerable variation in costs for each sequestration site. Injection of CO₂ into oil and gas reservoirs and unmineable coal seams can provide value-added byproducts that can offset the cost sequestration, and represent the best near-term option.

Fields suitable for enhanced oil recovery occur within the Powder River basin region at various stages of development. Available information suggests Hartzog Draw, 50 miles from Gillette, is a likely candidate for EOR. The vast CBM industry in Wyoming suggests many possibilities exist for ECBM recovery, but focus is particularly on those coal beds mined with a minimum of surface disturbance and unmineable seams due to thickness, surrounding geology and depth. Due to the extensive potential for ECBM recovery in Wyoming, site selection will be highly dependent on matching CO₂ source to sequestration sink.

Important pipeline considerations include operating pressure, temperature and the composition of the gas. A gas stream of 100% pure in CO₂ was considered. An 8-inch schedule 40 car-

bon steel pipeline with 8-inch valve sets and block valves, complete with pigging launch and receiving system was used to determine an estimated the capital cost for the pipeline design.

This report discusses:

- 1) Commercial technologies for separating CO₂ from flue gas (Chapter 3).
- 2) The selection of a representative point source of CO₂ emissions (Chapter 4).
- 3) The design of three potential separations technologies applied to the same CO₂ point source (Chapter 5).
- 4) An economic comparison of the three separation technologies (Chapter 5).
- 5) The design of a pipeline transportation system for separated CO₂ based on the output from the CO₂ separation technology (Chapter 6).
- 6) An economic analysis of the CO₂ pipeline (Chapter 6).

3 Analysis of CO₂ Separation Technologies

A number of different processes are available for carbon dioxide separation from gas streams. Five main types of processes exist: chemical and physical absorption, adsorption, membrane separation, and cryogenic distillation. In order for any one of these technologies to be feasible, there must be some form of economic or legal incentive. Economic incentives could arise from marketing the separated CO₂ to the food and beverage industry, for use in chemical processes such as urea production, and for use in enhanced recovery of oil and natural gas. Legal incentives are expected to arise from restrictions on carbon dioxide emissions resulting from measures such as the Kyoto protocol that are gaining worldwide support. Three currently commercial technologies were investigated that could be used to separate CO₂ from flue gas: chemical absorption, surface adsorption, and membrane separation.

3.1 CO₂ Separation by Chemical Absorption

Chemical absorption is perhaps the most widely used process for separation of carbon dioxide from gas streams. The process operates on the basic principle of mass transfer with reaction. Separation is achieved through a reversible acid-base reaction between carbon dioxide and an alkaline reagent. Carbon dioxide forms a slightly acidic solution when it is absorbed by water. With the addition of an alkaline reagent to the solution, carbon dioxide will react, increasing the driving force of CO₂ transfer from the gas phase into the solution. Once reacted, it can be regenerated by reversing the reaction.

3.1.1 Chemical absorption reagents and process reactions

The most common reagents used are alkanolamines, such as monoethanolamine (MEA), diethanolamine (DEA), and diisopropanolamine (DIPA). Potassium carbonate and sodium hydroxide are examples of non-amine reagents that can be used. Figure 1 shows the chemical structure of several selected alkaline reagents.

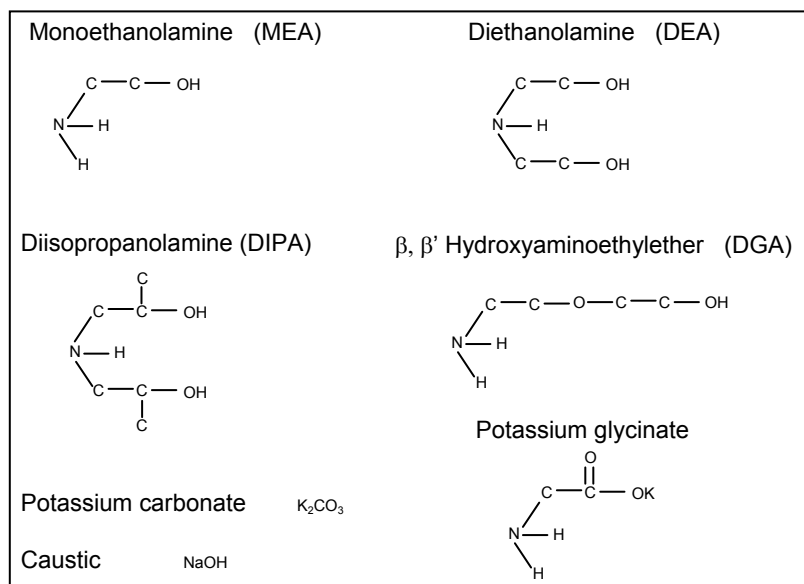
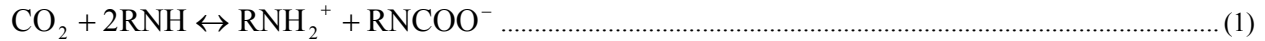


Figure 1. Common alkaline reagents.

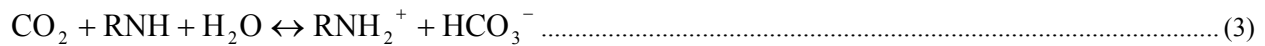
For an amine with functional group R, the basic reaction that occurs requires two moles of amine for every mole of CO₂. The products are a cationic amine species and an anionic carbamate species.



This reaction proceeds via a non-elementary mechanism and is approximately first order with respect to both CO₂ and amine concentrations. The mechanism involves a slower rate determining step in which a zwitterion, with both positively and negatively charged centers, is formed. Once formed, this zwitterion quickly reacts with a second amine to form the more stable ionic products.



Water can also be added to the above reactions as a reactant, forming a bicarbonate ion:



The overall forward reaction is exothermic, making it desirable to keep the temperature as low as possible for absorption. The opposite is true for regeneration of the solvent, as the reverse reaction is endothermic, requiring high temperatures. Typical operating conditions for the absorption step are 40°C and atmospheric pressure, while regeneration generally occurs at 120°C and slightly higher pressures (up to 2 atm).

3.1.2 Typical process flow

The process flow for a typical amine absorption process is fairly simple (see Figure 2). The operations of the main unit consist of an absorber tower, a stripper column, and a heat exchanger to recover heat from the hot regenerated solvent. The flue gas enters from the bottom of the absorber while the cooled lean solvent is pumped in from the top of the tower. The gas contacts the solvent and CO₂ is absorbed; the remaining flue gas leaves out the top of the tower, and the CO₂ rich solvent leaves out the bottom. From there, the solvent passes through the heat exchanger, where it is heated slightly before entering the stripper column. A reboiler at the bottom of the stripper column heats the solvent using steam to reverse the absorption reaction and regenerate the solvent. The hot lean solvent is recycled back through the heat exchanger and a cooler before it is pumped back to the top of the absorber. The CO₂ product stream from the top of the stripper column is saturated with water at nearly atmospheric pressure, and must be dehydrated and compressed if it is sent to a pipeline for transport to a consumer or sequestration site.

The majority of the operating cost is due to the large amount of steam required for regeneration of the solvent. Corrosion of the process equipment can be a serious problem. To prevent corrosion, it is necessary to either use stainless steel for construction or to add corrosion inhibitors to the solvent. Because the stripping process is carried out at around 120°C, which is above the boiling point of water, a large amount of solvent is lost to evaporation. The solvent also degrades in the presence of oxygen to form heat stable salts. Not only must fresh solvent be added to make up for this, the salts have to be periodically removed from the system to prevent clogging or other damage to the process equipment.

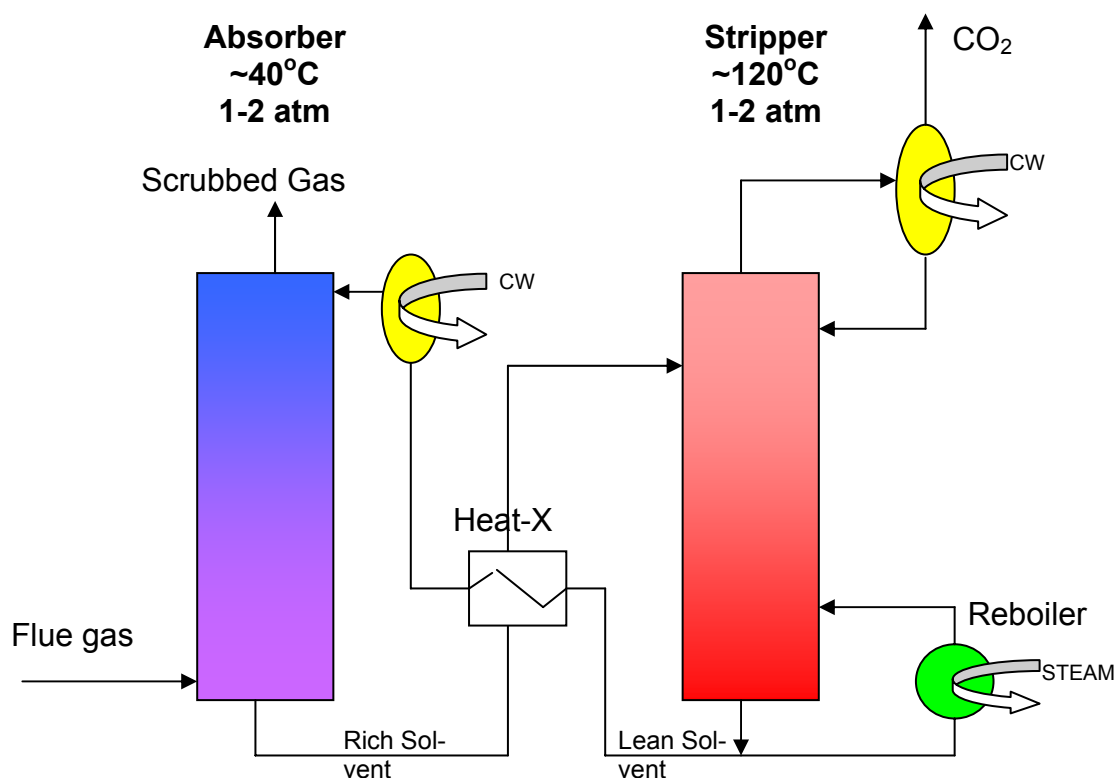


Figure 2. General flow sheet and operating conditions for amine CO₂ separation.

3.1.3 Commercial Technologies for Chemical Absorption

There are several commercial amine absorption processes available for CO₂ removal from flue gases. The most widely used, with 21 plants built, is the Fluor Daniel Econamine FG process. Second is the Kerr-McGee/ABB-Lummus Amine process with 3 plants. The most recently developed technology is the Mitsubishi Heavy Industries KS-1 process, which is used at a single plant built in 1999 [1].

The Fluor Daniel Econamine FG process uses an inhibited 30 wt% MEA solution. This process can recover between 85-90% of the CO₂ in a typical flue gas stream, while producing a product of >95% pure CO₂. Proprietary inhibitors are added to prevent corrosion and allow for the widespread use of carbon steel construction materials. Capacities of the 21 plants constructed range from 6 tones to 1200 tons of CO₂ recovered per day.

The Kerr-McGee/ABB-Lummus Amine process uses a 15-20 wt% MEA solution with no inhibitors. The lack of inhibitors requires the use of some stainless steel equipment, and limits the concentration of the amine in the solution to 20%. This process can recover >96% of the CO₂ from a typical flue gas, with a product stream of 95-98% pure CO₂. The plant capacities range from 200 tons to 800 tons of CO₂ per day.

Mitsubishi Heavy Industries KS-1 process was developed to address some of the key problems encountered with conventional amine absorption. The solvent is a proprietary sterically-hindered amine solution. The solvent can achieve 90% CO₂ recovery from a flue gas stream with a lower solvent circulation rate than other amine technologies. A lower solvent circulation rate means

that less energy is required to regenerate the solvent and the equipment sizes can be smaller leading to lower capital costs for the plant. The solvent also has low corrosivity and degradation rates without the use of inhibitors. The one plant in commercial operation has a capacity of 160 tons of CO₂ per day.

3.2 CO₂ Separation by Adsorption

Adsorption is the process by which a gas attaches to the surface of a solid material, called an adsorbent. This process is primarily attractive to perform separations from gas mixtures where one of the components is preferentially adsorbed on the surface of the adsorbent. Adsorption is a promising separation technique because it can successfully accommodate a wide range of inlet conditions, compositions and produce a varying degree of pure products [2,3]. The adsorption/desorption curve is commonly referred to as an isotherm and is modeled by a Langmuir equation [4]. Adsorption curves, which are defined as the amount of adsorbed gas per pressure unit, typically have steep positive slopes at low pressures and flatten to near horizontal at higher pressures.

3.2.1 Types of Adsorbents

The success of adsorption separation is greatly dependent on the physical characteristics of the adsorbent used. Theoretically, any micro-porous material could be successful for gas adsorption process. Common adsorbents contain well-structured tiny pores of consistent and precise size. There are four main adsorbents useful for separations: zeolites (synthetic and natural), activated carbon, silica gel, and activated alumina [4]. Activated carbon is primarily useful for the removal of non-polar gases and organic vapors; zeolites have been shown to be successful for drying, hydrogen purification, air purification, air separation, and gas chromatography; silica gel and activated alumina have both been shown to be useful for drying and gas chromatography. The main fundamental properties of the adsorbent to be considered are the selectivity and the working capacity (effective adsorption amount), which is defined as the moles of the adsorbed species per kilogram of adsorbent. Zeolites and activated carbon are discussed in further detail in this section.

Zeolites are crystalline aluminosilicates of alkali or alkali earth elements [4] with large surface areas onto which gases, such as CO₂, can adsorb. Zeolite cages are formed by heating the crystalline structure under vacuum to remove the water originally present, leaving an almost unaltered aluminosilicate skeleton with a void fraction between 0.2 and 0.5. The skeleton has a regular structure of cages, which are interconnected by six windows in each cage. Figure 3 is a line representation of a unit cell of types X and Y zeolite or faujasite. The cages can occlude large amounts of gas molecules in place of the original water. The BET surface area measured with nitrogen for zeolites falls in the range between 500 and 800 m²/g.

Carbon dioxide is preferentially adsorbed on the surface of zeolites over the competing nitrogen and oxygen. Water vapor, if available, will also be adsorbed; therefore, it is necessary to remove water vapor from the flue gas stream prior to passing the gas through a zeolite adsorption bed. Desorption of CO₂ is achieved by heating the zeolite bed under vacuum; however, compared to other sorbents, zeolite regeneration requires more energy because of the strong attraction of carbon dioxide to the zeolite substrate.

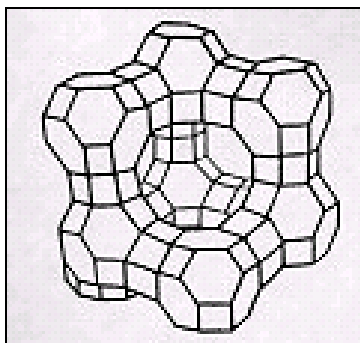


Figure 3. Molecular structure of zeolites.

Activated carbon is another adsorbent that can successfully remove carbon dioxide from a flue gas stream. In general, activated carbon has lower selectivity and lower working capacity than zeolites. However, activated carbon is useful for flue gas applications because the surface interaction between the activated carbon and carbon dioxide are much weaker than that for zeolites, implying a much lower desorption energy requirement. Activated carbon is also much less affected by water than other adsorbent materials and does not require stringent upstream moisture removal.

3.2.2 Regeneration Cycle of Adsorbents

The adsorption cycle is another important criterion used to select an appropriate adsorbent technology for CO₂ removal from flue gas. There are three popular methods to achieve adsorption-desorption among gas applications. Temperature-swing adsorption (TSA) is one of the oldest techniques for regeneration of the adsorbent. TSA uses an increase in temperature to achieve desorption and regenerate the adsorbent; however, the regeneration cycle is slower than other methods because the heating process is slow. Pressure-swing adsorption (PSA) is another adsorption technique useful for gas separation [5,6,7]. With PSA cycles, the adsorption step is performed under elevated pressures and the regeneration is achieved by subsequent pressure reduction. PSA cycle times are on the order of several minutes; whereas TSA cycles are measured on the order of hours or days. The third adsorption method is known as vacuum-swing adsorption (VSA). VSA cycle times are on the same level as the PSA cycle, but the regeneration is achieved by very low absolute pressures (vacuum).

The most basic PSA cycle was first developed and patented in 1958 by an individual named Skarstrom. The cycle was originally employed as a successful and very efficient technique for the drying of air, and became a large-scale commercial operation without any major modifications. The Skarstrom cycle utilizes a two bed design with an adsorption stage and a desorption stage. Simultaneously, one bed is in the adsorption stage and the other bed is in the desorption stage. When the first bed becomes full to capacity (and the other is theoretically completely empty at the same time), valves switch the two beds and bed one is in the desorption phase while the second is in the adsorption phase. Today, the Skarstrom cycle has been modified intensely to allow for a broader range of applications, such as flue gas. Virtually all PSA cycles in use today are extensions of the Skarstrom cycle.

The first adsorption process improvement proposed to further evolve the Skarstrom cycle was known as “cocurrent depressurization” (CD). This process improvement involves cutting the

adsorption phase short, well before the bed has reached full capacity, and employing the CD step. The composition of the void space in the adsorbed bed is equal to the composition of the inlet, and the total concentration of the strongly adsorbed species is not high. The CD step increases the concentration of the strongly adsorbed species in the bed by lowering the pressure in the voids which increases the concentration of the strongly adsorbed species in both phases. The higher concentration of the strongly adsorbed species allows for more adsorption downstream, as adsorption in the PSA process is related to the partial pressure of the species.

The second major adsorption process improvement that drastically improves efficiency of the process is known as “pressure equalization.” Pressure equalization (PE) involves adding an empty tank to the two beds in the Skarstrom cycle and increasing the pressures in the regenerated beds by utilizing the higher pressure gases emitted from the other beds in the cycle.

Another process improved that proved useful in increasing both product purity and recovery is known as “strongly adsorbed purge.” This method uses the high purity product to purge the cycle, effectively replacing the weakly adsorbed species in the bed. The problem with strongly adsorbed purge is that the pure product is produced at low pressures, and must be re-pressurized before being used as purge.

3.3 Membrane Separation

In membrane separation technology, a membrane separates two phases from each other and acts as a selective interface that separates preferred materials from the feed. The membrane material can consist of nonporous solid, microporous or macroporous solid with liquid or gas in the pores, gel, or a combination of these.

The basis for any membrane separation system is shown in Figure 4. One stream leaving the membrane system is enriched of CO₂ (the species that needs to be separated) while the other stream is depleted of CO₂.

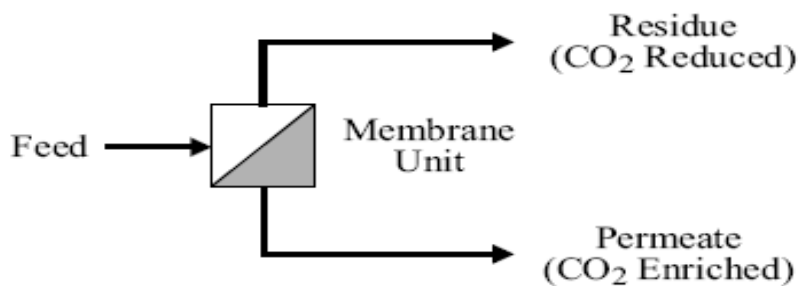


Figure 4. A single-stage membrane unit.

The movement of materials across the membrane can take place due to several driving forces. The driving force can either be due to a chemical potential or an electrical potential. A chemical potential can occur because of a concentration gradient or a pressure gradient or both. The gas separation membrane systems rely on the pressure gradient by setting the stream fed to the membrane at a higher pressure and by setting the stream leaving the membrane at a lower pressure. The pressure gradient provides the driving force for the materials to diffuse across the membrane.

Various species in the stream have different diffusivities through a given membrane material; therefore, each species diffuses at a different rate across the membrane and leads to separation into two streams. Molecules with larger molecular weights diffuse slower than molecules with smaller molecular weights. If the pores of the membrane are too large, convective flow of molecules will occur through the membrane and therefore, no separation will occur. If the pores of the membrane are quite small, larger molecules will not pass through the membrane and will be separated due to molecular sieving. If the pore diameters are smaller than the mean free path of the gas molecules entering the membrane, smaller molecules will pass through the membrane by Knudsen diffusion. In solution-diffusion membranes, selective gas molecules will dissolve into the membrane material and then diffuse across the membrane.

Cellulose acetate and polyimide membranes are more commonly used in the commercial world today for separation of carbon dioxide. Permeability of carbon dioxide across cellulose acetate and polyimide membranes are 10 and 13 Barrers^b respectively. These permeability values are very high compared to methane permeability of about 0.4 Barrers across cellulose acetate and polyimide membranes [8]. Separation of carbon dioxide using membranes is widely used for natural gas separation applications today. The main component in natural gas is methane and since cellulose acetate and polyimide membranes are very selective for carbon dioxide over methane, these membranes are very efficient to separate carbon dioxide from natural gas streams.

3.3.1 Types of Membrane Construction

Two common ways to construct membrane modules are the hollow fiber module and the spiral-wound module.

A hollow fiber module is a collection of cylindrically shaped fibers (see Figure 5). The feed flows through the cylinders and the selective material from the feed diffuses across the fiber material while the rest of the feed material keeps flowing through the cylinders. The material that diffuses across the fibers leaves as the permeate stream from one side and the material that do not diffuse leaves as the residue stream from another side of the membrane [9].

In a spiral-wound module (see Figure 6), the feed flows axially across the membrane envelope and the selective feed material permeates through the membrane envelope radially inward to the collection tube and exits. The material in the feed that is not selective to the membrane will continue to flow axially across the membrane envelope and exit from the opposite side [9].

3.3.2 Advantages and disadvantages of membrane separation

Some advantages of membrane separation systems over absorption and adsorption processes include lower operating costs because membranes do not need to be regenerated. In addition, the operation is simple and can be left for long periods of time. The capital costs for the membrane module itself is lower than other separation processes, but large pre-treatment units can increase the total capital costs for these installations. Some membrane units require more than one stage with pretreatment units and compressors [10].

^bA Barrer is a non-SI unit of gas permeability and is a bulk property of the material the gas is passing through. The permeability of a membrane would equal one Barrer if 846 standard mL of gas passed through a membrane 1 mm thick and 1 cm² in area under one bar of differential pressure in one day.

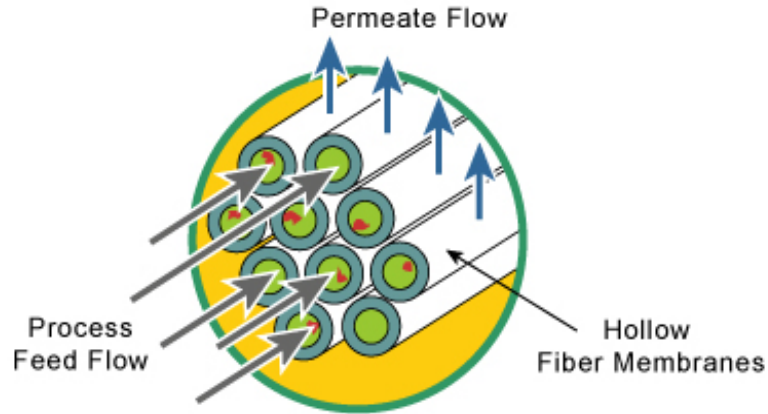


Figure 5. Hollow fiber membrane separation module.

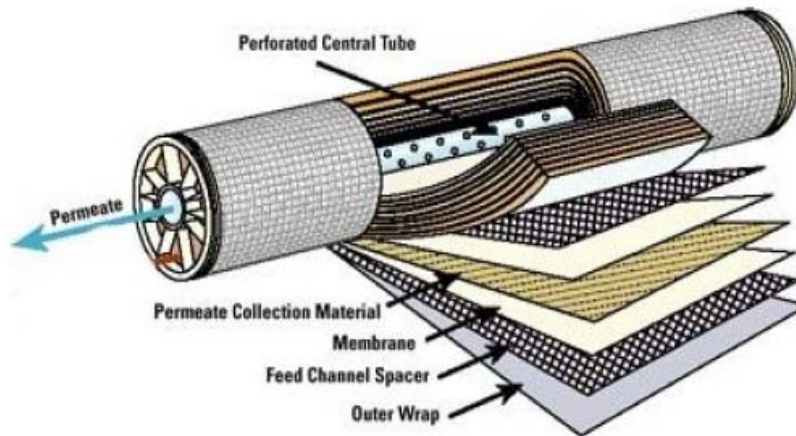


Figure 6. Spiral-wound membrane separation module.

Membrane units do not require as much space as the other types of acid gas removal systems because the equipment used for membrane units are comparatively smaller. Figure 7 shows the relative size of a membrane system in the left hand side, an amine unit, and a glycol plant [10].

3.3.3 Commercial membrane separation systems

Of the several commercial membrane separation systems available today, Cyanara and Separex membranes are discussed below.

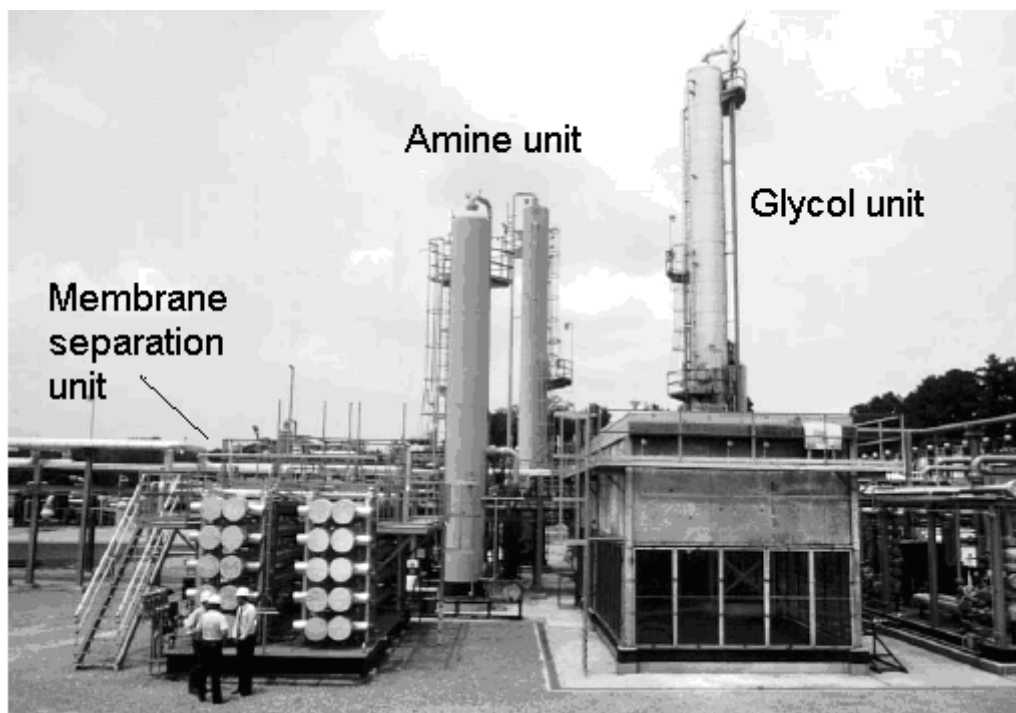


Figure 7. A membrane unit versus an amine and glycol unit.

3.3.3.1 Cyanara Membrane Technology

Cyanara membrane technology (NATCO Group) uses cellulose acetate membranes with hollow fiber modules for carbon dioxide separation processes (see Figure 8).

The first Cyanara membrane separation unit was installed in 1983 and these membrane units can currently process up to 700 MMscfd of inlet feed rate and can handle carbon dioxide inlet content varying from 5% to 85% and reduce the outlet residue stream carbon dioxide content from 1.5% to 50%. Most of the Cyanara membrane systems that are currently in use are for separating carbon dioxide from raw natural gas streams [10].

3.3.3.2 Separex Membrane Technology

Separex membrane technology is owned by Universal Oil Products (UOP). This technology uses cellulose acetate and polyimide as the membrane material with hollow fiber modules. A membrane unit was installed in Kadanwari, Pakistan in 1995 and was the largest membrane based natural gas processing plant in the world at that time. This is a two stage system that processes 210 MMscfd of inlet flow rate at 90 bars and reduces the carbon dioxide content from 12% to less than 3%. The membrane material used for this system is cellulose acetate with a spiral wound module [11].

Another unit in Qadirpur, Pakistan treats 265 MMscfd of natural gas at 59 bars. The carbon dioxide content in this process is reduced from 6.5% to less than 2%. This system has been constructed in two 50% membrane trains. Each of these trains consists of a pretreatment section and a membrane section. This system also uses cellulose acetate as the membrane material with a spiral wound module [11].

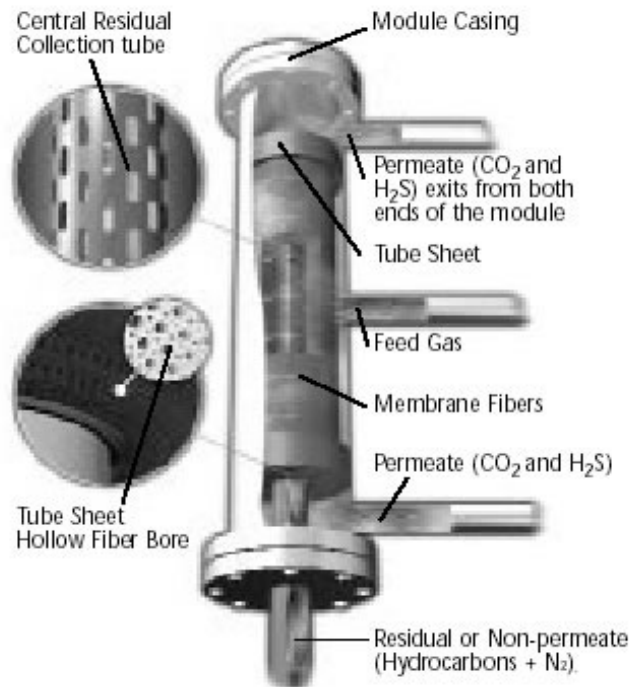


Figure 8. Cyanara membrane unit.

The membrane unit in Taiwan treats a natural gas flow rate of 30 MMscfd at 42 bar. Carbon dioxide content in this process is reduced from 12% to 3%. This membrane unit uses polyimide as the membrane material with a hollow fiber module unlike the units in Pakistan [11]. The membrane separation unit installed in west Texas treats a feed gas flow rate of 30 MMscfd at 42 bar. This process reduces the inlet carbon dioxide content from 30% to 10% [11].

4 Selection of an Anthropogenic Point Source for CO₂

Because this interim report supports the Big Sky Carbon Sequestration Project, only point sources for CO₂ emissions within the region's boundary were considered. Further, point sources near possible geologic sequestration sites were given priority. Because Wyoming has perhaps the best opportunity for geologic sequestration of the states within the region, the selection of a specific anthropogenic CO₂ point source was limited to those within Wyoming.

A cursory investigation revealed that Wyoming has a number of large point source emitters of carbon dioxide, with total emissions of about 63 million tons of carbon dioxide per year. The coal fired power plants listed in Table 1 account for over 80% of this total. The balance of emissions comes from trona processing, petroleum refining, and cement manufacture. Additional CO₂ is available as a byproduct of natural gas processing, but these facilities currently have separation technology in place, and therefore do not fit the criteria of this report.

Table 1. Wyoming coal-fired power plants.

Plant	County	Capacity (MW)	CO ₂ Emissions (Tons/yr)
Jim Bridger	Sweetwater	2,120	18,576,558
Laramie River Station	Platte	1,650	14,442,863
Dave Johnston	Converse	762	7,362,207
Naughton	Lincoln	700	6,012,586
Wyodak	Campbell	335	3,762,075
Neil Simpson II	Campbell	114	1,264,726
Wygen1	Campbell	90	~900,000

In order to compare the CO₂ separation technologies discussed in the previous sections, a flue gas CO₂ point source was selected to which the technologies could be designed. Any one of the above power plants could be used as all of the power plants in the state of Wyoming were adequate for this analysis, but, the Wyodak facility was selected because of its proximity to potential geologic sequestration sites and its relative average size.

The Wyodak pulverized coal power plant was selected as the point source for application of the three separation processes for further design and analysis. The Wyodak power generating facility is a 335 MW net power plant that consumes approximately 6050 tons/day of Powder River Basin coal.^c The plant was built in 1978, with an expected plant life of 45 years. The composition of the flue gas produced at the Wyodak facility is shown in Table 2.

The flow rate of flue gas through the stack is about 78 million ft³/hr at 185°F and atmospheric pressure. Calculated from the composition of the flue gas stream, the amount of CO₂ released by the Wyodak facility is about 10,300 tons/day. The steam generated by the facility is produced at a temperature of 1000°F and a pressure of 1800 psi. The facility utilizes air cooling systems. The heating value of the coal used in the plant averages 7727 Btu/lb and the average carbon content is 46.2% [12]. The plant thermal efficiency was calculated to be 29.3%.

^c Personal communication with Glen Fosher and Bernedett Henshaw – employees of the Wyodak power plant.

Table 2. Flue Gas Composition of Wyodak PC power plant.

Flue gas component	Concentration
N ₂	67.0%
CO ₂	11.8%
O ₂	12.0%
H ₂ O	8.0%
CO	300 ppm
SO ₂	180 ppm
NO _x	150 ppm

5 Design and Analysis of CO₂ Separation Processes

Facilities for each of three CO₂ separation technologies (absorption, adsorption, and membrane) were designed to fit the flue gas output specifications for the Wyodak coal-fired power plant and then compared to each other economically and technically in order to select the most appropriate technology.

The economics done for each technology was based on a discounted cash flow net present value approach. The analysis assumed a 10 year plant life, 10 year MACRS depreciation schedule, an interest rate of 15%, and a tax rate of 35%. Each process was compared by calculating a break-even CO₂ price, which is the calculated CO₂ market price that would return a rate of return equal to the discount rate of 15%. The following sections describe the design and comparisons of the separation processes previously discussed.

5.1 Chemical Absorption Simulation

Simulation of a chemical absorption process was done using Aspen Plus by members of a student team at the University of Wyoming as part of their Chemical Engineering Senior Design class.^d A 30 wt% MEA solution was chosen as the solvent. MEA was chosen because the power plant flue gas used as the input for this process has a low CO₂ partial pressure, and MEA is more reactive at these conditions than other solvents. The flow diagram can be found in Appendix A.

The flue gas composition shown in Table 2 was not simplified; NO_x, SO₂, and CO were included in the analysis. The flue gas entered the process at 185°F and atmospheric pressure with a volumetric flow rate equal to the Wyodak facility's flue gas flow rate of 78 million ft³/hr.

Although the actual process contains a recycle loop for the solvent, the simulation was done with an open loop and iterated by hand until the initial solvent stream was identical to the final solvent stream. This was done to avoid convergence difficulties encountered while attempting to run the simulation with a recycle loop.

The absorber was modeled using a RadFrac column with 40 equilibrium stages and no condenser or reboiler at a pressure of 1.2 bar. The stripper was modeled using a RadFrac column with 15 equilibrium stages including a partial condenser and a reboiler at a pressure of 2 bar. The following two reactions were used in both the stripper and absorber models:



The heat-exchanger was specified to have a 10°F hot outlet-cold inlet temperature approach. The solvent pumps and coolers were specified to operate at the expected operating temperature and pressure conditions of the process at their respective points.

5.1.1 Results

Once a solvent stream match was achieved, the process produced a 97% pure CO₂ stream with 90% recovery of the CO₂ in the flue gas. With the incoming flue gas flow rate, this corresponds

^d Members of the Spring 2006 team were: Ryan Hart, Zachary Tyrrell, Siluni Wickramathilaka, Melissa Binder, Melissa Gangl, and Ashley Madrid. Faculty advisors were: Gordon Harris, David Bell, Maciej Radosz, and George Hu.

to about 5500 tons of CO₂ per day. These results compare favorably with the values reported in the literature.

Using the results from the simulation, the major operating costs for the process were determined. The calculations were compared with values for two commercial processes reported in the literature as shown in Table 3.

Table 3. Operating cost of amine absorption processes in \$/ton of CO₂ recovered.

	Econamine FG Plus [13]	KS-1	Aspen simulation
Fixed Costs	4.56	4.50	4.50
Regeneration Energy (Steam)	7.20	6.00	5.96
Electricity	2.50	3.15	3.59
Cooling Water	0.31	0.50	0.87
Chemicals	2.41	2.10	1.88
Total	16.98	16.25	16.80

Integrating an amine scrubbing process into the steam-cycle of a power plant decreases the overall efficiency of electricity generation. The loss of electricity production must be accounted when calculating the overall operating cost of the process. This loss of efficiency, sometimes referred to as the ‘energy penalty’, is reported to be around 30% for a coal-fired power plant. In the case of Wyodak, this corresponds to a loss of about 100 MW of power production, or about \$17.50/ton of CO₂ recovered, which must be added to the total in Table 3. The total operating cost, excluding compression, is therefore \$34.30/ ton of CO₂.

5.1.2 Equipment Costs and Capital Investment

The major unit operations for the process were sized and priced using a combination of Aspen Plus and tables in Peters et al. [14]. A summary of the major equipment sizes and prices are shown in Table 4.

Table 4. Summary of major equipment sizing and costs for chemical absorption process.

Equipment	Key Sizing Data	Cost
Absorber	42.5 ft (Dia)	\$4,000,000
Stripper Tower	33 ft (Dia)	\$2,000,000
Cond & Reb.	15,600 sq. ft (HT area)	\$500,000
Solvent Pumps (2)	14,000 gpm (flow rate)	\$150,000
Total Heat X	45,000 sq. ft. (HT area)	\$2,000,000
Blowers (10)	100,000 cu ft/min	\$100,000
Total		\$8,750,000

The total capital investment was estimated using scaling factors based on the total equipment cost as described by Peters et al. [14]. The fixed capital investment for the plant was estimated to be \$45 million. Adding in a factor for working capital, the total capital investment was estimated at \$53 million.

5.1.3 Economic Results

The “break-even” price for carbon dioxide was calculated to be \$42.40/ton or \$2.45/Mscf. This price does not include the cost of downstream compression and transportation, which typically add about 20% to the total cost.

5.2 Physical Adsorption Simulation

The first step in analyzing the adsorption process for separating CO₂ from a flue gas stream is to identify the appropriate adsorption isotherm. Plugging the Langmuir parameters into the dual-site Langmuir isotherm equation a curve was generated (shown in Figure 9) showing the capacity of both carbon dioxide (top curve) and nitrogen (lower curve) versus the system pressure.

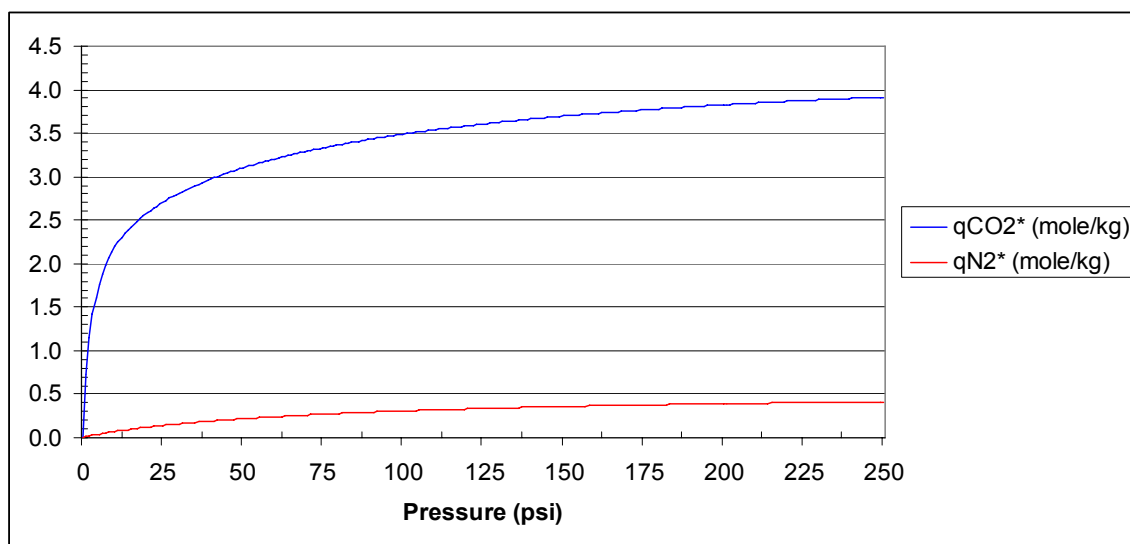


Figure 9. Dual-site Langmuir isotherm for competitive gas adsorption on zeolite 13x.

To evaluate the optimal pressure range, it was necessary to look at the ratio in the change of capacity for carbon dioxide to the change in pressure over the pressure range. This curve is shown below in Figure 10.

From observation of the above curve, it can be seen that below 50 psi the slope of the curve is quite steep, which implies that there is an advantage in increasing the capacity of the adsorbent by increasing the system pressure. Above 100 psi the curve flattens out, which implies that there is a diminishing advantage in increasing the capacity of the adsorbent by increasing pressure range. The optimal adsorption pressure range for the treatment of flue gas with zeolite 13X lies between 50 psi and 100 psi. This agrees well with the fact that typical PSA systems adsorb at pressures of four to six times ambient pressure [15]. For simplicity of design, desorption pressure was taken as ambient pressure, or 14.7 psi.

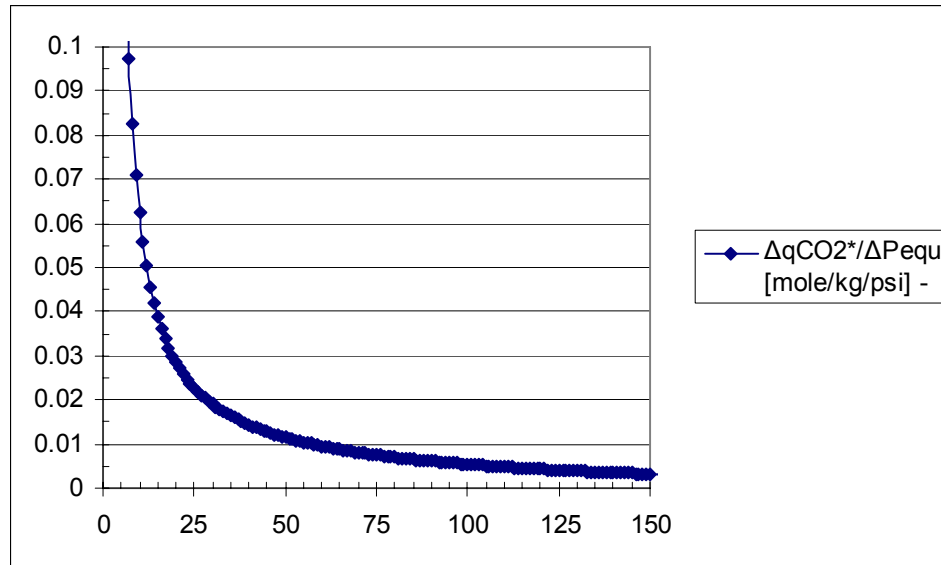


Figure 10. Ratio of capacity to pressure change versus system pressure on zeolite 13x.

The input specifications were evaluated first to determine the amount of carbon dioxide to be removed per cycle. The cycle times were chosen as 25% longer than reported values in order to obtain the most conservative approximations, and recovery of 50% of the inlet concentration of CO_2 was assumed. The input specifications and results are given below in Table 5.

Table 5. Design parameters for of carbon dioxide removal.

$t_{ads} = t_{purge}$ (°F)	$t_{press} = t_{blow}$ (°F)	Cycle time (s)	Flow rate (kmol/s)	y_{CO_2}	Recovery	CO_2 re- moved per cycle (kmole)
300	60	720	27.342	0.0755	50%	743.16

The target carbon dioxide removal was found to be about 740 kmol per cycle. Ten beds were assumed in the process, which is a very typical number for large-scale applications. The lowest bed density for beds containing zeolite 13x was used in order to obtain the most conservative estimate of the process. The results for the bed sizing and zeolite requirement are given below, for the upper and lower limits of the pressure range, in Table 6.

Table 6. Resulting zeolite requirement and bed volumes for zeolite 13x.

Pressure (psi)	q_{CO_2}	Zeolite re- quired (kg)	Density (kg/m^3)	Number of beds	Bed volume (m^3)
50	3.101	240,000	650	10	37
100	3.490	210,000	650	10	33

An important noticeable trend for the zeolite requirement is that the amount of zeolite decreases as the pressure increases because the capacity increases and more carbon dioxide can be removed. This plays an important role in the economic analysis because the synthetic zeolite is quite costly. However, the bed volumes are fairly similar and that equipment cost is relatively small.

With the optimal pressure range identified, it is now necessary to evaluate the compression system. A recovery of 50% was specified for all three separation processes, and because it was assumed that complete adsorption would take place, the flue gas flow rate was cut in half. An upper limit on the compressor duties was taken as 10,000 hp, as that is the upper limit of the appropriate cost curves found in Peters et al. [14]. The compressors were evaluated in Aspen and specified as “polytropic using ASME standard,” with 85% polytropic efficiency and 90% mechanical efficiency. For the 50 psi design, six identical parallel units were used with the flow diagram for one of the parallel trains shown below in Figure 11.

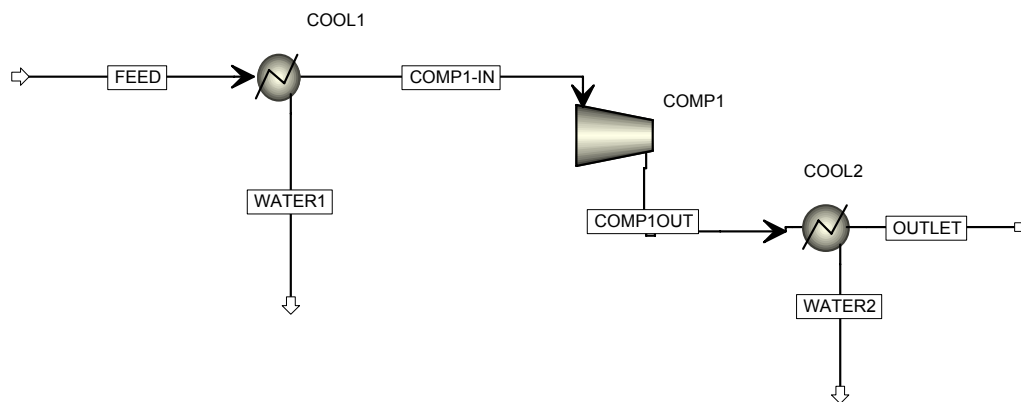


Figure 11. Compressor and heat exchanger unit for 50 psi process.

The outlet pressure of 50 psi could easily be achieved with one compressor in each parallel unit, using specified pressure ratio of each compressor equal to 3.4. Because of the high flue gas temperature, an initial heat exchanger was required to cool the temperature to 80°F, and a second heat exchanger after compression was required to cool the outlet to 77°F before entering the adsorption process. The power requirement of the compressor was found to be 9650 hp. The duty of the heat exchangers was first found assuming a simple cooler, and then the heat exchangers were analyzed in separate Aspen programs for ease of convergence.

5.2.1 Equipment Costs and Capital Investment

Carbon steel was used for all adsorber materials because all of the water that is condensed out of the flue gas stream in each heat exchanger. For the adsorber beds, a diameter of one meter was assumed as this value produced the most realistic bed lengths. The equipment pricing tables in Peters et al. [14] were used to price the equipment. The adsorber beds were priced as distillation towers and factored by 1.6 for high pressure units, found in Figure 15-11 on page 793. The compressors were priced as the rotary type by the hp requirement, using Figure 12-28 on page 531. The heat exchangers were priced based on area, found in Figure 14-16 on page 681. The resulting 50 psi equipment price can be found below in Table 7.

Table 7. Purchased equipment cost for 50 psi process.

	Characterization	Number of units	Cost per unit (2002 dollars)	Total cost
Adsorbers	Distillation towers	10	\$40,000	\$400,000
Adsorbent	Zeolite 13X (lbs)	528,285	\$32.65	\$17,200,000
Cooler 1	8220 ft ²	6	\$90,000	\$540,000
Cooler 2	2760 ft ²	6	\$30,000	\$180,000
Compressor	9650 hp	6	\$1,500,000	\$9,000,000
Total (adsorbent cost not included)				\$10,100,000
2005 total cost				\$12,100,000

The adsorbent was priced using “Fischer Scientific Online,” and the cheapest price for the zeolite 13x that could be found was used. It should be noted that this price is very high, and when buying in the large amounts that are required for this process the price could decrease substantially. The adsorbent price was not included in the purchased equipment cost, as this cost is used to determine the direct and indirect costs. The adsorbent cost was instead included in the working capital, to be discussed later. The 2002 purchased equipment price is shown in Table 8 [16].

Table 8. Purchased equipment cost for 100 psi process.

	Characterization	Number of units	Cost per unit (2002 dollars)	Total cost
Adsorbers	Distillation towers	10	\$40,000	\$400,000
Adsorbent	Zeolite 13X (lbs)	469,468	\$32.65	\$15,300,000
Cooler 1	8220 ft ²	6	\$90,000	\$540,000
Cooler 2	2270 ft ²	6	\$30,000	\$180,000
Cooler 3	1870 ft ²	6	\$20,000	\$120,000
Compressor 1	7220 hp	6	\$1,000,000	\$6,000,000
Compressor 2	7070 hp	6	\$1,000,000	\$6,000,000
Total (adsorbent cost not included)				\$13,200,000
2005 total cost				\$15,800,000

The adsorbent price was not included in the purchased equipment cost, but rather as working capital. The 2002 price was escalated to 2005 price using the same price indices. The purchased equipment costs for the two processes do not differ greatly. However, the major difference comes from calculating the operating costs associated with the two situations. The hours of operation were found by assuming a down time of two weeks per year for maintenance. The yearly cost of compression for the 50 psi process was found to be \$25.4 million, with the yearly cooling water costs nearly \$7.5 million. The cost of compression for the 100 psi process were found to be \$37.6 million, with cooling water costing \$8.1 million.

The yearly operating cost for the 100 psi process is nearly \$13 million more than for the 50 psi process. Because the two process both produce the same amount of carbon dioxide, found to be 1.2 million tons per year, they both make the same profit. By assuming that carbon dioxide sells

for \$0.35 per thousand cubic feet, the profit for both processes was found to be \$7.2 million per year.

The capital investment for the two processes was found by applying scaling factors based on purchased equipment cost, the same method as was applied to the chemical absorption simulation. The fixed capital investment for the 50 psi process was found to be \$53.8 million, with a total capital investment that includes the adsorbent price in the working capital found to be \$82.8 million. The fixed capital investment for the 100 psi process was found to be \$70.2 million, with a total capital investment found to be \$101 million. By inspection, it is known that the operating costs for both processes greatly exceed the profit from carbon dioxide, so a break-even point was calculated for both.

5.2.2 Economic results

Using the same assumptions as for the chemical absorption process, the break even point for the two processes was calculated. The break-even point for the 50 psi process was found to be about \$49 per ton of CO₂, or about \$2.80 per thousand cubic feet. The break-even point for the 100 psi process was found to be about \$64 per ton of CO₂, or about \$3.70 per thousand cubic feet. These values are much higher than even the most liberal estimations of the current price of carbon dioxide, at about \$1.00 per thousand cubic feet. These prices don't include transportation costs, which generally constitute 20% of the total price for carbon dioxide.

5.3 Membrane Separation Simulation

The four main models used in membrane calculations are: complete mixing, cross flow, countercurrent flow, and cocurrent flow. The complete mixing model was used to simplify the membrane unit calculations and is shown in Figure 12. The assumptions made in this model are that the gas molecules entering the membrane unit mix completely within the unit and also that the gas molecules permeate through the membrane material uniformly. Since most of the water in the flue gas stream was condensed in the compressor and heat exchanger unit before sending it to the membrane unit, the gas stream was treated as a three component system where component A was CO₂, component B was O₂, and component C was N₂ and the remaining H₂O. The selectivity of carbon dioxide with respect to water is very large; therefore, it was assumed that all of the remaining water that enters the membrane unit does not permeate and goes to the residue stream and is purged into the air.

An iterative procedure was followed to calculate the necessary parameters for the membrane unit [17].

Recall that each system (chemical absorption, physical adsorption, and the membrane system) was to achieve a 50% recovery of the CO₂ in the flue gas stream to be able to compare the economics of each of the technologies. Accordingly, a membrane system that achieves a 50% recovery was designed first and is shown in Figure 13.

The calculated parameters for the membrane system with 50% recovery are as given below. The required membrane area to achieve the 50% recovery of carbon dioxide is about 36,500 m² with a 15% stage cut. However, the purity of carbon dioxide in this permeate stream was only about 30%, which is too low for sequestration purposes.

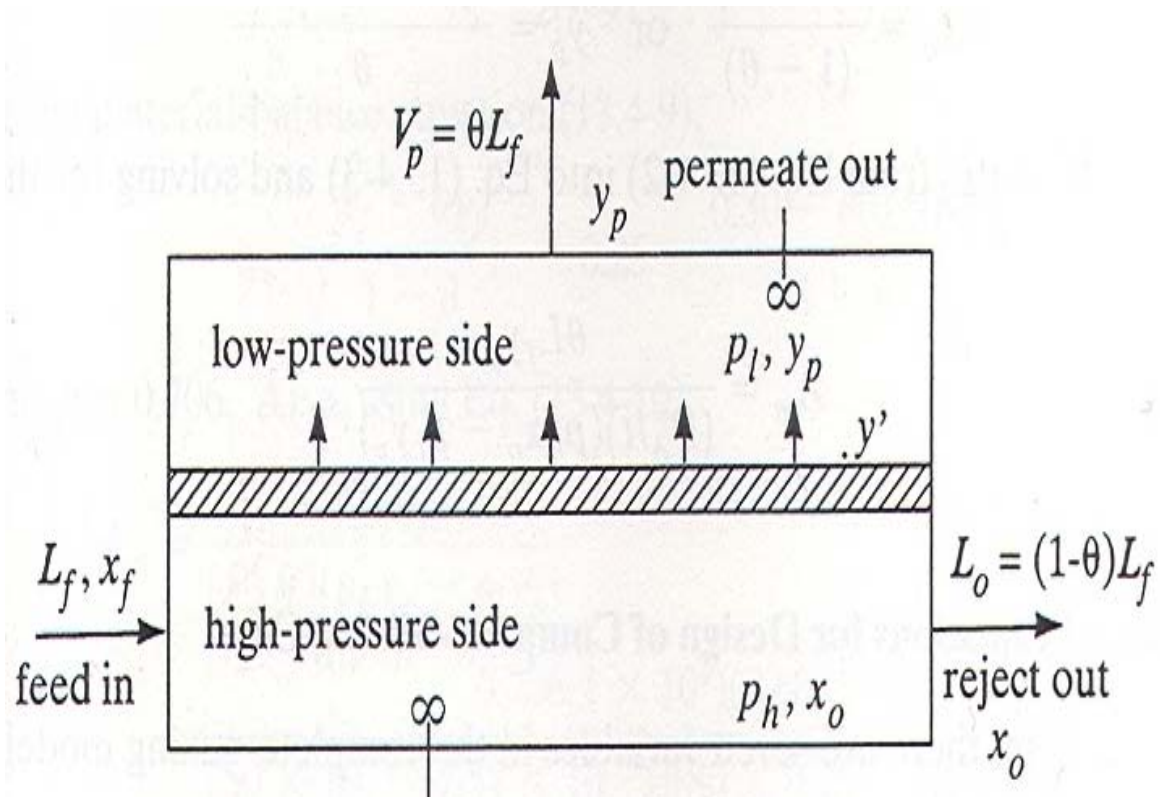


Figure 12. Complete mixing module.

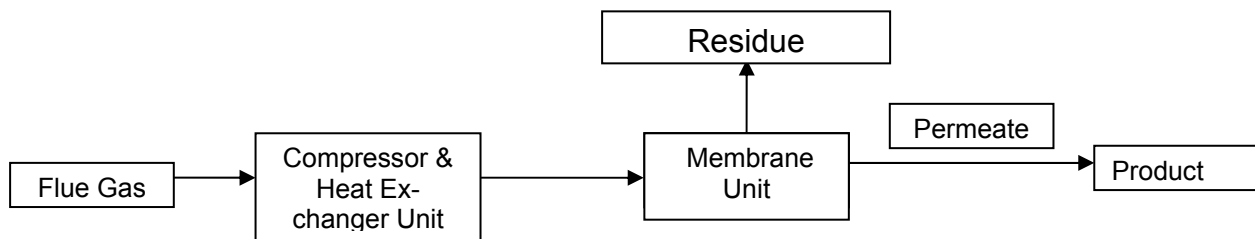


Figure 13. Membrane system for 50% recovery.

The project actually needs a resulting CO₂ stream of at least 98% purity. An additional membrane system that can achieve 98% purity was designed for comparison as well. The schematic of this system is shown in Figure 14.

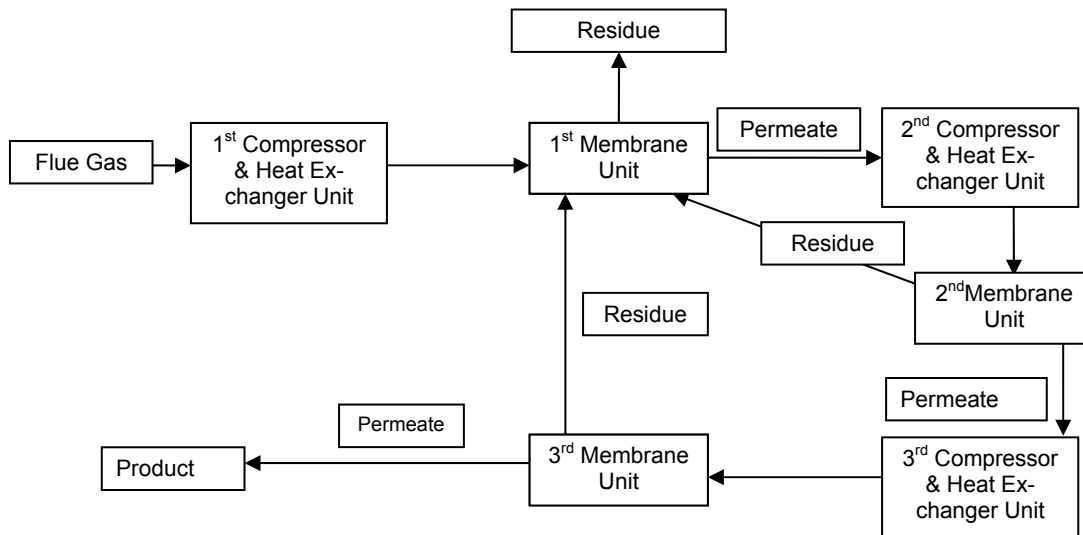


Figure 14. Membrane system for 98% purity.

The required membrane areas for the three membrane units to achieve the 98% purity of carbon dioxide are about 53,900 m², 1,440 m², and 50.4 m² respectively with 20%, 10%, and 10% stage cut each. The recoveries from the three units were about 59%, 24%, and 12% achieving an overall recovery of only about 2% of the CO₂ of the flue gas stream.

5.3.1 Design of Membrane Systems

The flue gas stream was cooled from 185°F to 120°F and split into six streams. A 1:3 pressure ratio was used for each of the six compressors, which were configured into 2 parallel sets for both membrane options, to achieve the pressure needed for the membrane separation unit. The final stream leading into the membrane system was cooled to 77°F. Subsequent compressor and heat exchange units were designed in the same manner, but were somewhat smaller units because of lower flow rates.

The main capital components of the membrane systems were the membrane casing, compressors, and heat exchangers. The membrane material costs were not included in the equipment cost, but were included in the working capital. The compressor costs were estimated using the compressor horsepower requirement for centrifugal rotary compressors. Carbon steel was used for the compressors because no water was anticipated to condense. The heat exchangers were assumed to be shell and tube heat exchangers.

5.3.2 Equipment Costs and Capital Investment

Equipment costs for the mixers and splitters were small compared to the other equipment costs and are included in the piping costs calculated as a factor of the total purchased equipment cost. The total purchased equipment costs were estimated using the Chemical Engineering Index in 2005\$. The total equipment cost for the 50% recovery membrane system was \$26.1 million and it was \$26.3 million for the 98% purity membrane system. The compressor cost for the first system was \$25.2 million and it was \$25.3 million for the second system. So the majority of the total equipment costs were due to the compressor costs.

The first compressor and heat exchanger unit has a high purchased equipment cost due to the high horsepower requirements in the compressors and the high cooling requirements in the heat exchangers. This was mainly due to the very high volumetric flow rate of flue gas that requires more compression and cooling. Since part of the flue gas stream was purged into the air as the residue stream after the first membrane unit, the second and third compressor and heat exchanger units had smaller equipment costs.

The capital investment for the membrane plant was calculated using the same method as the chemical absorption plant. The fixed capital investment needed for the first membrane system was about \$134 million while the fixed capital investment needed for the second system was \$135 million. The working capital was estimated by multiplying the total purchased equipment cost by a factor of 0.89 and adding the membrane material costs. The membrane material costs were estimated to be \$75/m² [18]. The total capital investment was calculated by summing the fixed capital investment and the working capital. So the total capital investment needed was \$160 million and \$162 million for each of the membrane systems.

5.3.3 Economic Results

The same assumptions were used as with the chemical absorption process to complete the break-even analysis. The main operating costs for the membrane systems were the electricity and cooling water costs. The electricity costs were estimated by using the kW requirements from the compressors and then calculating the kWh requirement for the year and using a value of \$0.0525/kWh. The electricity costs for the first membrane system was about \$M63.4/year and \$M63.6/year for the second membrane system. The cooling water costs were estimated using the cooling water flow rate (m³/s) and then calculating the flow rate (m³) requirement for the year and using a value of \$0.045/m³. The cooling water costs were \$10.3 million per year for each of the membrane systems. This resulted in a total operating cost of about \$73.7 million per year for the first unit and about \$73.9 million per year for the second unit. These operating costs were included in the net present worth calculations for each year. The sales for each year were calculated by multiplying the carbon dioxide production rate in tons/ year by the carbon dioxide sale price in \$/ton

A random carbon dioxide sale price was chosen initially and the sales were estimated by multiplying the particular cell with the sale price by the production rate. The cash flow and the net present worth were calculated in an Excel™ spreadsheet and then the summation of the net present worth was set to be zero in the “Goal Seek” function by changing the cell containing the carbon dioxide sale price. The above method was followed to find the carbon dioxide break – even point for each of the membrane systems designed. The first membrane system had a production rate of 922,000 tons/year and a break – even point of \$109/ton or \$6.30/Mscf. The second membrane system had a production rate of 48,200 tons/year and a break – even point of \$2,090/ton or \$121/Mscf.

5.4 Conclusions

Process designs were successfully carried out for chemical absorption, adsorption, and membrane separation systems. Investigation revealed that while all processes removed substantial amounts of CO₂ from the large point source emitter, only the chemical absorption and adsorption processes were found to be technically feasible for flue gas application at this point. The economic analyses are shown in Table 9.

Table 9. Comparison of separation process simulations.

Separation Process	CO ₂ Production (t/yr)	Capital Invest- ment (\$ million)	CO ₂ Break-Even Price (\$/t)
30 wt% MEA Absorption	2,000,000	53	42
Zeolite 13X Adsorption-50 psi	1,200,000	83	49
Zeolite 13X Adsorption-100 psi	1,200,000	101	64
Membrane-50% recovery	922,000	160	109
Membrane-98% purity	48,000	162	2090

Economic results showed that the chemical absorption process produces the least costly CO₂, with the membrane separation process producing the most expensive. It seems fairly clear, however, that CO₂ sales alone will be insufficient to make any CO₂ flue gas separation economically attractive. As the technology of emerging processes such as adsorption and membrane separation is improved through research efforts, the cost of separation could potentially be driven down to levels that would make separation economical in some situations.

6 Analysis of CO₂ Pipeline Transportation

The second goal of this interim report was to analyze the technical issues and costs associated with the construction of a CO₂ pipeline from a point source emitter to a suitable geologic sequestration site. The hypothetical pipeline would be designed for an actual route leading from the Wyodak coal-fired power plant to a real potential sequestration site near the power plant. The pipeline would be designed to handle the CO₂ produced from the separations work reported earlier in this report.

6.1 Pipeline Route Selection

For this scenario, a 47-mile pipeline is assumed and based on a route from the Wyodak power plant near Gillette, Wyoming to Hartzog Draw oil field – a potential CO₂ sequestration site. Figure 15 is a map showing the selected route of the pipeline. The pipeline route was chosen to have as few major pipeline and paved highway crossings as possible. The selected route passes through Burrowing Owl habitat, which would probably require an impact review to the U.S. Fish and Wildlife service to get construction approval. The pipeline route is about 47 miles with an elevation gain of about 690 ft.

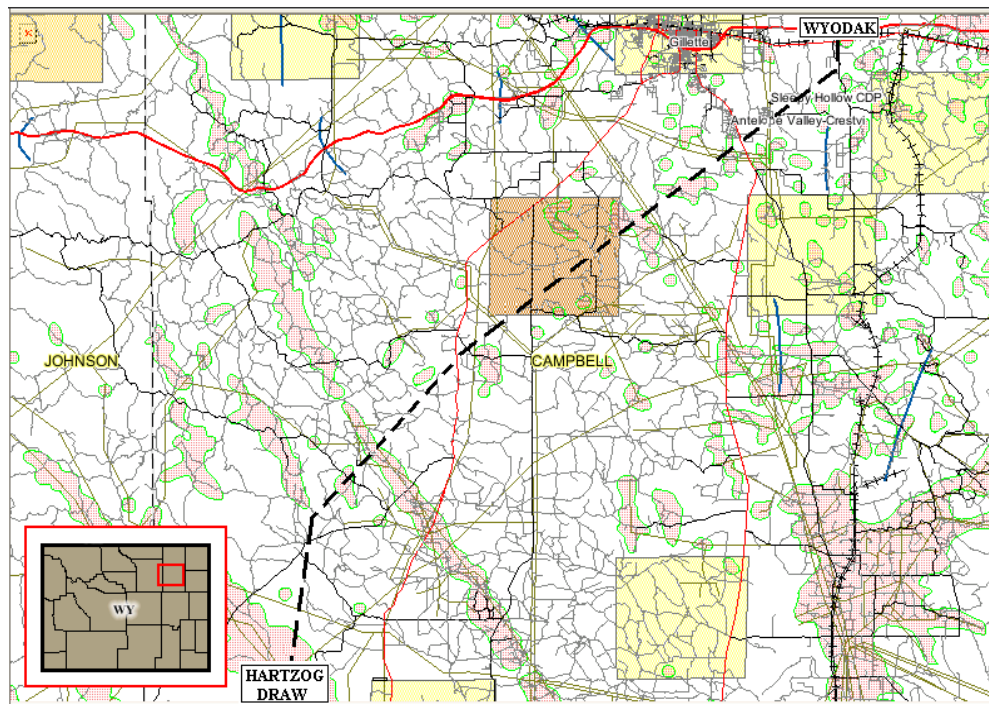


Figure 15. Pipeline route selection. The map shows major pipelines and roads. The shaded boxes indicate areas of threatened or endangers species.

After the initial pipeline route selection has been completed, the next step is to survey the entire route to make sure that there are not any major obstructions that would require changing the entire route selection. This survey is usually done by flying over the intended route to see what obstructions may occur. If obstructions are going to be a problem, a closer survey will be done on site to see if it is possible to move the center of the pipeline left or right by no more that

0.5 miles to get around the obstruction. If this is not possible then a new route selection is required before the permitting process can begin [19].

To transport the CO₂ from the point of capture to the site of storage, high-pressure pipelines are typically used. Carbon dioxide pipelines have been in use for many years. One of the most recent constructed was the Weyburn CO₂ pipeline from Beulah, North Dakota, to the Weyburn oil field in Saskatchewan [20].

6.2 Pipeline Design Simulation

For efficient pipeline transportation, supercritical CO₂ will be needed over the entire distance of the pipeline. A large compressor system will be required to compress the source CO₂, (leaving the point source at atmospheric pressure). A 4-stage reciprocating compressor was selected for this requirement based on the facilities at use for transporting CO₂ from the LaBarge, Wyoming CO₂ treatment plant. Inter-stage air-cooling is also utilized for the design specifications. This cooling process will allow the compressors to operate at a lower temperature therefore increasing the work life and also the overall efficiency. Cooling will also allow the exit temperature of CO₂ to be reduced which will help with the economics of the pipeline materials.

The 4-stage compressor cycle was simulated using Aspen Tech using a typical industry compression ratio of 3.25. At the inlet of the compressor train there is 8.9 million ft³/hr CO₂ with a pressure and temperature of 14.7 psi and 77°F respectively. CO₂ is then compressed to 0.0317 million ft³/hr with a pressure and temperature of 1,640 psi and 120°F respectively. Figure 16 shown below is the flow sheet from Aspen Tech.

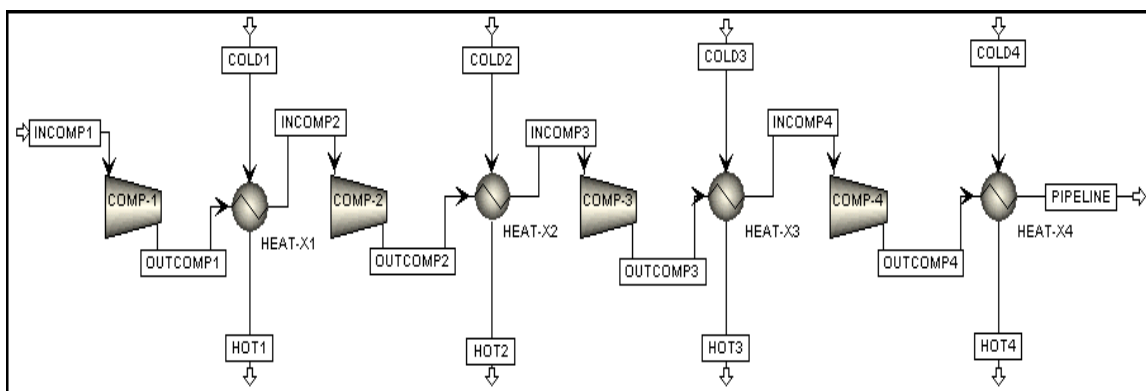


Figure 16. Process design flow chart for a 4-stage reciprocating compressor with air-cooled inter-stage cooling.

Using correlations of cost per kW, the PEC with delivery of the 4-stage reciprocating compressor is estimated to be \$40 million. To estimate the FCI, a cost based on the PEC was used. Indirect costs include engineering and supervision (32% PEC), construction expenses (34% PEC), legal expenses (4% PEC), contractor's fees (19% PEC), and contingency (37% PEC) [21]. With these additional costs, the FCI is estimated to be \$339 million.

The four interstage air-coolers were priced separately. Each cooling stage required a different duty for cooling and therefore prices of the four coolers vary. To price the coolers accurately,

the heat transfer surface area was hand calculated based on values from the Aspen Tech and using Eq. 5 results and using a tabulated heat transfer coefficient of air.

$$Q = UA\Delta T \dots\dots\dots (5)$$

Where:

- Q = heat generated (BTU/hr)
- U = heat transfer coefficient (BTU/ft²-°F-hr)
- A = required heat transfer area (ft²)
- ΔT = (T₂-T₁) = change in temperature (°F)

Cost analysis was correlated using a cost per ft² index. The PEC with delivery of all four cooling stages is estimated to be \$480,000. To estimate the FCI, a cost based on the PEC was used. Indirect costs include engineering and supervision (32% PEC), construction expenses (34% PEC), legal expenses (4% PEC), contractor's fees (19% PEC), and contingency (37% PEC) ¹². With these additional costs, the FCI is estimated to be \$3.1 M.

6.2.1 Pipeline Design Considerations

This part of the state does not have large mountain ranges that the pipeline needs to go through, so this helped lower the cost of constructing the pipeline because intermediate recompression was not required based on the Pipesoft simulation and hand calculations.

There are many design parameters that need to be considered in designing a pipeline. One of these is the operating pressure of the pipe. The temperature, gas composition, gas corrosivity, ambient temperatures and pipeline control are also important design considerations. The CO₂ will be transported as a supercritical fluid in order to maximize the volume that can be transported. The critical point for CO₂ is 88°F and 1070 psia. Typical pressures in other CO₂ pipelines are 1500 to 1800 psia [22]. Figure 25 shown below, shows the pressure distribution of the pipeline. This plot was created using the Pipesoft software after a successful simulation of the pipeline. The operating temperature of the pipe would be ambient conditions of 65°F.

The gas composition depends on how good of a separation can be achieved from the flue gas of the power plant. Using MEA for separating, the CO₂ can be brought to a 98% pure stream. Contaminants include H₂O, SO_x, NO_x, methane, and traces of other gases. For our design we assumed 100% pure CO₂ for the entire stream. We would expect to have about 95% CO₂ with a make up of 5% N₂ if the pipeline was actually built. It would be very unlikely that there would be any H₂O in the stream because of the strict contracts between companies that prohibit H₂O in the pipeline. This is because CO₂ is an acid gas and it will react with water to form carbonic acid. Carbonic acid is corrosive to carbon steel pipes, and causes damage referred to as sweet gas corrosion. If this corrosion is not caught as soon as possible, the lifetime of the pipeline will be greatly reduced and it will have a negative impact on the overall economics. Because of these problems, the DOT has set additional corrosion guidelines that must be followed. Any buried pipeline installed after July 31, 1971 must have an external protective coating. Also, a cathodic protective system must be designed, installed and operating within one year after completion of construction [19]. Proper precautions for internal pipeline must also be considered. Because an acid gas is being transported through the pipeline, corrosion must be checked biannually at intervals no greater than 7.5 months [19]. Adding inhibitors or preventative coating to the inside surface of the pipeline via pigging can reduce this corrosion.

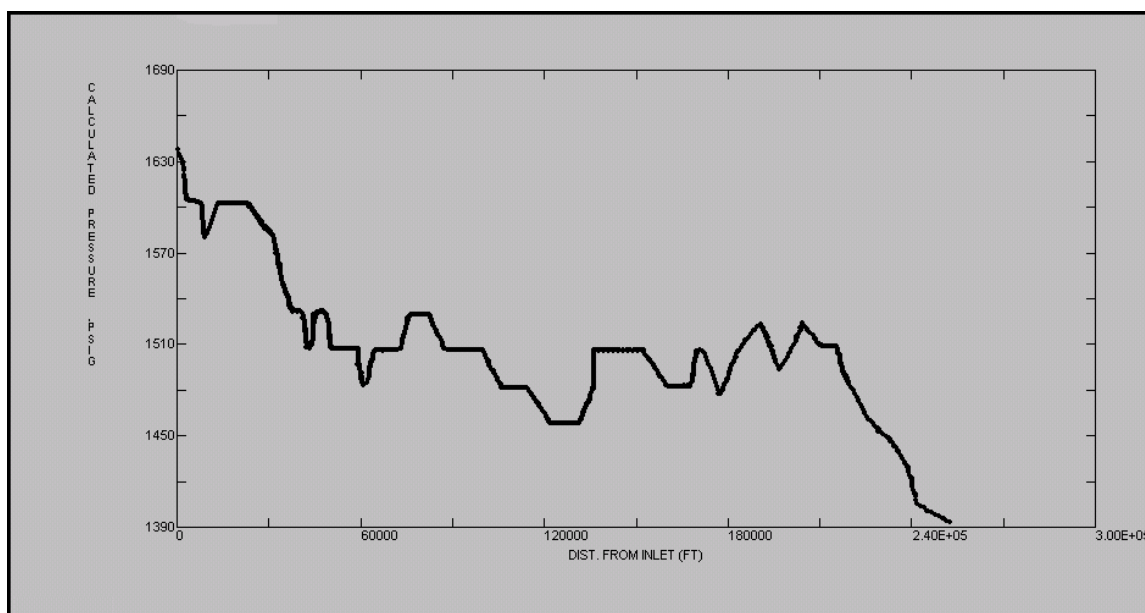


Figure 17. Pressure distribution as a function of distance from pipeline inlet created using Pipesoft simulation engine.

Pipeline control is important for providing continuous monitoring of the pipeline system. The system should be able to respond quickly to emergency situations. It should monitor key parameters such as the operating temperature and pressure of the pipeline, as well as the flow rate of gas through the pipeline system. The control system should be able to detect leaks and also provide predictions on the fatigue of the system [19]. In order to provide for safety, block valves should be placed along the pipeline about every 10 miles per DOT regulations [19]. This prevents loss of pressure integrity of the entire pipeline, whereby sections can be isolated, limiting the amount of CO₂ released into the atmosphere. The ability of CO₂ to collect in depressions and other low-lying areas near the pipeline present a significant hazard if leaks continue undetected. CO₂ can cause asphyxiation if in high enough concentrations. Normal CO₂ concentrations are on the order of 300 ppm.

Another concern in pipeline design is the formation of hydrates in the pipeline. These can build up in the pipeline and cause the flow rate to drop or block off the pipe. With low percentage of water and methane in the system, this should not be a major problem but it must be addressed. Scrapers or “pigs” can be used to clean out the inside of the pipe. Pig launchers and receivers have been placed every 10 miles in order to speed up the pigging process and make shut down as short as possible. Injecting inhibitors such as methanol or salt solutions during the pigging process can minimize the hydrate formation.

Transportation of CO₂ is an important focus for determining the cost and economic feasibility for sequestration. Transportation costs of CO₂ often depend on matching CO₂ sources and storage locations, generating considerable variation in costs.

The optimum pipeline design under the required conditions was calculated to be an 8 inch pipe using an economic optimum Eq. 6, shown below.

$$D_{i,opt} = 0.363 m_V^{0.45} \rho^{0.13} \dots\dots\dots (6)$$

Where:

- $D_{i,opt}$ = economic inside diameter (inches)
- m_V = required volumetric flow rate to Harzog Draw (MMscf/day)
- ρ = supercritical density of CO₂ (lb_m/ft³)

The Barlow and Lamé equation [23] (below) was used to calculate the required wall thickness under the required condition. The wall thickness was then calculated to be 0.312 inches yielding a schedule 40 pipe.

$$t_m = \frac{PD_o}{2(SE + PY)} + C \dots\dots\dots (7)$$

Where:

- t_m = pipe wall thickness (inches)
- P = design pressure (psia)
- D_o = outside diameter of pipe (inches)
- S = basic allowable stress
- E = quality factor
- Y = yield strength

The Weyburn project used carbon steel pipe and used schedule 40 for most of the pipeline and schedule 80 for special cases such as under roads [19]. The same considerations will need to be taken for this pipeline as it will be crossing some major highways and pre-existing pipelines. Using correlations for cost per foot of pipe, the PEC with delivery of using schedule 40 pipe totaled \$6.9 M. These costs included 4 block valves for the pipeline, 2 valves for take-off and receiving and all additional valves needed for the six pigging and receiving systems. The cost of scrapers, pigs, pig barrels and additional piping are negligible compared to the total piping and instrument costs, and therefore neglected in the economic analysis. To estimate the FCI, a cost based on the PEC was used. Indirect costs include engineering and supervision (32% PEC), construction expenses (34% PEC), legal expenses (4% PEC), contractor's fees (19% PEC), and contingency (37% PEC) [21].

6.2.1.1 Permitting

The final part of the design requires investigation of the regulatory aspects and permit process of pipeline design. Permission must be obtained based on what kind of land the pipeline crosses. The land could BLM, state, or privately owned. The BLM requires addressing of certain Critical Elements of the Human Environment that will be affected. Some of these include air and water quality, threatened or endangered species, farmlands, paleontological resources, Native American religious concerns, scenic rivers and wilderness, and other areas of critical environmental concern. These issues will be formally addressed in future work.

The permitting, approval or action which must be granted to either sequester CO₂ geologically or construct a pipeline requires an extensive effort. The designed pipeline from Wyodak to Harzog Draw, and CO₂ sequestration in both the Fox Hill and the Wyodak-Anderson coal zone

would make use primarily of state, private, and Bureau of Land Management (BLM) Wyoming lands. All three aspects of this paper's research have some permitting and approval in common. Pipeline construction, coal, and saline aquifer disposal would need to complete environmental impact statements of the affected areas. Use permits would need to be obtained, lending access to roads, allowing oversized vehicles access to roads and construction to occur. Decision Record for Proposed Action would need to be on file at multiple agencies. None of the aforementioned projects require any permitting or action through the U. S. Forest Service, Army Corps of Engineers, or Bureau of Indian Affairs. Pipeline construction and drilling permitting can be very different because they create very different types of disturbance. Approval of Plan of Development (APD) is required for drilling. Right-of-way to lands and roads is needed for pipeline construction. Water disposal and treatment is an issue for CBM production but not for either pipeline construction or brine injection. Oil and gas leases, flaring/venting gas authorization, and permits to appropriate groundwater would be required for ECBMR/CO₂ sequestration.

6.2.1.2 Safety

Major safety concerns with transportation of CO₂ are leaks, either eruptive releases or slow seepages over time. Failing valves or line blockages also pose a threat of disrupting flow. Geologic sequestration of CO₂ poses safety concerns associated with leaks into unintended zones and induced seismic activity. CO₂ which leaks into unintended zones may leak back into the atmosphere, threatening human health and safety. Hazard and operability studies (HAZOPS) were constructed for the compression with cooling and the pipeline. HAZOPS were not used for analyzing risk of CO₂ injection. A Monte Carlo simulation is commonly the way risk in sequestration is analyzed because there are so many failure scenarios and contributing factors, and should therefore be used to assess CO₂ sequestration risk in future work.

6.3 Discussion

If the proposed pipeline system to transport CO₂ from Wyodak to Hartzog Draw is constructed in the future, there will be some important considerations to be addressed. Although the route was selected entirely on economics, it would be helpful at least to the permitting process if the pipeline is re-routed around the "Burrowing Owl" habitat. This would most likely be more expensive in the long run, but considering the surrounding habitats pose no problems for the pipeline, it would be just as easy to build around the habitat.

The compressor system that is being used for the pipeline is very large and expensive. A smaller design may be able to compress the CO₂ to supercritical conditions and transport it the entire 47 miles without the need of recompression. If another stage of compression is needed along the pipeline the economics could still be reduced if the compressors are smaller than the proposed compressor currently in use. The economics that have been reported for this design are probably worst case figures. If other possibilities are researched the project would be much more attractive from a financial standpoint. However, the design that was used for the compressor is typical in the CO₂ industry.

For the material used on the pipeline, a carbon steel X-52 was selected based on its ability to withstand high temperatures and pressures. The only problem with using carbon steel for the transportation of CO₂ is the potential for corrosion and eventual failure along the pipeline. The pipeline that was designed utilized and corrosion factor of 10%. This assumption could be too high or too low. The economics of the purchased steel are a strong function of wall thickness

and diameter. Figure 26 below shows how sensitive the wall thickness is to different assumed corrosion factors.

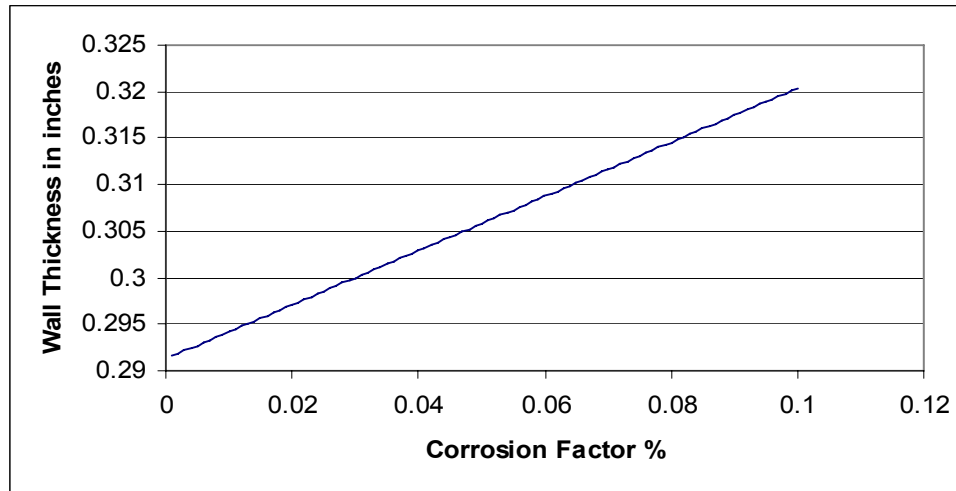


Fig.26 -Sensitivity plot of corrosion versus wall thickness

6.4 Summary and Recommendations

With these additional costs the FCI for a schedule 40 pipeline totals \$39.5 M. The final design would most likely incorporate both schedule 40 and schedule 80 types of piping due to highway and pipeline crossings; therefore these numbers represent a lower and upper limiting case for the overall pipeline design.

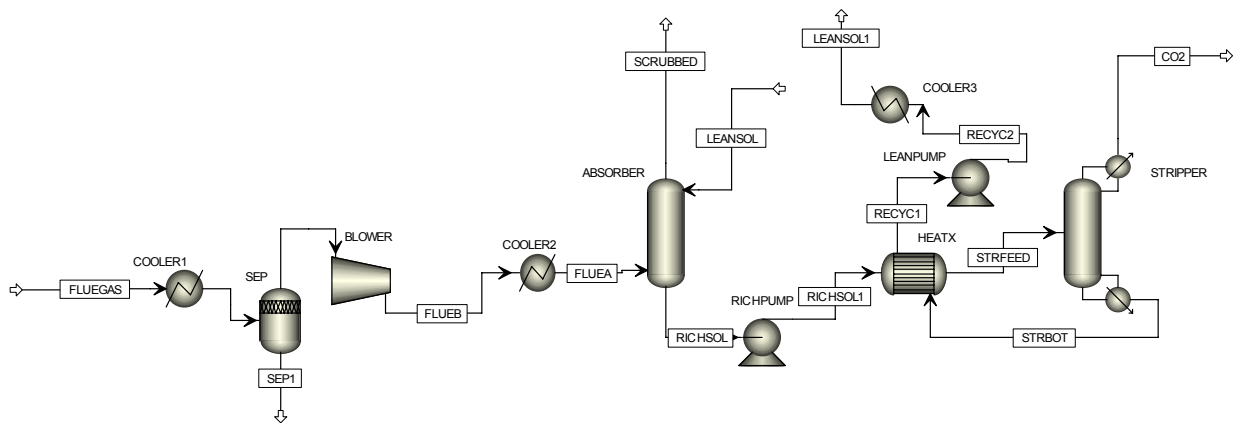
7 References

1. Barchas, R.; Davis, R. The Kerr-McGee/ABB Lummus Crest Technology for the Recovery of Carbon Dioxide from Stack Gases. *Energy Conversion and Management* (1992), 33(5-8), 333-40.
2. Siriwardane, Ranjani, et al. "Adsorption and Desorption of CO₂ on Solid Sorbents." U.S. Department of Energy web page (2001).
3. Suzuki, Motoyuki. *Fundamentals of Adsorption*. Japan: Kodansha Ltd., 1993.
4. Yang, Ralph. *Gas Separation by Adsorption Processes*. Stoneham, MA: Butterworth Publishers, 1987.
5. Jain, S., et al. "Heuristic Design of Pressure Swing Adsorption: A Preliminary Study." *Separation and Purification Technology* **33** (2003): 25-43.
6. Kikkinides, E.S., et al. "Concentration and Recovery of CO₂ from Flue Gas by Pressure Swing Adsorption." *Ind. Eng. Chem. Res.* 32(1993): 2714-2720.
7. Ko, Daeho, et al. "Optimization of Pressure Swing Adsorption and Fractionated Pressure Swing Adsorption Processes for CO₂ Sequestration." 2004 AIChE Annual Meeting (2004).
8. Seader, J. D., and Ernest J. Henley. *Separation Process Principles*. United States: Wiley, 1998.
9. Baker, Richard W. *Membrane Technology and Applications*. Chapter 8, John Wiley & Sons, Ltd. (2nd Edition).
10. NATCO (National Tank Company), NATCO Canada, Cyanara, Mozley, TEST Automation & Controls: <http://www.natcogroup.com/Content.asp?t=CyanaraMembraneSeparationSystems>.
11. Universal Oil Products web sites: <<http://www.uop.com/objects/84CO2RemvbyMembrn.pdf>> and <<http://www.uop.com/objects/98%20Separex.pdf>>.
12. Robertson, E.P.: "Measurement and Modeling of Sorption-Induced Strain and Permeability Changes in Coal," Ph.D. Thesis, Colorado School of Mines, Golden, Colorado (December 2005).
13. Chapel, D.; Mariz, C. Recovery of CO₂ from Flue Gases: Commercial Trends. Presented at the Canadian Society of Chemical Engineers Annual Meeting. October 4-6, 1999.
14. Peters, Max, Klaus Timmerhaus, and Ronald West. *Plant Design and Economics for Chemical Engineers*, 5th edition, New York: McGraw-Hill, 2003.
15. Chue, K.T., et al. "Comparison of Activated Carbon and Zeolite 13x for CO₂ Recovery from Flue Gas by Pressure Swing Adsorption." *Ind. Eng. Chem. Res.* **34** (1995): 591-598.
16. AIChE Journal, November 2005.
17. Geankoplis, Christie John *Transport Processes and Separation Process Principles*. Chapter 13, Prentice – Hall, 2003 (4th Edition)
18. Jain, R. 1989. Method for economic evaluation of membrane – based air separation. *Gas Sep. Purif.* 3 (3):123-127.
19. U.S. DOT, 2004, Transportation of Natural and Other Gas by Pipeline: Minimum Federal Safety Standards, Title-49, Part 192, Volume 3.
20. U.S. DOE Office of Fossil Energy, 2004, Weyburn carbon dioxide sequestration project, Project Facts, www.netl.doe.gov.
21. Peters, M., Timmerhaus, K., and West, R., 2003, *Plant Design and Economics for Chemical Engineers*, Fifth Edition, McGraw-Hill Chemical Engineering Series.
22. Petro Source, 2005, Petro Source Carbon Dioxide Pipeline Project Environmental Assessment, <http://www.wy.blm.gov/nepa/cfodocs/petro>

-
23. Perry, R., Green, D., 1997, Perry's Chemical Engineers' Handbook, Seventh Edition, McGraw-Hill Chemical Engineering Series.

8 Appendix A

8.1 Chemical Absorption Flow Diagram



8.2 Membrane Process Flow Diagram (1st Unit)

