Carbon Dioxide Capture from Large Point Sources

Phase I Final Report

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Compact Membrane Systems, Inc.
Wilmington, DE

Abstract

Capture of carbon dioxide at the source of its emission has been a major focus in greenhouse gas emission control. Current technologies used for capturing CO_2 suffer from inefficient mass transfer and poor economics. There is a need for compact and economic technologies for removal of CO_2 from stack gases.

The present technology proposed a stable amine based membrane contactor for CO_2 removal system from flue gas streams from large point sources with improved mass transfer rates and lower cost compared to conventional technologies.

Phase I program significantly exceeded objectives. Accomplishments included demonstration of both CO_2 absorption and desorption into diethanolamine in membrane contactors at both high temperature (60C) and low temperature (25C). Integrated system successfully absorbed and desorbed CO_2 simultaneously. Preliminary testing indicated the membranes are chemically compatible with amines. Economic analysis showed Compact Membrane Systems' amine membrane contactor had significantly superior economics than commercial amine scrubbing towers.

1. Introduction

Fossil fuels take care of more than 85% of world's energy needs. However, combustion of fossil fuels generates about 80% of all anthropogenic emissions of CO_2. CO_2 as greenhouse gas is considered to be a major contributor to global warming. There is a concerted effort towards capturing CO_2 at source of its large scale generation and sequestering. However, with existing technologies for CO_2 capture, cost of energy generation would increase significantly if CO_2 capture is mandated. For example, capturing CO_2 could push up the cost of generating electricity from 3.3 ¢/kWh to 5.2 ¢/kWh at a natural gas plant and from 4.6 ¢/kWh to 6.0 ¢/kWh at a coal plant based gasification (Herzog, 1999). Hence, there is a great potential for any technology which can capture CO_2 in an economic way. For CO_2 sequestration to be economical, one must produce a relatively pure, high pressure stream of CO_2. The main reasons are: the economics of transporting CO_2 any distance will favor concentrated CO_2; sink capacity is better utilized by injecting pure CO_2 and some impurities may be harmful to the operations of certain sinks or may have adverse environmental effects (Herzog, 2000). Strategies for reducing the cost of CO_2 sequestration include reducing the energy requirements for CO_2 capture and new innovative technologies (Herzog, 2000).

Membranes to separate CO_2 from gas streams are commercially available. Membrane based processes offer some unique advantages, namely compactness, low energy and labor requirements, ease of scale up. However, their selectivities for CO_2 do not generate high purity CO_2 streams. There is need for efficient membrane based process which can economically generate high CO_2 concentration streams.

Carbon dioxide at high concentrations is known to plasticize many polymers. The plasticization changes its permeation characteristics irreversibly by reducing their CO_2 selectivity. Condensation of liquids or fouling on the membrane surface can wash away some types of existing membrane coatings, thereby destroying the separation characteristics of membrane. Chemical attack or weakening of polymer membrane can also bring about mechanical separation of separating device. There are reports of large commercial installations that failed because of unanticipated poisoning by presence of hydrocarbons (either as condensates or vapor components). Simultaneous development of improved membrane technology and large (successful) demonstration plants will advance viability and acceptance of membrane separations throughout the industry (ATP, 1997).

A significant advantage of membrane technology is size of the treatment unit. Currently used amine absorbers of CO_2 typically expose 250 square meters of surface area of absorbent to gas stream per cubic meter of the absorber (typically a steel tower). However, hollow fiber technology has permitted membrane surface areas of 6000 or more square meters per cubic meter of membrane unit to be presented to gas streams. This results in greatly reduced costs associated with size and weight of absorber units, which can be doubly important in applications where space or weight of unit is critical. Examples of these include retrofits of existing gas production sites and also installation on offshore platforms. Membrane technology in the form of membrane contactors has become a commercial reality for gassing or degassing of various liquid streams. Examples include introduction of ozone into semiconductor water, and CO_2 and O_2 removal from semiconductor water.

This program focuses on high gas flux non-porous hollow fiber (HF) membrane contactors for removing CO_2 from flue gas streams into amine solution and its subsequent desorption. We utilize existing or modified microporous HF and coat them with our non-porous high gas flux chemically resistant perfluoromembranes. The system operates by having currently commercial amine absorbents flow on coated side of the membrane contactor. CO_2-laden flue gas will be introduced to non-membrane side of CMS coated membrane contactor in a counter-current manner. The amine solution provides selective absorption of CO_2. The membrane contactor provides design advantages associated with high pack density; flexibility in flow rates; no foaming, channeling, entrainment, corrosion**,** or flooding; insensitivity to motion and orientation and significant weight savings. The CMS perfluoromembrane provides wet-out resistance (wet-out floods membrane structure with liquid and gas transport is lost), maintenance of high gas transport and resistance to organic or water contaminant. The non-porous membrane minimizes loss of amine and water.

2. The Membrane Process

This program focuses on CO_2 removal from power plant flue gas streams. The determining factor for the economic viability of membrane separation processes is the product CO_2 concentration and CO_2 fluxes.

The concept of CO_2 separation from nitrogen, methane or oxygen with polymeric membranes has been commercial. However, most of the applications where polymeric membranes are employed are targeted towards obtaining high purity nitrogen or oxygen methane. In some cases, multistage membrane systems are employed to obtain required purity. In multistage membrane treatment system, feed gas at intermediate stages may require an expensive pretreatment step. This pretreatment is designed to remove condensable gases in order to prevent condensation on membrane surfaces. For most membrane materials, hydrocarbon condensation could result in serious reduction in permeance or selectivity and shorter membrane life. Gas pretreatment normally includes particulate filter to remove particulates. For treating gas streams from power plants, particularly coal-based, an extra step of particulate removal may be necessary as pretreatment. This step is necessary for all CO_2 removal technologies, and particularly applicable for membrane based technologies. The present proposal assumes particulate-free flue gas for the phase I study. Data in related research indicates our membranes can handle particulate.

Recently, large scale membrane absorbers were demonstrated for their efficiency to remove CO_2 and water from natural gas streams (Falk-Pedersen et al., 2001) with applicability to capture of CO_2 from coal-fired combustion. They have shown significant improvements are possible with membrane contactors over conventional technologies.

3. Conventional Absorption Process

The most widely used process for separation of acid gases CO_2 or hydrogen sulfide (H_2S) from flue gas is absorption using an appropriate solvent (Kohl and Nielsen,

1997). Chemical solvents (such as diethanolamine) and physical solvents (such as Morphysorb®) are used for acid gas removal. In typical acid gas absorption process for example, the sour gas enters bottom of absorber column or tower and flows upward in counter-current contact with aqueous solvent solution. The absorber column contains bubble trays or packing material designed to ensure proper distribution of liquid throughout column, providing for greatest possible degree of contact between gas and solvent. The solvent absorbs CO_2 and H_2S on its way down. The height of absorber column is dictated by required purity of gas leaving at top. The diameter is dictated by maximum allowable gas velocity before gas starts to entrain liquid. Large gas flow rates with high levels of CO_2 in feed make for large absorbers. The contaminant-rich solvent flows from the bottom of the absorber through a heat exchanger to the top of a desorber, and then flows downward in counter-current contact with vapor generated in a reboiler. The reboiler vapor strips the CO_2 from the rich solution. The CO_2 and steam leave top of desorber and pass overhead through condenser, where the major portion of steam is condensed and cooled. The CO_2 is separated and either vented or compressed. The lean solvent is returned to top of absorber.

Triethanolamine (TEA), which was the first to become commercially available, was used in early gas-treating plants. As other members of the alkanolamine family were introduced into market, they were also evaluated as possible acid-gas absorbents. As a result, sufficient data are now available on several of the alkanolamines to enable design engineers to choose most suitable compound for each particular requirement. The two amines which have proved to be of principal commercial interest for gas purification are monoethanolamine (MEA) and diethanolamine (DEA). Triethanolamine has been displaced largely because of its low capacity (resulting from higher equivalent weight), its low reactivity, (as a tertiary amine), and its relatively poor stability. Each has at least one hydroxyl group and one amino group. The hydroxyl group serves to reduce the vapor pressure and increase the water solubility, while the amine group provides the necessary alkalinity in water solutions to cause the absorption of acidic gases. The principal reactions occurring when solutions of a primary amine, such as monoethanolamine, are used to absorb CO_2 and H_2S. These amine complexes have appreciable vapor pressures under normal conditions so that the composition of the equilibrium solution varies with the partial pressure of the acidic gases over it. As the vapor pressures of these compounds increase rapidly with temperature it is possible to strip the absorbed gases from the solution by the application of heat. While the alkanolamine absorption processes are often effective, these are capital intensive, large, corrosive and prone to foaming.

4. Membrane Gas-Liquid Contactor

The overall absorption process remains the same when gas-liquid membrane contactor is used in place of a conventional column. The CO_2-laden gas enters contactor, where it is kept separate from lean solvent solution by membrane that is highly permeable to component for which removal is desired (e.g., CO_2, H_2S or water). The CO_2 diffuses through membrane into lean amine solution where it is chemically absorbed. Because membrane provides large contacting area per unit volume, size of equipment can be made smaller than with conventional units. A highly selective

separation can be achieved through suitable choice of absorption liquid. The major driving force for separation is absorption in liquid, not pressure differential. This is different from gas/gas membrane approach, which relies on the membrane alone to separate contaminants from the sour gas stream. The use of a gas-liquid absorption membrane contactor has several advantages over conventional amine absorption contacting equipment (Reed et al., 1995; Kovvali and Sirkar, 2002):

- higher packing density – 500 to 6000 m^2/m^3 (150 to 1800 ft^2/ft^3) versus 100 to 250 m^2/m^3 (30 to 75 ft^2/ft^3) for a conventional column
- greater flow rate flexibility (liquid to gas ratios) and solvent selection
- very low liquid flow rates, leading to very high G/L ratios, are possible without entrainment and associated problems
- no foaming, channeling, entrainment or flooding
- insensitivity to motion and flexibility regarding unit orientation (important offshore)
- significant savings in weight and area (also important at remote locations).

5. CMS Membrane Contactor

Membrane contactors have routinely been used for moving gases into and out of liquids, mainly water. The early membrane contactors were made from microporous polypropylene HF. The HF design provides very high surface area per unit volume and microporous polypropylene minimizes or eliminates membrane wet out when using water. However, when amines such as MEA and DEA are added to water, these microporous polypropylenes wet out easily after few hours. This wetting-out is detrimental to successful operation of membrane contactors as it reduces gas removal rates. A viable technique will be to provide non-porous, non-wetting high gas flow coating layer on substrate membrane to prevent wet-out and provide chemical resistance. This layer should provide minimal resistance to gas transfer and avoid wet out and drastically reduce permeation or evaporation of liquid and its components (e.g. MEA). One can use microporous HF (e.g. polypropylene, polysulfone) and maintain existing module designs and coat it with appropriate non-porous, non-wetting material. So far no such composite membranes are commercially developed. We propose to develop such composite membrane with a perfluorinated polymer coating on conventional HF membranes. Our composite membrane will have a) very low resistance to gas transfer across coating, b) exceptional chemical resistance, c) ease of formation on a wide variety of substrates to tailor to individual applications and d) have higher mass transfer than other flat sheet based contactors. CMS has exclusive right to one such polymer, PDD-TFE (perfluoro dioxole-tetrafluoroethylene), from DuPont. This polymer satisfies all the criteria listed above which are needed in a successful membrane contactor.

A recent competing approach has been to utilize microporous PTFE support (Falk-Pedersen et al., 2001) as in Kvaerner-Gore process. Even though this approach employs all perfluorinated materials for the membrane and/or module making, it has certain limitations. They are a) PTFE support is available only in flat sheet form, resulting in lower surface area/unit volume than HF membranes (less than 1000 m^2/m^3 for spiral wound modules versus 6000 m^2/m^3 for HF modules). This would result in bulky modules compared to HF modules, b) mass transfer in flat sheet membrane based

modules is usually inferior to that in HF modules, c) flat sheet membranes, particularly in spiral wound form are more prone to particulate fouling and excessive pressure buildup than H modules. Because of these reasons and economics, our CMS membrane contactor technology will be superior to all PTFE based membrane contactors in Kvaerner-Gore process.

CMS fabricates unique high gas flux perfluorinated composite membranes with a fully perfluorinated polymer deposited as thin (~ 2 μm or less) layer on porous membrane-based substrate. These membranes will preserve gas flux. The non-porous perfluorocarbon membrane while preventing wet out will also:

1. prevent significant decrease in mass transfer of CO_2, as a result of the microporous membrane being wetted by the liquid absorbent
2. prevent liquid absorbent loss as result of weeping though a wetted microporous membrane and also simultaneously reduce mist formation in the flue gas stream
3. reduce the degradation of absorbent by reducing the transfer of O_2.

6. Chemical and Fouling Resistance

The CMS membranes are based on copolymers of tetrafluoroethylene (TFE) with perfluorodimethyldioxole (PDD) (see Figure 6-1). Two copolymers that have been available are CMS-7 which has 87% PDD/13% TFE and a glass transition temperature of 240°C and CMS-3 which has a composition of 65% PDD/35% TFE and a glass transition temperature of 160°C.

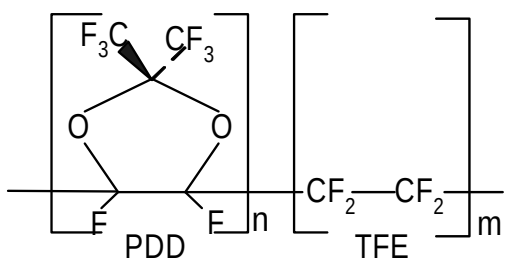

Figure 6-1. TFE perfluorodioxole copolymer

The perfluorinated nature of the CMS copolymers is such that they have excellent chemical and fouling resistance. Discussed below are specific examples that indicate the fouling and chemical resistance of these materials. CMS perfluorinated non-porous membranes have excellent non-wetting characteristics. This is due to a combination of the non-porous surface and the perfluorinated nature. The CMS polymer has CF_3 groups off the backbone (see Figure 6-1). These CF_3 groups are more hydrophobic and organophobic than other conventional perfluoro polymers (e.g. PTFE) which have only CF_2 off the polymer backbone. Table 6-1 shows chemical inertness of these materials to various acids, bases and organics. None have any significant effect on the perfluorinated CMS polymer.

TABLE 6-1: CMS Membrane Chemical Resistance

Reagent	Temp., °C	D wt %	Appearance Change
Acetone	23	0	None
Carbon Tetrachloride	23	0	None
12 N HCl	60	0	None
Hexanes	23	0	None
MEK	23	0	None
44% NaOH	60	0	None
Perclene®	23	-0.1	None
Toluene	23	0	None

7. Fabrication Flexibility

CMS composite membranes can be fabricated using a variety of porous substrates for support. Table 7-1 shows that we are able to coat various hollow fiber and flat sheet supports. This includes polypropylene, PVDF, cellulose acetate, expanded polytetrafluoroethylene (ePTFE) and polysulfone. It should be noted that the ideal selectivity for the CMS-7 membrane is 1.95. Membranes with selectivities approaching this value indicate successful coating.

TABLE 7-1: Composite Membrane Fabrication

Porous Support (HF**)	Nitrogen Flux* (GPU)	Selectivity* (O_2-N_2)
Polysulfone	4900	1.85
Cellulose Acetate	3000	1.80
Polypropylene	2400	1.75
PVDF (vinylidene fluoride)	300	1.70
e-PTFE	500	1.80

GPU = cc/cm^2-sec-cmHg x 10^{-6}

* Based on high flux membrane/similar results for high selectivity membrane

** Results for inside coated hollow fiber/similar results for both outside coated hollow fiber and flat sheet

8. Phase I Study Objectives

The Phase I research plan consisted of four major objectives::

1) Fabrication of perfluorinated membrane on appropriate hydrophobic hollow fiber membrane contactors such as microporous polypropylene or polysulfone with coatings on one side.
2) Demonstrate carbon dioxide removal from simulated flue gas mixture via amine absorption using membranes fabricated in above task.
3) Examine chemical compatibility of CMS membranes with amines and demonstrate enhanced stability of the perfluoro coated membranes.
4) Perform economic analysis and demonstrate that the perfluoro coated hydrophobic HF membrane contactors are superior to existing commercial carbon dioxide removal technology.

We have met and exceeded all goals associated with above objectives in Phase I. Although it was not part of a Task in our Phase I study, we have also demonstrated feasibility of simultaneous absorption and stripping of carbon dioxide from simulated flue gas using CMS hollow fiber membrane contactors.

9. Phase I Study Results

Fabrication of CMS membrane modules

Membrane preparation has gone very well. We have fabricated two custom tube and shell hollow fiber modules for CO_2 absorption. One module contains polypropylene hollow fiber with a CMS-7 PDD-TFE coating and a surface area of 826 cm^2. The second module contains polysulfone hollow fiber with a CMS-3 PDD-TFE coating and a surface area of 1440 cm^2. A larger surface area module was fabricated when the CMS-3 coating was used because it has lower gas permeability than the CMS-7 coating (but has a higher selectivity). Each membrane module was tested for initial performance by measuring the single gas permeances of nitrogen, oxygen, and helium using a pressure decay method**.** This was done to provide reference baseline permeation rate data for comparison with post amine exposure data. Both modules demonstrated the expected relative gas permeation rates, indicating good membrane integrity.

Three hollow fiber membrane modules for amine regeneration (CO_2 desorption) were procured from our commercial partner. The hollow fiber used in these modules is polysulfone with a CMS-3 coating. The membrane surface areas in these modules vary from 1860 cm^2 or 4000 cm^2. A higher membrane surface area was specified for CO_2 desorption compared to absorption because desorption is expected to be liquid side controlled. The hollow fibers in desorption modules are arranged in a bundle woven around a central perforated mandrel where the liquid amine would enter. This configuration provides better liquid mixing and contact with the membrane and will promote more rapid desorption than would be obtained with the simple shell and tube arrangement used in the absorption modules.

Investigate carbon dioxide removal rates

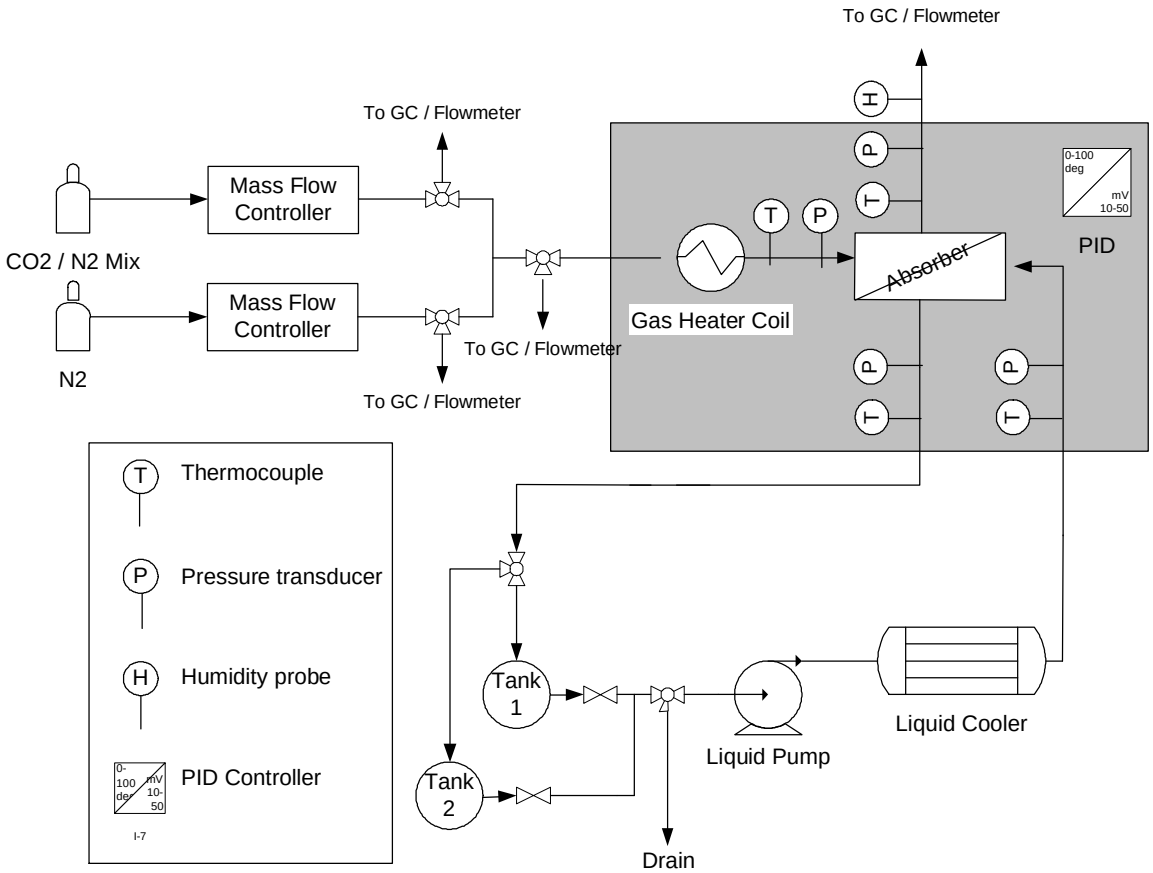

Figure 9-1. Temperature controlled membrane module testing apparatus

A schematic diagram of the apparatus is shown in Figure 9-1. The module was set up in a temperature controlled incubator oven (Despatch, model LEB1-28) in each experiment. In order to ensure that both the membrane module and the feed gas entering the lumen side of the module remained at consistent, equal temperatures, a simple heat exchanger consisting of coiled copper tubing was constructed and placed in line. Temperature controls consisted of a dual loop proportional-integral-derivative (PID) controller (Love Controls, model 85113-0), which was used to maintain the incubator oven temperature, and a fan cooled heat exchanger (Lytron, model ASOG-16) used to cool the liquor to room temperature. Flow control and mixing of the feed gases was accomplished using a mass flow controller unit consisting of mass flow meters and computer interface module (Porter Instrument Company, Inc., Series 200 MFC and PCIM4 CIM). Liquid flow control was provided by a digital pump drive (Cole-Parmer, model EW-75211-30) and miniature pump head (Micropump, model 81980). Additional instrumentation included thermocouples (Omega, model TC-K-NPT-G-72), pressure transmitters (Reotemp, model TH2-P16-3E4B), and a humidity sensor (Vaisala, model HMP233). Data acquisition and logging was provided by USB data acquisition board (iOtech, personal daq/56) and notebook computer (Sony, model PCG 5322) using DaisyLab acquisition software.

The use of two liquor tanks (Alloy Products, model 72-01) and a simple valve arrangement added versatility and expandability to the testing apparatus; allowing the

operator to run either once through absorption only experiments by pumping liquor through the system from one tank to the second, and easily upgrade the system to run continuous combined absorption and desorption experiments by installing a desorption module, liquid heater and vacuum pump prior to the absorption unit and then recirculating the liquor through the system using only one tank. During absorption only experiments a 20% by weight diethanolamine-water solution was pumped from tank #1 through the liquid cooler, which removed residual heat imparted by the pump drive, and then through shell side of the absorption module and into tank #2. If the operator wished to install the required equipment, during continuous combined absorption and desorption experiments, the amine solution would be pumped from tank #1 through a temperature controlled liquid heater where it would be heated to 60°C. The liquid would then enter a desorption module where vacuum would be used to remove carbon dioxide from the stream, and from here the liquor would be cooled to room temperature in the liquid cooler, passed through the absorption module and then back into tank #1 for recirculation. A schematic diagram of such a system is shown in Phase II work plan under Task 2.

Each experiment began by flushing the apparatus with the appropriate bottled feed gas or mixture of gases. Feed gases consisted of certified standard mixtures (MG Industries) of 20% CO_2 by volume in nitrogen and pure nitrogen. After flushing the apparatus for several minutes, the mass flow controller was used to generate the desired feed flow rate and feed gas concentration by blending additional nitrogen into the bottled CO_2 mixture when necessary. The flow rate exiting the mass flow controller was verified by temporarily diverting the feed stream to an acoustic displacement flow meter (J&W Scientific, model ADM2000 220-1171) by actuating an in-line 3-way valve. At this time, the concentration of the feed gas was also confirmed using the gas chromatograph's thermal conductivity detector (Hewlett Packard 5890 using an 8'x1/8" 60/80 Haysep column, Supelco model 09072004) and HP3396 Series II integrator combination. The GC was calibrated using certified gas mixtures of know concentration.

Next, the appropriate oven temperature was set, and a solution of 20% diethanolamine in water was poured into tank#1. The concentration of this solution was confirmed by measuring the solution's specific gravity with a hydrometer (VWR, model 34627-231). Finally, the liquor flow rate was set and the data logger was started.

After setting up the appropriate process conditions, the system was allowed to run for at least ten minutes before any measurements were made. The pressure and temperature of each stream and the flow rate and relative humidity of the exit gas stream were recorded at regular intervals. Samples of the treated gas stream were injected into the GC at regular intervals as well until repeated injections differed by no more than 5%, indicating that steady state operation had been achieved. The feed flow rate was verified again at the end of each experiment, before proceeding to the next randomly determined set of process conditions.

After recording each run, the steady state data was normalized to standard temperature and pressure (STP) conditions and averaged. Data differing by more than three times the standard deviation of the data set was discarded. The percent removal of carbon dioxide was then calculated at each test condition. After all experiments had been completed the membrane modules were flushed thoroughly with water, dried and retested for single gas permeability.

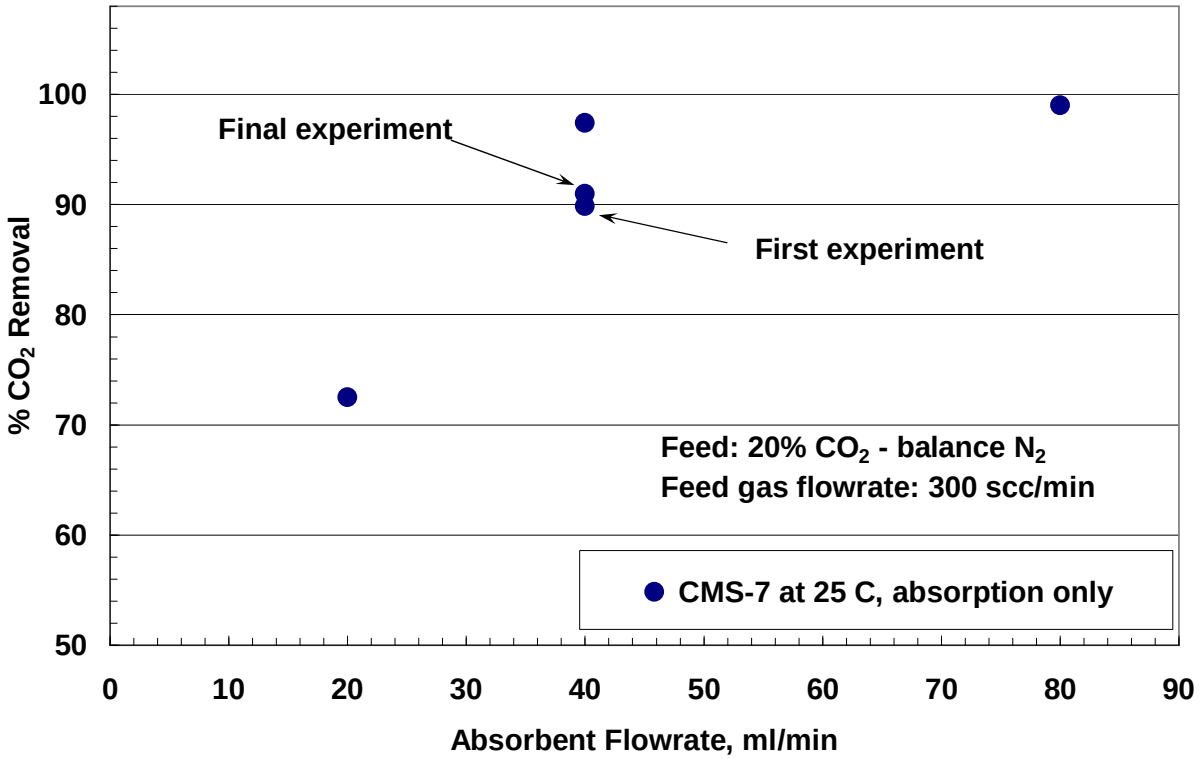

Figure 9-2. Effect of absorbent liquor flow rate on CO_2 removal

Effect of 20 wt% DEA absorbent flow rate were studied on carbon dioxide removal at a constant flue gas flow rate (Figure 9-2). As expected, a larger percentage of CO_2 is removed at higher liquor flow rates. Removal approaches 100% at a liquor flow of 80 ml/min, corresponding to a gas-liquid flow ratio of 3.75. Note that, the module used for absorption studies was straight parallel flow module and was not optimized for even liquid distribution on the shell side. Even better results are expected at higher gas flow rate in Phase II when we will use optimized commercial modules.

The effect of feed gas flow rate was also examined in both the CMS-7 and CMS-3 modules under a variety of process modes and conditions (Figures 9-3 and 9-4). No significant effect on carbon dioxide removal was observed for the CMS-7 module over a feed gas flow range up to 300 ml/min suggesting that CO_2 permeation through CMS-7 membrane is quite fast and the membrane resistance does not control the overall transport. However, the CMS-3 module showed a marked loss of nearly 25% in the percentage of CO_2 removed when the feed gas flow rate was increased from 100 to 200 ml/min (Figure 9-3). This indicates that the membrane resistance in CMS-3 device is not negligible for this application. As we will show later in Table 9-1, gas flux through CMS-7 membranes are much higher than CMS-3 membranes.

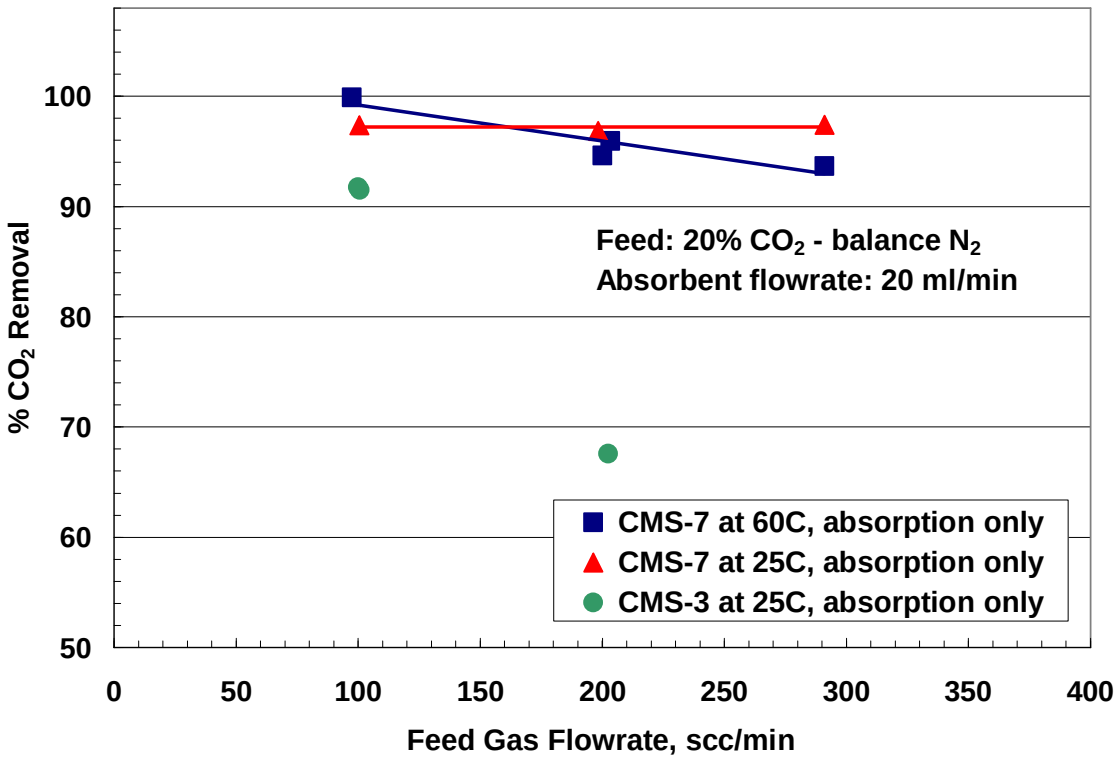

Figure 9-3. Effect of feed gas flow rate and temperature on CO_2 removal

Although higher temperature lead to a faster reaction rate (but the thermodynamic equilibrium will be less favorable), we did not observe much change in CO_2 removal rate when absorption experiments were carried out with CMS-7 module at 25 and 60°C,

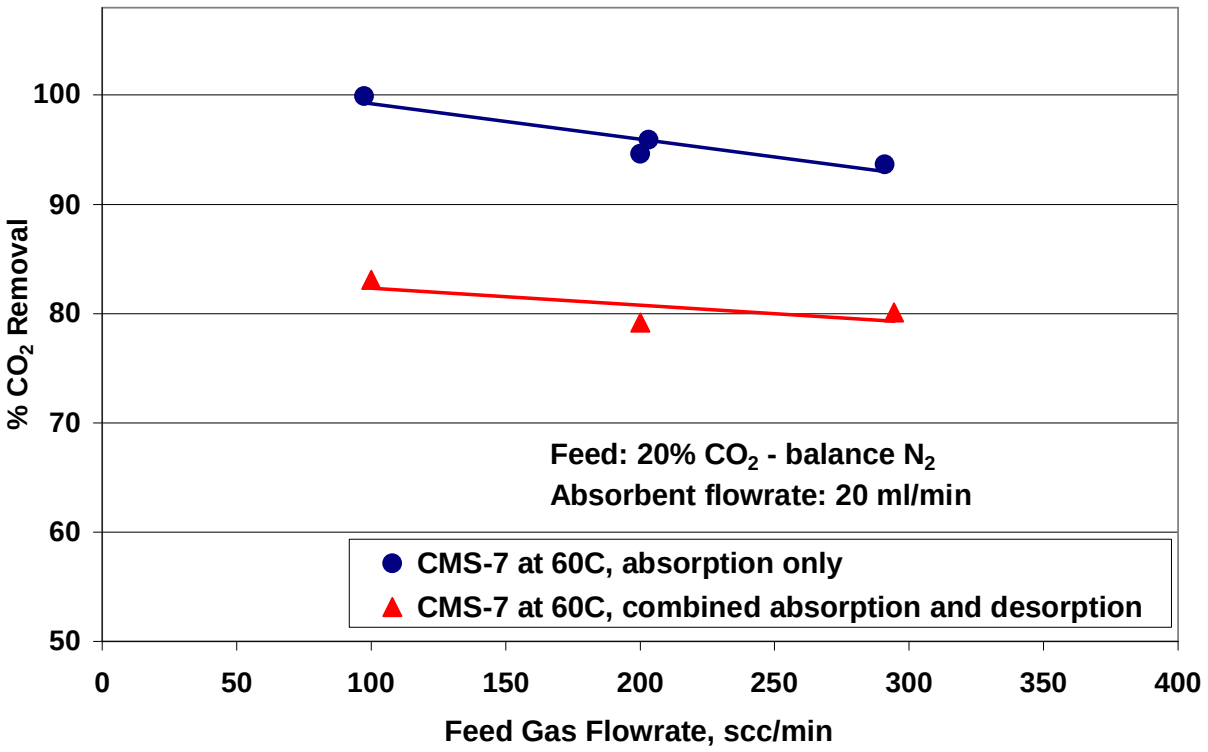

Figure 9-4. Absorption only vs. combined absorption and desorption operation

Though not required in Phase I study, we have also demonstrated feasibility of simultaneous absorption and stripping of carbon dioxide from simulated flue gas using CMS hollow fiber membrane contactors.

Data from combined absorption and desorption experiments are compared with the absorption only experiments for the CMS-7 module at 60°C in Figure 9-4. We have observed close to 80% CO_2 removal in a combined mode under steady state conditions. Obviously the removal rates are lower than absorption only situation as only fresh DEA solution was used in absorption only mode. On the other hand in combined mode of operation, the liquid already has a considerable CO_2 content. This will limit the CO_2 flux due to the increased back pressure and the reduced concentration of active components. Performance in combined absorption and desorption mode could be improved by increasing the size of the desorption module or carrying out desorption at a higher temperature. An optimized regeneration process is therefore required, which we will address in Phase II of the program.

To find out the level of removal under very high gas flow rate, we carried out absorption experiments at 1 and 2 l/min flow rate keeping the absorbent flow constant. The reduction in removal rates are shown in Figure 9-5. (Note that this module has only 300 fibers and it was not designed for such a high flow operation.)

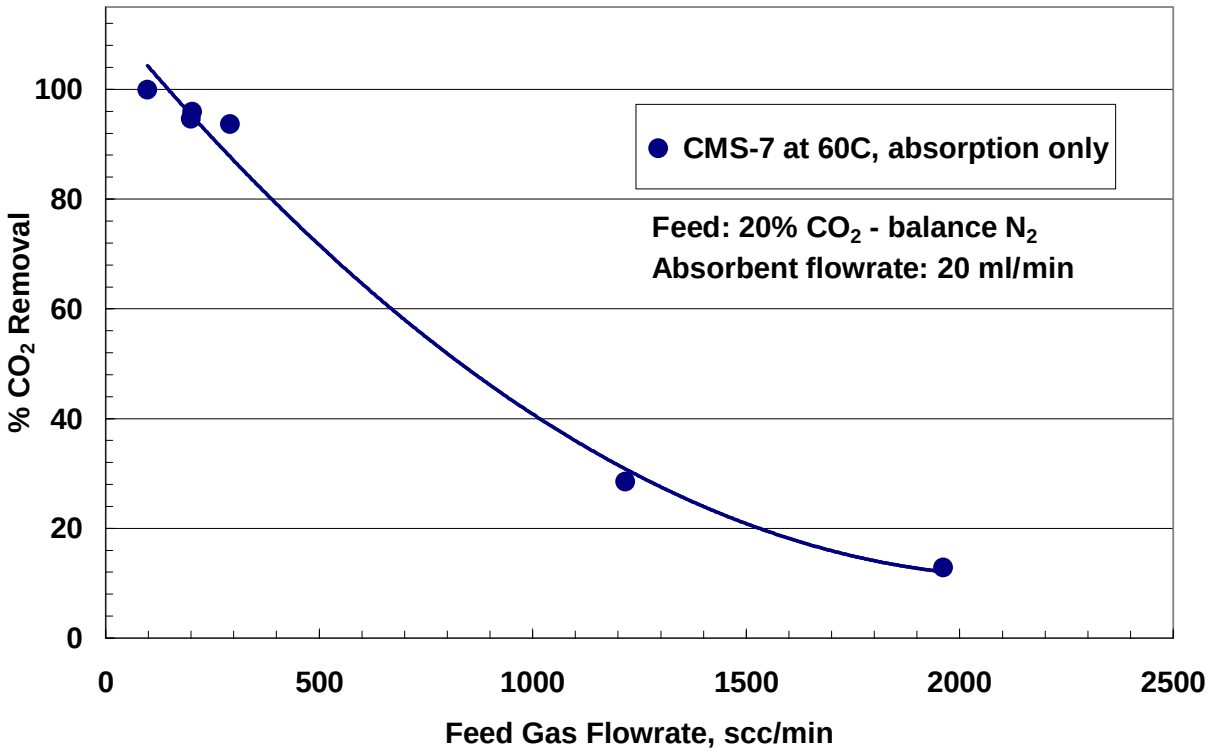

Figure 9-5. Effect of high feed gas flow rate on CO_2 removal

The effect of carbon dioxide concentration in the feed gas was also examined. Again, the removal rates are very high and no significant decrease in the percentage of carbon dioxide removed by the CMS-7 module was observed over the range of increasing concentrations examined for the given gas and liquor flow rates (Figure 9-6).

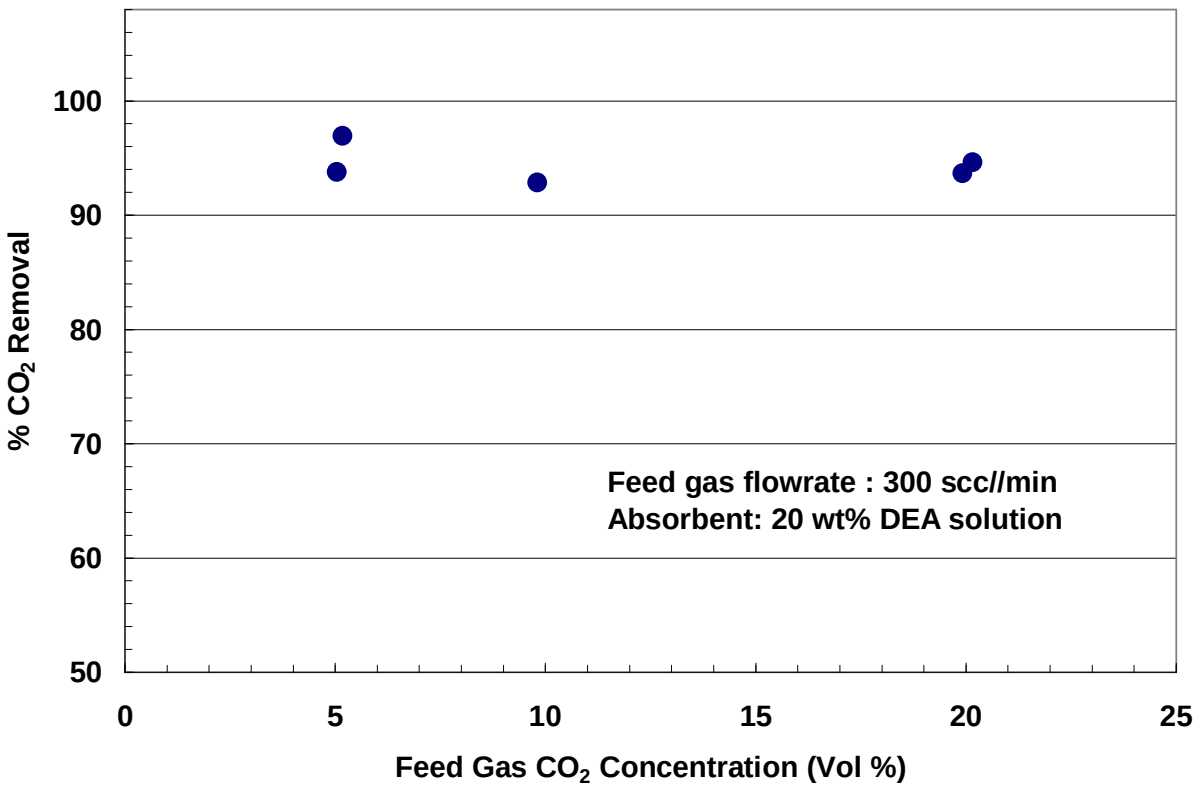

Figure 9-6. Effect of feed gas concentration on carbon dioxide removal

The amount of water removed from the amine solution during processing was monitored by recording the humidity, temperature and flow rate of the treated gas stream. The specific gravity of the amine solutions were measured before and after the experiments as well, but the hydrometer used was not precise enough to gauge or confirm water loss rates. As expected, water lost from the amine solution as humidity in the treated gas stream increases with increasing gas flow rate. Relative humidity levels in the treated gas stream averaged approximately 60%, regardless of gas flow rate, liquor flow rate, and membrane temperature or membrane material. Thus the amount of water removed from the liquor stream appears contingent only on the amount of feed gas processed and overall surface area of the membrane module. Water loss via the treated gas stream was modest, never exceeding 0.25g/hr/m^2 even when the feed gas flow rate was increased to 2,000 scc/min, which equates to a gas-liquid flow rate ratio of 100.

Chemical compatibility of CMS membranes with amine solutions and demonstration of improved stability at module level

For chemical compatibility tests of CMS membranes with amine solutions, six thin film specimens of CMS-3 and six thin film specimens of CMS-7 were prepared and annealed via proprietary CMS standard operating procedures. The films were nominally 25 micron thick. Each film was gas permeation tested with helium, nitrogen, oxygen

and carbon dioxide to establish baseline permeation rate data for comparison with post amine exposure data and to establish that each film was free of perforations. One of each type of film was immersed in 10%, 20% or 50% (by weight) aqueous diethanolamine solution at room temperature or 60º C. (Diethanolamine has been chosen as the primary candidate absorption solution over monoethanolamine because its lower corrosivity and lower vapor pressure make it a safer material) Gas permeation rates were measured again after 17 days or longer exposure. The permeation test data is presented in Table 9-1. Most of the films exhibited changes in the absolute gas permeation rates. Through past experience we know that it is not uncommon for gas permeances of new membranes to change over time as the membrane ages. This is a common phenomenon for glassy membranes.

The more important measurement is the gas selectivity (calculated as the ratio of permeances of two gases). As expected, in general neither CMS-3 nor CMS-7 membranes showed any significant loss in selectivity. The temperature and concentration of the amine solution did not appear to have a significant effect on membrane selectivity. A substantial decrease in selectivity for all three gas combinations would be an indication of membrane perforation as a result of chemical attack. Alternatively, mechanical damage of the membranes during handling and gas permeation testing would have the same effect. The data from these experiments were analyzed in Design Expert (by StatEase®) software in a 2x2x3 full factorial design. The factors (variables) and levels are listed below in Table 9-2.

Table 9-2. Factors and Levels

Variable	**Level 1**	**Level 2**	**Level 3**
CMS Membrane Type	CMS-3	CMS-7	
DEA Solution Temp (C)	23	60	
Weight % DEA	10	20	50

The response that was measured and analyzed was the % change in selectivity of O_2/N_2, He/N_2, and CO_2/N_2 before and after exposure to the DEA solution. As shown in Table 9-1 each of the samples had a different exposure time in DEA solution. This therefore limits the value of the statistical analysis, since it is not possible to account for this variation. However, the analysis as it was performed suggests that there is no significant relationship between any variable or combination of variables and the observed changes in selectivities.

Table 9-1. Effect of aqueous diethanolamine solutions exposure on CMS membranes

Sample #	Material	Exposure Duration (Days)	wt% DEA	DEA Solution Temp, C	Permeances Before Exposure (GPUs)				Permeances After Exposure (GPUs)				% Change in Selectivity After Exposure		
					N2	O2	He	CO2	N2	O2	He	CO2	O2/N2	He/N2	CO2/N2
093-08-05	CMS-3	17	50	60	3.7	9.7	50.5	23.5	3.8	10.0	50.6	23.9	1.21	-1.24	0.35
093-08-04	CMS-3	20	20	60	3.4	9.0	43.0	21.4	3.7	9.5	47.7	21.9	-1.67	3.22	-4.67
093-08-02	CMS-3	24	10	60	3.7	9.7	52.9	22.4	3.9	10.2	52.4	23.2	-0.24	-6.02	-1.74
093-08-06	CMS-3	31	50	23	3.4	9.2	49.9	22.0	4.4	11.6	58.1	26.7	-2.99	-10.03	-6.22
093-08-03	CMS-3	34	20	23	4.1	10.7	55.3	25.4	3.8	9.9	49.5	22.7	0.65	-2.23	-2.29
093-08-01	CMS-3	35	10	23	3.3	8.8	48.7	20.7	3.4	9.1	47.2	20.8	0.37	-5.93	-2.71
093-08-12	CMS-7	19	50	60	35.4	68.5	159.0	161.9	29.1	57.6	135.6	130.2	2.20	3.75	-2.21
093-08-10	CMS-7	20	20	60	37.2	71.3	164.9	166.0	32.8	63.5	147.2	143.6	1.16	1.36	-1.77
093-08-08	CMS-7	24	10	60	36.1	73.2	170.6	171.6	33.6	64.9	150.4	149.5	-4.67	-5.17	-6.29
093-08-11	CMS-7	26	50	23	35.8	69.4	160.8	160.3	30.6	59.7	142.7	141.7	0.72	3.96	3.55
093-08-09	CMS-7	33	20	23	32.7	65.0	150.7	150.2	29.7	57.0	135.8	135.9	-3.37	-0.62	-0.25
093-08-07	CMS-7	40	10	23	35.3	70.3	161.1	162.1	31.1	60.3	142.5	139.4	-2.57	0.56	-2.27

1 GPU = 10^{-6} cm^3 (STP)/cm^2.s.cm Hg

As an additional verification that chemical attack is not eroding the membranes, those films exhibiting decreases in all three selectivities were examined were examined under Scanning Electron Micrograph (SEM). SEM work was performed on an Amray 1200C Scanning Electron Microscope with the assistance of a Princeton Gamma Tech (PGT) Prism 2000 energy dispersive spectrometer (EDS) with a PGT Avalon Digital detector. Samples were prepared by mounting thick film membrane to a stainless steel stub with carbon tape and using a Denton Vacuum, Desk II gold sputter coater. Each sample was sputtered with ~100Å of gold coating. Sample image analysis was done at 250X at a 30 degree tilt. Images were examined for deformities in the surface compared to a standard membrane sample. Quantitative spectrum analysis was performed on each sample looking for levels of nitrogen (indicative of presence of DEA), fluorine (indicative of presence of membrane) and gold (sputter coating) and compared to a standard membrane sample. Two sample SEM photographs are shown in Figures 9-7 and 9-8. SEM examination of membrane samples and spectral analysis before and after exposure showed no notable defects or damage to the exposed membranes. The membrane structure was unaffected and was not attacked by aqueous DEA solution. The chemical exposure tests will be continued for longer duration with the remaining films.

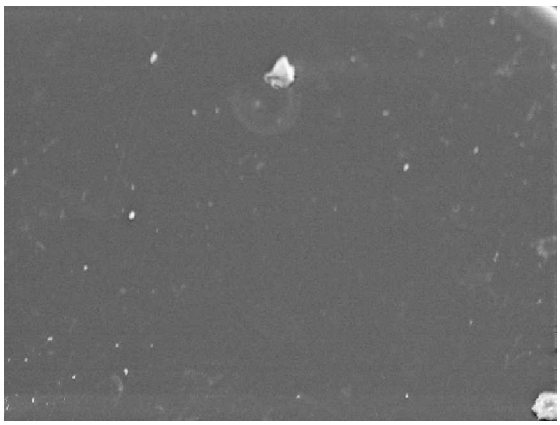

Figure 9-7. SEM photograph of CMS-7 membrane prior to exposure to amine solution (magnified 250x)

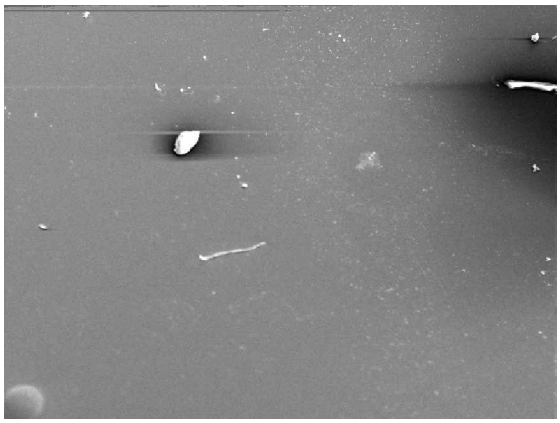

Figure 9-8. SEM photograph of CMS-7 membrane after exposure to 50% diethanolamine in water solution at 60°C for 17 days (magnified 250x)

Chemical compatibility data is available for the CMS-7 coated polypropylene membrane module as well, since it was tested for single gas permeances before and after exposure to 20% diethanolamine solutions during experimentation. The results are presented in Table 9-3.

Table 9-3. Effect of exposure to 20% diethanolamine solution on CMS-7 coated polypropylene hollow fiber module

	Prior to DEA exposure		After DEA exposure	
Gas	**Permeance (GPU)***	**Selectivity to N_2 -**	**Permeance (GPU)***	**Selectivity to N_2 -**
N_2	487.2	1.00	343.1	1.00
O_2	913.6	1.88	646.8	1.89
He	2206.3	4.53	1764.0	5.14

1 GPU = 10^{-6} cm^3 (STP)/cm^2.s.cm Hg

No detrimental effects to membrane selectivity were observed over the span of 19 days, indicating that the membrane was not damaged by exposure to the amine solution or cycling oven temperatures between room temperature and 60°C. Although, significant losses in permeability were noted, the selectivity values remained unaffected. As shown earlier in Figure 9-2, a comparison of carbon dioxide removal performance between the first and last experiments (under identical operating conditions) confirms that there was no decrease in membrane performance during the progression of experiments.

Economic Evaluation of CMS Membranes

Quantitative Economic Evaluation of CMS Membranes Relative to Tower Contactors

The cost saving offered by a CMS membrane contactor carbon dioxide capture system relative to a tower contactor CO_2 capture system was estimated. For the purposes of this Phase I program, the cost analysis was limited to the application of CMS membranes only in the CO_2 absorption step of the CO_2 sequestration process. The application of membranes in the amine regeneration step was not analyzed. (Analysis of the regeneration step will be completed in the Phase II program.) By focusing only on absorption, the economic comparison of the base case tower-only system (towers used for absorption and stripping) vs. the membrane/tower system (membranes used for absorption and towers used for stripping) is greatly simplified. The only components that are different in the two systems are the CO_2 absorption contactor and the flue gas blower. The specification for the flue gas blower changes as a consequence of the lower pressure drop afforded by the membrane absorber. The flue gas cooler upstream of the absorber is unchanged because both the flue gas feed

temperature to the absorber and the flue gas flow rate are unchanged. The balance of the two CO_2 capture systems can be assumed to be identical since the membrane absorber and the tower absorber are specified to provide the same CO_2 removal efficiency. This means that the CO_2 loading in the amine liquid will be identical for equal amine flow rates and since mass balance requirements dictate that the stripping towers in both systems reject CO_2 at equivalent rates, the stripping towers, reboiler, pumps, heat exchangers, and all the other components associated with the two kinds of CO_2 capture systems will be identical. If instead the analysis had included a system with amine regeneration by membrane contactors, most of the process components would have had different specifications from the base case tower-only process because the membrane stripper would operate at different process parameters than the stripper tower. Consequently, all the heat transfer equipment, pumps, etc. would be different from those in the base case system. An analysis as extensive as this was not considered essential for the Phase I program because demonstration of a cost saving by using CMS membranes in the absorption step is indicative of a cost saving in the stripping step since the stripping step is simply the reverse of the absorption step. It would therefore be expected that even greater cost savings would be realized if the membranes are applied in the stripping step. It should also be noted that a system that uses membranes only for absorption and a tower for stripping is not an artificial analysis since it is representative of a possible real world system configuration.

The results of the quantitative economic analysis are summarized in Table 9-4 and Figure 9-9. More detailed data from the economic analysis is presented in Table A-1 available in the Appendix. In order to determine the net cost saving offered by the membrane CO_2 capture system, only the cash flows that differ between the membrane and tower-only systems being compared are accounted for. All other cash flows are ignored since they balance out. The following assumptions, data and correlations were used in preparing the analysis.

<u>Assumptions</u>

- The cost of capital is 15%.
- The CO2 capture system has a useful life of 20 years and an end of life salvage value of $0.
- The Lang factor for fixed capital investment is 4.6. This factor was applied only to the flue gas blower and membrane housing. It was not applied to the membrane cartridges or the towers. (Tower costs shown are either an estimated installed cost or an installed cost taken from the reference cited. See Correlations below for details of tower cost estimation.)
- Supervisory labor is charged as 20% of maintenance labor costs.
- The purchased cost of membrane cartridges are $2.60/ft^2. This cost is consistent with current market prices of membranes that are manufactured in high volume.
- Based on our experience, we have estimated the labor cost to install the membrane cartridges into the housing at 1% of the purchase price.
- See Table A-1 for other assumptions related to costs.
- Costs are stated in 2001 dollars.
- NO_x, SO_2 and particulates are removed upstream of the CO_2 capture system.
- Flue gas fan design pressure drop is assumed to be 1 psi + the absorber pressure drop for natural gas power plants and 2 psi + the absorber in pressure drop for coal

plants. This difference is due to differences in requirements for upstream particulate control and flue gas desulfurization.

Data

- The flue gas stream conditions cited in Table 9-4 were taken from the references cited.
- Absorption tower design data was taken from the references cited in Table 9-4.
- The installed cost of the absorption tower in Base Case 2 was taken from the reference cited in the table. All other capital costs are estimated using the correlations stated below.

Equations & Correlations

- The flue gas fan power consumption was estimated using standard engineering equations for adiabatic compression power with an adjustment for adiabatic efficiency.
- Flue gas blower costs were estimated using blower cost correlations on page 577 of Douglas (first edition, 1988).
- Tower costs were estimated using the cost correlation for columns on page 574 of Douglas (first edition, 1988).
- Tower costs for Base Case 3 were estimated using the following relations and proportioning the height required in Base Case 3 relative to Base Case 2.

  NTU = ln (Concentration in/Concentration out) and Cost $\propto$ Height$^{0.802}$
- The membrane housing costs were estimated based on the cost of floating head heat exchangers because of their similarity in design. Heat exchanger costs were estimated using the cost correlation on page 572 of Douglas (first edition, 1988) with an additional 20% cost multiplier included to cover additional design requirements for membranes cartridges.

As can be observed from the data presented in Table 9-4 and Figure 9-9, the application of CMS membranes for absorption reduces costs for treatment of both natural gas plant and coal plant emissions. The total annual cost savings offered by the CMS membrane absorption system compared to a tower absorption system is about 50% at a CO_2 capture efficiency of 50%. At a CO_2 capture efficiency of 90%, the potential savings in total annual cost is about 25%.

We will first consider the economics of the system operating at 90% CO_2 capture. About 40% of the total annual savings previously stated can be attributed to the lower capital cost of the membrane absorber relative to the tower absorber. About 30% of the absorber tower capital cost was saved. It should be noted that the absorber tower typically represents a substantial portion of the total capital cost of a CO_2 capture plant (Undrum et. al., 2000) and therefore this savings represents a significant portion of the capital cost total for the entire capture plant. The balance of the savings is primarily attributable to a lower flue gas blower operating cost that is partially offset by higher maintenance costs for the membranes. The blower energy savings is a result of the lower pressure drop afforded by the membrane absorber compared to the tower absorber. The literature indicates that typical tower absorber pressure drops are in the range of 1.5 to 4.5 psi (Ciferno et. al. 2005; Gambini et. al., 2000; Mathieu 2003). We can design the membrane absorber to operate at a lower pressure drop since the gas side pressure drop is independent of the liquid amine flow rate. Our expectation is that the pressure drop in the membrane contactor will be in the range of 0.1 to 0.5 psi,

depending on the gas flow rate per cartridge. The substantial cost saving is not surprising since the flue gas blower is typically the primary consumer of power in CO_2 capture systems (Chapel et. al., 1999).

If the efficiency of the CO_2 capture plant is reduced to 50%, the plant with the membrane absorber gains an even greater advantage, in this case the total annual cost savings increases to about 50%. This savings includes approximately a 70% savings in absorber capital costs and approximately a 40% savings in blower operating costs. The capital cost savings on the absorber again represent about 40% of the total annual cost savings. The balance again comes from reduced blower operating costs. The offsetting effect of the membrane maintenance costs are now much smaller because of the reduced membrane surface area compared to the 90% efficient capture plant. The improved cost savings at lower capture efficiency illustrates a significant advantage of the membrane absorber. Because the gas and liquid streams flow independently, we can vary the gas flow rate across a wide range in order to attain the desired capture efficiency. At a given flue gas flow rate, we can reduce the membrane surface area to reduce capture efficiency and substantially reduce capital cost. Towers can be reduced in height or process a slip-stream rather than the total flow but the cost benefit is much smaller.

Qualitative Economic Evaluation of CMS Membranes Relative to Kvaerner Gore Process

The Kvaerner-Gore process represents another competing approach to CMS membranes for CO_2 capture. Due to the proprietary nature of this process and the lack of published performance and cost data this discussion of the relative economics will be qualitative. The Kvaerner-Gore process uses a microporous PTFE membrane that has certain disadvantages. First, the CMS hollow fibers membranes are believed to be much lower in cost than the membranes used in the Kvaerner-Gore process. Microporous PTFE is available only in flat sheet form which Kvaerner fabricates into a “ribbon-tube” by bonding two flat sheets together. We believe that membranes fabricated in this manner would be much more costly than membranes supported on hollow fiber that is spun by standard manufacturing techniques. Based on our membrane industry experience, hollow fiber module prices are around \$2-\$3 per square foot. By contrast, flat sheet ePTFE coated membranes are \$25/ft^2. The second limitation of the ribbon-tube membrane is its lower surface area/unit volume compared to CMS hollow fiber (HF) membranes. (A flat sheet spiral wound module would have less than 1000 m^2/m^3 versus 6000 m^2/m^3 for a HF module of similar external dimensions.) This would result in bulky modules compared to CMS HF modules, requiring larger, more costly housings. The third disadvantage is the lower mass transfer rate provided by the ribbon-tube design. With the ribbon-tube design, one loses the advantage of being able to flow liquid on the outside of the hollow fiber to minimize liquid side mass transfer resistance. Historically blood oxygenators have about 25% of the productivity when blood is fed to the inside of a fiber versus the outside of the fiber. Using that as an analogy, in the Kvaerner ribbon-tube configuration, we would expect that the Kvaerner system would have a quarter of the overall efficiency per unit membrane area compared to our hollow fiber design. This will translate into higher surface area requirements and consequently higher membrane costs. (This also

adversely impacts the cost of the housing.) Finally, the microporous nature of the Kvaerner membrane will allow more rapid transport of amine solution across the membrane. This will result in greater amine makeup costs and/or the need for costly amine recovery systems. For these reasons, the CMS membrane contactor technology will be superior to all PTFE based membrane contactors in Kvaerner-Gore process from operational and economic standpoints.

Table 9-4. Summary of economic analysis

General System Data	**Case 1**	**Base Case 1**	**Case 2**	**Base Case 2**	**Case 3**	**Base Case 3**
Carbon Dioxide Capture System Type	Membrane	Tower	Membrane	Tower	Membrane	Tower
Feed Stream Data Source Reference (See Bibliography)	Chinn, et. al.	Chinn, et. al.	Singh, et. al.	Singh, et. al.	Singh, et. al.	Singh, et. al.
Power Plant Type	Natural Gas	Natural Gas	Pulv. Coal	Pulv. Coal	Pulv. Coal	Pulv. Coal
Power Plant Electricity Output (MegaWatts)	400	400	400	400	400	400
Flue Gas Volumetric Flow Rate (scfm)	1,100,000	1,100,000	775,684	775,684	775,684	775,684
Carbon Dioxide Concentration Vol%	4	4	14.6	14.6	14.6	14.6
Carbon Dioxide Removal Efficiency %	90	90	90	90	50	50
Purchased Equipment Costs						
Total Installed Equipment Cost	$38,560,481	$51,018,769	$27,376,073	$36,255,767	$5,136,421	$17,996,601
Amortized Total Installed Equipment Cost ($/yr)	$6,160,479	$8,150,834	$4,373,642	$5,792,275	$820,602	$2,875,163
Total Annual Operating Cost ($/yr)	$7,679,032	$10,793,555	$6,989,064	$8,878,963	$4,771,809	$8,148,596
Total Annualized Cost ($/yr)	$13,839,512	$18,944,388	$11,362,706	$14,671,237	$5,592,411	$11,023,759
Annual Contributory Cost of CO2 Capture ($/Te)	$14	$19	$4	$6	$4	$8

Costs are stated in 2001 Dollars.
NA = Not Applicable

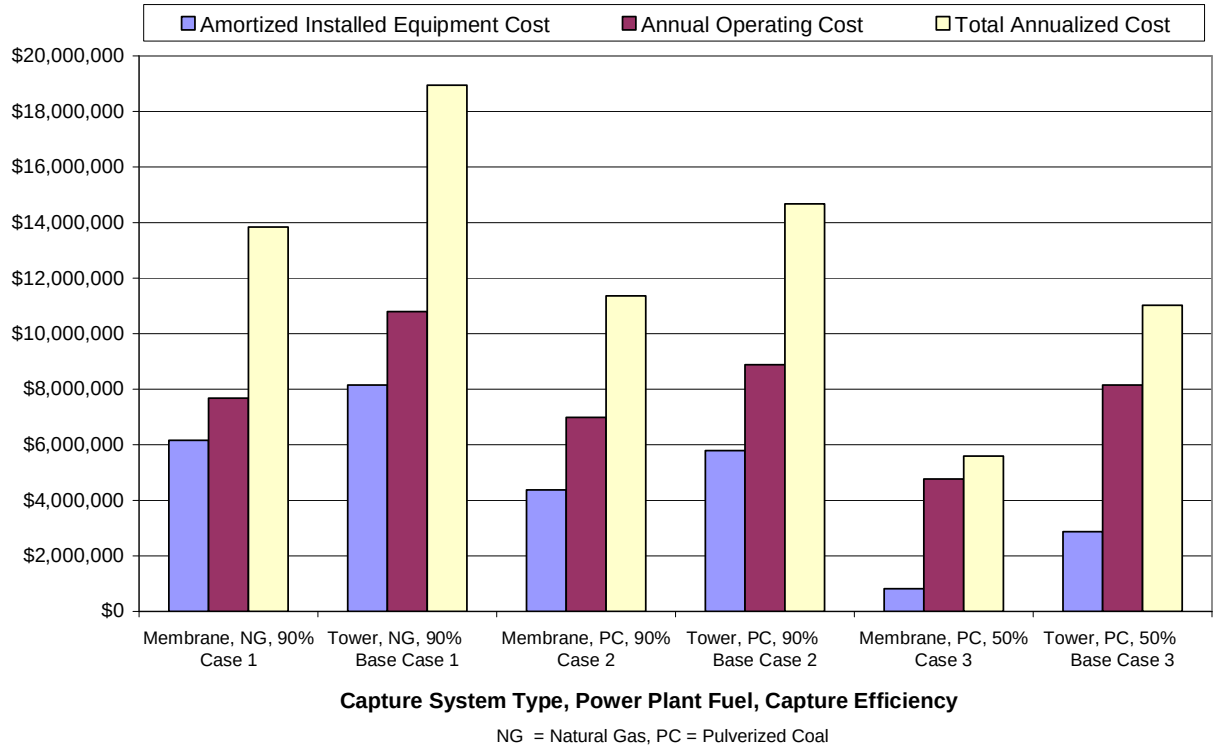

Figure 9-9. Total annual cost contribution of carbon dioxide absorption components

Related Research or R&D

Praxair-Pilot Tests: CMS has in collaboration with Praxair and Caterpillar done extensive related research. Praxair and CMS did initial ruggedness testing of large (1200 ft^2) CMS membrane on microporous polysulfone hollow fiber modules. Figure 9-10 shows that these modules maintained excellent stability over 2500 hours at 85ºC and 25 psi.

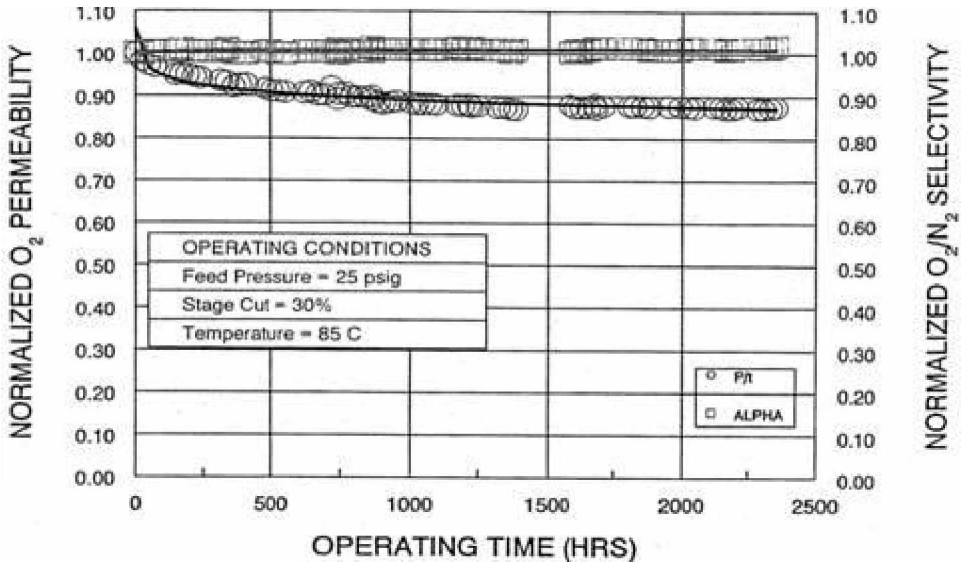

Figure 9-10. Long-term operation

Caterpillar On-Road Tests: This work was then continued by Caterpillar who took five on-road diesel trucks (see Figure 9-11) using a bank of 5-modules in each truck and ran this fleet for approximately one million road miles with minimal loss in performance. This later work was operated under very dynamic conditions and showed the membranes ability to handle these rapidly changing operating conditions.

Figure 9-11. Membrane housing above a truck engine

Caterpillar Particulate Tests: Caterpillar then challenged these membranes with a high load of contamination and once again the membranes showed their ruggedness. Dust fouling tests were conducted on three air separation membrane cartridges at Southwest Research Institute using their air filter performance evaluation test rig. 1000 grams of ultra fine Arizona road dust having a size range of 0-5 microns was passed through each cartridge over a 17 hour period. This mass of dust simulated 435,000 miles in a very dusty environment, and was considered the appropriate amount of dust that would pass through the air filter.

Test results showed that pressure drop through the cartridge doubled over the test period, but the permeation flux rate across the membrane wall was only slightly affected. The test results suggest that the membranes are inherently abrasion resistant and will not be affected by the soot or other abrasives in the flue gas. While management of pressure drop is critical this can be managed by normal flue gas filtration. These results strongly suggest that the CMS membranes are rugged enough to withstand an actual flue gas environment.

CMS Transformer Oil Tests: Lastly CMS-Praxair demonstrated the CMS membrane-polysulfone support ruggedness when exposed to transformer oil degassing for extended periods of time. CMS-Praxair has sold over 200 membrane systems used to measure decomposition gases in transformer oil. In support of this business CMS has established extensive liquid degassing facilities. We have done long-term degassing of transformer oil in our lab. These results over an 18-month period are shown in Figure 9-12. These results show membranes are quite stable to transformer oil after initial break in and therefore suggest comparable stability with amines.

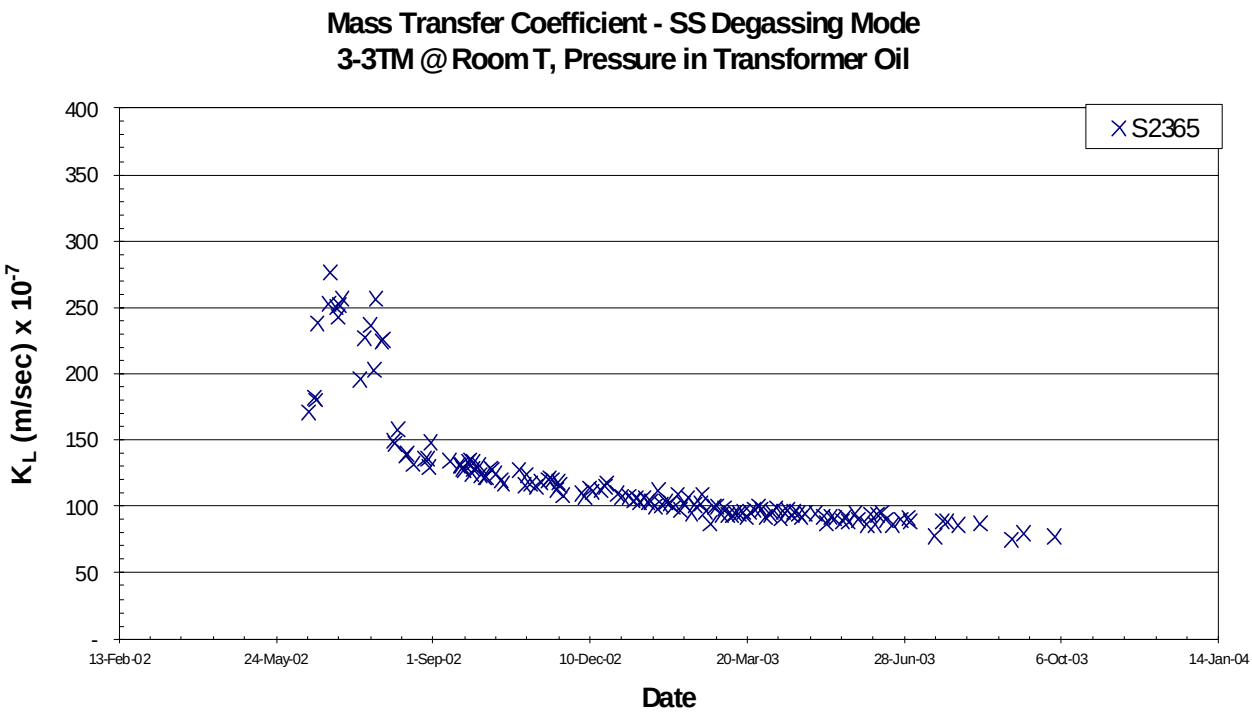

Figure 9-12. Long-term degassing of transformer oil

10. Concluding Remarks

Capture of carbon dioxide at the source of its emission has been a major focus in greenhouse gas emission control. Current technologies used for capturing CO_2 suffer from inefficient mass transfer and economics.

In this phase I project, Compact Membrane Systems, Inc. fabricated and tested a membrane-based absorption system for the removal of carbon dioxide from a simulated power-plant flue gas. The stability of the membrane system under various operating conditions and chemical environments were also tested.

Phase I accomplishments included:

- Demonstration of CO_2 absorption into aqueous diethanolamine solution and subsequent regeneration of the absorbent liquor in membrane contactors at both high temperature (60°C) and low temperature (25°C).
- Integrated system successfully absorbed and desorbed CO_2 simultaneously.
- Preliminary testing indicated the membranes are chemically compatible with amines.
- Preliminary economic analysis showed amine membrane contactor had economic advantage compared to conventional amine scrubbing towers.

The United States has set a goal of reducing the CO_2 emissions intensity of economic activity (pounds of CO2 emitted per dollar of GDP) by 18% by 2012. In order

to meet this goal, we must improve existing technology for capture of CO_2 from flue gas. Existing processes are technically feasible, but economically unsatisfactory. This project has the potential to move us forward toward the goal of an economically feasible process for capture of CO_2 from stack gases.

11. References

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Appendix

Table A-1. Detailed economic analysis

General System Data	**Case 1**	**Base Case 1**	**Case 2**	**Base Case 2**	**Case 3**	**Base Case 3**
Carbon Dioxide Capture System Type	Membrane	Tower	Membrane	Tower	Membrane	Tower
Feed Stream Data Source Reference (See Bibliography)	Chinn, et. al.	Chinn, et. al.	Singh, et. al.	Singh, et. al.	Singh, et. al.	Singh, et. al.
Power Plant Type	Natural Gas	Natural Gas	Pulv. Coal	Pulv. Coal	Pulv. Coal	Pulv. Coal
Power Plant Electricity Output (MegaWatts)	400	400	400	400	400	400
Flue Gas Volumetric Flow Rate (scfm)	1,100,000	1,100,000	775,684	775,684	775,684	775,684
Carbon Dioxide Concentration Vol%	4	4	14.6	14.6	14.6	14.6
Carbon Dioxide Removal Efficiency %	90	90	90	90	50	50
Absorption System Operating Factor (hrs/year)	8000	8000	8000	8000	8000	8000
Approximate MEA Concentration (wt %)	30	30	30	30	30	30
Total gas flow (sm3/min)	31,149	31,149	21,965	21,965	21,965	21,965
Actual Gas Temperature C	50	50	50	40	50	40
CO2 in feed flow kg/min	2272.607	2272.607	5849.374	5849.374	5849.374	5849.374
CO2 in effluent flow kg/min	227.261	227.261	584.937	584.937	2924.687	2924.687
required CO2 removal rate kg/min	2045.346	2045.346	5264.437	5264.437	2924.687	2924.687
Absorption Tower Data						
Number of Absorption Towers	NA	2	NA	4	NA	4
Column Internal Materials	NA	Stainless St.	NA	Stainless St.	NA	Stainless St.
Column Shell Materials	NA	Steel	NA	Steel	NA	Steel
Absorption Tower Diameter (m)	NA	11.3	NA	10	NA	10
Absorption Tower Height (m)	NA	27.9	NA	?	NA	?
Flue Gas Pressure Drop Across Absorb Tower (psi)	NA	3	NA	2.9	NA	2.9
Absorption Membranes Data						
m2 area required	931,457	NA	656,833	NA	80,555	NA
Absorption Membranes Flue Gas Pressure Drop (psi)	0.08	NA	0.08	NA	0.68	NA
Flue Gas Fan Data						
Assumed Fan Design Flue Gas Pressure Drop (psi)	1.08	4.00	2.08	4.90	2.68	4.90
Fan Overall Efficiency	0.65	0.65	0.65	0.65	0.65	0.65
Fan Input Power Requirement kW	5778	20038	7661	17007	9739	17007
Annual Flue Gas Blower Power (kW-hr/yr)	46,222,373	160,307,765	61,288,315	136,057,362	77,910,258	136,057,362
Financial Data						
Maintenance labor rate including overhead ($/hr):	57.69	57.69	57.69	57.69	57.69	57.69
Maintenance labor factor (hr/sh):	1.00	0.00	0.71	0.00	0.09	0.00
General and Administrative Charges %	35	35	35	35	35	35
Property Tax and Insurance	0.04	0.04	0.04	0.04	0.04	0.04
Electrical Energy Cost ($/kW-hr)	0.0546	0.0546	0.0546	0.0546	0.0546	0.0546
Membrane Cost per Unit Area ($/ft2)	$2.60	NA	$2.60	NA	$2.60	NA
Cost of Capital %	15	15	15	15	15	15
Plant Economic Life (years)	20	20	20	20	20	20
Amortization Factor	0.1598	0.1598	0.1598	0.1598	0.1598	0.1598
Lang Factor for Fixed Capital Investment	4.6	4.6	4.6	4.6	4.6	4.6
Purchased Equipment Costs						
Total Cost for Absorption Towers & Packing (Installed)	NA	$42,738,401	NA	$29,536,400	NA	$11,277,234
Total Cost Membrane Cartridges for Absorption	$26,067,897	NA	$18,382,229	NA	$2,254,431	NA
Steel Housings for Adsorption Membranes	$2,139,421	NA	$1,492,619	NA	$199,016	NA
Steel Flue Gas Blowers	$519,689	$1,800,080	$422,603	$1,460,732	$422,603	$1,460,732
Total Purchased Equipment Cost	$28,727,007	$44,538,481	$20,297,451	$30,997,132	$2,876,049	$12,737,966
Total Installed Equipment Cost	$38,560,481	$51,018,769	$27,376,073	$36,255,767	$5,136,421	$17,996,601
Amortized Total Installed Equipment Cost ($/yr)	$6,160,479	$8,150,834	$4,373,642	$5,792,275	$820,602	$2,875,163
Operating Costs						
Annual Electrical Power Expense ($/yr)	$2,523,742	$8,752,804	$3,346,342	$7,428,732	$4,253,900	$7,428,732
Annual Membrane Cartridge Replacement Expense ($/yr)	$2,606,790	NA	$1,838,223	NA	$225,443	NA
Annual Maintenance Labor ($/yr)	$57,843	$0	$40,789	$0	$5,002	$0
Annual Supervisory Labor ($/yr)	$11,569	$0	$8,158	$0	$1,000	$0
Annual Property Tax & Insurance ($/yr)	$1,542,419	$2,040,751	$1,095,043	$1,450,231	$205,457	$719,864
Annual General & Administrative Expense ($/yr)	$936,670	$0	$660,509	$0	$81,006	$0
Total Annual Operating Cost ($/yr)	$7,679,032	$10,793,555	$6,989,064	$8,878,963	$4,771,809	$8,148,596
Total Annualized Cost ($/yr)	$13,839,512	$18,944,388	$11,362,706	$14,671,237	$5,592,411	$11,023,759
Annual Contributory Cost of CO2 Capture ($/Te)	$14	$19	$4	$6	$4	$8

Costs are stated in 2001 Dollars.
NA = Not Applicable