Reducing the Cost of CO 2 Capture from Flue Gases Using Membrane Technology

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SEPARATIONS

Reducing the Cost of CO_2 Capture from Flue Gases Using Membrane Technology

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Studies of CO_2 capture using membrane technology from coal-fired power-plant flue gas typically assume compression of the feed to achieve a driving force across the membrane. The high CO_2 capture cost of these systems reflects the need to compress the low-pressure feed gas (1 bar) and the low CO_2 purity of the product stream. This article investigates how costs for CO_2 capture using membranes can be reduced by operating under vacuum conditions. The flue gas is pressurized to 1.5 bar, whereas the permeate stream is at 0.08 bar. Under these operating conditions, the capture cost is U.S. \$54/tonne CO_2 avoided compared to U.S. \$82/tonne CO_2 avoided using membrane processes with a pressurized feed. This is a reduction of 35%. The article also investigates the effect on the capture cost of improvements in CO_2 permeability and selectivity. The results show that the capture cost can be reduced to less than U.S. \$25/tonne CO_2 avoided when the CO_2 permeability is 300 barrer, CO_2/N_2 selectivity is 250, and the membrane cost is U.S. \$10/m^2.

Introduction

The abatement of greenhouse gases is becoming increasingly important. In 2004, the Australian government issued a White Paper on Energy outlining the national strategies for abating greenhouse gases.[1] One of the key options highlighted was the capture of CO_2 from stationary emissions and subsequent storage in geological reservoirs (carbon capture and storage − CCS). In Australia, net CO_2 emissions exceeded 550 million tonnes in 2003.[2] Of these, over 45% were from post-combustion coal-pulverized power plants and because of this it is likely that these will be the initial focus of CCS in Australia.

Several technologies have been proposed to capture CO_2 from power-plant flue gas including absorption, adsorption, cryogenic distillation, and membrane gas separation.[3] The technology examined in this article is polymer-based membrane gas separation. Although membranes have not yet been used commercially to recover CO_2 from flue gas, they are used extensively for CO_2 recovery in natural-gas processing and show promise for other applications.[4−6] One of the advantages of a gas membrane separation process is its simplicity; there is no need to add chemicals or to regenerate an absorbent/adsorbent. In addition, membranes are compact, relatively simple to operate, and can be retrofitted easily onto the tail end of power-plant flue gas streams without requiring complicated integration.

The driving force across a gas-separation membrane is the pressure differential (ΔP) between the feed side and the permeate side, as shown by Fick's Law.

$$J_i = \frac{P_i^*}{\delta} A_m(x_i \mathrm{P}_f - y_i \mathrm{P}_p) \quad (1)$$
$$= \frac{P_i^*}{\delta} A_m \Delta P$$

where J_i is the flux across the membrane [kmol/s], P_i* is the permeability value for component i [kmol·m/(s·m^2·bar)], δ is the membrane thickness [m], A_m is the membrane area [m^2], x_i and y_i are the mole fraction of the component i in the feed and permeate sides, and P_f and P_p are the pressures in the feed side and permeate side |, respectively. The permeability P_i* is also commonly expressed in barrer, where 1 barrer is equivalent to 3.348×10^{-12} kmol·m/(s·m^2·bar).

One option for achieving the driving force across the membrane is to set the permeate stream at atmospheric pressure and compress the feed gas to a higher pressure. This has been the assumption of the majority of previous studies that have investigated the feasibility of membrane technology for CO_2 capture.[7−11] These studies showed that the cost for CO_2 capture was at least 30% higher than for CO_2 recovery using amine chemical absorption. In our previous work,[10,11] we have shown that the cost is high because of the high capital costs associated with compressors needed to compress the low-pressure flue gas and the low CO_2 purity product stream. Under these operating conditions (where the feed gas is compressed to 15 to 20 bar),

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Table 1. Processing Conditions and Composition of a Typical Supercritical Bituminous Power-Plant Flue Gas

parameter	value
power plant size (net)	500 MW
boiler type	supercritical
coal type	Australian bituminous
thermal efficiency (LHV)	40%
temperature	110 °C
pressure	0.95 bar
volumetric flowrate, mole fraction of component gases in the flue gas stream	670 m^3/s 484 m^3(STP)/s
CO_2	0.13
N_2	0.75
O_2	0.05
H_2O	0.07
SO_x	220 ppm
NO_x	<50 ppm
initial CO_2 emission	0.83 kg/kWh

the feed-gas compressors and post-separation compressors account for over 50% of the capital and operating costs. In addition, we also showed that relative improvements in membrane permeability and selectivity resulted in only a small reduction of the capture cost. This is because the membranes are less than 10% of the total capital cost.

This article investigates the effects of using vacuum permeate pumping rather than compressing the feed stream to very high pressures. The pressure in the permeate stream is below atmospheric, whereas the feed gas is close to atmospheric pressure. Vacuum permeate conditions have been successfully implemented in pilot and commercial-scale projects for oxygen enrichment[12,13] and membrane distillation.[14] The same principles can be applied to CO_2 recovery systems.

Experimental work undertaken by Ge et al.[15] investigated vacuum conditions using an enzyme-based facilitated transport membrane to recover CO_2 from a respiratory gas stream in advanced life support applications. They found that vacuum permeate conditions enhanced the overall recovery of low-concentration CO_2. In addition, Hagg and Lindbrathen[16] have investigated CO_2 capture from a natural-gas fired power-plant flue gas. Their results showed that to recover 90% of the CO_2 from a low CO_2 concentration of flue gas (less than 4%), the membrane process required a feed pressure of at least 4 bar if the downstream (permeate) pressure is 0.1 bar.

No economic evaluations have yet been reported for the cost of CO_2 recovery using gas-separation membranes under vacuum conditions. The analysis in this article compares the cost of capturing CO_2 using gas-separation membranes under vacuum conditions with previous studies. It also examines to what extent improvements in permeability, selectivity, membrane unit costs, and operating parameters under vacuum permeate conditions can reduce the capture cost.

Methodology

The Flue Gas. This study examines the post-combustion capture of CO_2 from the flue gas of a supercritical bituminous black-coal power plant. The CO_2 capture facility is integrated into a newly built power plant with a net power output of 500 MW.

The assumed characteristics of the flue gas are given in Table 1. The flowrate, composition, and conditions are based on the flue gas of an east coast Australian power plant examined in the study by Dave et al.[17]

CO_2 Capture. This analysis assumes that at least 85−90% of the CO_2 contained in the flue gas is recovered in the permeate stream. The separated enriched CO_2 stream is compressed to 100 bar for transport. For cases where the purity of the CO_2 in the product stream is less than 90%, because the solubility of gases such as nitrogen and oxygen in supercritical CO_2 is very low, these gases can be separated from the CO_2 prior to transport in the post-capture compressor. This is achieved using compression and cooling of the product via vapor−liquid separation gas to remove (or reject) the gaseous components and any CO_2 vapor.

The analysis also assumes that water vapor is removed from the flue gas by molecular sieves, that the SO_x content of the flue gas is reduced to 10 ppm[18] in a flue gas desulphurisation (FGD) unit, and that NO_x is removed in the power plant before CO_2 recovery.

Three process layouts were chosen to evaluate the cost and performance of CO_2 capture using membranes. These include a single stage membrane system (SMS) and a two-stage cascade membrane system (TCMS, TCMS-RR) with and without retentate recycle.[11] Figures 1 and 2 show the simplified schematics of the SMS and TCMS-RR processes.

For the baseline economic evaluation of the vacuum permeate system, the feed gas is set at 1.5 bar and the permeate pressure is 0.08 bar. Vacuum conditions are obtained using positive displacement pumps, which have operating pressures down to 0.04 bar. Accounting for leaks in the system, the minimum operating permeate pressure in this analysis is assumed to be 0.05 bar.

For the high-pressure feed operation, the feed-gas pressure is 15 bar and the permeate pressure is set at atmospheric.

The poly (phenylene oxide) (PPO) membrane manufactured by Delair (now Aquilo Gas Separation BV) is selected as the baseline membrane because it is a membrane with a moderate CO_2 permeability of 70 barrer and a reasonably good CO_2/N_2 selectivity value of 20.[7] The thickness of the membrane is assumed to be 125 μm.

Simulation Program. Calculations for the membrane process were carried out using an in-house technoeconomic model

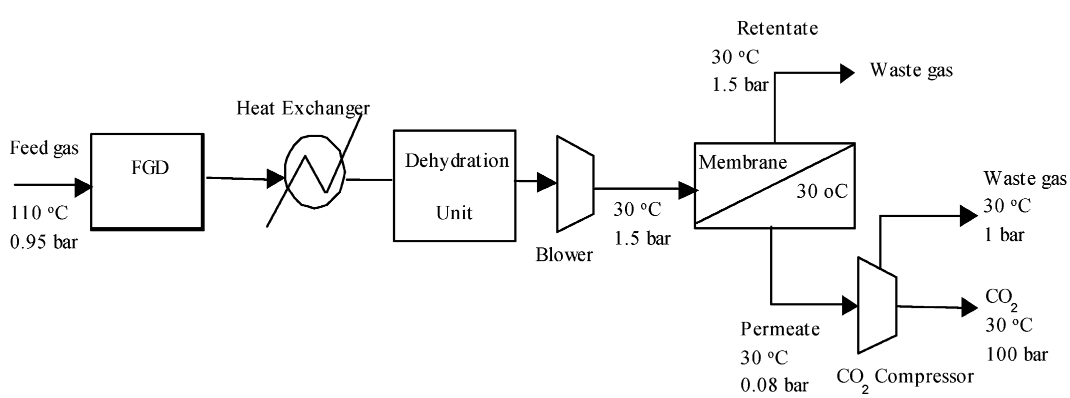

Figure 1. Simplified diagram of SMS with vacuum permeate conditions.

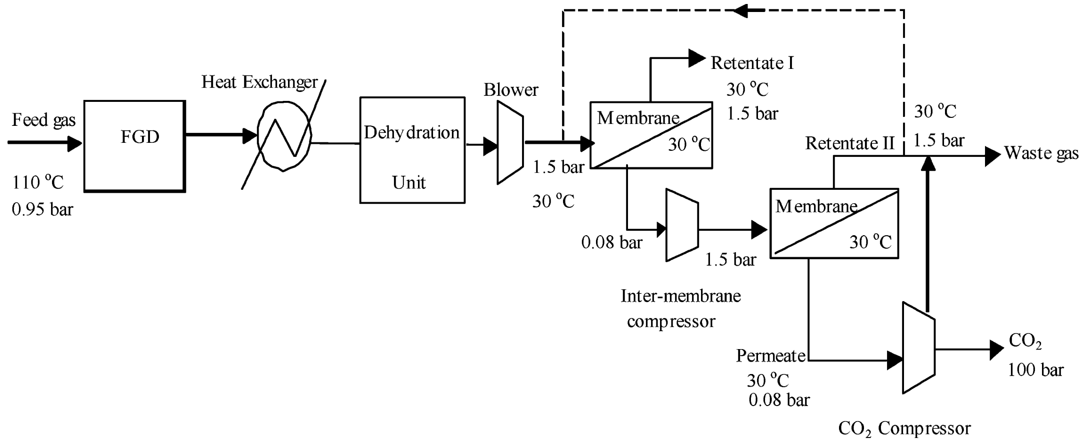

Figure 2. Simplified diagram of TCMS with vacuum permeate conditions. The dotted lines show the additional retentate recycle in the TCMS-RR process.

Table 2. Summary of Economic Inputs

cost year	2005
discount rate	7% pa (real)
project life	25 years
construction period	2 years
plant operating capacity	85%
currency	US$
cost of fuel (coal)	1.5 $/GJ LHV
baseline cost of electricity	34 $/MWh
membrane price	50 $/m^2

developed by the University of New South Wales for the CO2CRC.[19] The model adopts the cross-flow model described by Shindo et al.[20] to describe the CO_2 recovery across a single membrane. The input data for the model include the feed composition and volume, process conditions such as pressure and temperature, and membrane characteristics such as the membrane thickness and component permeability values. The membrane model calculates the required membrane area, stage cut, and volumes and compositions of the gas components in the permeate and retentate streams. For simplicity, we assume that plastization effects can be neglected and that the flue gas consists of only CO_2, N_2, and O_2.

The power consumption required for compression, expansion, pumping, and cooling is calculated using general mass and energy balances. For the high-pressure feed operation, we assume that the retentate stream is expanded from 15 bar to atmospheric pressure. The energy gained from the expansion is used to offset some of the energy consumed by the system for compression. The efficiency of the compressors, expanders, and vacuum pumps is assumed to be 85%.

Economic Evaluation. The economic assumptions used in this analysis are listed in Table 2 and are based on the CO2CRC guidelines.[19] The reference plant is the same plant without an integrated CO_2 capture facility. Capital and operating costs are estimated for CO_2 recovery, preseparation treatment, and CO_2 compression. The methodology for estimating capital equipment, operating, and capture costs are outlined in Tables 3−4. The total capital cost includes all of the process equipment shown in Figures 1 and 2, plus a general facilities cost. The general facilities cost includes ancillary equipment such as storage tanks, spare pumps, valves, and the control system. The operating cost includes the fixed general maintenance cost comprising of labor, nonincome government taxes that may be payable, and the general insurance cost. The variable operating costs include costs for the flue gas desulphurisation, cooling water and membrane replacement. It is assumed that this cost is incorporated in the general maintenance cost. The costs are based on current membrane technologies.

Table 3. Membrane System Capital Costs: Model Parameters and Nominal Values

	capital cost elements	nominal value
A	process equipment cost (PEC)	sum of all process equipment
B	general facilities	10−20% PEC
	total equipment cost (TEC)	A + B
C	instrumentation	10% TEC
D	piping	10% TEC
E	electrical	5% TEC
F	total installed cost (TIC)	A + B + C+ D + E
G	start-up costs	1% TIC
H	engineering	5% TIC
I	owners costs	7% (F + G + H)
J	engineering, procurement, construction and owner's cost (EPCO)	F + G + H + I
K	project contingency	10% EPCO
	total capital cost (CAPEX)	= J + K

Table 4. Membrane Systems Operating Costs: Model Parameters and Nominal Values

operating cost elements	nominal value
Fixed Operational and Maintenance (FOM) Cost	
general cost	4% Capex
insurance cost	2% Capex
Variable Operational and Maintenance (VOM) Cost	
FGD operating cost	\$100/tonne SO_x removed
cooling water	\$0.01/m^3
membrane replacement price	\$50/m^2
expected membrane life	3 years

The feasibility of a CO_2 recovery technology is measured in terms of cost per tonne of CO_2 avoided (U.S. \$/tonne CO_2 avoided), which is the CO_2 avoidance cost. This is the additional cost of establishing a CO_2 capture facility for an industrial plant or power plant. It can be calculated using

$$\$/\text{tonne CO}_2\text{ avoided} = \frac{\text{COE}_0 - \text{COE}_{\text{CCS}}}{\text{CO}_{2,0} - \text{CO}_{2,\text{CCS}}} \qquad (2)$$

where COE is the cost of electricity (\$/MWh), CO_2 is the rate of CO_2 emissions (tonnes/MWh), and the subscripts 0 and CCS denote without and with CO_2 capture, respectively. Further details are reported by Allinson et al.[19]

In this analysis, many simplifications have been made including neglecting the effects of plasticisation and membrane degradation and the use of a simplified membrane design. In addition, the capital costs are highly uncertain. Therefore, the results are not accurate on an absolute basis but are indicative of the possible cost reductions achievable through technology improvement.

Results for Vacuum Permeate Conditions. Table 5 compares the difference in capture costs for the PPO membrane

Table 5. Comparison of Capture Cost for Gas-Separation Membranes under Vacuum Conditions

membrane system	membrane area $\times$ 10^3 (m^2)	% CO_2 in the permeate	energy penalty (%)	capture cost (U.S. \$/tonne CO_2 avoided)
high-pressure feed SMS	200	43	52	82
vacuum SMS	1850	45	28	54
vacuum TCMS	2800	74	34	75
vacuum TCMS−RR	2550	77	33	71

under vacuum permeate conditions with costs using a high-pressure feed. In our previous work,[11] we showed that the capture cost for gas-separation membranes with the TCMS configuration and a high-pressure feed is approximately 15% higher than that for the SMS configuration. As the objective of this article is to compare membrane systems under high-pressure feed conditions with vacuum permeate conditions, for simplicity only the capture cost for the high-pressure feed with the SMS layout is shown.

The cost of capturing CO_2 using gas-separation membranes under vacuum permeate conditions is U.S. \$54/tonne CO_2 avoided. This is 65% of the cost using a pressurized feed. Vacuum permeate conditions also yield lower costs for both of the two-stage membrane systems (TCMS and TCMS-RR). The costs are lower because:

(1) In the vacuum process, there is no need for a feed compressor, thus reducing the capital cost.

(2) The total energy consumption for the vacuum process is less than that for the high-pressure feed system, with energy penalties below 35% for the vacuum process compared to 50% for the high-pressure feed process, thus reducing the operating cost.

Together, these factors lower the overall cost and increase the amount of CO_2 avoided.

By operating the membrane system under vacuum permeate conditions, the contribution of the membrane system to the overall capital cost is much higher than that for the high-pressure feed process. As shown in Figure 3, for vacuum permeate systems the membranes are the largest cost, accounting for over 40% of the total equipment costs. In comparison, for the high-pressure feed process, the membrane cost accounts for less than 10% of the total equipment costs. There are two reasons for the increase in membrane costs for vacuum permeate processes. First, without the feed-gas compressor, the cost contribution of the compressors to the overall total capital expenditure decreases. Second, the transmembrane pressure difference in vacuum permeate systems is less than that for the high-pressure feed process. In this analysis, the amount of CO_2 recovered per second (J_i) is fixed at 85−90% of the flue gas or 2.49 kmol/s. In addition, P_i* and δ are fixed. Because ΔP is reduced for the vacuum permeate systems, Fick's law (eq 1) shows that the required membrane area (A_m) must increase to maintain a fixed value for J_i. Hence, the cost of the membrane unit increases under vacuum conditions.

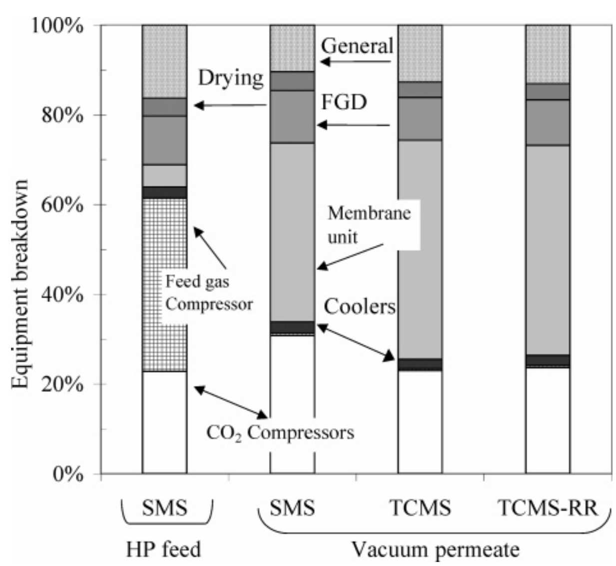

Figure 3. Percentage breakdown of equipment cost for gas-separation membrane systems under vacuum permeate and high pressure (HP) feed conditions.

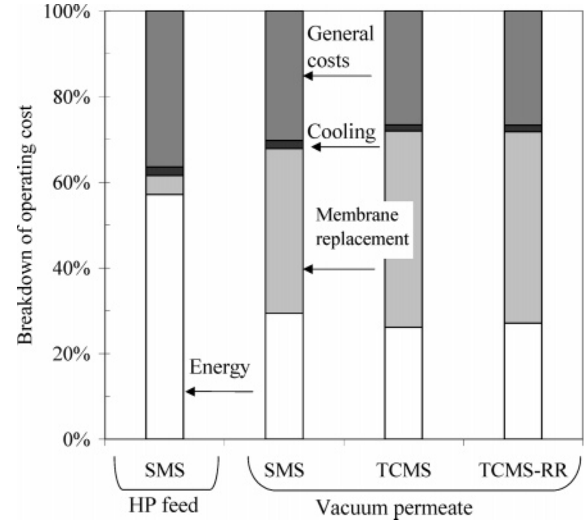

Figure 4. Breakdown of operating costs for gas-separation membranes operating under vacuum permeate and high pressure (HP) feed conditions.

The second-largest capital cost item is the post-separation compressor, also shown in Figure 3. The membrane unit and CO_2 compressor together account for approximately 70% of the total equipment costs of vacuum permeate membrane systems. Figure 4 illustrates that the two most expensive items are replacing the membranes and the energy required for compression. The results from Figures 3 and 4 indicate that capital and operating costs savings could be achieved by:

(1) Reducing the cost of the membrane unit by lowering its unit cost and/or making the membrane area smaller.

(2) Reducing the cost of the CO_2 compressor by increasing the purity and decreasing the flowrate of the permeate.

The Effect of the Membrane Price. Figure 5 shows the changes in the CO_2 capture cost as the price of the PPO membrane increases for a membrane with a high-pressure feed and a membrane with vacuum permeate conditions. In regard to the membrane with high-pressure feed, because the membranes account for less than 10% of the overall capital cost (Figure 3), reductions in the membrane price do not have a significant effect. In comparison, in the vacuum permeate membrane systems, reducing the membrane price from U.S. \$50/m^2 to U.S. \$10/m^2 lowers the capture cost to almost U.S. \$35/tonne CO_2 avoided. However, if the membrane cost is very high (above U.S. \$130/m^2) it would be more economical to operate with a high-pressure feed than under vacuum permeate conditions. This is because the membrane area is substantially higher in the vacuum permeate systems (as shown in Table 5).

The Effect of CO_2 Permeability. The CO_2 permeability will influence the rate at which CO_2 is removed from the feed gas. For a fixed flux of CO_2 across the membrane, increasing the membrane's CO_2 permeability will decrease the required membrane area and thus reduce the capital cost. Figure 6 shows

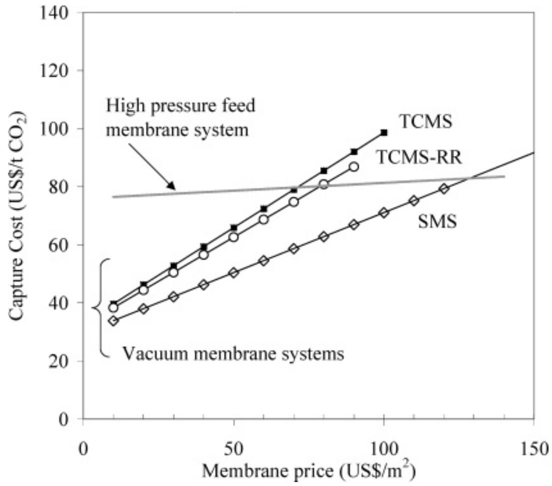

Figure 5. Change in the capture cost (U.S. \$/tonne CO_2 avoided) as a function of the membrane price for the vacuum membrane systems SMS (◇), TCMS (■), and TCMS-RR (○), and the SMS high pressure feed membrane system (—).

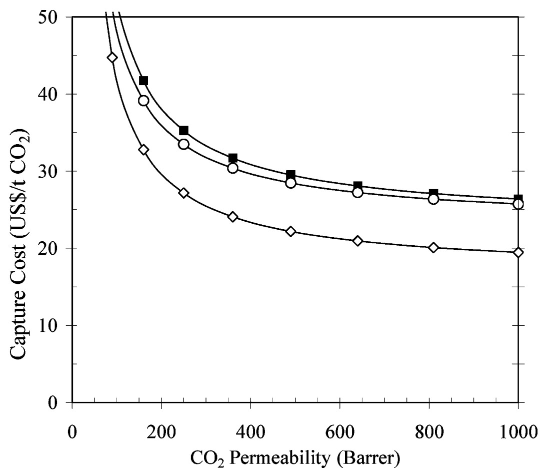

Figure 6. Effect of CO_2 permeability on the CO_2 capture cost (U.S. \$/tonne CO_2 avoided) for different vacuum membrane systems; SMS (◇), TCMS (■), and TCMS-RR (○).

the change in capture cost for the vacuum membrane systems as the CO_2 permeability increases from 10 to 1000 barrer. The CO_2/N_2 selectivity of the membrane is assumed to be constant at 20 and is not affected by changes in the permeability.

Increasing the membrane CO_2 permeability from 70 to 350 barrer lowers the capture cost from U.S. \$50 to \$25 per tonne CO_2 avoided for the SMS process layout and from approximately U.S. \$75 to \$30 per tonne CO_2 avoided for the two-stage process layouts. This is because the membrane area is almost 5 times smaller and the total capital cost decreases by 50%. However, the capture cost does not decrease significantly for CO_2 permeabilities greater than 500 barrer because beyond this point the required change in membrane area is not significant.

The Effect of CO_2 Selectivity. By changing to vacuum permeate conditions, the capture cost of recovering CO_2, especially for the SMS configuration, is significantly reduced. However, one disadvantage is that the purity of the CO_2 in the permeate stream is low. It is less than 50% for the SMS and less than 80% for the TCMS-RR. For gas-separation membranes, to compete with other CO_2 capture technologies such as chemical absorption and cryogenic distillation where the CO_2 purity in the product is high (greater than 98%), they should also produce high purity CO_2. The following analysis examines

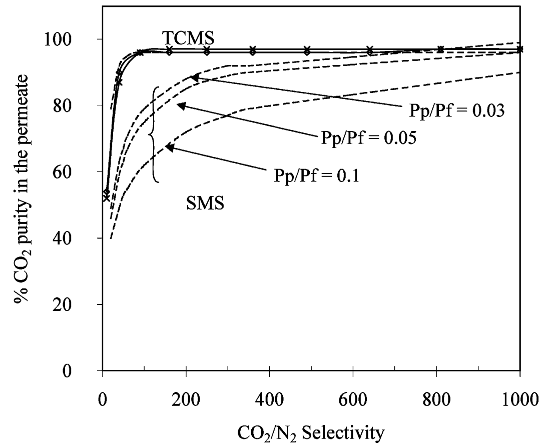

Figure 7. Effect on permeate CO_2 purity with changes in the CO_2/N_2 selectivity at pressure ratios of 0.01 to 0.1.

the parameters that affect the CO_2 purity and the impact on capture costs.

The parameters that affect the CO_2 purity are the pressure ratio and the membrane CO_2/N_2 selectivity. Using Fick's law, the ratio of the fluxes for the CO_2 to N_2 is

$$\frac{n(\mathrm{CO_2})}{n(\mathrm{N_2})} = \frac{P^*_{\mathrm{CO_2}}}{P^*_{N_2}} \frac{\left(x_{\mathrm{CO_2}} - y_{\mathrm{CO_2}}\dfrac{P_p}{P_f}\right)}{\left(x_{\mathrm{N_2}} - y_{\mathrm{N_2}}\dfrac{P_p}{P_f}\right)} \tag{3}$$

Thus, the ratio of the flux of CO_2 compared to the flux of N_2 (or the purity of CO_2 in the permeate stream) can be increased by (1) increasing the CO_2/N_2 selectivity and/or (2) decreasing the pressure ratio (ratio of the permeate pressure to the feed pressure).

Figure 7 summarizes the effect on the permeate CO_2 purity of changes in the CO_2/N_2 selectivity at different pressure ratios. A selectivity of 1000 is required to obtain 90% CO_2 in the permeate stream at a pressure ratio of 0.1. At a lower pressure ratio of 0.03, a selectivity of 250 is required to obtain a CO_2 purity of 90%.

Figure 7 also shows the relationship between the CO_2 purity in the permeate stream and pressure ratio and selectivity combinations for the two-stage membrane systems. At a pressure ratio of 0.05, 0.07, or 0.1, to obtain 90% CO_2 purity the selectivity is approximately 40 for both of the two-stage processes. Increasing the membrane selectivity beyond 40 does not significantly improve the CO_2 purity. This is because the two-stage configurations are designed to obtain a high CO_2 purity product. Therefore, improvements in selectivities do not produce noticeable changes in the concentration of CO_2 in the permeate stream.

Figure 8 illustrates how capture cost changes with changes in CO_2/N_2 selectivity for the high-pressure feed SMS and both the vacuum permeate-based TCMS and SMS. In this analysis, it is assumed that the selectivity of CO_2 compared to N_2 increases along with the selectivity of CO_2 compared to O_2. Alentiev and Yampolskii[21] have shown that the relative permeability of N_2 to O_2 is five in the majority of glassy polymer membranes. Thus, the CO_2/O_2 selectivity is determined by dividing the CO_2/N_2 selectivity by five. The pressure ratio for all of the systems is 0.03, and the CO_2 permeability is assumed to be constant at 70 barrer.

The results show that for all of the systems, as the CO_2/N_2 selectivity increases from 10 to 40, the capture cost decreases.

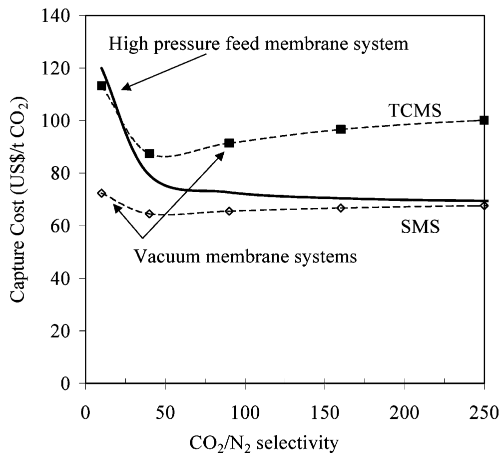

Figure 8. Capture cost as a function of CO_2/N_2 selectivity for the high-pressure feed SMS system (—) and for different vacuum membrane systems (- -); SMS (◇) and TCMS (■).

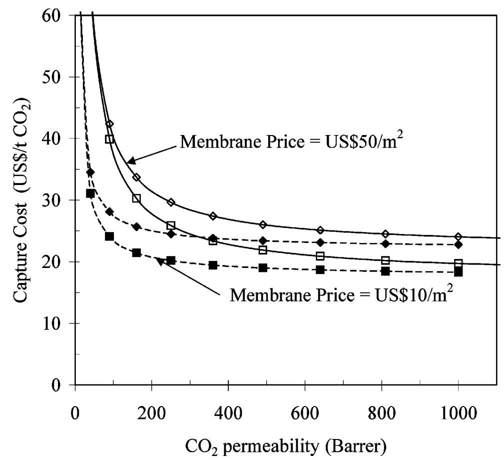

Figure 9. Capture cost as a function of improved membrane characteristics at membrane prices U.S. \$50/m^2 (—) and U.S. \$10/m^2 (- -) and CO_2/N_2 selectivity of 40 (◇) and 250 (□).

However, for CO_2/N_2 selectivities above 40, the capture cost decreases for the high-feed pressure system but increases for the vacuum permeate systems.

From Fick's law (eq 1), increasing the selectivity of CO_2/N_2 increases the mole fraction of CO_2 in the permeate (y_i) and decreases the mole fraction of CO_2 in the retentate (x_i). Consequently, the driving force across the membrane is also reduced. To obtain the same amount of CO_2 recovered (J_i), the membrane area (A_m) increases and thus the capital costs for membranes also increases. However, as the CO_2 purity of permeate stream (y_i) increases, the flowrate of the permeate decreases, resulting in a smaller post-separation compressor. It is the balance between the cost savings generated by the post-separation compressor and the increase in membrane costs that influences the cost trends.

For the high-pressure feed membrane system, with increasing CO_2/N_2 selectivity the costs of the CO_2 compressor decrease while the costs of the membranes increase. However, because the membranes account for only a small component of the capital cost and the post-separation compressor accounts for a much larger proportion, the cost benefits of the smaller compressor outweighs the increasing membrane cost.

For the vacuum membrane systems and CO_2/N_2 selectivity values up to 40, the cost reduction for the CO_2 compressor also outweighs the cost increase for the membranes; however, for selectivity values above 40, the very large membrane area required and the associated costs exceed the cost reductions achieved by having a smaller CO_2 compressor. To capitalize on the benefits of the higher CO_2 permeate purity, improvements in CO_2 selectivity must be coupled with permeability improvements.

Improved Membrane Characteristics and Lower Membrane Cost. Using the results from the above analysis, the following section investigates the possible cost reductions that can be achieved when improved membrane characteristics and vacuum permeate conditions are coupled with membrane price reductions.

Figure 9 illustrates for the SMS configuration how the capture cost of CO_2 decreases with increasing CO_2 permeability at different membrane prices. The results show that the capture cost can be reduced to almost U.S. \$20/tonne CO_2 avoided when the CO_2 permeability is 300 barrer; the CO_2/N_2 selectivity is 250 and the membrane price is \$10/m^2. The reductions in the membrane price have a more significant effect at low permeabilities than at high permeabilities. The results show that for CO_2 permeabilities of up to 500 barrer, reducing the membrane price will have a greater impact on reducing the overall capture cost rather than increasing the CO_2/N_2 selectivity at a fixed CO_2 permeability. However, for CO_2 permeabilities above 500 barrer, the combined effect of CO_2 permeability and CO_2/N_2 will have a greater impact. This occurs because at permeabilities above 500 barrer, the total membrane area is a smaller contributor to the total capital costs. Similar trends are observed for the other process configurations.

The results indicate that for a membrane with a CO_2 permeability of approximately 200 barrer, reducing the membrane price through, for example, reducing membrane production costs will result in a lower capture cost than will efforts at improving the CO_2 selectivity. Conversely, if a membrane had an exceptionally high CO_2 permeability value (greater than 800 barrer), the capture system would be made less costly by improving the CO_2/N_2 selectivity than by lowering the membrane price.

Conclusion

The recovery of CO_2 from post-combustion flue gas using polymeric gas-separation membranes can be achieved at a cost of between U.S. \$20 and \$40 per tonne CO_2 avoided. This is accomplished with a vacuum membrane system, improvements in membrane CO_2 permeability, CO_2/N_2 selectivity, and membrane cost reductions. Currently, there exist no commercial membranes with these combinations of high permeabilities and selectivities for CO_2/N_2. Additionally, current commercial membranes suffer from degradation of performance over time due to a variety of factors, and operation is limited to near-ambient temperature. These are the challenges that will be needed to be overcome to make CO_2 capture using membrane technology a competitive option.

Currently, CO_2 capture by gas-separation membranes is not as effective as other CO_2 recovery methods because of the low permeability and selectivity of commercially available membranes. The results show that if the membrane price is moderate to high (U.S. \$40−50/m^2) then developing new membranes for vacuum permeate membrane systems should focus on creating membranes with very high permeability (300−550 barrer) and moderate CO_2/N_2 selectivity (40−60). However, if the membrane price is low (U.S. \$10−30/m^2) then it would be better to use existing membranes with permeabilities on the order of 200 barrer coupled with CO_2/N_2 selectivities of 40−60.

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