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Economic Evaluation of Membrane Systems
for Large Scale Capture and Storage of CO₂
Mixtures

Minh T. Ho*

Dianne E. Wiley†

*UNESCO Centre for Membrane Science, The University of New South Wales

†Australian Cooperative Research Centre for Greenhouse Gas Technologies (CO₂CRC)

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Abstract

The capture and storage of CO₂ (CCS) as a greenhouse mitigation option is becoming an increasingly important priority for Australian industry. Membrane based CO₂ removal systems can provide a cost effective, low maintenance approach for removing CO₂ from gas streams. This study examines the effect of membrane characteristics and operating parameters on CCS costs using economic models developed by UNSW for any source-sink combination. The total sequestration cost per tonne of CO₂ avoided for separation, transport and storage are compared for the separation of CO₂ from coal fired power plants and natural gas processing. A cost benefit analysis indicates that sequestration of gases of high purities are dominated by compression costs which can be off-set by utilising membranes of higher selectivity coupled with higher permeability to reduce the required transmembrane pressure.

ECONOMIC EVALUATION OF MEMBRANE SYSTEMS FOR LARGE SCALE CAPTURE AND STORAGE OF CO₂ MIXTURES

Minh T. Ho^{1,2} and Dianne E. Wiley^{*1,2}

1. UNESCO Centre for Membrane Science, The University of New South Wales
2. Australian Cooperative Research Centre for Greenhouse Gas Technologies (CO2CRC)

* T: +61 (2) 938 55541, F: +61 (2) 938 55456, E: D.Wiley@unsw.edu.au

ABSTRACT

The capture and storage of CO₂ (CCS) as a greenhouse mitigation option is becoming an increasingly important priority for Australian industry. Membrane based CO₂ removal systems can provide a cost effective, low maintenance approach for removing CO₂ from gas streams. This study examines the effect of membrane characteristics and operating parameters on CCS costs using economic models developed by UNSW for any source-sink combination. The total sequestration cost per tonne of CO₂ avoided for separation, transport and storage are compared for the separation of CO₂ from coal fired power plants and natural gas processing. A cost benefit analysis indicates that sequestration of gases of high purities are dominated by compression costs which can be off-set by utilising membranes of higher selectivity coupled with higher permeability to reduce the required transmembrane pressure.

INTRODUCTION

CO₂ capture and storage (CCS) is an important short-term lever for addressing climate change. Over the course of the century, CCS could account for 30% or more of all climate mitigation beyond “business as usual” technology improvements (1). To meet reduction targets, CCS technologies are likely to be deployed on a large scale around the globe.

Early studies of CO₂ capture with membranes indicated that the cost was 30% higher in cost than amine chemical absorption (1, 2). The limitations of the studied membranes was due to the high cost of compressing low pressure flue gas and the low purity of permeate which results in the need for multi-stage processing to achieve the most economic arrangement for CCS. The objective of this study is to examine the effect of the membrane characteristics of permeability and selectivity and operating parameters such as trans-membrane pressure on the total cost of CCS.

METHOD

The economic analysis for this paper examines the cost of CO₂ capture and storage of mixtures for both a 500 MW coal-fired power plant and a 35 MMSCFD natural gas processing facility. The analysis for the coal-fired power plant assumes that the flue gas is from a typical 500 MW Australian pulverised black-coal power plant. The analysis for the natural processing facility

assumes a typical offshore Australian facility, with processing conditions that meet the Australian pipeline specifications (3). It is assumed that the feed gas for both processes is dehydrated and all contaminant gases such as NO_x, SO_x and H₂S have been removed. The conditions for the geological reservoir are the same as those in Ho et al (4). A summary of the input parameters is given in Table 1. The specifications of the two polymeric membranes considered in this study are shown in Table 2. For convenience, the power plant flue gas is referred to as the CO₂/N₂ system and the natural gas facility is referred to as the CO₂/CH₄ system.

In this study, the effect of removal efficiencies and the purity of product using gas separation membranes for CO₂ recovery and hence the subsequent effect on storage of a mixed gas product is evaluated. Both 1-stage and 2-stage membrane systems were investigated. To simulate a 2-stage system, the output from the first stage is taken as the input for the second stage simulation. The membrane was modelled using a modified cross-flow model with no recycle stream as described by Shindo et al (5). This cross-flow model enables a multi-component gas mixture to be examined and provides a reasonable first approximation of real systems which have many membranes modules operating in series and parallel in each stage.

One of the consequences of using gas separation membranes is that the permeate or retentate streams contains other component gases as well as the desired gas because membranes are not perfect separators. To increase the concentration of CO₂ in the gas stream to a level which is economically viable for storage, the permeate stream from the first membrane, which is enriched in CO₂, is recompressed and passed through a second membrane (6). This configuration is referred to as a two-stage cascade membrane system (TCSM) as shown in Figure 1.

The second 2-stage membrane system examined in this study was the two-stage series membrane system (TSMS), as shown in Figure 2. In this process set-up, the retentate stream from the first membrane is passed through a second membrane, removing further CO₂ (7). This configuration was used for the recovery of CO₂ from natural gas streams to achieve the desired pipeline specifications of less than 2% CO₂.

In this study, it was assumed that the power requirement needed for the CO₂ separation process and compression stages is provided from a supplementary power supply. A standard assumption made purely for the purposes of this study is that the supplementary energy will come from a new natural gas combined cycle power plant ('NGCC'). The CO₂ emission from the NGCC power plant is assumed to be 0.4 kg CO₂ per kWh¹⁵ (4). Due to the lower concentrations of CO₂ in NGCC flue gases, we assume that such CO₂ emissions are vented to the atmosphere and not captured. Therefore, they contribute to the total CO₂ emissions of the system. The net tonnes of CO₂ avoided is the difference between the tonnes of CO₂ stored and the tonnes of CO₂ emitted after capture. The percent CO₂ avoided is calculated as:

$$\% \text{ CO}_2 \text{ avoided} = \frac{\text{CO}_2 \text{ captured} - \text{CO}_2 \text{ emitted from supplementary power}}{\text{CO}_2 \text{ original emission from source}} \quad (1)$$

In assessing the economic feasibility of membranes for the purposes of CCS, the economic indicator of \$/tonne CO₂ avoided or cost of CO₂ avoidance is widely used. The \$/tonne CO₂ avoided describes the total capital and operational investment needed for the purposes of CCS and is calculated as:

$$\text{Cost of CO}_2\text{ avoided} = \frac{\sum_{i=1}^n \frac{K_i + O_i}{(1+d)^i}}{\sum_{i=1}^n \frac{(\text{CO}_2 \text{ avoided})_i}{(1+d)^i}} \quad (2)$$

where K_i and O_i are the real capital and operating costs (US\$ million) in the i th year of operation, d is the discount rate (% pa) and CO_2 avoided is the annual amount of CO_2 avoided in million tonnes. The economic assumptions used in this study are listed in Table 3.

RESULTS AND DISCUSSION

Relative cost breakdown

For low pressure feed gas mixtures such as the flue gas from a power plant, results in Figure 3 shows that the largest contributor to capital costs for the membrane system is the cost of the compressor required to elevate the pressure of the inlet feed gas to a suitable level before separation, coupled with the cost of the expanders needed to de-pressurise the retentate waste gas stream. The compressor and expanders account for approximately 80% of the total investment, while the membrane and membrane housing account for less than 15% for both a 1-stage and 2-stage membrane system (8).

For the natural gas facility, the largest cost item is for the membrane and associated housing, which accounts for 62% to 85% of the total capital cost as shown in Figure 4. This is because there is no high compression cost associated with the natural gas feed, which is already available at a high pressure. For a 1-stage membrane system there is no compression cost, however for a 2-stage membrane system such as a TCMS configuration an intermediate compressor is used. From the cost analysis, the cost of the compressor required to recompress the CO_2 enriched stream accounts for approximately 30% of the total cost. This is significantly less than for the low-pressure flue gas system.

The effect of permeability and thickness

The CO_2 permeability will influence the rate at which CO_2 is removed from the feed gas by the membrane. Higher CO_2 permeabilities will reduce the size of the membrane needed for separation. Figure 5 shows that increasing the CO_2 permeability by factors of 10 from 1 Barrer to 1000 Barrer results in significant reductions in the cost of capture for both the power plant flue gas and natural gas processing feed gas. This cost reduction is most noticeable between CO_2 permeabilities of 1 and 100. This reduction occurs due to the reduced area of membrane needed for CO_2 separation. Figure 5 also shows the effect on cost by changing the permeance; the ratio of the permeability divided by the membrane thickness. For the same CO_2 permeability, if the membrane thickness is halved from 0.1 μm to 0.05 μm , the cost of capture for both CO_2 feed gases also reduces. This reduction is most noticeable between permeabilities of 1 and 100 and is also a result of the reduced area of membrane needed. However, as the permeability increases to values greater than 100, the economic gain in reducing the membrane area is not as significant. This is due to the small cost contribution of membranes to the overall capital

investment as shown in Figures 3 for the power plant flue gas system, and the low membrane cost assumed in Table 3.

The capture cost for the natural gas feed gas is substantially less than for the power plant flue gas at the same CO₂ permeability values. This is because the natural gas feed is already provided at a high pressure eliminating the need for a feed gas compressor. The economic analysis assumes that compressors are more expensive than membranes. The capital expenditure for the natural gas processing facility is therefore less than for the power plant flue gas and hence the capture cost is also less.

The effect of selectivity

The CO₂ selectivity of a membrane represents the ratio of the CO₂ permeability over the permeability of other component gases in the mixture. Figure 6 shows the changes that occur in the concentration of CO₂ in the permeate stream as a function of the CO₂ selectivity for both the power plant flue gas and natural gas streams. Figure 7 shows the effect on the capture cost for the two CO₂ feed gases with changes to the CO₂ selectivity from a value of 20 to 100. As shown in Table 2, the CO₂/N₂ selectivity of the PPO membrane used for the flue gas separation is 20 (1), and the CO₂/CH₄ selectivity of the cellulose acetate membrane considered for natural gas processing is 20 (7). For the power plant flue gas, the CO₂ recovery rate is fixed at 70%, while for the natural gas processing stream the level of CO₂ recovery is varied to achieve pipeline specifications.

From the results, for the flue gas stream, increasing the CO₂ selectivity from 20 to 50 results in an increase in permeate CO₂ concentration from approximately 50% to 65%. Further increases in selectivity from 100 and 200 can achieve CO₂ concentrations of 75% or 80% respectively, as shown in Figure 6. Meanwhile, the results in Figure 7 show that the CO₂ storage cost reduces with the improved selectivity due to the increased CO₂ purity in the permeate stream. It has been shown by Allinson et al (4) that it is more cost effective and requires less power to compress high CO₂ purity streams than streams with low CO₂ content. However, improving the CO₂ selectivity does not have significant impact on the capture cost of the flue gas stream. The capture cost actually increases slightly with improved CO₂ selectivity. This occurs because for this analysis, the amount of CO₂ removed and CO₂ permeability are fixed. According to Fick's law as shown in Equation 3, by increasing the CO₂ selectivity, the mole fraction of CO₂ in the permeate increases and the mole fraction of CO₂ retentate is reduced. Consequently the driving force across the membrane is also reduced and to obtain the same effective CO₂ recovery; that is the same number of moles of CO₂ removed, the membrane area increases. This increase in membrane area increases the total capital cost.

$$n_{CO_2} = \frac{\bar{Q}_{CO_2}}{t} \times A_{membrane} \left(P_{feed} x_{CO_2} - P_{permeate} y_{CO_2} \right) \quad (2)$$

where n_{CO_2} is the number of moles of CO₂ removed, $A_{membrane}$ is the membrane area (m²), Q is the CO₂ permeability, t is the thickness, P_{feed} and $P_{permeate}$ is the pressure in the feed and permeate streams respectively, and x_{CO_2} and y_{CO_2} is the mole fraction of CO₂ in the retentate and permeate respectively.

For the natural gas feed, improving the CO₂ selectivity from 20 to 60 results in an increase in

CO₂ permeate concentration from approximately 60% to 77% as shown in Figure 6. The methane gas in the permeate stream is considered as lost product. Therefore any improvements in CO₂ selectivity for a natural gas processing facility have the twin benefits of reducing the required membrane area resulting in less capital expenditure, as well as reducing the quantity of methane product lost. Figure 8 shows the changes in capture cost and quantity of methane product loss per annum at two CO₂ selectivity values. For a feed gas with a 10% mole fraction of CO₂, improving the CO₂ selectivity from 20 to 60 almost halves the quantity of methane product that is lost. Improving the selectivity results in better production rates at all CO₂ feed concentrations, from 10% to 50%. Improving the selectivity also decreases the capture cost even though the area of the membrane required increases. This is because the gains from the reduced methane loss compensates for the increase in capital cost. Improving the CO₂ selectivity for the natural gas feed will also reduce the storage cost due to the increase in CO₂ purity in the permeate stream (Figure 8). The storage cost for the natural gas facility is also less than the storage cost for the power plant flue gas because the facility is already located offshore. This has the benefit of reduced capital expenditure due to fewer infrastructure cost items such as pipelines.

Combined effects of permeability and selectivity

The results above indicate that improving the CO₂ permeability will reduce the capture cost, while improving the CO₂ selectivity will reduce the storage cost but may increase the corresponding capture cost for both CO₂ feed gases.

The cost analysis in Figure 9 examines the combined effect of improving both the CO₂ permeability and selectivity for CO₂/N₂ separation. This analysis considers doubling the permeability and selectivity of the PPO membrane listed in Table 2. By doubling the permeability from 75 Barrer to 150 Barrer, and doubling the CO₂/N₂ selectivity from 20 to 40, the storage cost reduces compared to the base case but the capture cost remains unchanged. The storage cost reduces due to the higher concentration of CO₂ in the permeate stream by utilising the higher CO₂ selectivity. However the capture cost remains unchanged because even though the improved permeability reduces the required membrane area, the improved selectivity increases the required membrane area. Thus there is no net change in the membrane area. Improving the permeability and selectivity in isolation only generates total CCS savings of 1-2%.

The effect of pressure ratio

The pressure ratio is the ratio of the permeate pressure to the feed pressure for the gas separation membrane stage. Keeping all other membrane characteristics constant, a high-pressure ratio will result in a lower membrane area for a fixed flux across the membrane. From the cost analysis shown in Figure 4, the biggest contributing factor in the both the capital and operating cost is the energy required to compress the inlet feed gas to a suitable pressure. The analysis in the previous sections show that improvements in the membrane properties can provide cost benefits, however if the feed pressure remains unchanged, improving the permeability and selectivity of the membranes considered in Table 2 only generate savings of less than 10%. By utilising advantages of improved selectivity and permeability in the relation to pressure ratio, the cost of mitigation may considerably be reduced. Figure 10 shows the capture, storage and total CCS cost for CO₂ recovery from power plant flue gas for two scenarios. The first considers CO₂ recovered at various recovery rates of 60 to 90% or CO₂ avoided rates of 35% to 55% using the PPO membrane characteristics listed in Table 2. The

second case examines a membrane system at the same CO₂ recovery rates but the permeability and selectivity values have been improved. The analysis has been set so that the mole fraction of CO₂ in the permeate stream is the similar to that as the first case, but the feed pressure has been reduced. This is achieved because at each CO₂ recovery rate, the amount of CO₂ is fixed and according to Fick's Law as shown in Equation 3, increasing the CO₂ permeability while maintaining similar membrane area results in a reduction in the driving force across the membrane. Due to the improved selectivity, the CO₂ mole fraction in the permeate stream increases and thus, the feed pressure can be reduced. The storage cost is the same for both scenarios because the concentration and flowrate of the CO₂ in the permeate stream is the same. However, the capture cost and hence total CCS cost for the membrane system with the improved selectivity, permeability and pressure ratio is much lower than the standard case. This is because a lower feed pressure requires a smaller compressor, reducing both the capital and operational costs. By being able to operate at a lower trans-membrane pressure when membrane selectivity and permeability are improved, the CO₂ capture and total CCS costs may be reduced up to 15%.

CONCLUSION

The analyses shows that considerable CO₂ removal rates can be achieved with gas membrane separation systems, and the economic competitiveness of the systems in comparison to amine based systems depends on both the membrane characteristics and characteristics of the feed gas. For low-pressure systems such as power plant flue gas, utilisation of reduced trans-membrane pressure through improvements to the selectivity and permeability can improve the capture, storage and total CCS cost. Improvements in permeability and selectivity for membrane systems recovering CO₂ from natural gas processing can result in substantial cost benefits through reduced capital expenditure and reduction in methane losses.

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KEY WORDS

Greenhouse mitigation, carbon capture, geosequestration, CCS economics, membrane systems, CO₂ capture

Table 1 Summary of the input parameters for CCS studies

Feed gas condition		500 MW Power Plant Flue gas (CO ₂ /N ₂)	Natural Gas Processing Feed gas (CO ₂ /CH ₄)
Flowrate	m ³ /s	500	11.5
Composition	% mol	CO ₂ 14	CO ₂ 10
		N ₂ 80	CH ₄ 90
		O ₂ 6	-
Temperature	°C	93	45
Pressure	bar	1	55

Table 2 Properties of the polymeric membranes considered in this study

Membrane properties	500 MW Power Plant Flue gas (CO ₂ /N ₂)	Natural Gas Processing Feed gas (CO ₂ /CH ₄)
Membrane material	Polyphenyleneoxide (PPO) (<u>1</u>)	Cellulose Acetate (<u>7</u>)
CO ₂ permeability (Barrer)	75	6
Selectivity	CO ₂ /N ₂ 20	CO ₂ /CH ₄ 20
	CO ₂ /O ₂ 4	-
Membrane temperature (°C)	30	30
Membrane feed pressure (bar)	20	55

Table 3 Summary of the economic parameters used for CCS studies.

Discount rate	7 % pa
Cost of external power	20 \$/MWh
Fixed annual operating cost	4% of total Capital Costs
Project life	20 years
Construction period	2 years
Membrane cost	80 \$/m ²
Production cost for Natural Gas Methane	1.5 \$/GJ

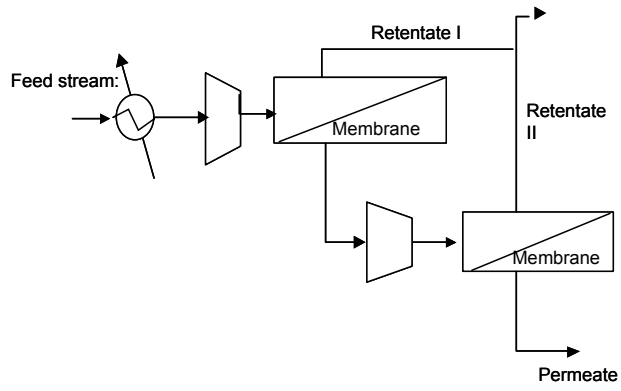


Figure 1 A two-stage cascade membrane system configuration (TCMS)

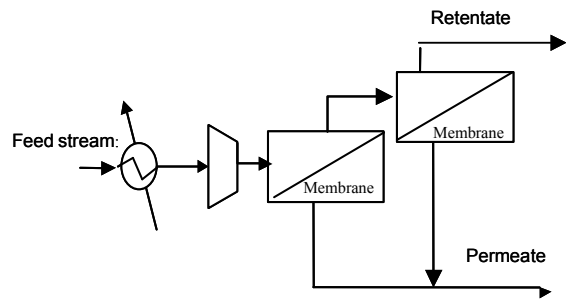


Figure 2 A two-stage series membrane system configuration (TSMS)

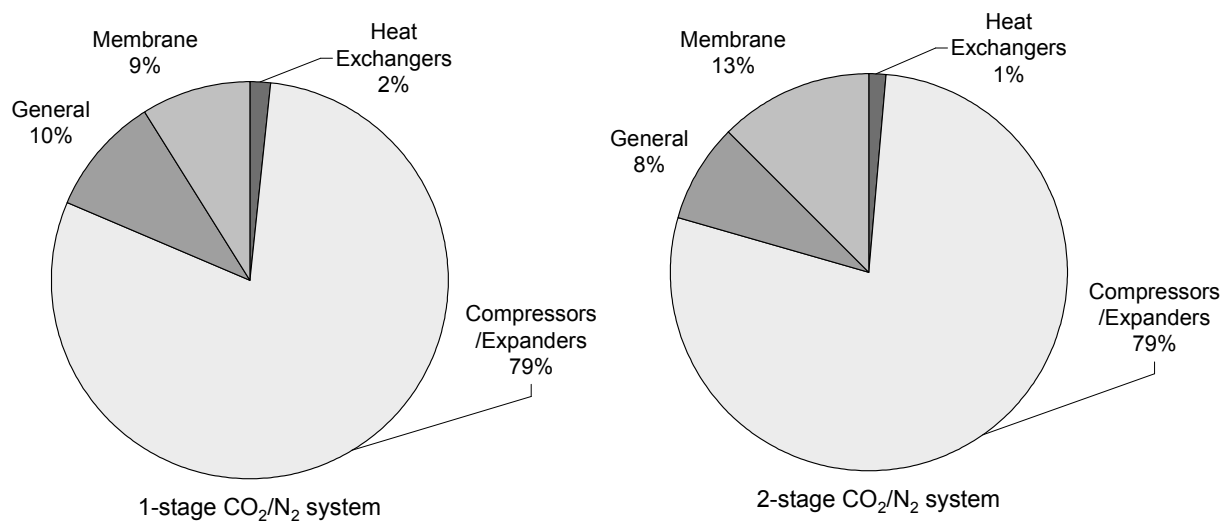


Figure 3 Relative capital expenditure cost breakdown for both a 1-stage and 2-stage CO₂/N₂ system

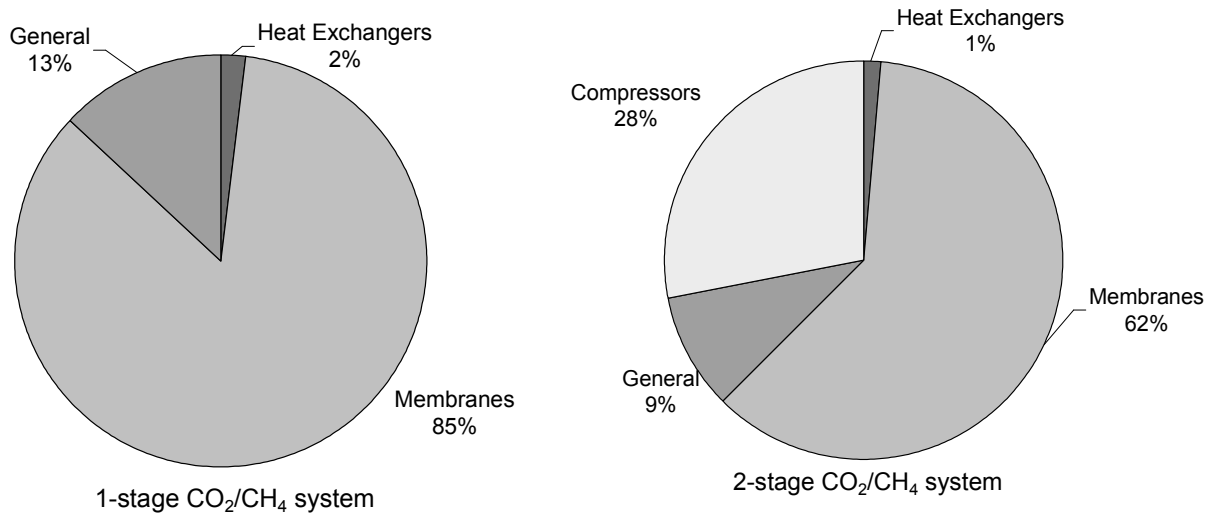


Figure 4 Relative capital expenditure cost breakdown for both a 1-stage and 2-stage CO₂/CH₄ system

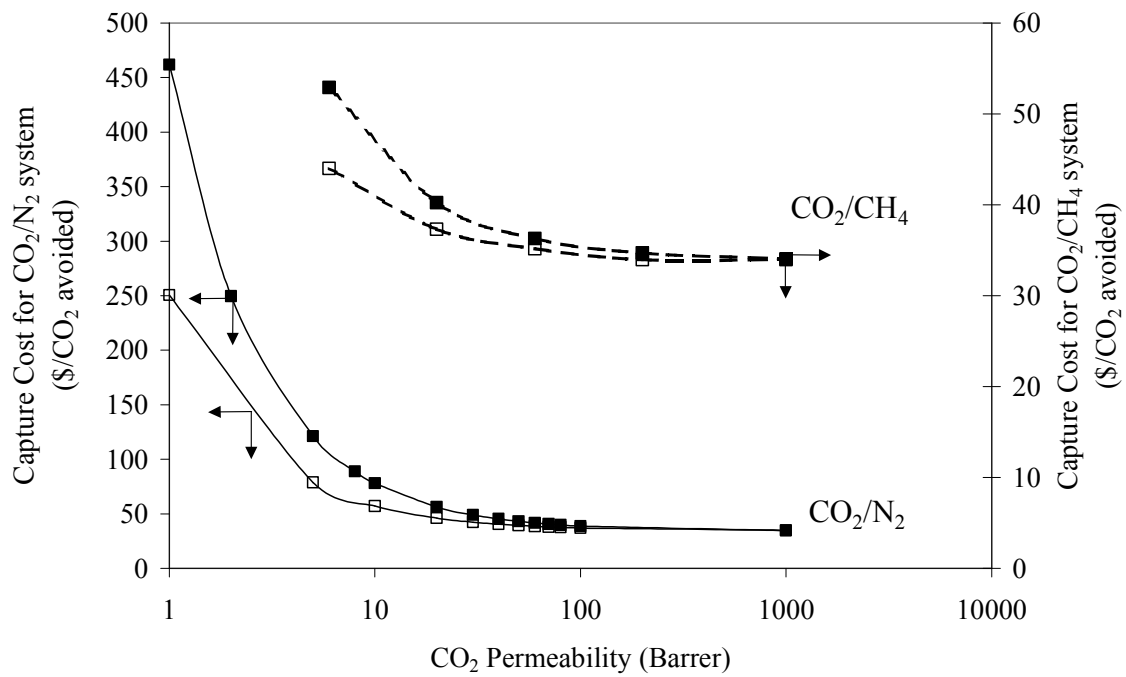


Figure 5 The effect on the capture cost for CO₂/N₂ and CO₂/CH₄ systems with changes to the CO₂ permeability at two thicknesses; ■ 0.1 μm and □ 0.05 μm

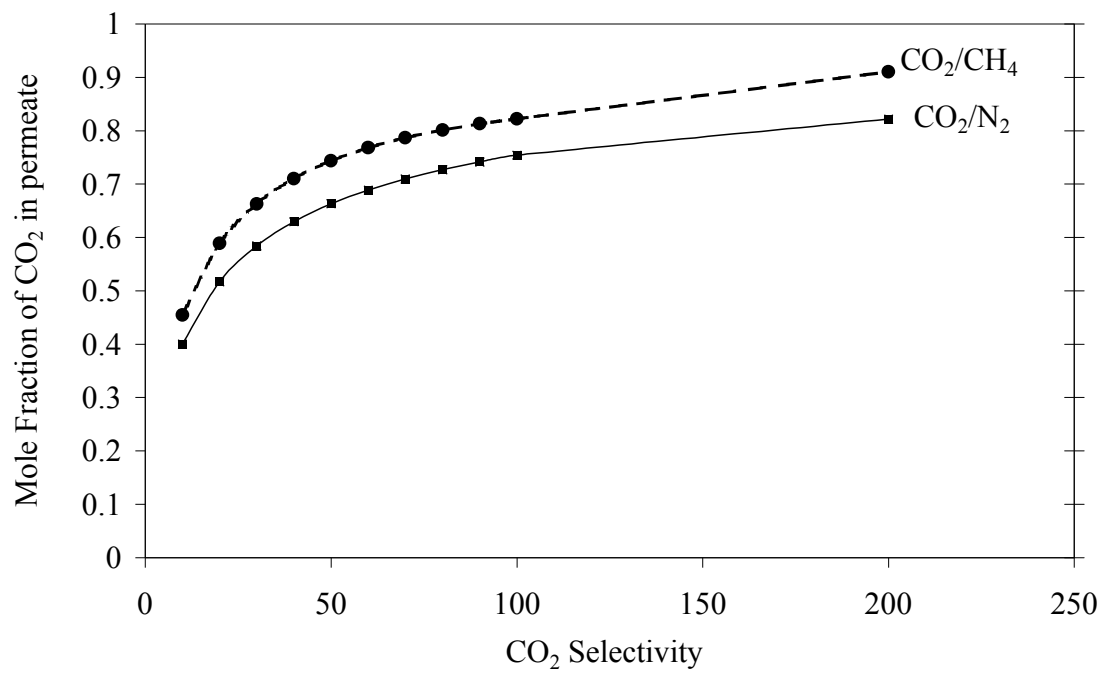


Figure 6 Effect of selectivity on CO₂ permeate purity

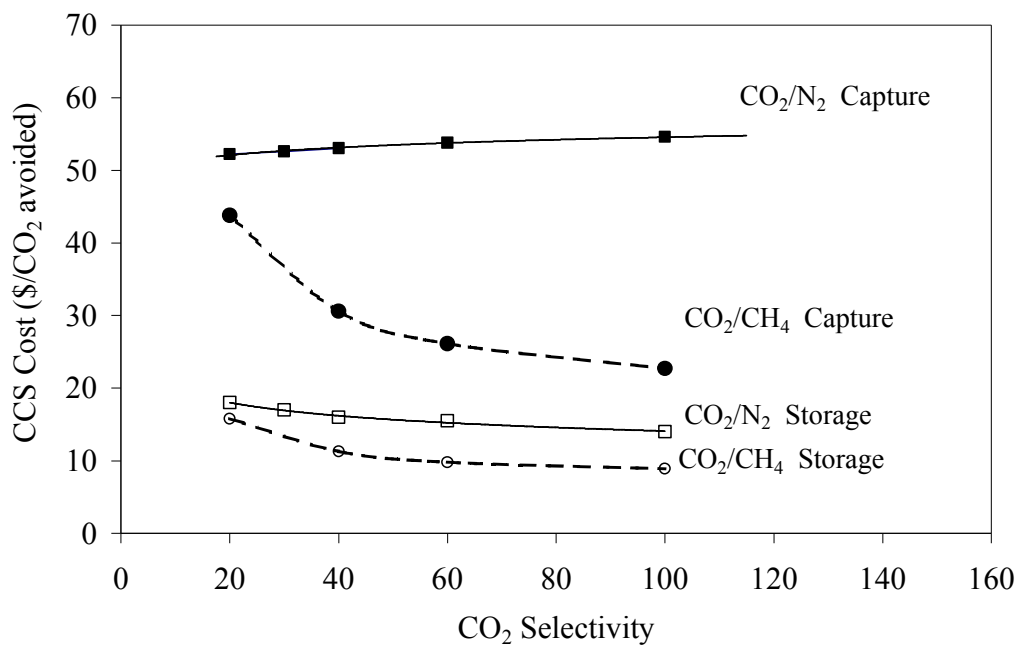


Figure 7 Selectivity effect on the capture and storage cost for CO₂/N₂ and CO₂/CH₄ systems

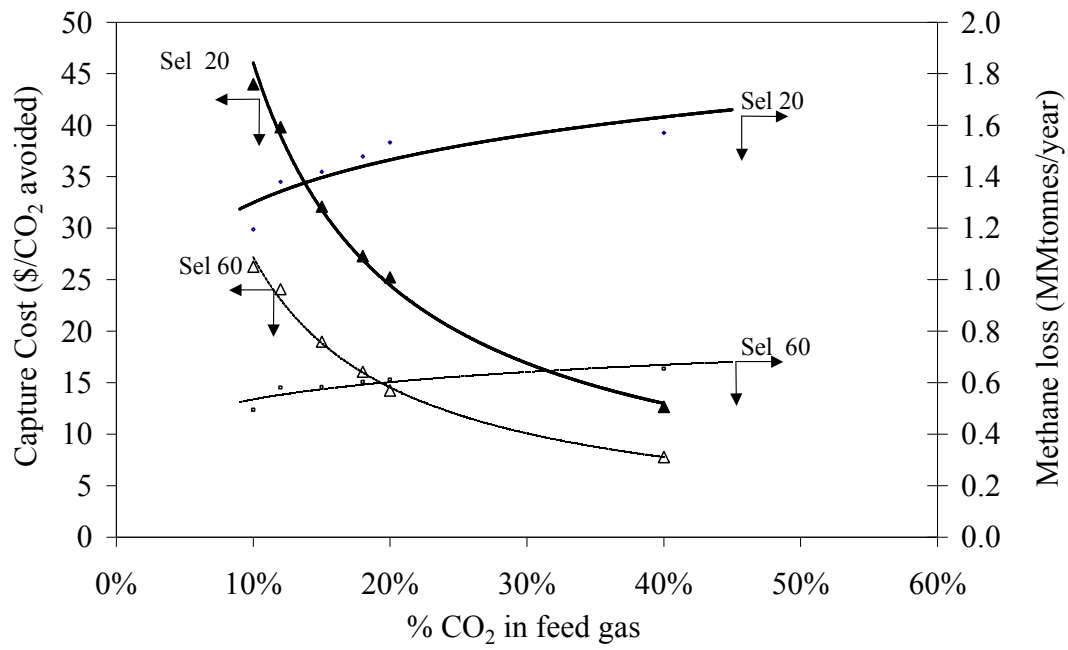


Figure 8 Selectivity effect on the capture cost and methane loss for the CO₂/CH₄ system

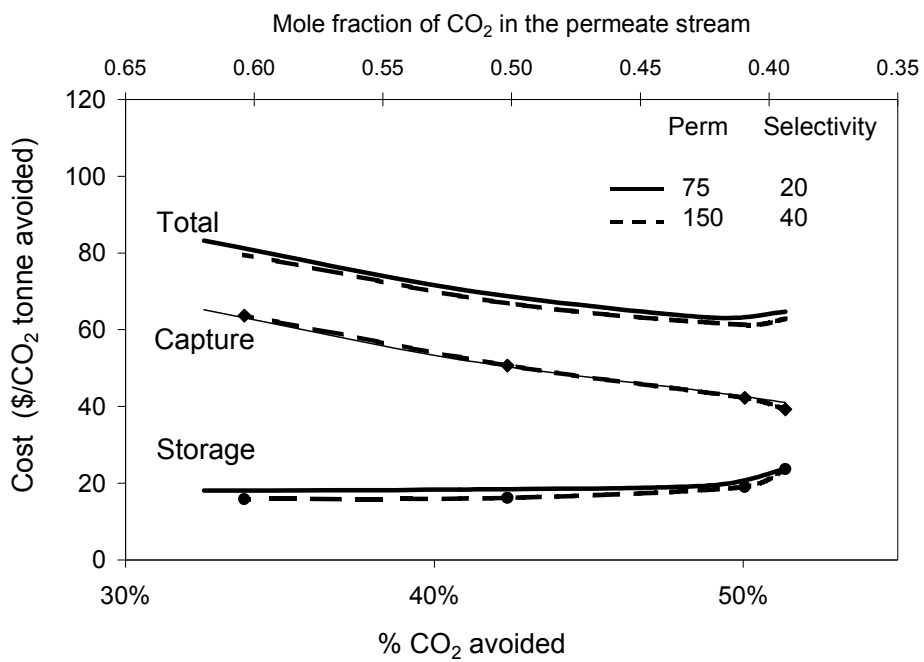


Figure 9 Combined effects of improved permeability and selectivity for CO₂/N₂ capture, storage and total CCS costs

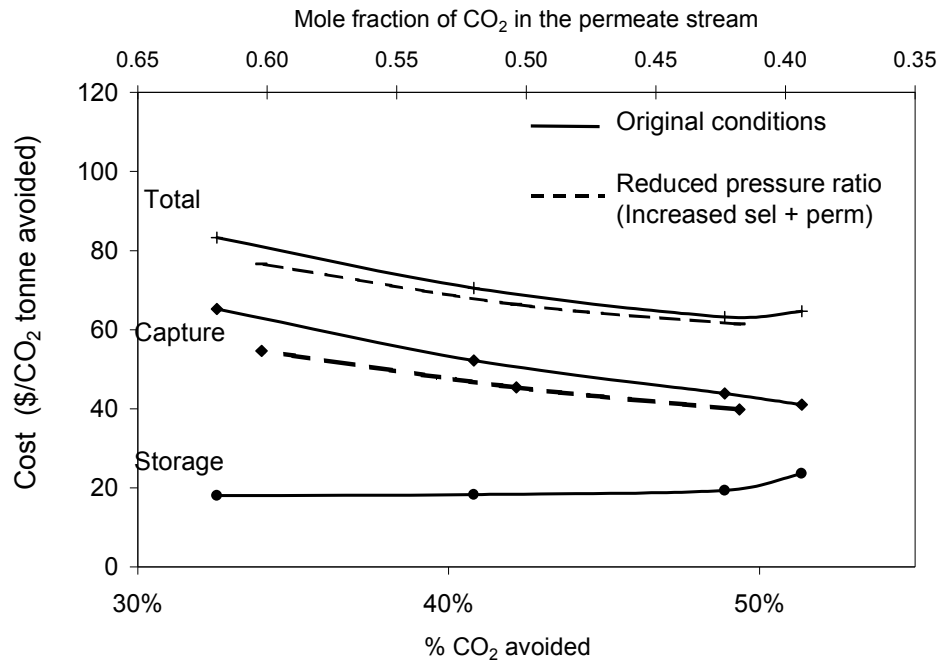


Figure 10 Effect of improved permeability, selectivity and reduced membrane pressure for CO₂/N₂ system