Technical and Economic Assessment of Membrane-based Systems for Capturing CO_2 from Coal-fired Power Plants

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Background

- Post-combustion carbon capture and storage (CCS) has been considered a key technology for deeply reducing carbon dioxide (CO_2) emissions from existing and new coal-fired power plants.
- Membranes have been used commercially for industrial gas separation, and have the potential for application to power plant flue gases.
- Membrane systems are among the advanced technologies being developed for more cost-effective CO_2 capture.

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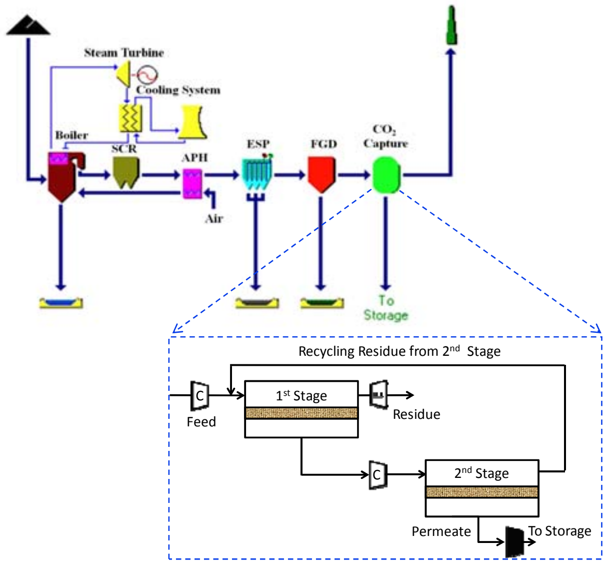
Postcombustion Membrane Capture Systems for Coal-fired Power Plants



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Research Objective

- The objective of this study is to evaluate the technical feasibility and cost of polymeric membrane systems for CO_2 capture at coal-fired power plants. More specifically,
  - To investigate the performance (e.g. membrane system size and energy penalty) and cost of different capture system configurations (e.g. single- and multiple-stage modules) to identify feasible membrane systems that are able to simultaneously achieve 90% CO_2 capture and above 95% purity of CO_2 product; and
  - To examine a range of key factors affecting the capture system performance and cost.

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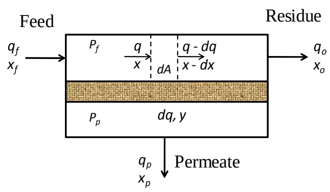
Membrane Gas Separation Basics

- The driving force for membrane gas separation is the partial pressure difference of a gas component between the feed and permeate sides. The flow flux of a gas component through membranes is:

$$J = \frac{P^*}{\delta}\left(xP_f - yP_p\right) = \overrightarrow{P^*}\left(xP_f - yP_p\right)$$

Where:

J = the volume flux of a component (cm³(S.T.P)/cm².s);
P^* = membrane permeability that measures the ability of the membrane to permeate gas (cm³(S.T.P).cm/(s.cm².cm Hg));
$\overrightarrow{P^*}$ = membrane permeance (cm³(S.T.P)/(s.cm².cm Hg))*;
δ = the membrane thickness (cm);
x = the mole fraction of CO_2 in the feed stream;
y = the mole fraction of CO_2 in the permeate stream;
P_f = the feed-side pressure (cm Hg);
P_p = the permeate-side pressure (cm Hg).



* Note:

Membrane permeance unit: 1 gas permeation unit (gpu)= 10^{-6} cm³(S.T.P)/(s.cm².cm Hg)

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Key Parameters for Membrane Gas Separation Process

- **Membrane selectivity**: the ratio of two gas permeabilities and the measure of the ability of a membrane to separate two gases, such as CO_2 versus N_2.

$$\alpha_{CO2/N2} = \frac{P^*_{CO2}}{P^*_{N2}}$$

- **Pressure ratio**: the pressure ratio for the feed versus permeate sides

$$\emptyset = \frac{P_f}{P_p}$$

- **Stage cut**: the flow fraction of the feed gas that permeates the membrane

$$\theta = \frac{q_f}{q_p}$$

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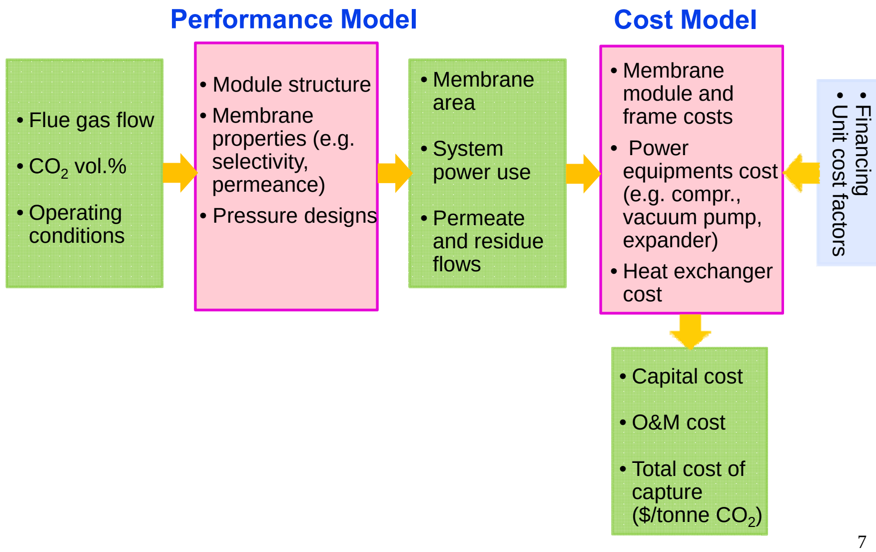
Technical and Economic Assessment Paradigm



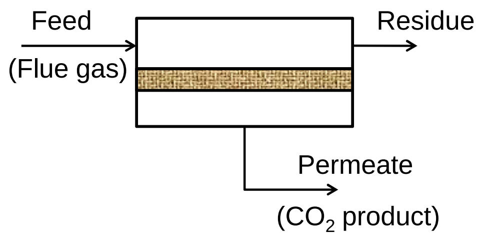
Technical and Economic Assumptions for Assessment*

Parameter	Value
Plant capacity factor	75%
Feed flue gas flow rate (S.T.P m^3/s)	500
Flue gas CO_2 concentration (volume)	13%
Flue gas pressure (bar)	1.0
Membrane CO_2/N_2 selectivity (S.T.P)	50
Membrane CO_2 permeance (S.T.P gpu)	1000
Compressor/ pump/expander efficiency	85%
Fixed charge factor	0.113
Electricity price($/kWh)	0.05
Compressor installed capital cost ($/hp)	500
Vacuum pump installed capital cost ($/hp)	1000
Expander unit capital cost ($/kW)	500
Membrane module capital price ($/$m^2$)	50

Major Inputs for Process Modeling (rows from feed flue gas flow rate to compressor/pump/expander efficiency)

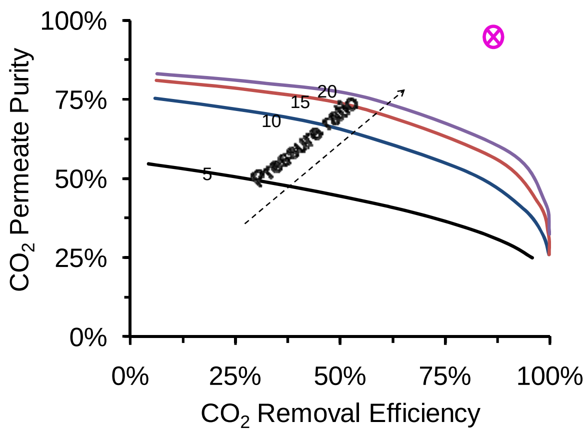
* Default values for case studies later, unless otherwise noted.

Single-Stage Membrane Systems

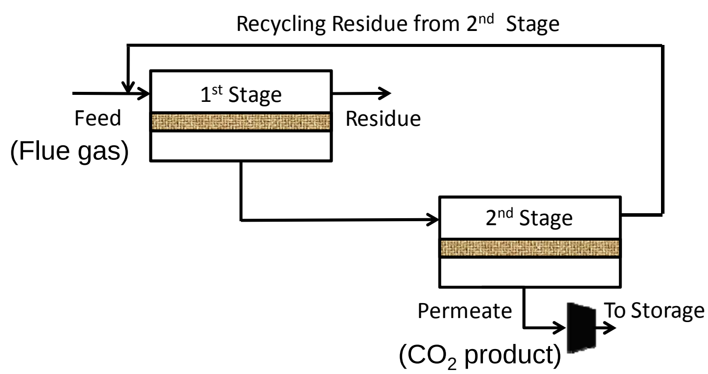


Single-Stage Membrane Systems Performance

- Given Inputs:
  - » Inlet CO_2 concentration = 13%
  - » CO_2/N_2 selectivity= 50

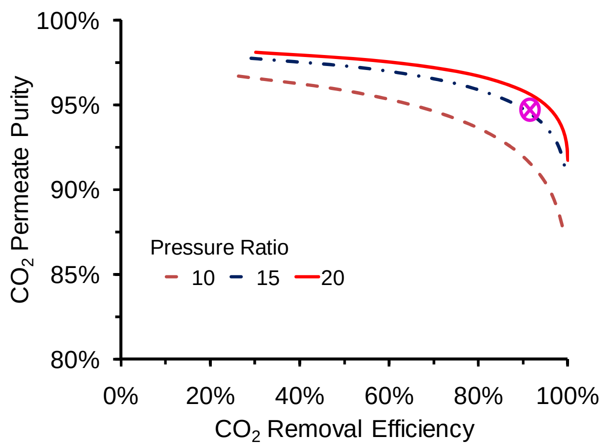


Two-Stage Ideal Cascade with Recycling



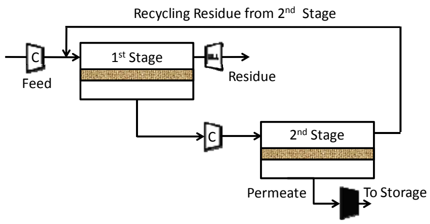
Two-Stage Membrane System: Pressure Ratio Requirement for Meeting Removal Targets

- Given Inputs:
  - » Inlet CO_2 concentration = 13%
  - » CO_2/N_2 selectivity= 50

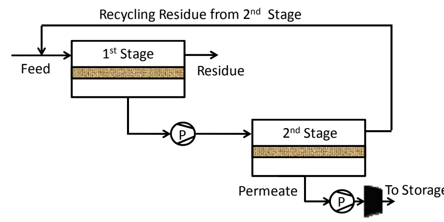


Two-stage Membrane System: Effects of Driving Force Design for Gas Separation

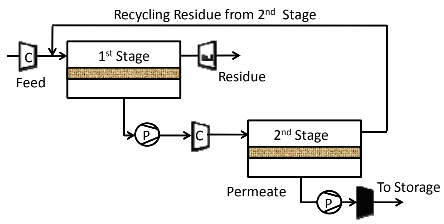
(a) Feed-side compression



(b) Permeate-side vacuum pumping



(c) Feed-side compression + permeate-side vacuum pumping



Note:

C = compressor
E = expander
P = vacuum pump

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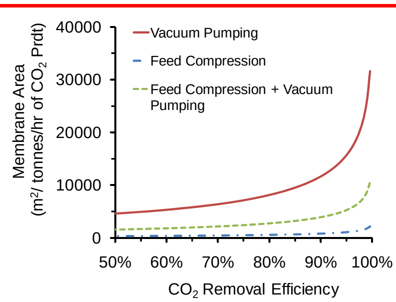
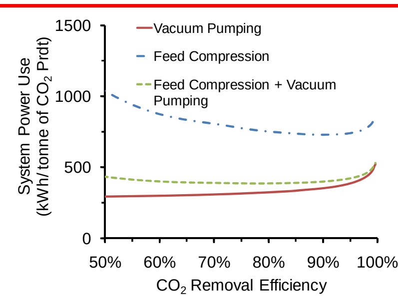
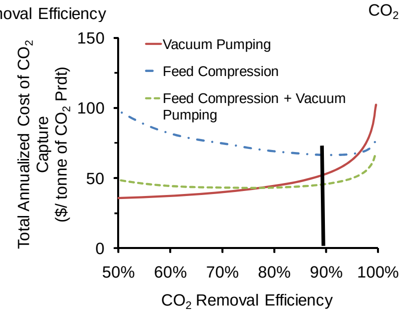
Two-stage Membrane System: Effects of Driving Force Designs (cont'd)







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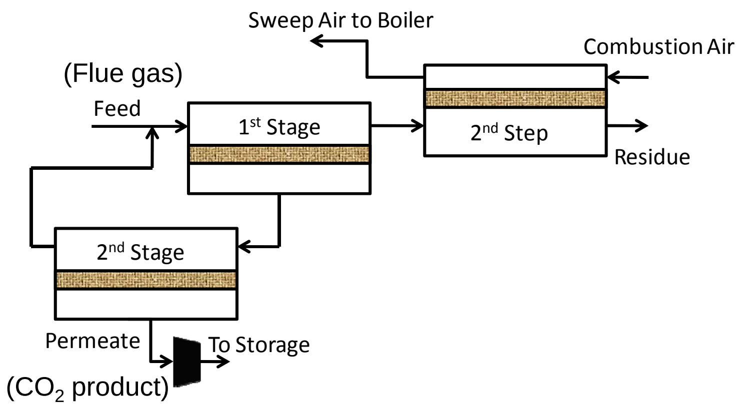
Two-Stage, Two-Step Membrane Systems with Combustion Air Sweep

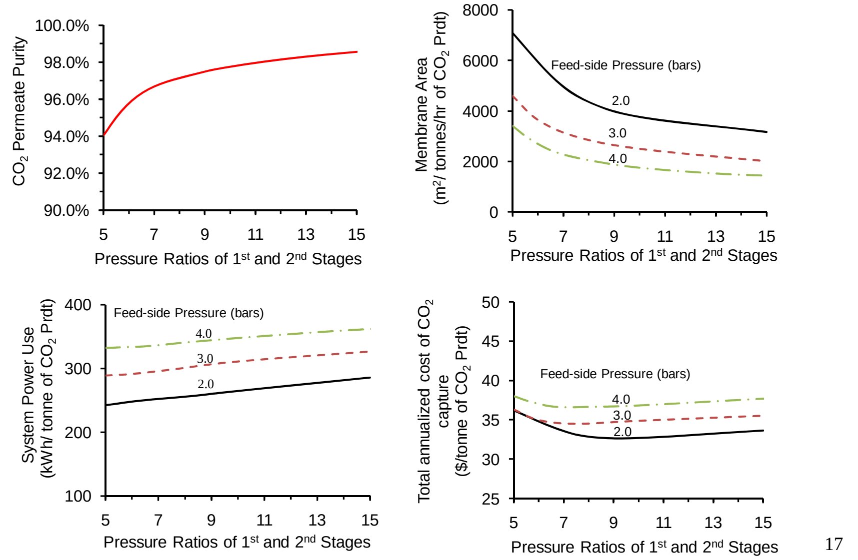
- Using boiler combustion air as a sweep gas may remarkably increase the feed CO_2 partial pressure and reduce the energy penalty of gas separation (Merkel *et al*., 2010).



Two-Stage, Two-Step Membrane Systems with Combustion Air Sweep: Effects of Factors

- Key factors considered:
  - Feed-side pressure design
  - Pressure ratio for feed- versus- permeate side
  - Membrane CO_2/N_2 selectivity and CO_2 permeability
  - Membrane facilities price

Two-Stage, Two-Step Membrane Systems with Air Sweep: Effects of Driving Force Designs @ 90% Removal Efficiency



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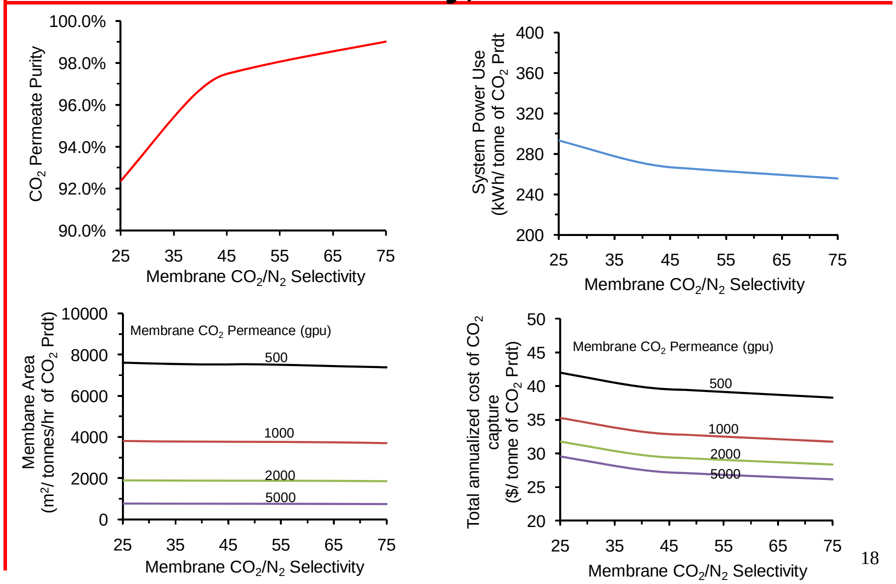
Two-Stage, Two-step Membrane Systems with Air Sweep: Effects of Membrane Properties @ 90% Removal Efficiency, Pressure Ratio 10



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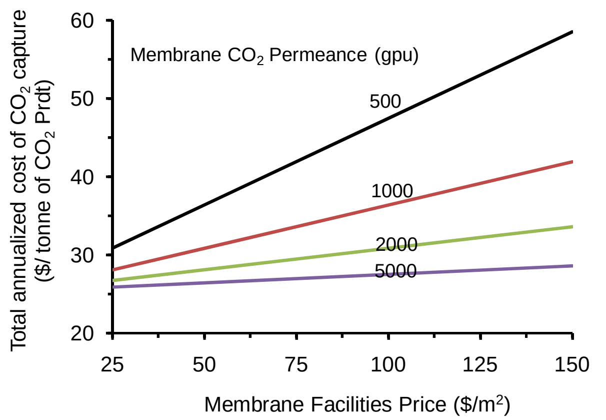
Two-Stage, Two-step Membrane Systems with Air Sweep: Effect of Membrane Facilities Price



Note:

Pressure ratio=10; Feed-side pressure = 2.0 bars; CO_2/N_2 selectivity=50; feed flue gas CO_2 concentration =13%; system CO_2 removal efficiency=90% and product purity =+95%.

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Comparisons between Multi-Stage Membrane Systems @ 90% CO_2 Capture

Variables	Two-stage system	Two-stage, two-step system with air sweep
Feed flue gas flow (m^3/s) (w/o CCS)	500	500
Flue gas CO_2 concentration (w/o CCS)	13%	13%
Membrane CO_2 permeance (gpu)	1000	1000
Membrane CO_2/N_2 selectivity	50	50
Feed-side pressure (bars)	3.0	2.0
Permeate-side pressure (bars)		
1st and 2nd stages	0.2	0.2
2nd step	-	1.0
CO_2 product purity	**95%**	**98%** ↑
Membrane area (m^2/tonnes/hr CO_2)	**3808**	**3766**
System power use (kWh/tonne CO_2)	**399**	**265** ↓
Total capture cost ($/tonne CO_2)	**45.6**	**32.7** ↓
Approximate cost of CO_2 avoided ($/t)	**100.0**	**62.3** ↓

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Conclusions

- Multiple-stage membrane systems are capable of meeting the CO_2 capture performance targets (90% capture and above 95% product purity) for given membranes.
- A hybrid design using both compressors and vacuum pumps for producing the separation driving force between feed and permeate sides is effective in reducing the energy requirements and cost of CO_2 capture.
- A combination of combustion air sweep with the two-stage, two-step membrane capture system would significantly reduce the energy use and cost of CO_2 capture.
- Future modeling efforts would take into account the effects of minor air pollutants of flue gas on the performance and cost of membrane capture system.
- We will continue to systematically investigate the effects of adding membrane capture systems on the power plant performance and cost of electricity.

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Disclaimer

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Thank You

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Appendix

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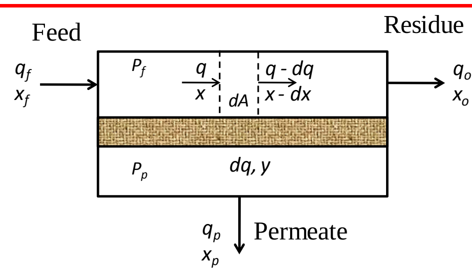
Major Assumptions of Polymeric Membrane Gas Separation Modeling

- Ideal gas behavior;
- Binary gas separation (CO_2 versus N_2);
- Constant permeability of each gas component that is independent of pressure and the same as the pure gas;
- Negligible pressure drop in both feed and permeate streams;
- Isothermal conditions.

Binary Gas Separation Models under Crossflow Pattern



The local permeation rate of a gas component in a binary (CO_2 and N_2) membrane system over a differential membrane area is described as (Geankoplis, 1993):

$$-ydq = J_{CO2}dA = \frac{P^*_{CO2}}{\delta}\left[xP_f - yP_p\right]dA \tag{1}$$

$$-(1-y)dq = J_{N2}dA = \frac{P^*_{N2}}{\delta}\left[(1-x)P_f - (1-y)P_p\right]dA \tag{2}$$

Binary Gas Separation Models under Crossflow Pattern (cont'd)

And then,

$$\frac{y}{1-y} = \frac{\alpha(1-y/\phi)}{(1-x)-(1-y)/\phi} \quad (3)$$

Weller and Steiner applied ingenious transformations to obtain analytical solutions for the governing equations (Geankoplis, 1993).

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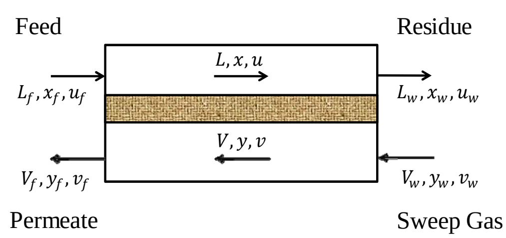
Binary Gas Separation Models under Countercurrent Flow Pattern

Pan and Habgood (1974) developed a unified theoretical framework for binary membrane separation systems with and without gas sweep. Major assumptions made are: (a) two permeable components plus the possible condition with a nonpermeable fraction in the feed side and a gas sweep stream in the permeate side; (b) the constant permeability; and (c) negligible pressure drop in both the feed and permeate sides.



$R^w = R_f^w$ $\qquad$ $R^w = 0$

Binary Gas Separation Models under Countercurrent Flow Pattern (cont'd)

For membrane separation systems with and without gas sweep, the mathematical models derived starting from mass balances are expressed as:

$$\frac{L}{L_w} = \frac{y - x_w + F_w(y - y_w)}{y - x} \qquad (1)$$

$$\frac{V}{L_w} = \frac{x_w - x + F_w(y_w - x)}{y - x} \qquad (2)$$

$$u = \frac{u_w(y - x)}{y - x_w + F_w(y - y_w)} \qquad (3)$$

$$v = \frac{v_w F_w(y - x)}{x_w - x + F_w(y_w - x)} \qquad (4)$$

$$\frac{dy}{dx} = \frac{y - x_w + F_w(y - y_w)}{x - x_w + F_w(x - y_w)} \times \left\{\frac{\alpha(1-y)(x-\gamma y) - y[(1-x-u) - \gamma(1-y-v)]}{\alpha(1-x)(x-\gamma y) - x[(1-x-u) - \gamma(1-y-v)]}\right\} \qquad (5)$$

Binary Gas Separation Model under Countercurrent Flow Pattern (cont'd)

The dimensionless membrane area is estimated as:

$$\frac{dR^w}{dx} = \frac{y - x_w + F_w(y - y_w)}{(x-y)\{\alpha(1-x)(x-\gamma y) - x[(1-x-u) - \gamma(1-y-v)]\}} \qquad (6)$$

The governing equations above are solved using 4th order Runge-Kutta approach.

When there is no sweep gas used in the permeate side, the permeate concentration at the residue end is determined as (Pan and Habgood, 1974):

$$\frac{y_w}{1 - y_w} = \frac{\alpha(x_w - \gamma y_w)}{1 - x_w - u_w - \gamma(1 - y_w)} \qquad (7)$$

Energy Use Estimation for Major Equipments

The energy use for the compressor and expander is estimated respectively as (Vallieres *et al*, 2003; Bounaceur *et al*, 2006; Favre, 2007; Yang *et al*, 2009):

$$E_{cp} = \frac{1}{\eta_{cp}} Q_{cp} \frac{\gamma RT}{\gamma - 1}\left[\left(\frac{P_h}{P_l}\right)^{\frac{\gamma-1}{\gamma}} - 1\right]$$

$$E_{ex} = \frac{1}{\eta_{ex}} Q_{ex} \frac{\gamma RT}{\gamma - 1}\left[1 - \left(\frac{P_l}{P_h}\right)^{\frac{\gamma-1}{\gamma}}\right]$$

When vacuum pumps are used in the permeate side, the energy use is estimated as (Vallieres *et al*, 2003; Bounaceur *et al*, 2006; Favre, 2007; Yang *et al*, 2009):

$$E_{vp} = \frac{1}{\eta_{vp}} Q_{vp} \frac{\gamma RT}{\gamma - 1}\left[\left(\frac{P_h}{P_l}\right)^{\frac{\gamma-1}{\gamma}} - 1\right]$$

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Cost Measure for Membrane Capture Systems

- Capital cost (CC):

Process Area Costs	**Plant Costs**
Membrane module	Process facilities capital (PFC)
Membrane frame	General facilities capital (10% of PFC)
Compressor	Eng. & home office fees (10% of PFC)
Expander	Project contingency cost (15% of PFC)
Vacuum pump	Process contingency cost (2% of PFC)
Heat exchanger	Other indirect cost (e.g. ower's cost)
CO_2 product compression	
Process facilities capital cost	**Total capital requirement (TCR)**

- Fixed O&M cost (FOM): estimated empirically as a percent of capital cost (Van der Sluus *et al*, 1992).
- Variable O&M cost (VOM): power use for major equipments.

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Total Levelized Cost (TLC) of CO_2 Capture

- Total levelized cost of CO_2 capture is defined as:

$$TLC = \frac{FCF \cdot TCR + FOM + VOM}{(m_{CO2} \cdot 365 \cdot 24) \cdot \mathrm{CF}}$$

Where:

CF	= capacity factor (%)
FCF	= fixed charge factor (fraction)
FOM	= fixed O&M cost ($/yr)
VOM	= variable O&M cost ($/yr)
m_{CO2}	= CO_2 capture product (tonne/hr)
TCR	= total capital requirement ($)
TLC	= total levelized cost of CO_2 capture ($/tonne)

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Process Facilities Capital (PFC) Estimates*

Cost Item	Method	Cost Item	Method
Membrane module	$CC_{mm} = A_m \cdot c_m$	Vacuum pump	$CC_{vp} = e_{vp} \cdot c_{vp}$
Membrane frame	$CC_{mf} = \left(\frac{A_m}{2000}\right)^{0.7} \cdot c_{mf}$	Heat exchanger	$CC_{exch} = \left(\frac{q_f}{400}\right) \cdot c_{exch}^{ref}$
Compressor	$CC_{cpr} = e_{cpr} \cdot c_{cpr}$	CO_2 prdt compr.	$CC_{mcpr} = e_{mcpr} \cdot c_{mcpr}$
Expander	$CC_{exp} = e_{exp} \cdot k_{exp} \cdot F_h$	PFC	Sum of all above

Note:

A_m: membrane area (m^2); c_{cpr}: installed unit cost ($500/hp); c_{exch}^{ref}: referred heat exchanger cost (M$3.5); c_m: membrane module price per unit ($/$m^2$); c_{mf}: referred frame cost (M$ 0.238); c_{mcpr}: compression unit cost ($902/kW); c_{vp}: installed unit cost ($1000/hp); e_{cpr}: compressor power use (hp); e_{exp}: expander power use (kW); e_{mcpr}: CO_2 compression power use per unit (93.0 kWh/tonne CO_2); e_{vp}: vacuum pump power use (hp); F_h: equipment cost factor for housing, installation, etc (1.8); k_{exp}: unit cost ($500/kW); q_f: feed gas flow (m^3/s).

* The process facilities capital cost for each of items is estimated referring to Van der Sluus *et al*(1992).

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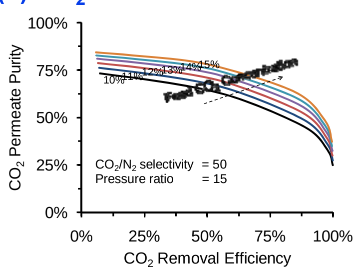
Operating and Maintenance Cost Estimates*

Cost Item	Method	Cost Item	Method
Membrane module	$COM_{mm} = 0.01 \cdot CC_{mm}$	Vacuum pump	$COM_{vp} = 0.036 \cdot CC_{vp}$
Membrane frame	$COM_{mf} = 0.01 \cdot CC_{mf}$	Heat exchanger	$COM_{exch} = 0.036 \cdot CC_{exch}$
Compressor	$COM_{cpr} = 0.036 \cdot CC_{cpr}$	CO_2 prdt compr.	$COM_{mcpr} = 0.036 \cdot CC_{mcpr}$
Expander	$COM_{exp} = 0.036 \cdot CC_{exp}$	Total O&M	Sum of all above

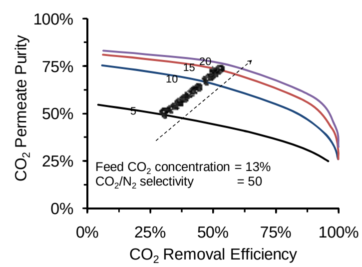
* The O&M cost for each of items is estimated empirically as a percent of capital cost (Van der Sluus *et al*,1992).

Single-Stage Membrane Systems: Effects of Key Factors on Performance (cont'd)

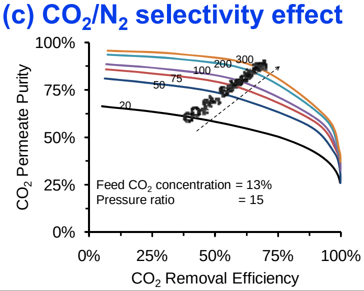
(a) CO_2 concentration effect



(b) Pressure ratio effect



(c) CO_2/N_2 selectivity effect



(d) Gas flow pattern effect