Process Design Analyses of CO_2 Capture from Natural Gas by Polymer Membrane

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Summary: Membrane-based natural gas separation has become one of the promising technologies due to its compactness, energy efficiency, environment friendliness and economic advantages. In this work, a three stage membrane process for the separation of CO_2/CH_4 is proposed based on a novel fixed site carrier membrane which has the potential to meet the CO_2/CH_4 separation and durability requirement. A simulation analysis, which utilizes the Aspen Hysys capabilities to calculate and couple energy balances in the process model, has been conducted to investigate the effect of process parameters on the gas processing cost. Two different natural gas mixtures containing 9.5% and 2.9% CO_2 have been simulated for various process conditions. This fixed site carrier membrane performs well when wetted with water. Therefore, natural gas feed streams are saturated with water. It is evident from the analysis that it is possible to maintain 2% CO_2 in retentate and methane loss in permeate below 2% by optimizing the process conditions. The analysis shows that fixed site carrier membrane offers a viable solution for natural gas sweetening.

Keywords: Polymer membrane, CO_2 capture, Natural gas sweetening, Techno-economic analysis.

Introduction

The severe implications of carbon dioxide emissions on public health and environment has ranked the control of CO_2 emissions as one of the most challenging issue being faced by the world [1]. Majority of these emissions come from the burning of fossil fuels and therefore CO_2 emissions to the atmosphere will continue to rise unless necessary measures are taken which reduce the net emissions by improving the energy efficiency, using clean energy alternatives and capturing the post-combustion CO_2 emissions. The demand for natural gas is constantly rising as it is cleaner and more efficient among the fossil fuels. Moreover, natural gas (methane) is a principle feedstock for the chemical industry. The global consumption of natural gas is projected to increase from 95 trillion cubic feet in 2003 to 182 trillion cubic feet in 2030 [2]. Raw natural gas varies substantially in composition depending on the source reservoir. Methane is always the major component, typically 75% - 90% of the total, but natural gas also contains significant amounts of ethane, some propane and butane, and 1%-3% of other higher hydrocarbons. In addition, natural gas contains undesirable impurities, such as water, carbon dioxide, nitrogen, and hydrogen sulfide which need to be removed before being stored as compressed natural or delivered to the pipelines (specifications for CO_2 are set to a maximum of 2 mol% [5]). The removal of CO_2 and other impurities from natural gas will increase the fuel calorific value, decrease the natural gas volume to be transported, reduce pipeline corrosion and minimize the atmospheric pollution [3].

From technical point of view, CO_2 removal from natural gas is a CO_2/CH_4 separation problem and conventionally amine absorption process is employed to remove carbon dioxide from natural gas [1-3]. Apart from number of technical challenges, amine absorption is still considered as the best available technology for CO_2 capture. Membrane based CO_2 capture processes are considered with lot of incertitude due to their sensitivity towards extreme process conditions; mainly temperature, pressure, high flow rates. However, despite these limitations, it is important to assert that membrane process yet offer a great potential for the CO_2 capture for their low capital investment, small footprints, low energy requirement, environment friendliness, and finally it is easy to incorporate modular up-scaling in the existing membrane process. An efficient design of membrane separation system depends on the implementation of precise mathematical models which enable the accurate simulation of process variables and predict the required membrane area, energy and gas processing cost [4].

Various studies have been reported so far on the optimization of membrane based gas separation processes by employing different polymeric membranes. Some suggested the use of multi-stage membrane configurations for the removal of CO_2 from methane [6], while others favored the use of a

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hybrid process comprised of a single stage membrane configuration coupled with amine absorption unit for further natural gas purification [7, 8]. Generally, in a two stage membrane configuration, both membrane modules contain the same membrane. However, employing different membrane in each stage has also been investigated to obtain high purity product by separating different gas components at different membrane stages [9-13]. Furthermore, the process design, optimization and economical aspects of natural gas separation by membrane process has been investigated by employing two different membranes (CO_2 selective and H_2S selective) in a single and two stage membrane configuration by assuming a tertiary gas mixture containing CH_4, CO_2, H_2S [5]. It was concluded in this study that a single stage configuration containing H_2S selective membrane is most economical for processing natural gas which contains small amounts of CO_2. Similarly, a single stage configuration containing CO_2 selective membrane is most economical if natural gas contains small amounts of H_2S. But if the natural gas contains considerable amount of CO_2 (~40 mol %) and H_2S (~10 mol %), then a two stage configuration, H_2S selective membrane in the first stage and CO_2 selective membrane in the second stage would be the economical way. Variety of polymeric membranes, namely polyimides [14], polyetherurethaneurea [15], polyphenylene oxide [16], cellulose acetate and polyamide ether [17], have been investigated for CO_2/CH_4 separations. The effect of permeate recycling on gas processing cost, membrane area and required energy was investigated and was concluded that the advantage of permeate recycling is relative and depends on feed composition and operating conditions [15, 16], while the influence of process parameters (pressure, temperature, feed composition) on gas processing cost and methane loss were evaluated in [14].

This work can be considered as a contribution in strengthening the belief in CO_2/CH_4 membrane separation system capable to meet natural gas pipeline specifications (2 mol %) at significant lower methane loss compared to other studies [14-16]. A novel Fixed-Site-Carrier (FSC) membrane, a PVAm/PVA blend membrane, has been used in this study. In this blend membrane, PVAm has a high density of fixed amino groups which act as CO_2 transport carriers and PVA, a mechanically robust polymer, enhances polymeric network with good membrane formation properties [18]. The reversible reaction of CO_2 with amino carriers in PVAm facilitates the CO_2 transport and enhances the CO_2 permeability and CO_2/CH_4 selectivity [19]. This ultrathin PVAm/PVA blend layer (0.3 μm) supported on a polysulfone membrane has been evaluated and the experimental results revealed its suitability for CO2/CH4 separations. The selectivity up to 45 and the permeance up to 0.3 m3 (STP)/m2.bar.h at 2 bar has been recorded but the experiments at higher pressures are ongoing to validate its reliability for natural gas sweetening. However, in this analysis a selectivity of 35 and permeance 0.3 m3 (STP)/m2.bar.h has been used. The technical viability of membrane based CO2 capture process have been investigated by using a three stage membrane process to meet natural gas pipeline requirement (2% CO_2) and confine the methane loss in permeate below 2%. Two different gas compositions containing 9.5% and 2.9% CO_2 has been considered in this analysis. The influence of CO_2 composition in feed (natural gas) and feed pressure on methane loss, membrane area, energy required and gas processing cost has been investigated.

Results and Discussion

This simulation analysis is started with the influence of permeate side pressure on the CO_2 purity and recovery for fixed membrane area. It can be seen in Fig. 1 that by increasing the permeate side pressure, both CO_2 purity and recovery decreases, however, later decreases significantly. This Fig reveals that for the specified area, CO_2 recovery and purity can be attained more than 80% only if the pressure on the permeate side is kept below 10 bar. Fig. 1 shows that by increasing permeate side pressure both specific required energy and Gas Processing Cost (GPC) decrease. In fact cost of required energy has considerable impact on the GPC. Hence, less energy consumption results in less Gas Processing Cost. Fig. 2 (which is a counterpart of Fig. 1) shows that demand for high CO_2 purity and recovery results in higher energy consumption (due to lower permeate side pressure) which yields the higher gas processing cost in turn.

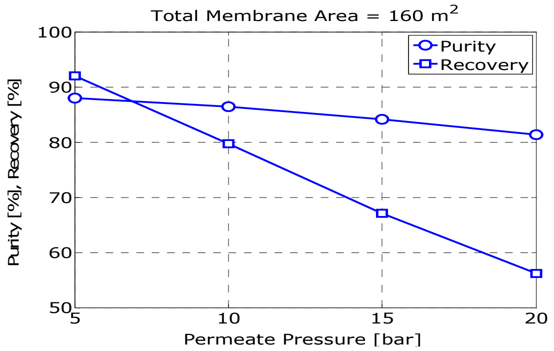

Fig. 1: CO_2 purity and recovery vs. permeate pressure (in first two stages) case A.

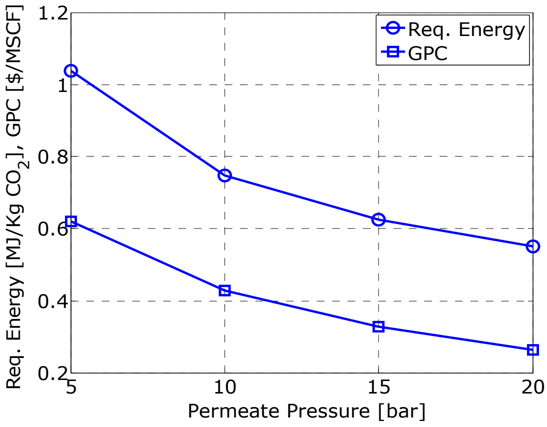

Fig. 2: Permeate side pressure vs. required energy and GPC (Gas Processing Cost).

If it is desired to achieve 90% CO_2 purity and recovery then as evident from Fig 1, it is only possible by either decreasing the pressure on permeate side or increasing the membrane area. Fig. 3 and 4 show the required membrane area, required energy and GPC plotted against pressure on the permeate side (permeate pressure only in first two stages as the pressure on stage 3 is kept 1 bar constant for all cases) respectively. It can be concluded from Fig. 5 and 6 that to achieve 90% purity and recovery, more membrane area and energy is required as the pressure on permeate side increases. The gas processing cost (GPC) is strongly influenced by the required membrane area and required energy. Therefore, a cumulative influence of membrane area and energy is obvious in terms of rise in gas processing cost (Fig 4).

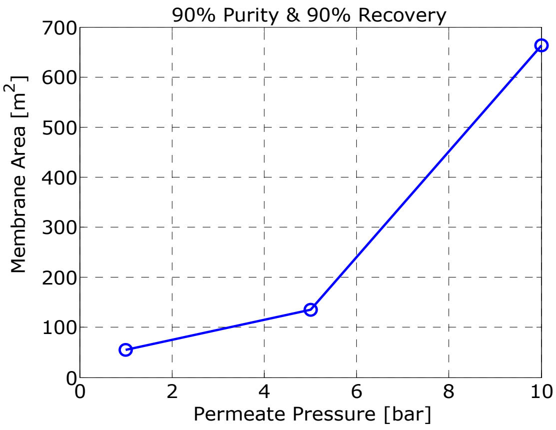

Fig. 3: Permeate side pressure vs. required membrane area to attain 90% purity & recovery.

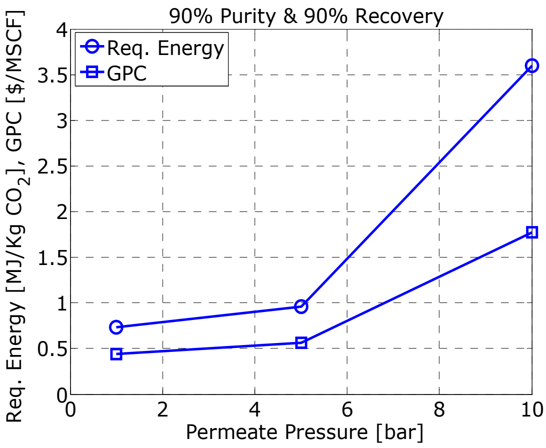

Fig. 4: Permeate side pressure vs. required energy and GPC to attain 90% purity & recovery.

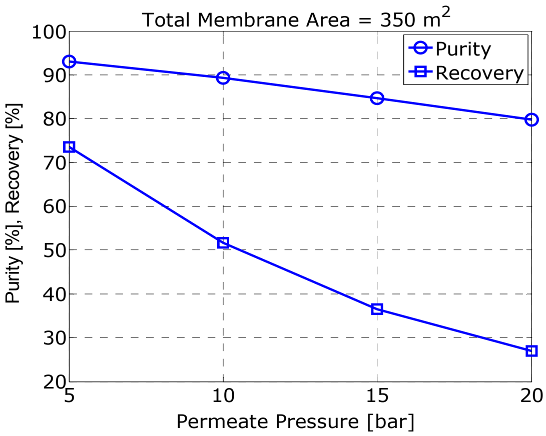

Fig. 5: CO_2 purity and recovery vs. permeate pressure (in first two stages) case B.

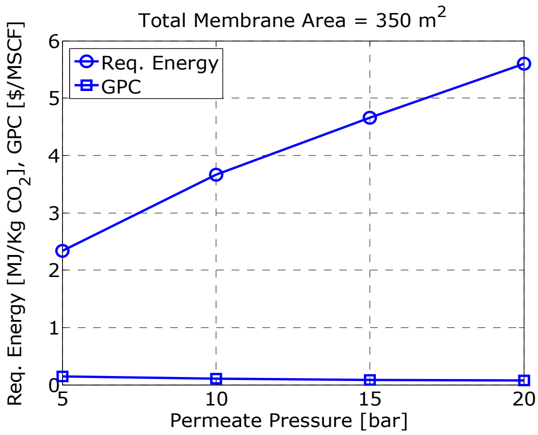

Fig. 6: Permeate side pressure vs. required energy and GPC for case B.

Similar analysis (like case A) has been carried out for case B by keeping the total membrane area constant (350 m^2). Fig. 5 shows that for this area, it is not possible to achieve 80% recovery even if the permeate side pressure is 5 bar (unlike Fig. 1). It is evident from Fig 6 that more specific energy is required for Case B. In contrary to Case A, natural gas in Case B has less CO_2 and higher feed flow rate (7 times higher than case A), which ultimately requires more membrane area. Though required specific energy has a strong influence on the gas processing cost but in Fig. 6, this influence is significantly dampened due to higher feed flow rates and lower CO_2 recovery, however as expected gas processing cost decreases as the permeate side pressure increases. For case B, it has been further investigated that how membrane area, specific required energy and gas processing cost would change if CO2 purity in permeate is kept constant (90%) and CO2 recovery is varied from 50% to 75%. As expected, Fig. 7 shows that more membrane area is required if CO2 recovery is set 75%. Similarly, Fig. 8 shows that for 75% recovery, less specific energy is required. Fig. 9 shows the comparison of GPC for 50% and 75% recovery. As mentioned earlier, required membrane area and specific energy have strong influence on the gas processing cost. This combined influence of required membrane area and specific energy on gas processing cost (GPC) is replicated in Fig. 9, showing that gas processing cost is bit higher if higher CO2 recovery is desired. Fig. 4 shows that for case A, GPC is about 0.5 $/MSCF when permeate side pressure is set around 5 bar and methane loss in permeate is about 0.5%. From Fig. 9 for case B, it can be seen that for permeate side pressure around 5 bar, GPC is about 0.14 $/MSCF and methane loss is about 0.25% for 75% recovery. These values are significantly lower than the results presented in [15]. It is worth mentioning here that feed flow rate taken in this analysis (Table-1) is far less than the flow rates taken in [15].

Table-1: Gas compositions and process conditions for Case A and Case B.

Case A	Case B
Feed Flow rate, MMSCFD = 0.35	**Feed Flow rate, MMSCFD = 2.48**
Feed pressure, bar = 90	**Feed pressure, bar = 115**
Feed temperature, °C = 60	**Feed temperature, °C = 8**
Feed composition, mol%	**Feed composition, mol%**
$CO_2 = 9.5$	$CO_2 = 2.9$
$CH_4 = 72.4$	$CH_4 = 97.1$
C_1-$C_6 = 18.1$	H_2O = Saturated
H_2O = Saturated	

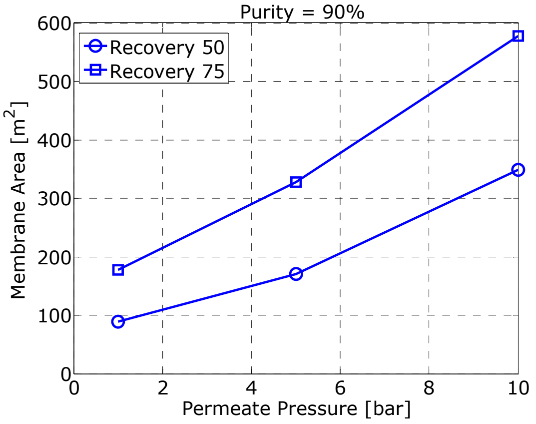

Fig. 7: Permeate side pressure vs. required membrane area for 50% & 75% recovery.

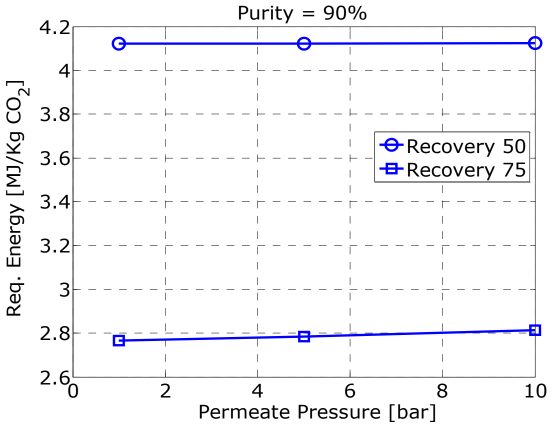

Fig. 8: Permeate side pressure vs. required energy for 50% and 75% recovery.

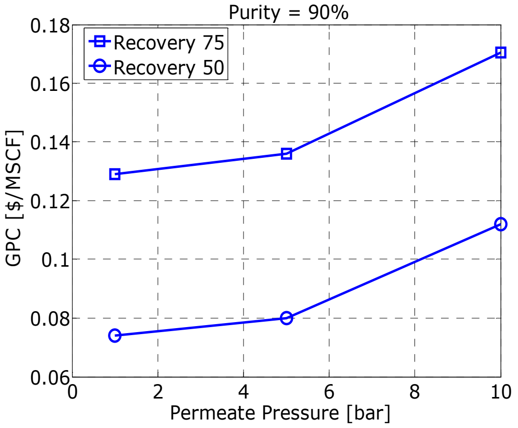

Fig. 9: Permeate side pressure vs. GPC for 50% and 75% recovery.

Mathematical Modeling of Gas Separation by Membranes

The principle of membrane gas separation depends on the membrane material, process conditions and the gas components in the mixture. The Solution-Diffusion Model is widely accepted transport mechanism for gas permeation through polymer membranes [20]. The governing flux equation for the gas permeation (eq. 1) is based on Fick's law where the driving force is the difference in partial pressures over the membrane. The flux, J (m3 (STP)/m2h), is expressed as,

$$\frac{q_{p,i}}{A_m} = \frac{q_p y_{p,i}}{A_m} = J_i = \frac{P_i}{l}\left(p_h x_i - p_l y_i\right) \quad (1)$$

where q_p is the volumetric flow rate of the permeating gas (i) (m^3(STP)/h), P_i is the permeability of gas component i (m^3(STP)m/(m^2.h.bar)), l is the thickness of the membrane (m), p_h and p_l are pressure on the feed and permeate sides (bar), x_i and y_i are the fractions of component i on the feed and permeate sides, respectively, and A_m (m^2) is the required membrane permeation area. The general definition of permeability (*P)* of gases through membrane is defined as the product of diffusion, *D (*m^2/s) and solubility, *S* (m^3(STP)/m^3.bar) coefficients for the gas in the membrane material.

$$P = D.S \quad (2)$$

The intrinsic membrane selectivity "α" is estimated by the ratio of the pure gas permeabilities (Pi , Pj),

$$\alpha = \frac{P_i}{P_j} \quad (3)$$

while the process selectivity (α*) also termed as separation factor can be calculated by eq. 4,

$$\alpha^* = \frac{y_i / y_j}{x_i / x_j}.$$

(Process selectivity depends strongly on the process variables) (4)

In this work, a facilitated transport membrane is considered which a polymer with a "carrier component" incorporated in the polymer structure itself. The membrane material, poly (vinylamine) (PVAm), contains the amino group which contributes to the transport of CO2 through the membrane as a bicarbonate ion (HCO3-) when the membrane is wet (swollen with water). Hence CO2 transport through membrane is attributed to this carrier effect along with the Fickian diffusion, as illustrated in eq 5,

$$J_i = \frac{D_i}{l}\left(c_{i,0} - c_{i,l}\right) + \frac{D_{ic}}{l}\left(c_{ic,0} - c_{ic,l}\right) \quad (5)$$

where $c_i = S_i.p_i$. Here Si and pi are the solubility and partial pressure of component i respectively. Standard polymeric membranes without carriers generally lose their separation efficiency when swollen, but for this specific facilitated (PVAm) membrane, these properties will be enhanced. Further details on this membrane can be found in the paper published by [19]. In eq. 5, Di is the Fickian diffusion coefficient while Dic is the carrier facilitated diffusion coefficient, while c is the concentration of component i and its complex ic at the interfaces of the membrane 0 and l on the feed and permeate sides, respectively. The concentration difference of the complex ic in eq 5 must be further expressed by an equilibrium constant of the complex reaction and a distribution coefficient. Further details on this are explained by Cussler [21]. It is important to mention here that CO2 transport due to carrier effect has been ignored in this analysis.

CO2 Capture by a Membrane Separation Process

A sketch of membrane separation process is depicted in Fig. 10. The process variables (flow rates, pressure, temperature, compositions) require optimum values of membrane parameters (permeability, selectivity, membrane area) to achieve the desired degree of separation (CO2 in retentate and CH4 in permeate). Apart from feed composition and temperature, pressure ratio (ψ), between the upstream (p_h) and downstream *(*p_l*)* pressure over the membrane is an important process variable when using polymeric membranes. The recovery of desired component (in this work *i* = *CO*$_2$) by a membrane separation process is calculated by eq 6,

$$R = \theta . y_i / x_i \quad (6)$$

wherey_i and x_i are the mole fractions of CO_2 in permeate and feed and stage cut ($\theta = Q_p / Q_f$) being the ratio of permeate flow rate to the feed flow rate.

Process Conditions and Simulation Method

The CO_2 concentration in natural gas can vary from 5-40% depending on the source [22], methane always being the major component. Natural gas also contains significant amounts of ethane, some propane, butane and 1–3% of other higher hydrocarbons [23]. In contrast to many other studies, where a binary CO_2/CH_4 [3, 6] or tertiary $CO_2/CH_4/H_2S$ [15] mixture is considered, two real natural gas mixtures (classified as case A and case B). The gas compositions and process conditions for both natural gas mixtures are given in Table-1 and 2.

However, the nominal concentrations of H2S,NOx and ash have been ignored in the gas mixture. The feed (natural) gas flow rates (in both cases) are relatively smaller compared to existing membrane systems (100-700 MMSCFD) for natural gas sweetening [24, 25]. It is worth mentioning here that membrane systems for natural gas sweetening are typically favoured for small size applications (less than 5 MMSCFD) and remote applications. This supports the idea of employing membrane system for natural gas sweetening in this analysis. Membrane and amine absorption system become competitive for a size of 5-50 MMSCFD [25, 26].

FSC membrane performs well when swollen with water; hence feed gas saturated with water will facilitate the CO_2 transport through membrane. It is desired that neither vacuum nor sweep gas on the permeate side should be applied in this process. Furthermore, it is assumed that the membrane permeability is independent of pressure, concentration and the pressure drop on the feed side is nominal. The heat integrated structure of simulated process is shown in the Fig. 11a. The feed gas initially at 90 bar and 50°C is mixed with retentate from stage 3 (Ret3) and then fed to the first membrane stage. First membrane stage enables 50% trimming of CO_2 and thus remaining CO_2 containing stream (Ret1) is fed to stage 2. The retentate from stage 2 (Ret 2) contains less than 2% CO_2 and is sent to pipeline as product gas at 89 bar. Pressure on the permeate side is varied to see its influence on the required membrane area and energy. The temperature of permeate from stage 1 and stage 2 decreases due to high pressure drop across the membrane (depending on the applied pressure ratio across the membrane) and fair amount of higher hydrocarbons in the permeate liquefy. Therefore, permeate from stage 1 and stage 2 are mixed in a mixer and fed to a flash vessel where higher hydrocarbons are collected at the bottom and the vapour product enriched with CO_2, methane and some amount of other hydrocarbons is compressed, cooled and then fed to stage 3 for final enrichment of CO_2 in permeate from stage 3. The retentate from stage 3 is compressed to 90 bars and recycled back to stage 1 as feed. Permeate from stage 3 contains 90% CO_2 which can be stored for further processing and storage. Heat exchangers are used to economize the energy consumption. Cooling water is circulated in the heat exchangers to lower the temperature of compressed gas streams. The process conditions and compositions of different gas/liquid streams related to Fig. 11a are shown in Table-2a.

It is obvious from Case B data (Table-1) that natural gas stream contains only methane and CO_2, however saturated with water like Case A. Hence the process flow diagram for Case B is similar to Case A, except without a flash vessel. Similarly, process conditions and compositions of different gas/liquid streams related to Fig. 11b are shown in Table-2b.

A one-dimensional isothermal model has been implemented to evaluate the performance of a hollow-fiber membrane module composed of the described (CO2- selective) membrane. In this work, a feasibility analysis has been carried out with an in-house membrane program interfaced within process simulation program (Aspen Hysys) to investigate the influence of permeate pressure on the energy demand and gas processing cost. The input data is mainly based on the feed composition, CO2 and CH_4 permeances (~ 0.3 m3 (STP)/bar.h.m2 and 0.00856 m^3 (STP)/bar.h.m2 respectively) and the process conditions. The simulations are based on the SRK equation of state and the codes were developed for co-current, cross-flow and counter-current flow. This program has the possibility to utilize Hysys capability to calculate and couple energy balances in the process model. It gives the possibility to introduce sweep gas (however, not used in this analysis) on the permeate side which enables to choose a spiral-wound or a hollow-fiber module with or without inclusion of sweep gas.

Gas Processing Cost (GPC) has been calculated by the same methodology as reported in [5], and it is based on the values assigned to the selected process/economic parameters which might differ considerably for different evaluators. In this analysis, GPC defined as the cost per MSCF (1000 Standard Cubic Foot) of product, is based on three cost components; Total Plant Investment cost (TPI), Variable Operating & Maintenance cost (VOM) and the methane lost in permeate (CH_4Loss). Table-3 shows the values of economic and process parameters along with calculation methodology. The gas processing cost does not include the cost of permeate (~ 90% CO_2) re-compression to desired pipeline pressure. It is worth mentioning here that gas processing cost can differ substantially by changing the process parameters given in Table-3. The membrane area and the required energy impart substantial influence on the gas processing cost which can be seen in the results presented in subsequent paragraphs for different process conditions.

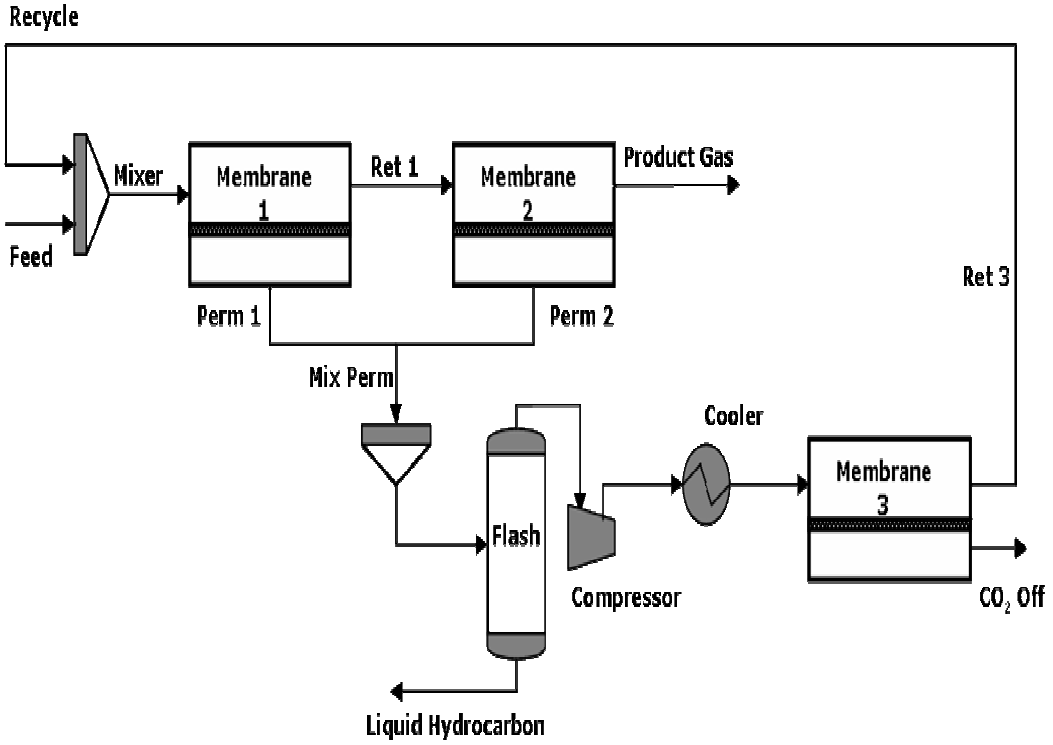

Fig. 10: Simplified two stage membrane separation process (without sweep flow).

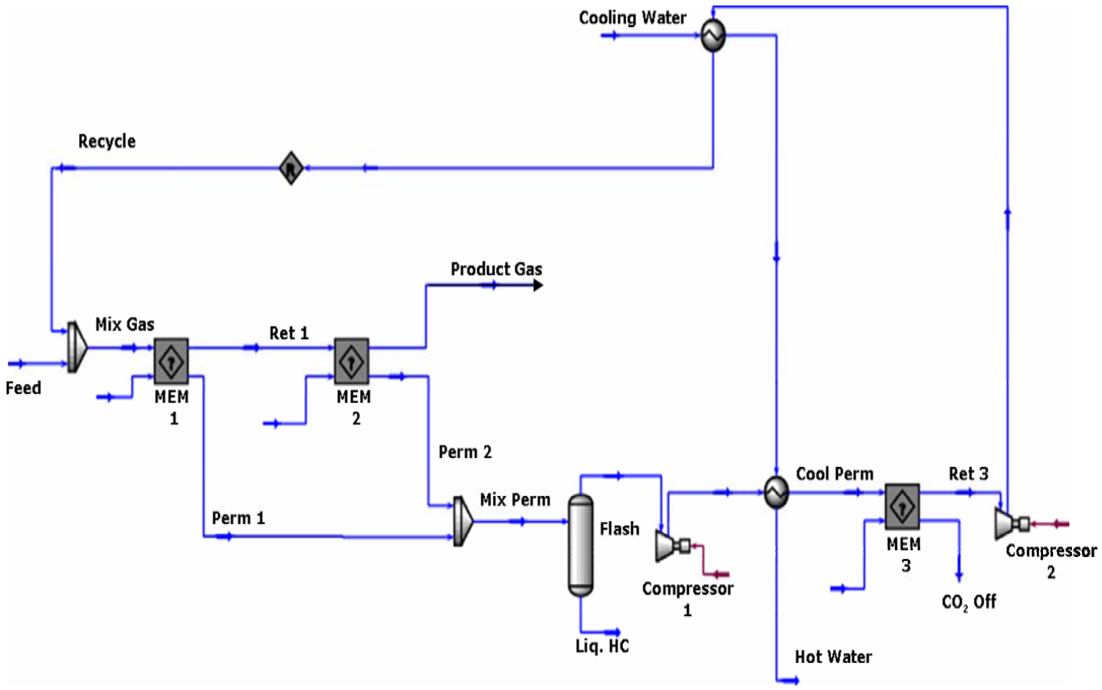

Fig. 11a: Hysys process flow diagram for 3-stage (Case A) membrane integrated natural gas sweetening without using any sweep or vacuum on the permeate side.

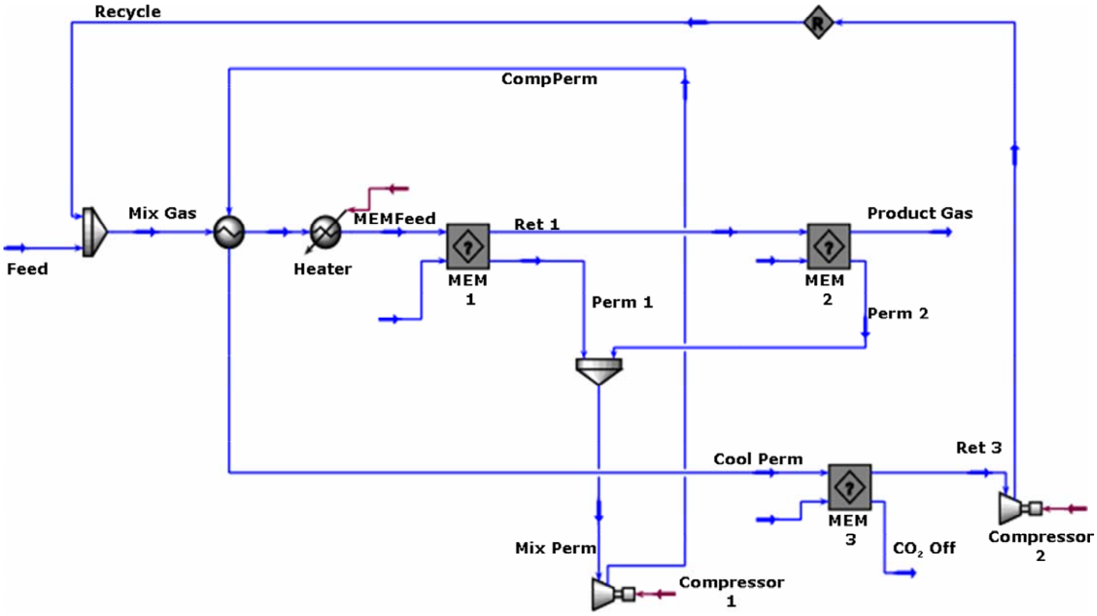

Fig. 11b: Hysys process flow diagram for 3-stage (Case B) membrane integrated natural gas sweetening without using any sweep or vacuum on the permeate side.

Table-2a: Gas compositions and process conditions related to Fig. 2a when permeate side pressure in stage1 and stage 2 is kept 5 bars constant. Whereas the permeate pressure in stage 3 is kept 1 bar constant.

Name	Feed	Mix Gas	Ret1	Product Gas	Perm1	Perm2
Vapour Fraction	1	1	1	1	0.989	0.989
Temperature, C	60	62.371	62.371	62.371	1.215	7.642
Pressure, bar	90	90	89.5	89	5	5
Molar Flow, m^3/d(gas)	0.354	0.480	0.404	0.319	0.076	0.086
Mass Flow, kg/h	410.673	603.281	459.454	331.751	143.827	127.703
Name	Liq.	Cool Perm	Ret3	CO_2 Off	Cool Water	Hot Water
Vapour Fraction	0	1	1	1	0	0
Temperature, C	4.732	67	67	44.789	8	72.830
Pressure, bar	5	30.05	30	1.1	2	1.9
Molar Flow, m^3/d(gas)	0.002	0.160	0.127	0.033	1.061	1.061
Mass Flow, kg/h	6.332	265.198	193.004	72.194	952.390	952.303
Component Name	Feed	Mix Gas	Ret1	Product Gas	Perm1	Perm2
CO_2 mole fraction	0.095	0.132	0.047	0.011	0.581	0.163
C_1 mole fraction	0.722	0.664	0.740	0.798	0.250	0.520
C_2 mole fraction	0.089	0.090	0.098	0.097	0.050	0.088
C_3 mole fraction	0.052	0.060	0.062	0.056	0.048	0.067
C_4 mole fraction	0.029	0.039	0.038	0.029	0.042	0.072
C_5 mole fraction	0.006	0.008	0.008	0.005	0.013	0.016
C_6 mole fraction	0.005	0.006	0.005	0.002	0.013	0.014
H_2S mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
H_2O mole fraction	0.002	0.001	0.002	0.002	0.003	0.060
Component Name	Liq.	Cool Perm	Ret3	CO2 Off	Cool Water	Hot Water
CO2 mole fraction	0.031	0.374	0.23	0.897	0	0
Cl mole fraction	0.012	0.46	0.499	0.042	0	0
C2 mole fraction	0.01	0.074	0.093	0.012	0	0
C3 mole fraction	0.002	0.006	0.083	0.015	0	0
C4 mole traction	0.205	0.066	0.08	0.019	0	0
C5 mole fraction	0.2	0.014	0.006	0.007	0	0
C6 mole fraction	0.4	0.006	0.009	0.008	0	0
H2S mole fraction	0.14	0	0	0	0	0
H2O mole fraction	0	0	0	0	1	1
Unit Operations	Duty (MJ/h)					
Compressor 1	22.90			MEM 1	MEM 2	MEM 3
Compressor 2	44.08		Area [m2]	38	84	13

Table-2b: Gas compositions and process conditions related to Fig. 2b when permeate side pressure in stage1 and stage 2 is kept 5 bars constant. Whereas the permeate pressure in stage 3 is kept 1 bar constant.

Name	Feed	Mix Gas	MEMFeed	Ret1	Product Gas	Perm1
Vapour Fraction	1	1	1	1	1	1
Temperature, C	8	24.81	60	60	60	7.96
Pressure, bar	115	115	115	114.9	114.8	5
Molar Flow, m^3/d(gas)	2.48	2.78	2.78	2.61	2.42	0.18
Mass Flow, kg/h	2079.02	2393.22	2393.22	2158.55	1955.39	234.67

Name	Perm2	Mix Perm	Comp Perm	Cool Perm	Ret 3	CO_2 Off
Vapour Fraction	1	1	1	1	1	1
Temperature, C	15.28	11.72	205.30	60	60	35.92
Pressure, bar	5	5	35.1	35.05	35	1
Molar Flow, m^3/d(gas)	0.19	0.37	0.37	0.37	0.31	0.06
Mass Flow, kg/h	203.16	437.83	437.83	437.83	313.88	123.95

Component Name	Feed	Mix Gas	Ret1	Product Gas	Perm1	Perm2
CO_2 mole fraction	0.029	0.044	0.044	0.021	0.007	0.383
C_1 mole fraction	0.971	0.956	0.956	0.979	0.993	0.617
C_2 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
C_3 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
C_4 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
C_5 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
C_6 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
H_2S mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
H_2O mole fraction	0.000	0.000	0.000	0.000	0.000	0.000

Table-2b continue

Component Name	Liq.	Cool Perm	Ret3	CO_2 Off	Cool Water	Hot Water
CO_2 mole fraction	0.193	0.284	0.284	0.284	0.163	0.899
C_1mole fraction	0.807	0.716	0.716	0.716	0.837	0.101
C_2 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
C_3 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
C_4 mole traction	0.000	0.000	0.000	0.000	0.000	0.000
C_5 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
C_6 mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
H_2S mole fraction	0.000	0.000	0.000	0.000	0.000	0.000
H_2O mole fraction	0.000	0.000	0.000	0.000	0.000	0.000

Unit Operations	Duty (MJ/h)					
Heater	126.4					
Compressor 1	122.4			MEM 1	MEM 2	MEM 3
Compressor 2	65.7		Area [m^2]	125	175	28

Conclusion

The results of this analysis indicate that a three stage membrane separation system for CO_2 capture from natural gas (without using any sweep gas or vacuum on the permeate side) is viable. Simulation analysis is based on two cases describing two different natural gas (feed) flow rates and compositions. Analysis shows that it is possible to attain 90% CO_2 recovery and 90% purity at considerably lower gas processing cost, especially for case A. Natural gas in case B contains lower concentration of CO_2 hence 90% recovery is not feasible in this case. However, for 90% CO_2 purity and 75% recovery, it is feasible to employ this process at lower gas processing cost. The gas processing cost does not include any cost of feed pre-treatment such as removal of higher hydrocarbon and moisture and the recompression of permeate (containing 90% CO_2) up to desired CO_2 pipeline pressure.

Table-3: Economic and Process parameters for Gas Processing Cost (GPC).

Total Plant Investment (TPI)	
Total membrane Module Cost (MC)	$\$5/ft^2$
Installed Compressor Cost (CC)	$\$8650 \times (HP/\eta)^{0.82}$
Fixed Cost (FC)	MC + CC
Base Plant Cost (BPC)	1.12 × FC
Project Contingency (PC)	0.2 × BPC
Total Facilities Investment (TFI)	BPC + PC
Start-up Cost (SC)	0.10 × VOM
Total Plant Investment (TFI)	TFI + SC
Annual Capital related Cost (CRC)	0.2 × TPI
Annual Variable Operating & Maintenance Cost (VOM)	
Contract & Material Maintenance Cost (CMC)	0.05 × TFI
Local Taxes & Insurances (LTI)	0.015 × TFI
Direct labour Cost (DC)	$15/h
Labour Overhead Cost (LOC)	1.15 × DL
Membrane Replacement Cost (MRC)	$\$ 2.5/ ft^2$ of membrane
Utility Cost (UC)	$ 0.07 /kWh
Annual variable Operating & Maintenance Cost (VOM)	CMC+LTI+DL+LOC+MRC+UC
Annual Cost of CH$_4$ Loss in Permeate (CH$_4$ LS)	
Annual Natural Gas Lost (NGLS)	$365 \times OSF \times (Q_1 \times x\_CH_4\_Q_1) \times (Q_P \times x\_CH_4\_Q_P)$
Methane Loss (CH$_4$ LS)	NGLS × NHV × NWP
Gas Processing Cost ($/MSCF), GPC	(CRC+CH4LS+VOM)/[365×OSF×Qf×(1-SCE)×1000]
Other Assumptions	
Membrane Life (t)	4 years
Wellhead price of crude natural gas 9NWP)	$ 4/MMBTU
Heating Value of Natural gas (NHV)	1066.8 MMBTU/MMSCF
Stage Cut Equivalent (SCE)	Permeate flow rate(Q_P)/Feed Flow rate (Q_f)
On-Stream Factor (OSF)	96%
Compressor efficiency (eta)	0.85

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