Vol. 15, No. 2 (2016) 611-624

Revista Mexicana de Ingeniería Química

SIMULATION OF THE NATURAL GAS PURIFICATION PROCESS WITH MEMBRANE TECHNOLOGY. TECHNICAL AND ECONOMIC ASPECTS

SIMULACIÓN DEL PROCESO DE PURIFICACIÓN DE GAS NATURAL CON TECNOLOGÍA DE MEMBRANA. ASPECTOS TÉCNICOS Y ECONÓMICOS

J.F. Palomeque-Santiago[1]* J. Guzmán[1], A.J. Zuñiga-Mendiola[2]

[1]*Instituto Mexicano del Petróleo, Programa de Ingeniería Molecular, Eje Central Lázaro Cárdenas Norte, No. 152, Col. San Bartolo Atepehuacan, México, D.F. C.P. 07730*

[2]*Instituto Tecnológico de Cd. Madero, Av. 1o. de Mayo esq. Sor Juana Inés de la Cruz s/n Col. Los Mangos C.P.89440, Cd. Madero Tamaulipas, México*

Received July 31, 2014; Accepted March 10, 2016

Abstract

The economy of the sweetening process of natural gas with membrane technology was analyzed in this study based on an own synthesized polyimide membrane. ProII commercial software was utilized to simulate the process with three different configurations: a) single stage, b) double stage with permeate recycle and c) triple stage with retentate recycle, using $CH_4/CO_2/H_2S$ as a ternary mixture with different compositions. The economic analysis was made taking the single stage configuration as a reference in relation to multiple step configurations. Gas processing cost (GPC) is mainly affected by three parameters: total plant investment (TPI), annual variable operating and maintenance costs (VOM) and annual cost of methane lost in the permeate (CH_4LS). These three parameters behave in opposite trends as feed pressure and feed flow increase.

Keywords: membrane processes, simulation process, natural gas purification.

Resumen

En este estudio se analiza la economía del proceso de endulzamiento de gas natural con tecnología de membranas, basado en una membrana de poliimida de desarrollo propio. Se utilizó el software comercial ProII simulando diferentes configuraciones: a) etapa simple, b) doble etapa con recirculación de permeado y c) triple etapa con recirculación de retenido utilizando una mezcla ternaria $CH_4/CO_2/H_2S$ con diferentes composiciones. El análisis económico fue realizado tomando la configuración de etapa simple como referencia en relación a las configuraciones en etapas múltiples. El costo de procesamiento de gas (GPC) es afectado principalmente por tres parámetros: inversión total de la planta (TPI), costos anuales de operación y mantenimiento (VOM) y los costos anuales por pérdida de metano en el permeado (CH_4LS). Estos tres parámetros se comportan de manera opuesta con el incremento en la presión de la alimentación y el incremento en el flujo de alimentación.

Palabras clave: procesos de membrana, simulación de procesos, purificación de gas natural.

1 Introduction

The composition of natural gas varies depending on its source, and therefore gases with high levels of pollutants such as acid gases (H_2S and CO_2), nitrogen, condensable hydrocarbons, aromatics (benzene, toluene, xylene, ethylbenzene) among others, can be found. The conventional natural gas sweetening is a treatment in which the acid gases are removed. This treatment is necessary because it prevents corrosion in the distribution lines, increases the calorific value of the gas and reduces the transport volumes, among other advantages (Bhide *et al.,* 1998). The sweetening process consists in the use of an aqueous solution of amines which absorbs acid gases; subsequently, these amines need to be cleaned by using activated carbon, generating large amounts of waste matter. Foaming, heat-stable salts, amine degradation products and carbon steel corrosion triggered by the amine solution are some problems present in these systems. For these reasons, it is necessary to develop alternative technologies to carry out the efficient purification of natural gas through environmentally friendly processes at low costs (Peters *et al.,* 2011).

Within the alternative procedures for the

* *Corresponding author. E-mail*: jpalomeq@imp.mx
Tel. 91757398

purification of natural gas, the membrane separation process has shown to be competitive in terms of cost and efficiency of the separations. Permeability and selectivity are important properties, since the higher the permeability, the less membrane area is required for a given separation, thus reducing the system cost. At higher selectivity, the hydrocarbon losses are reduced because acid gases are removed more efficiently, and therefore an increased volume of valuable product is recovered. Unfortunately, these properties move in opposite trends in membranes: as permeability increases, selectivity decreases and vice versa.

Different types of membranes have been commercialized for natural gas sweetening such as cellulose acetate, polyimides, silicon rubber, polysulfone, poly(phenylene oxide), ethyl cellulose, etc. (Baker and Lokhandwala, 2008; Yampolskii, 2012). The research for development of new polymeric materials with better transport properties is very wide; several review articles have reported the progress in the polymer science as potential future applications (Sanders *et al.,* 2013; Zhang *et al.,* 2013; Adewole *et al.,* 2013; Rufford *et al., 2012*; Scholes *et al.,* 2012). Polyimides have rigid chains and strong interchain interactions which limit their solubility and processability, but by incorporating bulky pendant groups, these disadvantages can be overcome resulting materials with chain packing efficiency, good permeability with only a minimum selectivity loss and with high glass transition temperature (Ayala *et al.,* 2003; Liaw *et al., 2012*; Xiao *et al.,* 2009). The behavior of a new polyimide like the one reported in the present work in the sweetening process can be predicted by simulation studies.

Among the many variables that affect the economy of the membrane process, the feed stream conditions and product requirements (flow rates, pressures, and temperatures), membrane properties (permeability, selectivity, and membrane life), local costs (labor, taxes, and energy costs) and local and international costs of the final products are very important. Some studies have been published regarding the optimization of the membrane area and process variables; the simulation of the natural gas sweetening has been studied by several authors: Qi determined that a two-stage configuration with permeate recycle and a three-stage configuration with residue recycle are suitable for natural gas treatment because they have the lowest process costs (Qi and Henson, 1998). Bhide described different process configurations related to the cost of removing CO_2 from a natural gas feed containing 5-40 mol% CO_2; they found that a two-stage configuration minimizes hydrocarbon losses and operating costs (Bhide and Stern, 1993). Among the optimization variables in processing plant designs, the membrane area and system pressure are key points for the operating and investment costs (Lababidi *et al.,* 1996), however, flow rates, stream compositions and number of stages must also be taken into account. Datta reported that for low CO_2 concentrations, a two-stage and a three-stage configuration are optimum because minimum gas processing costs can be achieved, but the choice of one configuration depends on carbon dioxide concentration in the feed and natural gas price (Datta and Sen, 2006). Hao showed that a two-stage configuration with recycle is optimum for upgrading a $CO_2/CH_4/H_2S$ mixture containing 0-10 mol% H_2S and up to 20 mol% CO_2, highlighting the importance of the selectivity of the membrane in the processing costs (Hao, *et al.,* 2008). The effects of the feed flow rate, feed pressure, feed composition and wellhead price of natural gas on the processing cost were reported. Ahmad studied different plant configurations, suggesting a two-stage configuration with permeate recycle as optimal for a CH_4/CO_2 separation system because the gas processing cost is minimum, however, a two-stage configuration with retentate recycle had a better CH_4 recovery, but compressor power, membrane area and gas processing costs are higher (Ahmad *et al.,* 2012).

Gas processing cost (GPC) has been defined as the parameter to optimize the process configuration; this parameter is reported to be mainly integrated by three components: total plant investment (TPI), annual variable operating and maintenance costs (VOM) and annual cost of methane lost in the permeate (CH_4LS); the influence of the process variables on GPC has also been studied (Qiu *et al.,* 1989; Bhide and Stern, 1993; Qi and Henson, 1998; Hao, *et al.,* 2008; Ahmad *et al.,* 2012; Ahmad *et al.,* 2013), but the behavior of TPI, VOM and CH_4LS with respect to such variables has not been reported, and depending on the application or the particular stream to purify, these components, as well as methane purity or methane recovery may become important in making the decision to incorporate additional stages. The aim of the present work is to study the influence of the process variables on gas processing cost and its components as a function of design parameters such as membrane area, compression power and stage cut, in a membrane separation processes to achieve a goal of 2 mol% CO_2 in product stream by using $CH_4/CO_2/H_2S$ as a ternary mixture. Single and multiple steps configurations were analyzed. This simulation study is based on the transport properties of a new polyimide not yet reported in the literature.

2 Simulation study

The membrane considered in the present study was synthesized in our laboratory. The structure of this polymer is shown in scheme 1 (Guzmán-Lucero D. J. *et al.,* 2014). Gas permeation data were obtained using a constant volume, variable pressure apparatus. Three process configurations were considered, the flow diagrams are illustrated in Fig. 1: a) single stage (1 stage), b) double stage with permeate recycle (2 stage PR), and c) triple stage with retentate recycle (3 stage RR). The natural gas composition depends on the stream source: a ternary $CH_4/CO_2/H_2S$ gas mixture containing 5-40 mol% CO_2 and 2-8.5 mol% H_2S were used. Table 1 shows the feed compositions, which were set as typically used in gas processing plants.

Operating conditions:

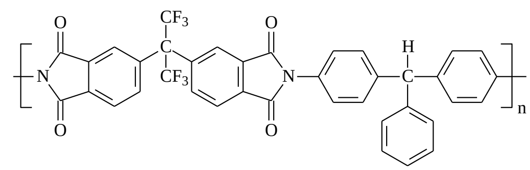
- Processing capacity: 60 MMSCFD.
- Feed pressure: 70 Kg/cm^2.
- Permeate pressure: 3 Kg/cm^2.
- Feed temperature: 25°C.

Membrane properties:

- CH_4 permeance: 2.83 GPU.
- CO_2 permeance: 116.64 GPU.
- H_2S permeance: 93.34 GPU.

Commercial softwares are useful tools to simulate membrane processes (Chowdhury *et al.,* 2005; Scholes *et al.,* 2012; Zhao *et al.,* 2012). This study was performed with the commercial software ProII 9.0. The target was set to obtain a CO_2 composition of 2 mol% in product stream as commercial methane is desired. ProII simulates the fractionation of components across a semi-permeable membrane. The model incorporated by ProII is applicable to high flux asymmetric membranes in any flow pattern provided the assumptions described by Pan, 1983. Our results may have a deviation of about ±15% because of the assumptions listed in the next section.

Proper pretreatment design is critical to the performance of all membrane systems. Typical pretreatment includes a coalescing filter, an adsorbent guard bed, a particle filter and a heater. This pretreatment must remove efficiently: liquids which cause swelling of the membranes and destruction of membrane integrity, heavy hydrocarbons which slowly coat the membrane surface, thus decreasing permeation rate, particulate materials that can block the membrane flow area and certain corrosion inhibitors that can be destructive to the membrane. Other elements can be added according to the feed characteristics, such as: a chiller to reduce the dew point of the gas and also the heavy hydrocarbon content, a turboexpander for the same purpose as a chiller, with the benefit of being a dry system, or a glycol unit to prevent hydrate formation or freeze-up. The cost of the pretreatment strongly depends on the composition of the feed, therefore, in this work, pretreatment was not considered.



Scheme 1. Polyimide structure 6FDA-DTM

Table 1. Characteristics of the feeds used in the simulation study

Run	Molar composition, %		
	CH_4	CO_2	H_2S
1	93.0	5.0	2.0
2	87.5	9.7	2.8
3	81.8	14.5	3.7
4	74.5	20.0	5.5
5	65.3	27.4	7.3
6	51.5	40.0	8.5

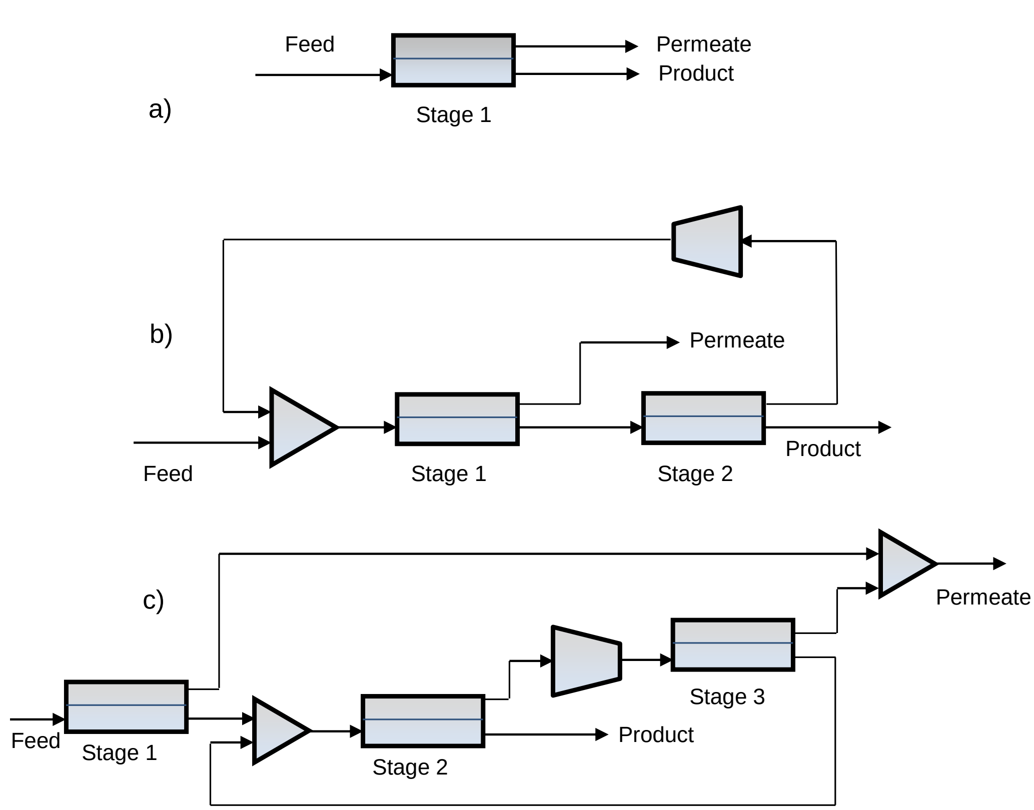


Fig. 1. Flow schemes: a) single stage, b) double stage with permeate recycle, c) triple stage with retentate recycle.

2.1 *Model description*

The controlling equation for the model is:

$$F_i = K_i \times \text{Area} \times (P_i, \text{retentate} - P_i, \text{permeate})$$

where:
F_i= Permeation flow of component i through the membrane (std volume/time)
K_i= Permeability constant of component i (volume/[area*time*pressure])
Area= Effective permeable area of the membrane
P_i= Partial pressure of component i.

Assumptions

The mathematical formulation for the permeation of the asymmetric membrane is based on the following assumptions:

1. The feed gas is on the skin side of the asymmetric membrane.
2. No mixing of permeate fluxes of different compositions occurs inside the porous supporting layer of the membrane.
3. The porous supporting layer has negligible resistance to gas flow, and diffusion along the pore path is insignificant due to high permeate flux.
4. The permeances of gases through the membrane are independent of pressure and mixed-gas effects.
5. Feed gas pressure drop is negligible.
6. There are no effects by CO_2 plasticization.

Economic parameters

Economic analyses of membrane systems vary from author to author, since many variables are taken into account. Prices of services and hydrocarbons change every day and comparisons of economic analyses are not easy. The cost of membrane modules is largely dependent on the type of module utilized: spiral wound or hollow fiber module, materials cost, pressure design, co-current or countercurrent flow, etc. Generally, high-pressure modules are more expensive than low-pressure or vacuum modules. Hollow fiber modules are significantly cheaper, per square meter of membrane than spiral-wound or plate-and-frame (Baker R. W., 2004). In this study, hollow fiber modules are considered, and for determining the membrane module cost we made an estimation of the cost of production of the synthesized polyimide; this value is in the range of high performance materials: 1-10 USD/m^2 (Baker and Lokhandwala, 2008). One gram of material covers around 1m^2 of membrane area.

The wellhead price of natural gas depends on the market conditions and economic assessments differ considerably depending on the evaluators' considerations. Nevertheless, such differences can be informative if the methodology is clearly described (Hao *et al.,* 2002). Significant differences also exist in the methods used by different industries to raise capital, interest payments, required internal rate of return, depreciation policy, marketing strategy, and in other local factors (Bhide and Stern, 1993). The gas processing cost for natural gas is a variable which may be expressed as the cost per MSCF of feed, but when this contains substantial amounts of CO_2, it is better to define GPC as the processing cost per MSCF of product. Some authors have found an optimal processing configuration depending on the natural gas price: at low natural gas price, a 2-stage configuration with no recycling is optimal, and a 3-stage configuration is optimal when natural gas price is high or at high CO_2 content in the feed (Datta and Sen, 2006). Other considerations associated to cost processing are, for example, facilities investment, labor costs, utility cost and price of crude natural gas, which vary from country to country. Membrane module cost, membrane replacement cost and membrane life, depend on the material and the manufacture process of the membrane. A single-stage process requires the minimum membrane area, there is no need of power requirement and it represents the lowest capital investment. Although hydrocarbon losses are very high, it is always the first configuration that every author studies, so a single-stage configuration can be used as a reference (Bhide and Stern, 1993).

The comparison of the cost ratios of multistage configurations relative to the simplest single-stage one is constant, once particular costs are identified, so the aim of this study is to compare the relative costs of a 2-stage PR and a 3-stage RR with a 1-stage one, since particular costs are closely related. Cost estimations in this study were performed by following methodology of Hao *et al.,* 2008 who assumed that gas processing costs are mainly determined from the total plant investment, annual variable operating and maintenance costs and annual cost of CH_4 lost in permeate. The economic parameters used in the present study are in Table 2.

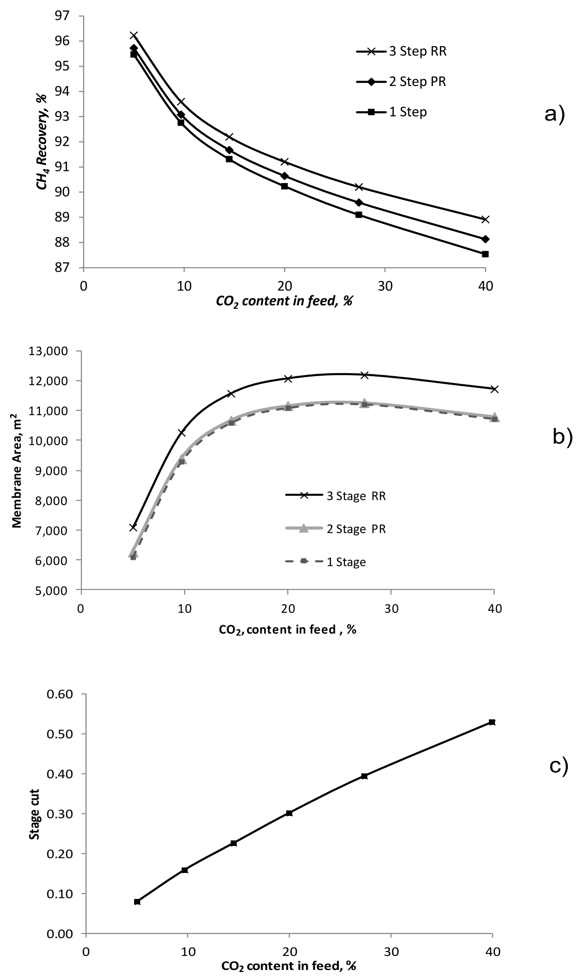


Fig. 2. Effect of CO_2 content in the feed on: a) methane recovery, b) membrane area, c) stage-cut.

Table 2. Economic parameters and assumptions

Parameter	Value
Total membrane module cost (MC)	\$10/ft^2 (includes the cost of the membranes)
Installed compressor cost (CC)	\$8650$\times$ (HP/η)$^{0.82}$
Fixed cost (FC)	MC + CC
Base plant cost (BPC)	1.12$\times$ FC
Project contingency (PC)	0.20$\times$ BPC
Total facilities investment (TFI)	BPC + PC
Start-up cost (SC)	0.10 $\times$ VOM (see below)
Total plant investment (TPI)	TFI + SC
Contract and material maintenance cost (CMC)	0.05 $\times$ TFI
Local taxes and insurance (LTI)	0.015 $\times$ TFI
Direct labor cost (DL)	\$15/h
Labor overhead cost (LOC)	1.15 $\times$ DL
Membrane replacement cost (MRC)	\$5/ft^2 of membrane
Utility cost (UC)	\$0.07/kW h
Annual variable operating and maintenance cost (VOM)	CMC + LTI + DL + LOC + MRC + UC
Annual natural gas lost (NGLS)	365 $\times$OSF $\times$ FN $\times$ XFNCH$_4$ $\times$ FLCH$_4$
Annual cost of CH_4 lost in permeate (CH$_4$LS)	NGLS $\times$ NHV $\times$ NWP
Annual capital related cost (CRC)	0.2 $\times$ TPI
Gas processing cost (GPC)	(CRC + CH$_4$LS + VOM)/[365 $\times$ OSF $\times$FN $\times$ (1-SCE) $\times$ 1000]
Membrane life (t)	5 years
Wellhead price of crude natural gas (NWP)	\$2.0/MMBTU
Heating value of natural gas (NHV)	1066.8 MMBTU/MMSCF
On-stream factor (OSF)	96%
Compressor efficiency (η)	0.8

3 Results and discussion

3.1 *Effect of feed composition*

Fig. 2 shows the effect of CO_2 content in the feed on the CH_4 recovery (Fig. 2a), membrane area (Fig. 2b) and stage cut (Fig. 2c). A maximum membrane requirement appears around 25 mol% of CO_2 in all cases (10,700 m^2 for 1 stage and 12,100 m^2 for 3-stage RR). 1-stage configuration needs less membrane area, but exhibits less CH_4 recovery in comparison with a 3-stage RR for achieving the target of 2% of CO_2 in product stream. Stage cuts for the three configurations follow the trend illustrated in 2c), where higher values of the stage cut are required as the CO_2 content in the feed is increased in order to get the product target. This means that both gases (CH_4 and CO_2) permeate in a higher extent, causing more methane loss and less CH_4 recovery. The compression power for a 3-stage RR is more than 4 times higher than that of a 2-stage PR, which is understandable since the last is processing more gas volume as seen in Fig. 3.

By comparing Figures 3b) and 3c), it can be seen that a 3-stage RR configuration, recovers additionally 250,000 SCFD of product, and 250,000 SCFD of permeate gas are additionally saved, but at the expense of 227.7 kW of additional compression power. These opposite variables affect global gas processing costs as discussed below. CO_2 and H_2S are removed at the same extent independently of the configuration used as shown in Table 3.

The ratios of the GPC components relative to 1 stage are presented in Table 4 for a 2-stage PR and a 3-stage as a function of feed composition. It can be seen clearly that the main impact from triple-stage configuration is on the plant investment due to the addition of another membrane module, which also impacts the compression power requirements as it can be seen in Fig. 3. The additional module investment increases the maintenance costs and lowers the costs of methane lost in permeate, which is also associated with high CH_4 recovery rates.

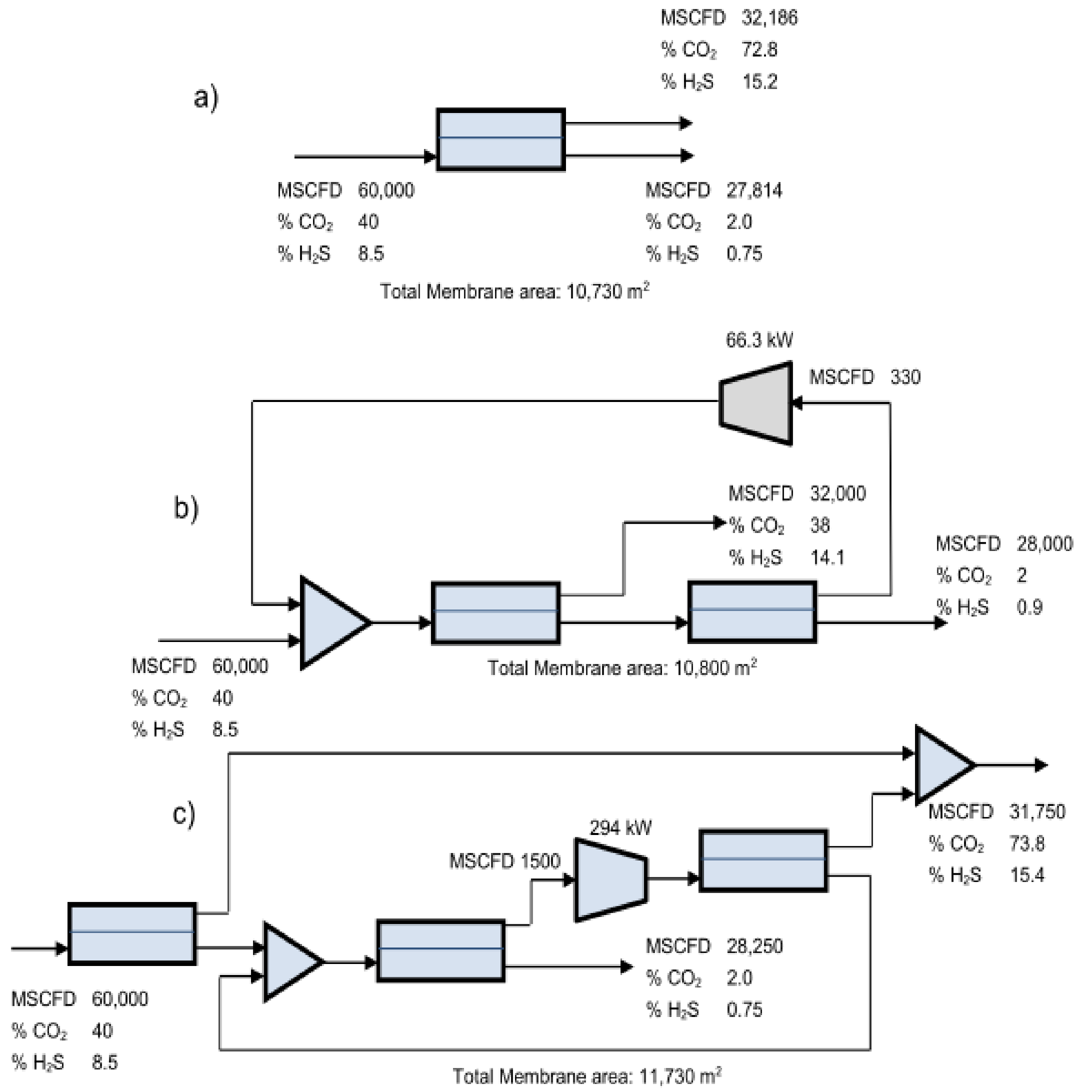


Fig. 3. Mass balances for the three configurations.

Table 3. Efficiency removal of acid gases as a function of feed composition for all configurations

Run	CO_2 removal (%)	H_2S in product (%)	H_2S removal (%)
1	63.4	0.90	59.0
2	82.7	0.73	78.5
3	89.3	0.69	85.7
4	93.0	0.80	89.8
5	95.6	0.84	93.0
6	97.7	0.75	95.9

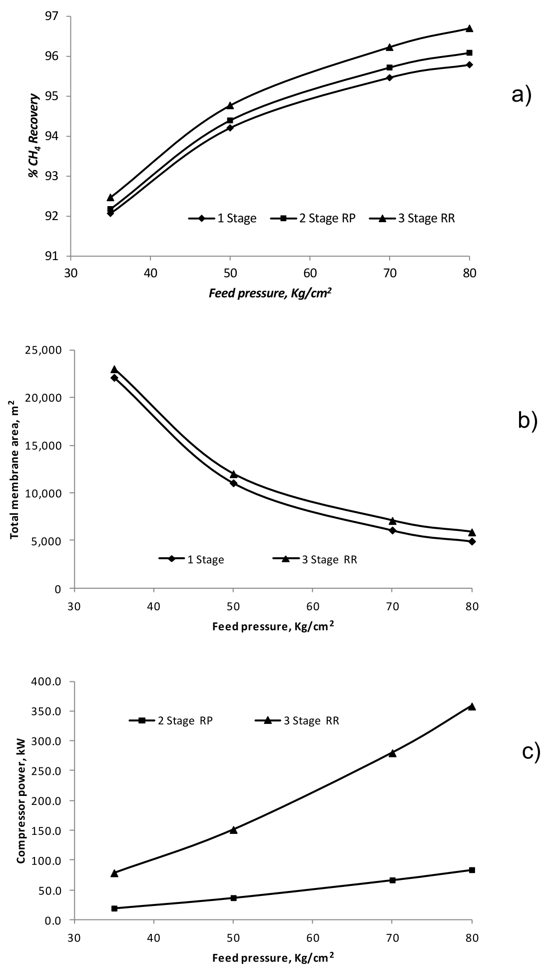


Fig. 4. Effect of feed pressure on: a) methane recovery, b) total membrane area, c) compressor power.

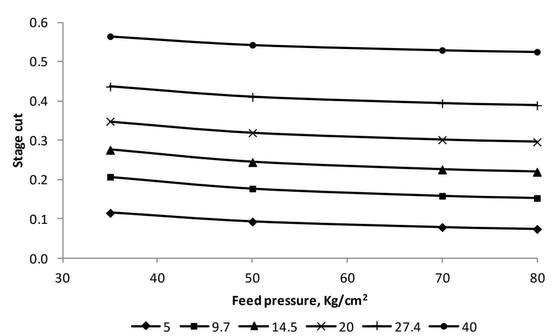


Fig. 5. Stage cut for triple configuration with permeate recycle, as a function of feed pressure for different CO_2 content in the feed (5-40 mol%).

3.2 *Effect of feed pressure*

The effect of feed pressure for a feed composition of 5 mol% of CO_2 (run 1, Table 1) on different configurations is shown in Fig. 4. A triple-stage configuration, as it is well known from the literature cited, results in a methane recovery rate that is higher than in other configurations, while the differences in membrane areas are relatively low; however, high investment in compressor power is needed; this means that an increment in plant investment is necessary for recovering more methane. Feed pressure has a negligible impact on the stage cut and removal of acid gases in the case of a 3-stage RR because of the low area required for getting the target (Figs. 5 and 6); H_2S removals are shown in Table 5. Madaeni modeled a mixture of CO_2 and CH_4 entering a hollow fiber module and found that feed pressure has little effect on CO_2 purity in the retentate stream (Madaeni *et al.,* 2010). High relative gas processing costs are obtained as feed pressure is increased, mainly by high requirements in total plant investment, as it can be seen in Table 6. Feed pressure and membrane area are critical design parameters because they are related to operational and capital expenditures respectively (Khalilpour *et al.,* 2013); as seen in Fig. 8, GPC lowers in function on feed pressure.

3.3 *Effect of feed flow*

2-stage PR and 3-stage RR show a linear increase in membrane area as a function of processing capacity for the conditions of Run 1, Table 3; increments in feed flow imply slightly changes in power requirements (not shown); this is because the compression flow changes relatively little. Slight decrements of 1.2% on CH_4 recovery and a slight raise of 0.012 in stage cut were found with feed flows from 20-80 MMSCFD (Fig. 7). Despite the increments in membrane area and compression power, the GPCs relative to the single-stage configuration were continuously decreasing (Table 7). This behavior is mainly related to the decreasing of the relative total plant investment (RTPI), which is correlated with the increments in membrane area and compression power, but in spite of the punctual increments, these are compensated by a high flow capacity. These results are in good agreement with the results shown by Bhide for a 3-stage permeate recycle configuration (Bhide *et al.,* 1998; Bhide and Stern, 1993).

In summary, low GPCs can be obtained by increasing the feed pressures, and at low CO_2 content in the feed, while it is slightly affected by feed flow. Fig. 8 illustrates the results of the gas processing costs for single stage as a function of the studied variables. The positive effect of feed pressure on the GPC is correlated with the effect of the driving force across the membrane, which lowers the membrane area requirements (Ahmad *et al.,* 2012), rising methane recovery and therefore, lowering the costs of CH_4 lost in permeate. Due to the elevated compressor power requirements, total plant investments rise in recycle configurations, and this effect is much stronger than the VOM and CH_4LS effects, which affects the relative gas processing cost (RGPC) (Table 6 and Fig. 9).

Opposite trends in RTPI, RVOM and RCH_4LS were found when comparing changes in feed pressure versus feed flow (Table 6 versus Table 7); these components are related to membrane area, compressor power, and CH_4 recovery (Figs. 4 and 7). When feed flow is increased, there is a raise in membrane area, power requirement, and costs of CH_4 lost in permeate, but these changes are compensated by the increase in large volumes of processing gas.

The GPC is increased as the CO_2 content in the feed is also increased. By analyzing Fig. 2, it can be seen that the raise in membrane area requirements also cause the total plant investment to be increased; an increment in the stage cut lowers the CH_4 recovery and increases CH_4LS. The relative components in Table 4 are lowered as a function of the feed composition with the exception of RCH_4LS; this suggests that the costs of recycle configurations are decreased with respect to the costs of a single-stage one. GPC reduction is not always the main criterion for selecting a configuration; it is also advisable to consider more investment for achieving better CH_4 recovery and lower costs of methane lost in permeate, this tendency is shown in Fig. 10.

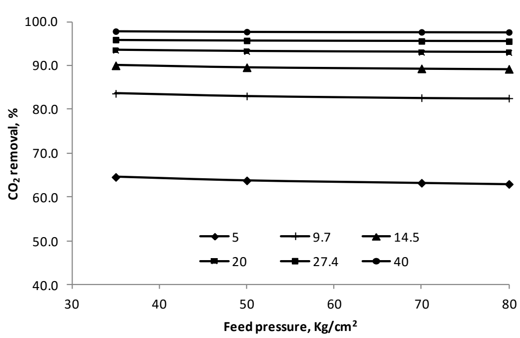


Fig. 6. CO_2 removals for triple configuration with permeate recycle, as a function of feed pressure for different CO_2 content in the feed (5-40 mol%).

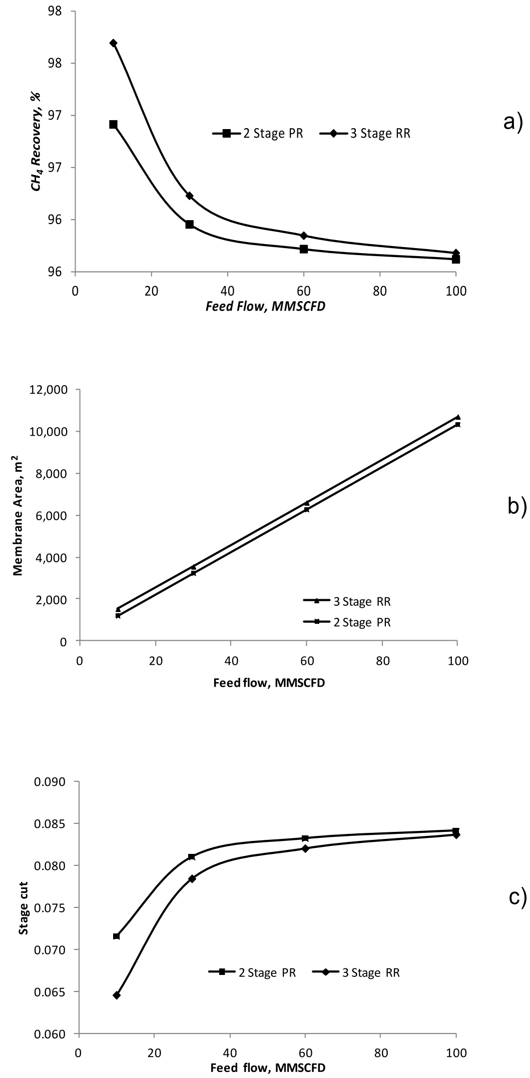


Fig. 7. Effect of feed flow on: a) CH_4 recovery, b) membrane area, c) stage cut.

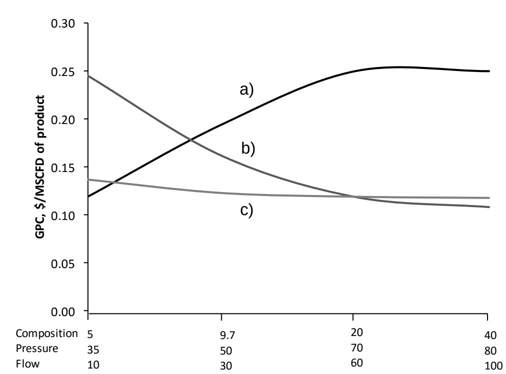


Fig. 8. GPC as a function of: a) feed composition, b) feed pressure and c) feed flow.

Table 4. Relative costs as functions of feed composition

Run	Relative Total Plant Investment (RTPI)	Relative Annual Variable Operating and Maintenance Cost (RVOM)	Relative Annual Cost of CH_4 Lost in Permeate (RCH_4LS)	Relative Gas Processing Cost (RGPC)
Double stage with permeate recycle				
1	1.65	1.23	0.94	1.03
2	1.41	1.16	0.96	1.01
3	1.36	1.14	0.96	1.00
4	1.35	1.14	0.96	1.00
5	1.34	1.14	0.96	1.00
6	1.36	1.14	0.95	1.00
Triple stage with retentate recycle				
1	3.09	1.77	0.83	1.10
2	2.41	1.58	0.88	1.06
3	2.25	1.53	0.90	1.06
4	2.21	1.52	0.90	1.06
5	2.21	1.52	0.90	1.06
6	2.29	1.55	0.89	1.06

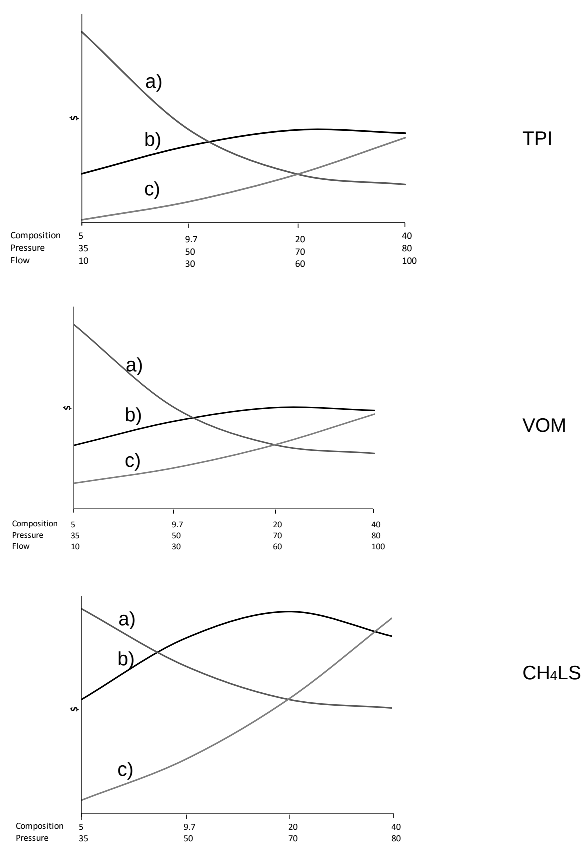


Fig. 9. Trends of TPI, VOM and CH_4LS as a function of: a) feed composition, b) feed pressure and c) feed flow.

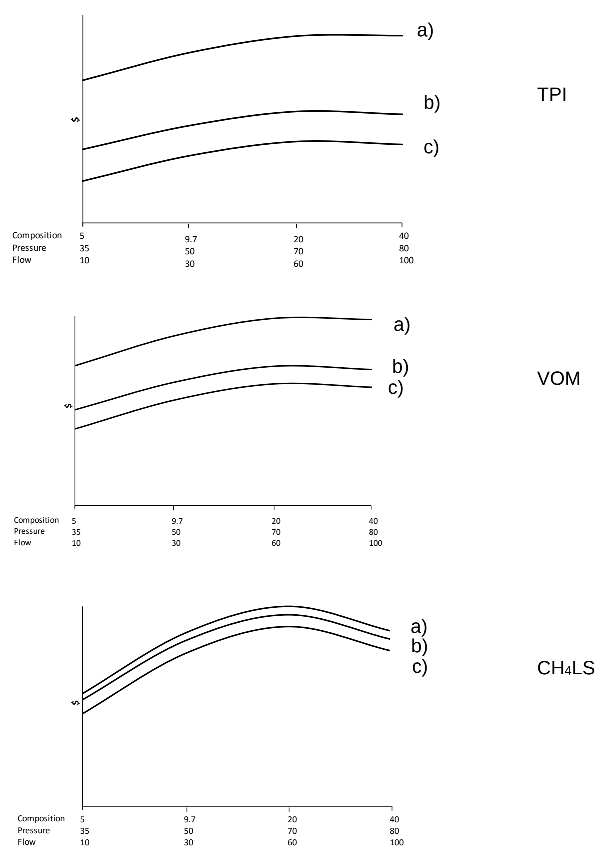


Fig. 10. Trends of TPI, VOM and CH_4LS for: a) single stage, b) double stage with permeate recycle and c) triple stage with retentate recycle

Table 5. Efficiency removal of acid gases as a function of feed pressure for 40 mol % CO_2 in the feed.

Feed pressure Kg/cm^2	CO_2 removal (%)	H_2S in product (%)	H_2S removal (%)
35	97.8	0.65	96.6
50	97.7	0.71	96.2
70	97.7	0.75	95.9
80	97.6	0.76	95.7

Table 6. Relative costs as functions of feed pressure

Feed pressure Kg/cm^2	Relative Total Plant Investment (RTPI)	Relative Annual Variable Operating and Maintenance Cost (RVOM)	Relative Annual Cost of CH_4 Lost in Permeate (RCH_4LS)	Relative Gas Processing Cost (RGPC)
Double stage with permeate recycle (PR)				
35	1.07	1.04	0.99	1.00
50	1.23	1.09	0.97	1.01
70	1.65	1.23	0.94	1.03
80	1.96	1.31	0.93	1.04
Triple stage with retentate recycle (RR)				
35	1.24	1.12	0.95	1.01
50	1.76	1.34	0.90	1.04
70	3.18	1.80	0.83	1.11
80	4.24	2.08	0.78	1.14

Table 7. Relative costs as functions of feed flow

Feed Flow MMSCFD	Relative Total Plant Investment (RTPI)	Relative Annual Variable Operating and Maintenance Cost (RVOM)	Relative Annual Cost of CH_4 Lost in Permeate (RCH_4LS)	Relative Gas Processing Cost (RGPC)
Double stage with permeate recycle (PR)				
10	5.04	1.54	0.68	1.17
30	2.30	1.34	0.89	1.05
60	1.65	1.23	0.94	1.03
100	1.39	1.16	0.96	1.01
Triple stage with retentate recycle (RR)				
10	8.28	2.00	0.51	1.38
30	3.42	1.66	0.83	1.13
60	2.19	1.43	0.92	1.07
100	1.71	1.30	0.95	1.04

Conclusions

Low gas processing costs can be obtained at low CO_2 content in the feed and at higher feed pressures and feed flows. The positive effect of pressure on GPC results in low membrane area requirements, rising methane recovery and therefore lowering the costs of CH_4 lost in permeate, however, compressor power requirements arise and total plant investments are also raised; in this case, the effect on total plant investment is much stronger than the effects on the annual variable operating and maintenance cost and the annual cost of CH_4 lost in permeate.

The effect of raising the feed flow on gas processing cost leads to an increase in the membrane area and the compression power, but these effects are offset by a decrease in the cost of CH_4 lost in permeate, a raise in CH_4 recovery, and the increasing volumes of processing gas.

The increasing CO_2 content in the feed raises the gas processing costs because of membrane area requirements and total plant investment; an increment in the stage cut lowers CH_4 recovery and increases the CH_4 lost in permeate; however, the relative components RTPI (relative total plant investment), RVOM (relative annual variable operating and maintenance cost) and RCH_4LS (relative annual cost of CH_4 lost in permeate) are lowered as functions of the feed composition; this suggests that the cost ratios of the recycle configurations are decreased with respect to the single-stage costs.

Acknowledgements

The authors acknowledge the Instituto Mexicano del Petróleo for supporting this research work through the project D.00264.

Abbreviations and symbols used

Area	effective permeable area of the membrane
BPC	base plant cost, $
CC	installed compressor cost, $
CH_4LS	annual cost of CH_4 lost in permeate $/year
CMC	annual contract and material maintenance cost, $/year
CRC	annual capital related cost, $/year
DL	annual direct labor cost, $/year
F	permeation flow through the membrane, std volume / time
FN	net feed flow rate, MMSCFD
FC	fixed cost, $
$FLCH_4$	fraction of CH_4 lost in permeate, mol-fraction
GPC	gas processing cost, $/MSCF of natural gas of product
HP	power requirement for compressor, hp
K	Permeability constant of component *i*, volume / [area*time*pressure])
LOC	annual labor overhead cost, $/year
LTI	annual local tax and insurance cost, $/year
MC	total cost of membrane module(s), $
MMBTU	10^6 BTU
MMSCFD	10^6 standard cubic feet per day (at 14.696 psia and 60 °F)
MSCFD	10^3 standard cubic feet per day (at 14.696 psia and 60 °F)
MRC	annual membrane replacement cost, $/year
NGLS	annual loss of natural gas, MMSCF/year
NHV	heating value of natural gas, 1066.8 MMBTU/MMSCF
NWP	wellhead price of crude natural gas, $/MMBTU
OSF	on-stream factor
P	partial pressure, lb/in^2
PN	net permeate flow rate, MMSCFD
PC	project contingency, $
RCH_4LS	relative annual cost of CH_4 lost in permeate (respect to single stage at each particular conditions)
RGPC	relative gas processing cost (respect to single stage at each particular conditions)
RTPI	relative total plant investment (respect to single stage at each particular conditions)
RVOM	relative annual variable operating and maintenance cost (respect to single stage at each particular conditions)
SC	start-up cost, $
SCE	stage-cut equivalent
SCFD	standard cubic feet per day (at 14.696 psia and 60 °F)
t	membrane life, years
TFI	total facilities investment, $
TPI	total plant investment, $
UC	annual utility cost, $/year
VOM	annual variable operating and maintenance cost, $/year
XFN	composition of net feed stream
Greek symbols	
η	compressor efficiency
Subscripts	
i	component *i*

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