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Design Simulation Analysis of Natural Gas Purification Mechanisms and the Economic Utilazation of Membrane Technology

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ABSTRACT

This research evaluated the economics of natural gas sweetening utilizing a polyimide membrane. The processes were simulated using Pro II commercial software in three configurations: single stage, double stage with retentate recycling, and triple stage with permeate recycle. The economic study compared the single stage setup to the multiple step configuration. The GPC is influenced by three factors: total plant investment, yearly variable operational and maintenance expenses, and annual methane loss in permeate (CH4LS). As input pressure and input flow rise, these three factors reverse.

Keywords:membrane mechanisms, simulation mechanisms, natural gas purification.

I. INTRODUCTION

Due to the fact that natural gas composition changes depending on the source, high pollutant gases such acid gases (H2S and CO2) might be discovered. The acid gases are eliminated in the typical natural gas sweetening. This treatment is required to avoid corrosion in distribution lines, boost gas calorific value, and minimize transit quantities (Bhide et al., 1998). This is done by soaking amines in water, which absorbs acid gases, and then cleaning them with activated carbon, which generates a lot of trash. There are issues with carbon steel corrosion caused by amine breakdown products and foaming in these systems. For these reasons, it is vital to develop low-cost alternative methods for effective natural gas purification (Peters et al., 2011).

Membrane separation mechanisms have demonstrated to be comparable in terms of cost and separation efficiency among various natural gas purification techniques. The greater the porosity, some less membrane area is needed for a larger fraction, lowering the system cost. With improved selectivity, less hydrocarbon is lost due to acid gas removal, and more value product is recovered. Unfortunately, in membranes, permeability rises while selectivity diminishes.

Membranes for natural gas sweetening include cellulose acetate, polyimides, silicone rubber, polysulfone, poly(phenylene oxide), and ethyl cellulose (Baker and Lokhandwala, 2008; Yampolskii, 2012). Several review papers have cited developments in polymer science as prospective future uses (Sanders et al., 2013; Zhang et al., 2013; Adewole et al., 2013; Rufford et al., 2012; Scholes et al., 2012). By including bulky pendant groups, these disadvantages may be overcome, resulting in materials with high chain packing efficiency, good permeability with little selectivity loss, and high glass transition temperature (Ayala et al., 2003; Liaw et al., 2012; Xiao et al., 2009). Simulation studies can anticipate the behavior of a novel polyimide like the one disclosed in this study in sweetening processes.

Local expenses (labor, taxes, and energy prices) are particularly essential in determining the economics of membrane processes. Several writers have explored the modeling of natural gas sweetening and optimized the membrane area and mechanism variables: Qi discovered that a two-stage with retentate recycle and a three-stage with residue recycle are acceptable for natural gas treatment (Qi and Henson, 1998). They observed that a two-stage system reduces hydrocarbon losses and operating costs while extracting CO2 from a natural gas source containing 5-40 mol% CO2 (Bhide and Stern, 1993). The membrane area and system pressure are important factors for operational and financing costs (Lababidi et al., 1996), but so are flow rates, stream compositions, and stage counts. A two-stage or three-stage system is optimal for low CO2 concentrations, according to Datta, although the decision relies on the feed carbon dioxide concentration and price of natural gas (Datta and Sen, 2006). Hao demonstrated the role of membrane selectivity in the processes costs by upgrading a CO2/CH4/H2S combination comprising 010 mol% H2S and up to 20 mol% CO2 (Hao, et al., 2008). The impacts of feed rate, pressure, feed mix, and natural gas wellhead price on mechanism cost were reported. For a CH4/CO2 separation system, Ahmad recommends a two-stage design with permeate recycle since the gas mechanisms cost is low. A two-stage setup with retention recycle had greater CH4 recovery, but higher compressor power, membrane area and gas mechanisms costs (Ahmad et al., 2012).

It has been reported that total plant investment (TPI), annual variable operating and maintenance costs (VOM), and annual cost of methane lost in the permeate (CH4LS) are the main components of GPC (Qiu et al., 1989; Bhide and Stern, 1993; Qi and Henson, 1998; Hao, et al., 2008; Ahmad et al., 2009). To generate a 2 mol% CO2 product stream utilizing CH4/CO2/H2S as a ternary combination, the current work investigates the impact of membrane size, compression power, and stage cut on gas mechanisms cost and components. Multiple step combinations were studied. This simulation work uses a novel polyimide with unreported transport characteristics.

II. SIMULATION STUDY

Our lab created the membrane used in this investigation. (Guzman-Lucero D. J. et al., 2014) It was measured and use a constant volume, constant pressure instrument. Three mechanisms designs were examined, as shown in Fig. 1: a) single stage (1 stage), b) double stage (2 stage PR), c) triple stage (3 stage PR) (3 stage RR). The natural gas composition varies depending on the stream source: 5-40 mol% CO2 and 2-8.5 mol% H2S were employed. Table 1 indicates the feed compositions utilized in gas mechanism plants.

Conditions of use:

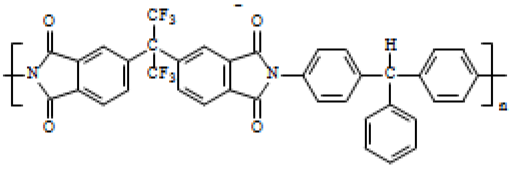
- Mechanisms capacity: 60 MMSCFD.
- Feed pressure: 70 Kg/cm^2.
- Permeate pressure: 3 Kg/cm^2.
- Feed temperature: 25°C.

Membrane properties:

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

Commercial software can model membrane processes (Chowdhury et al., 2005; Scholes et al., 2012; Zhao et al., 2012). This research used ProII 9.0 commercial software. The goal was to achieve a CO2 content of 2% in the product stream for commercial methane. ProII replicates semi-permeable membrane fractionation. ProII's model applies to high flux asymmetrical membranes in any fluid flow provided Pan's assumptions are satisfied. The assumptions given in the following section may create a 15% variance in our findings.

All membrane systems need proper pretreatment design. There are four main types of pretreatment: coalescing filters, particle filters and heaters. This pretreatment must effectively remove liquids that cause membrane swelling and degradation, heavy hydrocarbons that coat the membrane surface and delay permeability, particulates that may obstruct membrane flow, and corrosion inhibitors that can harm the membrane. To minimize the dew point and high hydrocarbon content of the gas, a turboexpander may be used instead of a chiller, and a glycol unit can avoid hydrate formation or freezeup. Pretreatment is expensive and relies on the feed composition, hence it was not addressed in this study..



Scheme 1. Polyimide structure 6FDA-DTM

Table 1.Characteristics of the simulated feeds

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

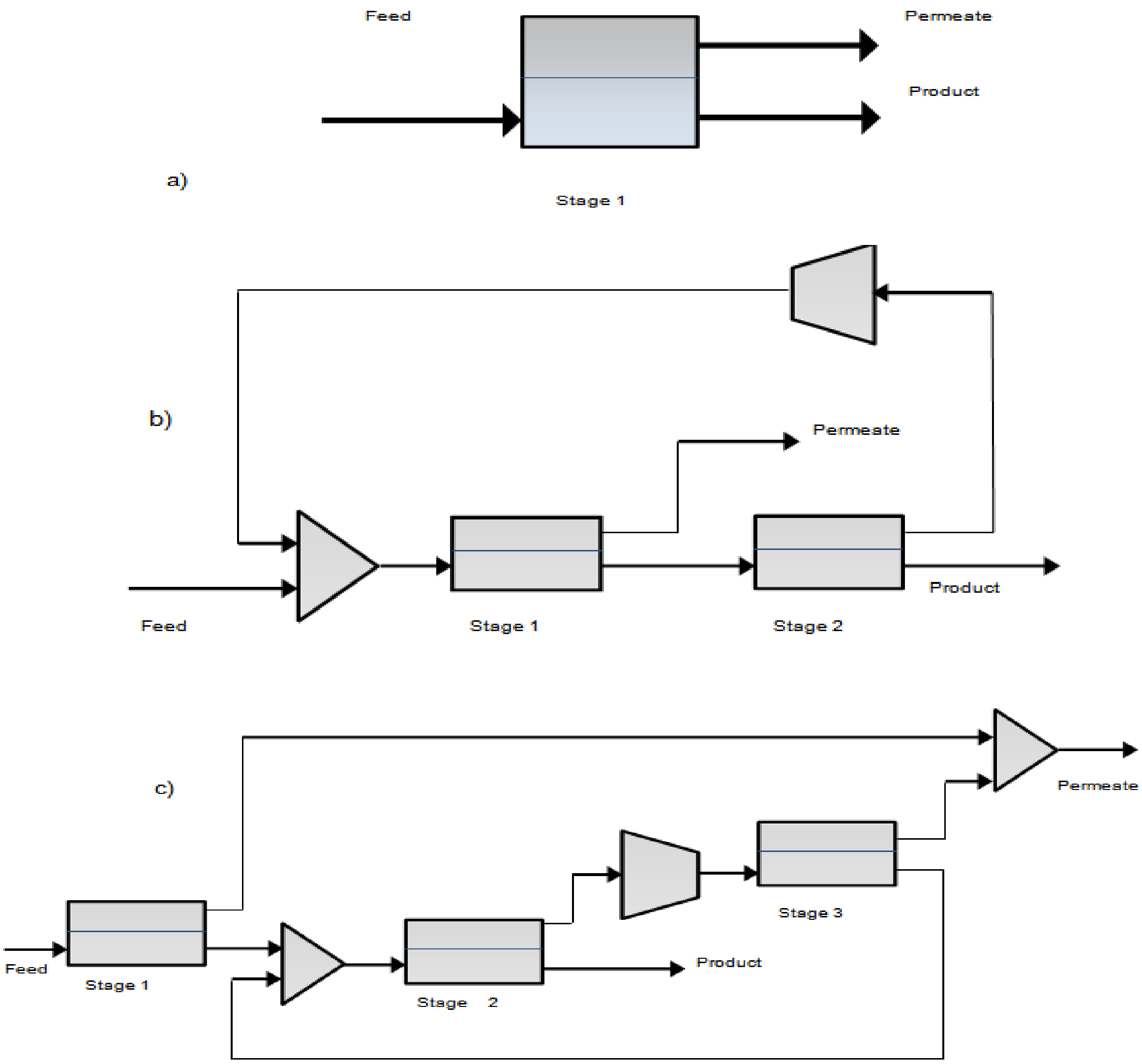


Fig. 1.Flow schemes: single (a), double (b), triple (c), with permeate recycling .

2.1 Model description

The controlling equation for the model is:

$F_i = K_i \times Area \times P_i,retentate - P_i,permeate$

where

Fi = Membrane permeation (volume/time) of component I Ki = volume/[area*time*pressure. permeability constant The membrane's effective permeable area Pi = component i partial pressure.

where:

Assumptions

The mathematical framework for asymmetric membrane permeability assume the following.

1. The feed gas is on the asymmetric membrane's skin.
2. The porous supporting layer of the membrane does not mix permeate fluxes of various compositions.
3. Due to strong permeate flux, the porous supporting layer has low gas flow resistance and low diffusion along the pore channel.

The membrane's gas permeance is independent of pressure and mixed gas effects.

5. No feed gas pressure decrease.
6. CO2 plasticization has no impact.

Economic parameters

Many factors are included in economic assessments of membrane systems. Prices of goods and fuels fluctuate daily, making economic comparisons difficult. The cost of membrane modules varies depending on the material, pressure, and flow direction. However, high-pressure modules tend to be more costly. Hollow fiber modules cost far less per square meter than spiral-wound or plate-and-frame (Baker R. W., 2004). In this research, hollow fiber modules are studied, and the cost of producing polyimide is estimated to be in the variety of high performance components: 1-10 USD/m2 (Baker and Lokhandwala, 2008). 1 gram material covers 1m2 membrane.

The wellhead cost of natural gas is determined by market circumstances, and economic judgments are based on the evaluators' perspectives. Such disparities might be instructive if the technique is communicated explicitly (Hao et al., 2002). Interest rates, necessary rate of return, amortization policy, business model, and other local considerations vary widely amongst businesses (Bhide and Stern, 1993). For natural gas, the mechanisms cost per MSCF of feed is a variable that may be stated as a cost per MSCF of product when the feed includes significant levels of CO2. Depending on the exposure gas price, a 2-stage system with no recycling is ideal, whereas a 3-stage configuration with high CO2 concentration in the feed is optimal (Datta and Sen, 2006). Other cost concerns include facility investment, personnel expenses, utility costs, and the price of oil natural gas, which varies by nation. Membrane module cost, replacement cost, and life depend on membrane material and manufacturing methods. A single stage mechanism needs the least membrane area, no power, and the least capital investment. Despite the substantial hydrocarbon losses, every author investigates the initial configuration, hence a singlestage setup may be used as a benchmark (Bhide and Stern, 1993).

Since specific expenses are closely connected, this research examines the relative costs of a 2-stage PR and a 3-stage RR to a 1-stage design. It was anticipated by Hao et al. (2008) that gas mechanisms costs are principally controlled by total plant investment, yearly variable operating costs, and annual cost of CH4 lost in permeate. Table 2 lists the economic variables considered in this investigation.

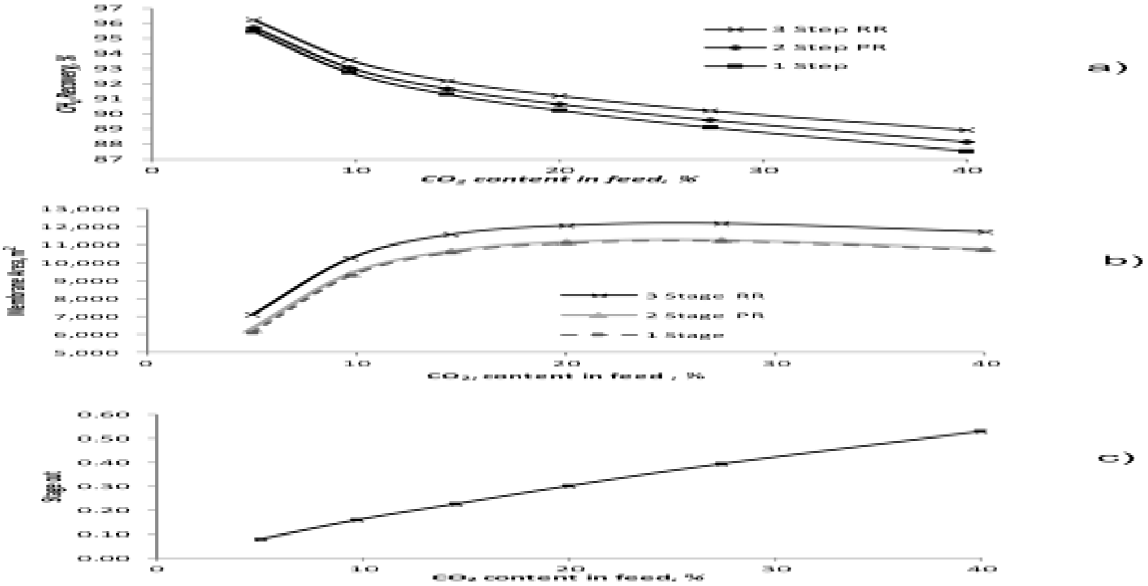


Fig. 2.CO_2 level in feed affects methane recovery, membrane area, and stage-cut.

Table 2. Economic assumptions

Parameter	Value
Total membrane module cost (MC)	\$10/ft^2 (includes the cost of the membranes)
Installed compressor cost (CC)	\$8650× $(HP/\eta)^{0.82}$
Fixed cost (FC)	+ CC
Base plant cost (BPC)	1.12× FC
Project contingency (PC)	0.20× BPC
Total facilities investment (TFI)	BPC + PC
Start-up cost (SC)	0.10 × VOM (see below)
Total plant investment (TPI)	TFI + SC
Contract and material maintenance cost (CMC)	0.05 × TFI
Local taxes and insurance (LTI)	0.015 × TFI
Direct labor cost (DL)	\$15/h
Labor overhead cost (LOC)	1.15 × 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)	×OSF × FN × XFNCH$_4$ × FLCH$_4$
Annual cost of CH$_4$ lost in permeate (CH$_4$LS)	NGLS × NHV × NWP
Annual capital related cost (CRC)	0.2 × TPI
Gas mechanisms cost (GPC)	(CRC + CH$_4$LS + VOM)/[365 × OSF ×FN × (1-SCE) × 1000]
Membrane life (t)	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

III. RESULTS AND DISCUSSION

3.1 Feed composition effect

Figure 2 demonstrates the influence of feed CO2 content on CH4 recovery (a), membrane area (b), and stage cut (c) (Fig. 2c). In all circumstances, the maximal membrane CO2 demand is roughly 25%. (10,700 m2 for 1 stage and 12,100 m2 for 3-stage RR). A 1-stage layout uses less membrane area but recovers less CH4 than a 3-stage RR to achieve 2% CO2 in the product stream. The stage cuts for the three configurations follow the pattern seen in 2c), where bigger stage cuts are needed as the CO2 level in the feed increases. As a result, more methane is lost and less CH4 is recovered. The compression power of a 3-stage RR is almost 4 times that of a 2-stage PR, which makes sense given the latter's larger gas volume.

Considering Figures 3b) and 3c), a 3-stage RR arrangement recovers 250,000 SCFD more product and saves 250,000 SCFD permeate gas, but at the cost of 227.7 kW more compression power. Lesser known aspects influence worldwide gas mechanisms costs. CO2 and H2S are eliminated to the same degree regardless of design (Table 3).

Table 4 shows the GPC component ratios compared to 1 stage for a 2-stage and 3-stage PR based on feed composition. As shown in Fig. 3, the major effect of a triple-stage system is on plant investment owing to the inclusion of another membrane module. The extra module investment reduces maintenance costs and boosts permeate methane collection rates.

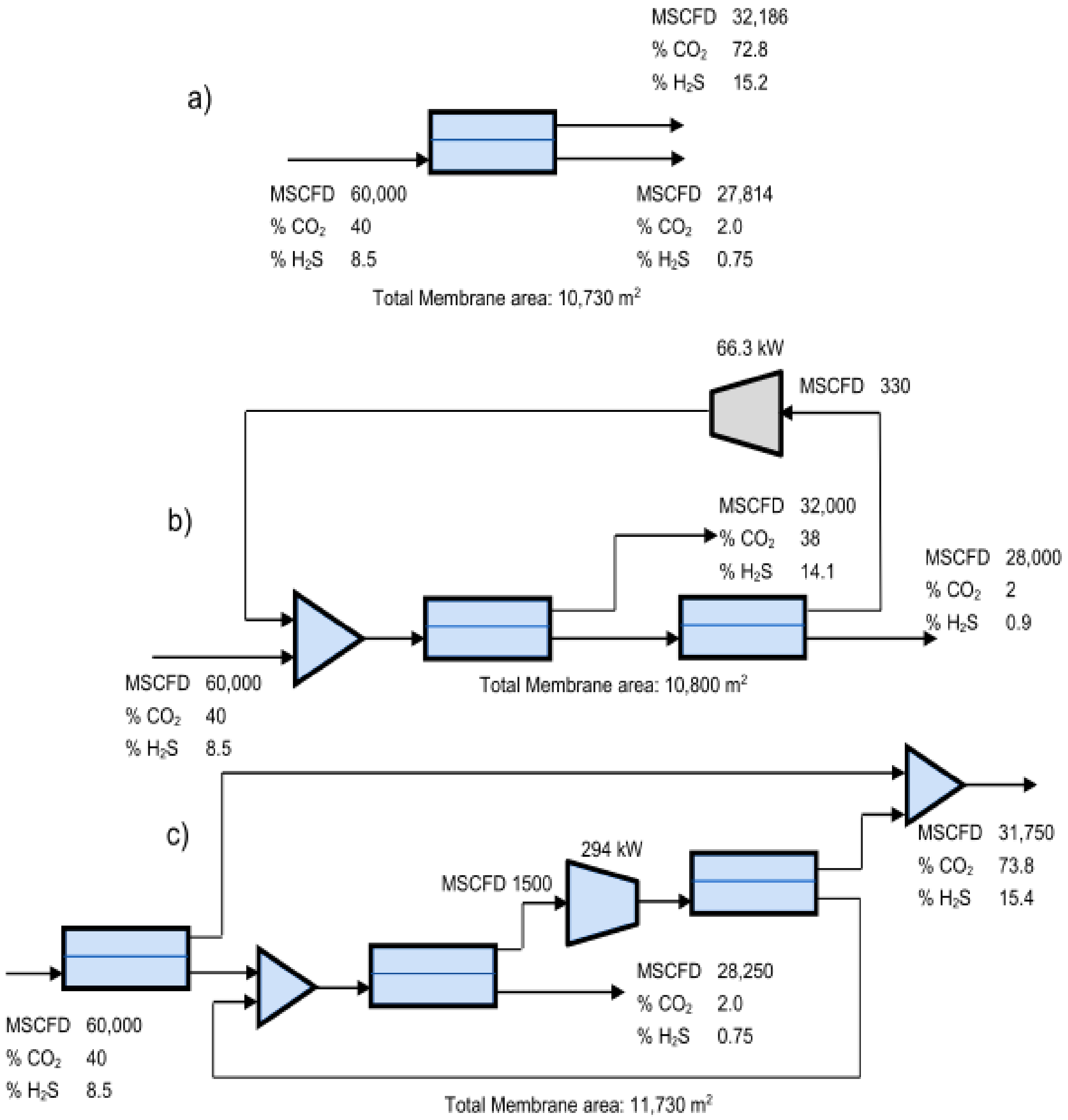


Fig. 3. The three arrangements' mass balan

Table 3.Acid gas removal efficiency vs. feed composition for all combinations

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

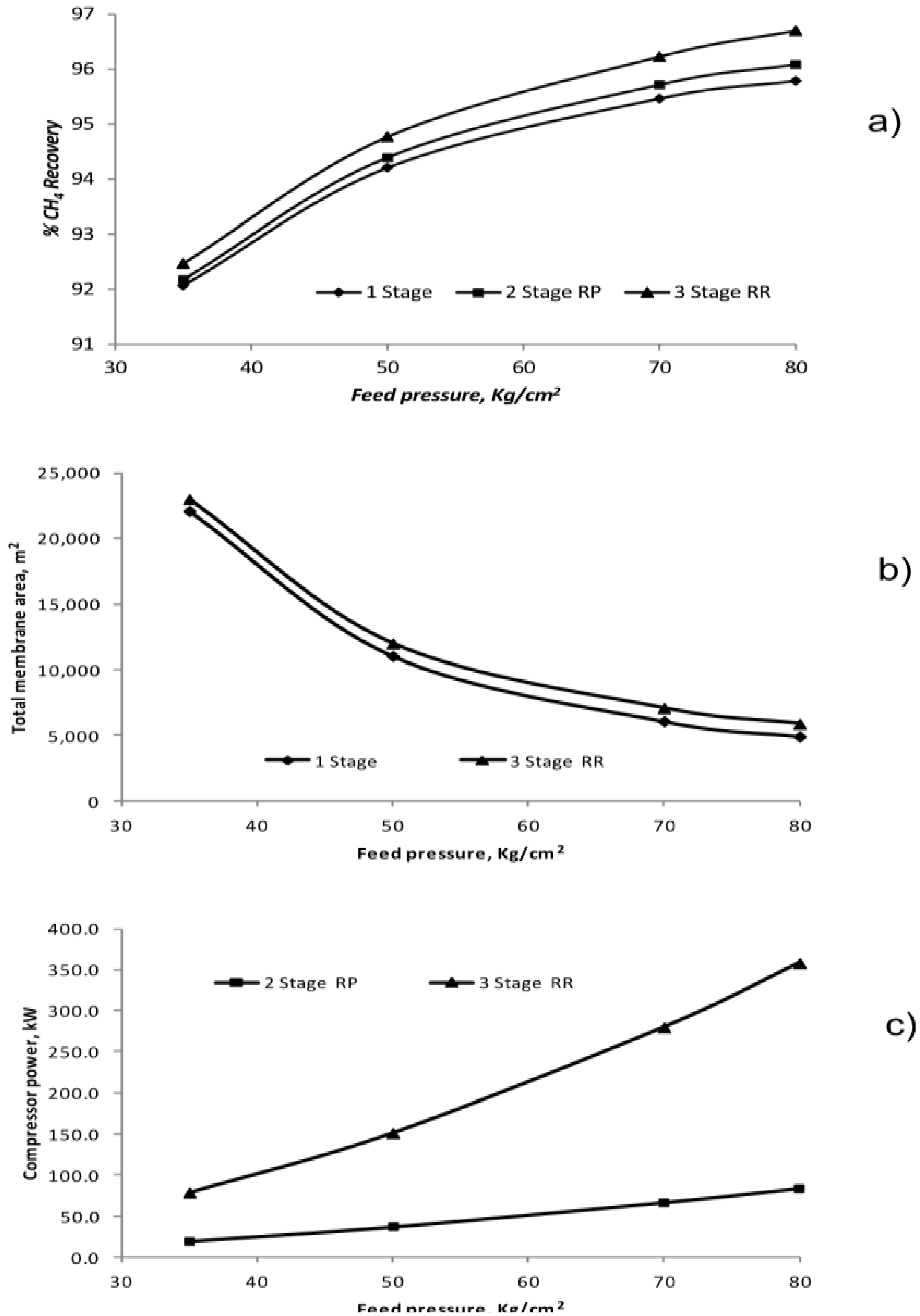


Fig. 4. Pressure effect on methane recovery (a), total membrane area (b), and compressor power (c).

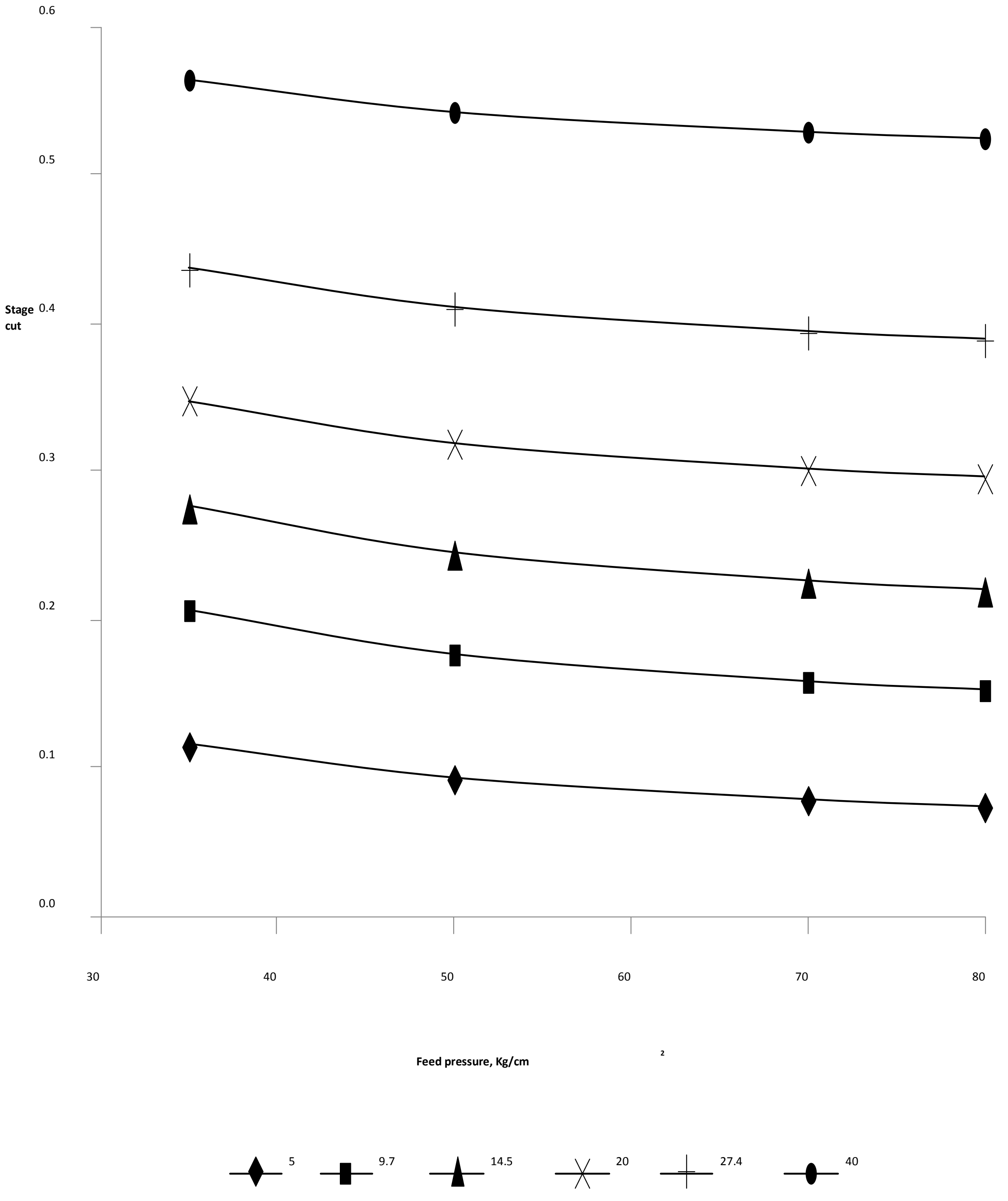


Fig. 5.Phase cut for triple design with retentate recycling, based on feed pressure and CO2 concentration (5-40 mol percent).

3.2 Feed pressure effect

Fig. 4 shows the influence of feed pressure for a feed moisture of 5% CO2 (run 1, Table 1) on various configurations. According to the referenced research, a triple-stage system yields a better methane recovery rate than other configurations with similar membrane areas; nonetheless, large compressor power investment is required, requiring an increase in plant expenditure. Because a 3-stage RR requires a small target area (Figs. 5 and 6), feed pressure has little effect on stage cut and acid gas removal (Table 5). A study by Madaeni demonstrated that feed pressure had no influence on CO2 purity in the permeate stream (Madaeni et al., 2010). As shown in Table 6, high relative gas mechanisms costs result from high overall plant investment needs. When seen in Fig. 8, GPC decreases as feed pressure increases (Khalilpour et al., 2013).

3.3 Feed flow effect

Variations in feed flow indicate modest varies in power needs (not illustrated), since the compression flow variations very little. With feed flows of 20-80 MMSCFD, small reductions of 1.2 percent in CH4 recovery and 0.012 in stage cut were recorded (Fig. 7). Though membrane area and compression power increased, GPCs compared to singlestage arrangement decreased (Table 7). This is due to the decrease in relative total plant input (RTPI), which is connected with increases in membrane area and compression power, but balanced by high flow capacity. These findings are in excellent accord with Bhide's 3 stage permeate recycling results (Bhide et al., 1998; Bhide and Stern, 1993).

In summary, raising feed pressures reduces GPCs, whereas decreasing CO2 concentration in the feed has a little effect. Fig. 8 shows the single stage gas mechanism expenses as a function of the analyzed variables. the pushing force across the membrane, decreasing the membrane area required (Ahmad et al., 2012), and therefore minimizing the costs of CH4 lost in retentate. Total plant investments grow in recycling setups because to higher compressor power needs, which influences the relative gas mechanisms cost (RGPC) (Table 6 and Fig. 9).

When comparing variations in feed pressure vs feed flow (Table 6 versus Table 7), opposite trends in RTPI, RVOM, and RCH4LS were identified (Figs. 4 and 7). Greater feed flow increases membrane area, power requirements, and permeate CH4 costs, but these changes are offset by increased amounts of mechanisms gas.

The GPC increases as the feed's CO2 concentration increases. As shown in Fig. 2, an increase in membrane size increases total plant investment; an increase in stage cut reduces CH4 recovery and boosts CH4LS. Except for RCH4LS, the relative components in Table 4 are lower as a function of feed composition, indicating that recycling setups are cheaper than single-stage ones. As illustrated in Fig. 10, improved CH4 recovery and reduced costs of methane lost in retentate are not necessarily the major criteria for choosing a design.

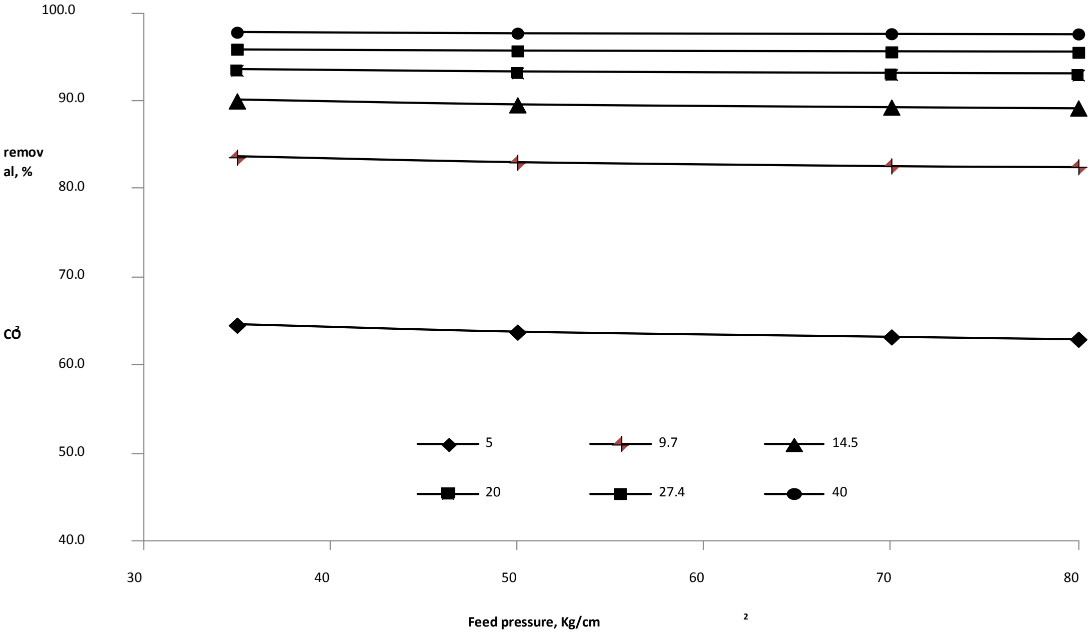


Fig. 6. CO2 removals for triple structure with retentate recycling, vs. feed pressure (5-40 mol percent).

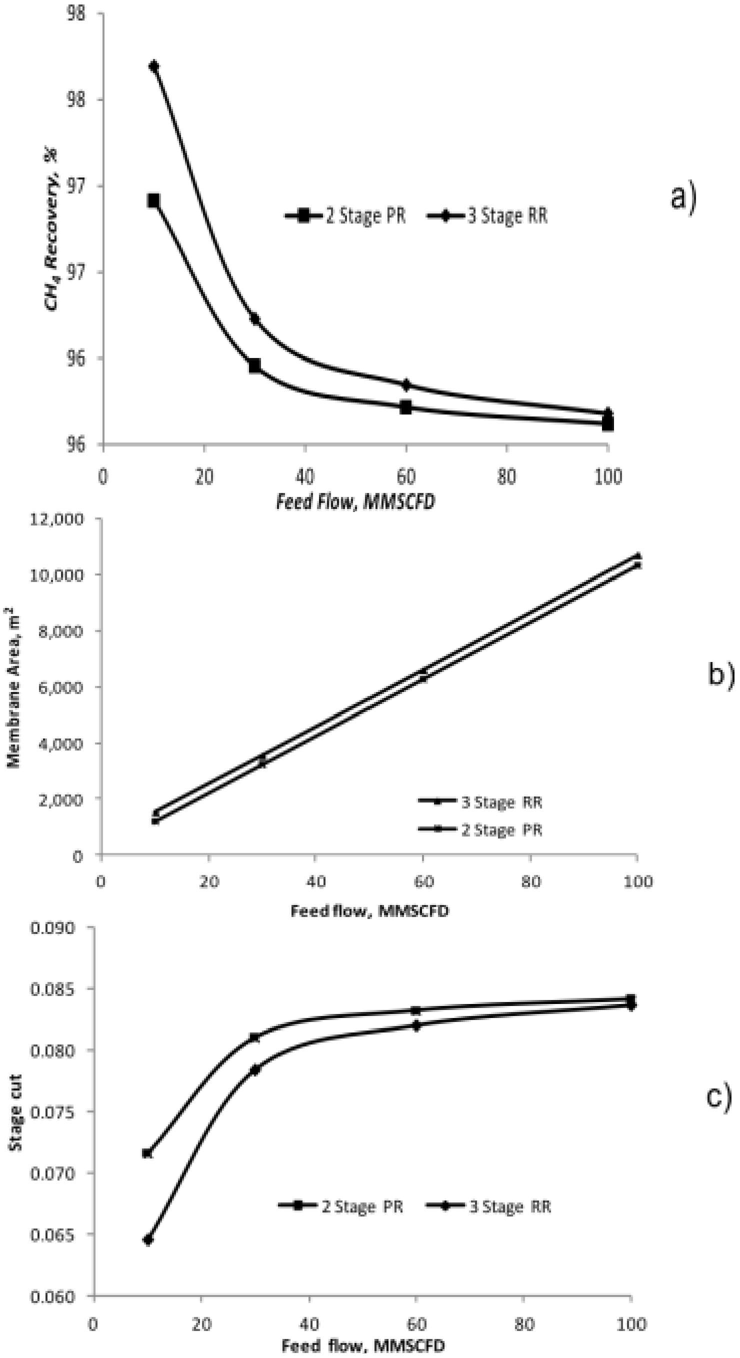


Fig. 7. feed flow effect on: a) CH4 recovery; b) membrane area; c) stage cut.

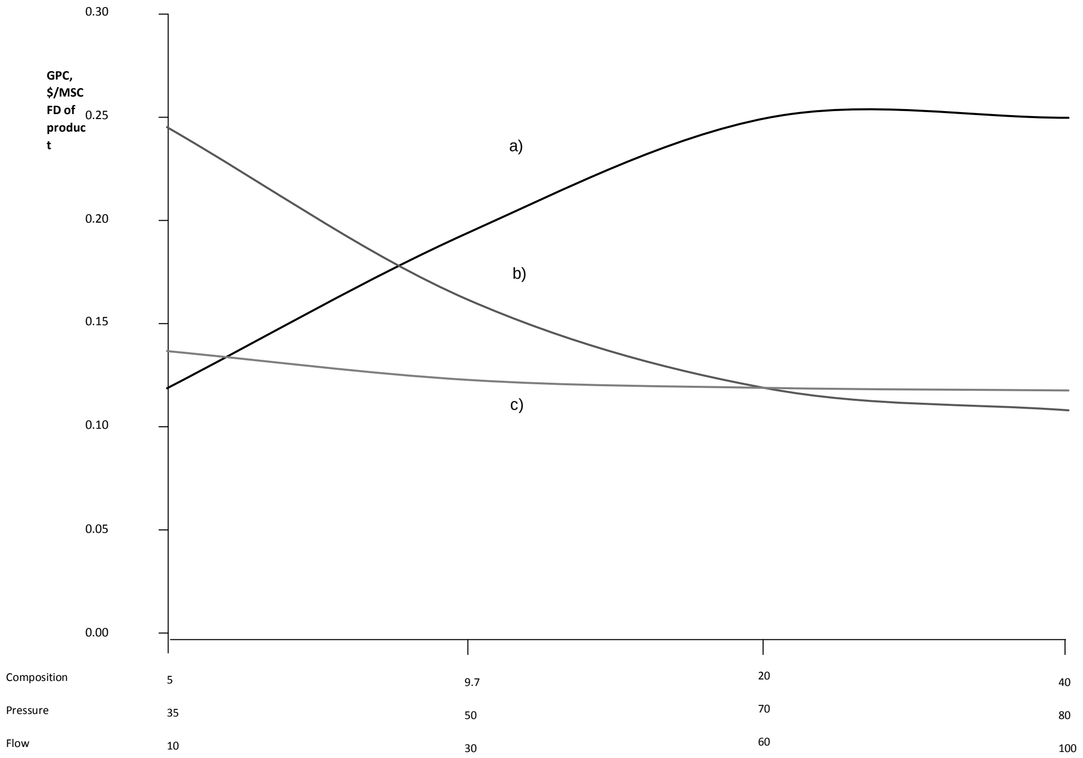
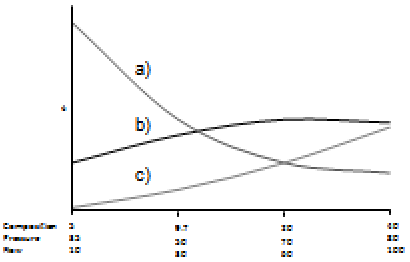
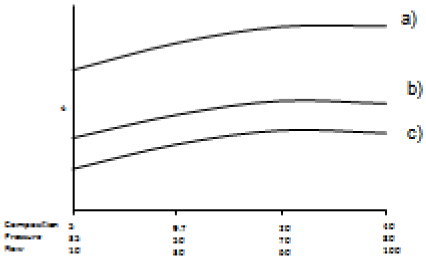


Fig. 8. GPC as a function of: a) feed composition, b) feed pressure and c) feed flow.

Table 4.Relative costs by feed composition

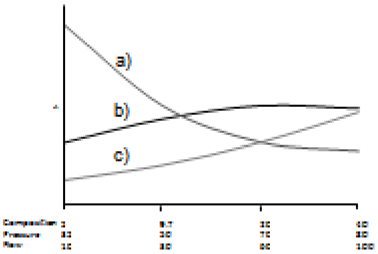
Run	Relative Total Plant Investment (RTPI)	Relative Annual Variable Annual Cost of Operating and Maintenance Cost (RVOM)	Relative CH_4 Lost in Permeate (RCH$_4$LS)	Relative Gas Mechanisms Cost (RGPC)
		Double stage with permeate recycle		
1	1.65	1.23	0.94	1.03
2	1.41	1.16	0.96	1.01
3	1.36	1.14	0.96	1.00
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

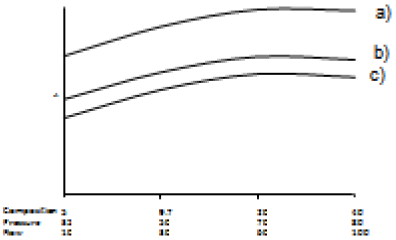




TPI

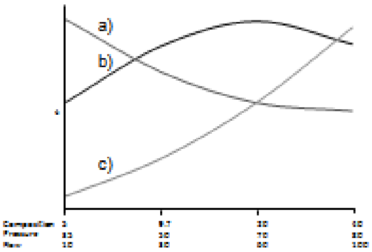
TPI



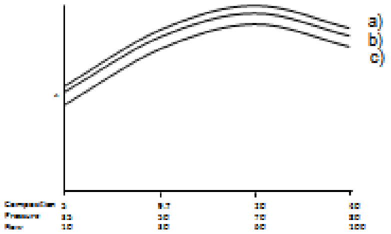


VOM

VOM



CH_4LS



CH_4LS

Fig. 9. Trends of TPI, VOM and CH9. Trends of TPI, VOM and CH LS as a function of4LS as a function: a) feed composition Fig. 10. Trends of TPI, VOM and CH4LS for: a) single of: a) feed composition, b) feed pressure and c) feed stage, b) double stage with permeate recycle and c)flow.triple stage with retentate recycle

Table 5. Acid gas removal efficiency vs feed pressure for 40 mol% CO2 in feed.

Feed pressure	**CO_2 removal**	**H_2S in product**	**H_2S removal**
Kg/cm^2	(%)	(%)	(%)
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.Costs relative to 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$_4$LS)**	**Relative Gas Mechanisms 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

Table7. Costs relative to feed pressure

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 Mechanisms 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

IV. CONCLUSIONS

Low gas mechanism costs can be achieved with low CO2 supply and high feed pressures and flows. The positive impact of pressure on GPC reduces membrane area requirements, increases methane recovery, and hence lowers permeate costs. However, it increases compressor power requirements and total plant investments.

Raising the feed flow increases the membrane area and compression power, but lowers the cost of CH4 lost in permeate, increases CH4 recovery, and increases the volume of mechanisms gas.

The feed composition affects the relative components of RTPI (relatively total plant input), RVOM (relative annual variation operating cost), and RCH4LS (relative annual cost of CH4 lost in permeate).

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