

Parametric Study of Single and Double Stage Membrane Configuration in Methane Enrichment Process

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Abstract— Operational study of a biogas upgrading plant with cleaning and methane (CH₄) enrichment has been presented in this study. Parametric study was conducted to investigate the effect of variation of process conditions for single stage without recycle (SSWR) and double stage with permeate recycle (DSPR) on product purity, CH₄ recovery and compression power requirement. In the study, achieving high CH₄ recovery and product purity simultaneously could not be attained in SSWR configuration. The performance of DSPR yielded a better result but with higher membrane area and compression power. DSPR configuration achieved high CH₄ recovery and purity at increasing feed pressure, selectivity and feed flow. The CH₄ losses increased in both configurations as %CO₂ increased in the feed. DSPR configuration is considered the best configuration due to the end use of the product, as vehicular fuel, which requires high product purity.

Keywords— Biogas, methane enrichment, parametric study, membrane configuration

I. INTRODUCTION

To reduce greenhouse gas emission and secure long-term energy supply globally, our exploitation of the earth's finite resources such as fossil fuel must be reduced, while renewable energy must be developed as alternatives. The proportion of energy generated from renewable resources is expected to increase to >20% by 2020 [1]. During the same period, the greenhouse gas should decrease by 14% in 2020 compared with 2005 [1]. The use of biogas is considered as one of the most efficient means of utilizing the renewable energy and reducing greenhouse gas emission. Biogas generated from the anaerobic decomposition of organic matter comprises mainly methane (CH₄), carbon dioxide (CO₂), smaller traces of acidic gases and impurities such as hydrogen sulphide (H₂S), nitrogen (N₂), water vapour (H₂O) and traces of other volatile organic gases (VOCs) [2]. The precise concentration of these gases in any particular biogas sample depends on the source of substrate and operating process conditions, however, the reported range is typically 40-70% vol. of CH₄ while CO₂ and other trace gases takes

up the remaining percentage volume [2]. Upgrading biogas to fuel grade biomethane involves two major processes; cleaning and CH₄ enrichment. The cleaning of the biogas consists of removal of acidic gases and impurities, while the enrichment process is for separation of CO₂ from biogas. The removal of trace impurities and the elimination of CO₂ from the biogas, which leads to a corresponding increase in CH₄ concentration will not only increase the calorific value of the biogas, but will also reduce corrosion of internal combustion engine parts caused by acid gas components. After upgrading, the final product is referred to as biomethane, typically containing 95-99% CH₄ and 1-3% CO₂ which makes it suitable as vehicular fuel. Current technologies for cleaning of biogas and its subsequent CH₄ enrichment are physiochemical processes which can be grouped as follows [3].

- Absorption process (physical and chemical absorption)
- Hybrid solution (mixed physical and chemical solvent)
- Physical separation (adsorption on solid surface; membrane; cryogenic)

Each of the cleaning and enrichment technologies, except for cryogenic separation processes, are in operation on large scale and can deliver biomethane that meets vehicular fuel standard [4]. The use of membranes, which is the main technology of interest in this study, exhibit many advantages in comparison to other conventional method of biogas upgrading, including lower operation cost, easier maintenance and greater process flexibility, no need for chemicals, the absence of phase and temperature changes, which results in low energy requirement [2], [5], [6]. Despite these advantages, it is difficult for membrane permeation systems to compete with the more common absorption based processes such as amine scrubbing, especially when processing high volume of biogas [7]. Higher CH₄ losses generated by membrane systems increases gas processing cost [7]. However, the CH₄ lost during the upgrading process of biogas obtained from anaerobic digesters, can be used as fuel for heat generation since anaerobic digestion typically requires higher than ambient temperature for optimal operation [7].

A membrane separation process is used to split a feed mixture containing two or more species through a semi-permeable barrier where the species that moves through the barrier is called permeate and the other that does not move through is called retentate [8]. Gases can be separated on two types of membrane; dense membrane (non-porous) and porous membrane [9]. The transportation of gases through dense membrane occurs via solution diffusion while porous membrane occurs via Knudsen flow, selective adsorption/diffusion and molecular sieving [9], [10]. The

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transportation of gases through membrane takes place when a driving force is applied to the gaseous species. This driving force is mostly due to pressure difference or concentration difference across the membrane [9]. Various membrane materials are able to separate CO₂ & CH₄ and both polymeric as well as inorganic materials can be used. However, in industrial scale gas separation, only polymeric membrane materials are applied due to their low manufacturing cost when compared to inorganic materials [11]. The efficiency of a given membrane is determined by two parameters; its selectivity and the flow through the membrane [9]. The latter often denoted as the flux or permeation rate, is defined as the volume flowing through the membrane per unit area and time [9]. In other literature, the flow through the membrane has been further divided into pressure ratio, which is the ratio of feed pressure to permeate pressure across the membrane, and stage cut, which is the fraction of the feed gas that permeates the membrane [12]. The selectivity of two gases 'a' and 'b' in a membrane is defined as the permeability of gas 'a' divided by permeability of gas 'b' which coefficient, sorption or diffusion depends on the type of material that is used in the membrane [13]. In order to apply membranes on an industrial scale, large membrane area are normally required [14]. Membranes are grouped together to form a bundle, since the permeability of a single membrane is not so high [14]. The bundle of membrane is installed into a pipe called module which is the smallest practical unit of membrane area [9]. Three types of module exist; hollow fiber modules; spiral wound modules & envelope type module [11]. Hollow fiber is commonly used in industry due its high packing density, low investment and operating cost [9], [11]. However, pre-treatment process is always required when hollow fiber is used because it is very susceptible to fouling and difficult to clean [9]. Fig. 1 shows a schematic diagram of a hollow fiber membrane [14].

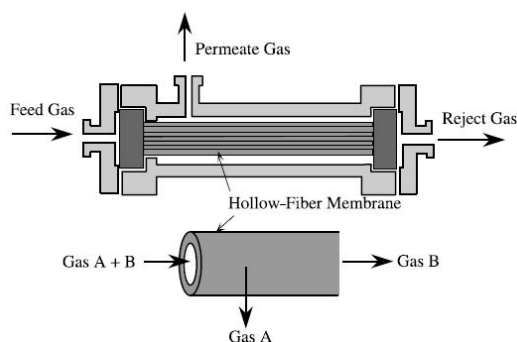


Fig. 1 Schematic diagram of a hollow fiber membrane

The arrangement of modules for separation to achieve a desired product purity and recovery of feed specie is called stage(s) [9]. This arrangement is based on economic consideration and end use of the product [9]. On economic consideration, three important elements are considered; the cost of membrane plant (membrane element and pressure housing); the capital and operating cost; and product losses [15]. The quality of the product depends on the end use. Critical operating parameters that affects the quality of upgraded biogas and CH₄ recovery in using membrane are the operating pressure, the raw biogas feed flow rate which is a function of the plant capacity, and composition of the

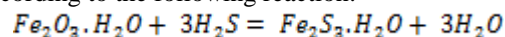
feed [1]. Membrane configuration has been reported to be of single stage without recycle, multistage single stage and multistage configurations [3], [9], [12]. A single membrane module or a number of such modules arranged in parallel or in series without recycle with a single pass flow constitute a single stage membrane separation process [16]. The extent to which a feed mixture can be separated in the single stage is limited. Thus, in order to achieve a higher degree of separation, cascades of membrane modules with recycle are often used. Multistage configuration of two stages or more has been reported to yield higher CH₄ recovery and product purity though with higher investment cost due to additional membrane area and compression cost [15].

The usual target of an enrichment process using membrane is to produce a retentate stream essentially stripped of CO₂ and with a low flowrate but highly CO₂ rich permeate stream. These two requirements cannot be met simultaneously; a trade-off must be made between CH₄ recovery and purity. Therefore, this work presents a simulation of two possible membrane configurations of an existing operational biogas upgrading plant that uses hollow fiber membrane module for CH₄ enrichment with descriptive detail of the cleaning process. The simulation was conducted to investigate the performance of the configurations on product purity, product recovery and compressor power requirement. A parametric study investigating the effects of variations of feed composition, operating pressure and feed flow rate was conducted, with a view to investigate the performance of the configurations from an operation point of view.

II. DESCRIPTION OF PROCESS PLANT

A. Biogas Cleaning Process

The raw biogas used in the upgrading process was generated from a closed landfill at Sebenza in the Ekurhuleni Municipality of South Africa. Pipes were bored into the landfill for extraction of the biogas generated by the natural decomposition of the organic fraction of waste. The biogas composition is as follows; CO₂ 38.9%; CH₄ 55%; O₂ 0.4%; N₂ 5%; H₂S 0.002% and H₂O 0.66%. The biogas stream was made to go through a vessel packed with high efficient iron-III-oxide (Fe₂O₃) adsorbent for preliminary desulphurization according to the following reaction:



The feed is at atmospheric pressure and operating temperature of 27°C. A blower was installed to increase the pressure to about 1.4bar. The temperature of the vessel was kept low to ensure the Fe₂O₃ was not dehydrated, which will reduce its reactivity. Similar recommendation for Fe₂O₃ was also given by Wang [17]. Fe₂O₃ is not regenerated in the plant, and has to be replaced once saturated with elemental sulphur. The resulting biogas stream, which was now almost completely free of H₂S, was then compressed to about 12 bar in a two stage compressor and stored temporarily in a vessel to ensure a uniform flow rate. Thereafter, the biogas was cooled to about 40°C with a heat exchange using cooling water. Possible particles and oil content in the biogas were removed by passing it through a mechanical filter and an oil separator because the compressor was oil

based for lubrication. A pressure swing adsorption (PSA) with activated alumina (Al_2O_3) was used for completed dehydration of the biogas. The N_2 , siloxane, possible H_2S and other traces of volatile organic gases (VOCs) were adsorbed using an activated carbon filter. All unit operations up to this point were considered part of the cleaning process. The resulting biogas has about <5% vol. N_2 , 1ppm H_2S and ppb levels for VOCs and siloxanes which is deemed not harmful to the membranes in the subsequent enrichment process.

B. Methane Enrichment

The most important task required to achieve fuel grade biomethane is the separation of CO_2 from the product gas stream. The plant studied used hollow fiber membrane for the separation of CO_2 from the biogas to achieve the desired fuel grade biomethane. The first membrane module was fed from the bore side at 10.311bar & $36^\circ C$. The permeate stream of the first membrane module which leaves at about 1.082bar was used to purge the PSA unit. The retentate stream leaves the membrane at 8.82bar and a slight temperature drop, with a high concentration of methane. This serves as the feed stream for the second stage separation. A sampling point is available for each membrane module. Gas species of the product stream can be randomly evaluated and the composition of gases through the membrane can be known. The permeate stream of the second membrane module, which contains significantly higher amounts of CH_4 as compared to permeate of the first stage, was recycled back for recompression through the suction of the compressor. Biomethane produce during the second stage separation is the desired product which is stored into a vessel after which it was odourised with mercaptan as a safety procedure and for leak detection. The last stage is compression of the biomethane to 220bar into high pressure cylinders.

III. METHODOLOGY

A. Process Simulation

The simulation of the upgrading process was done using ChemCAD, a steady state process simulator, from Chemstation. The PSA unit, a dynamic and unsteady state process, was represented with a component separator unit operation and was constrained to plant result for H_2O and H_2S removal. The cleaning process according plant data effectively eliminated vapour and H_2S content in the biogas while N_2 was reduced to acceptable limit; hence in the enrichment process simulated, CO_2 and CH_4 were only considered.

B. Data Collection

The process operating conditions were data collected from the plant. The membrane intrinsic properties (selectivity and permeability), configurations and module design data were not shared due to intellectual property infringement and confidentiality concerns. Intrinsic properties were obtained from literature while the module design data was provided by Almeesoft Engineering, USA. The module design data were verified by the plant manager

and was accepted to be inconformity with industrial standard.

C. Membrane Module Design and Simulation Validation

A built-in membrane model for gas permeation available in ChemCAD 6.5.5 was used for the CH_4 enrichment. Almeesoft Engineering gas permeation software was also interface with ChemCAD during sensitivity analysis to study the effect of varying the process conditions on CH_4 recovery and product purity. The important process parameter specified for the simulation of the membrane is given in TABLE I and represents base condition for the simulation.

TABLE I
 PROCESS PARAMETER FOR MEMBRANE SIMULATION

Parameter	
Feed pressure	9.9bar
Permeate pressure	1.08bar
Feed flow rate	80m ³ /h
Flow mode	Counter-current
Feed side	bore
Temperature	30°C
Compressor stages	2
Cooling medium	water
CO_2 permeability ¹	440 Barrer
CH_4 permeability ¹	28.2 Barrer
N_2 permeability ¹	35.6 Barrer
O_2 permeability ¹	111 Barrer
CO_2/CH_4 selectivity	15.6
CO_2/N_2 selectivity	12.4
CH_4/N_2 selectivity	0.79
CO_2/O_2 selectivity	3.96
Account for shell pressure drop	yes

¹Harasimowicz, et al. [18]

Two membrane configurations was considered for this study; single stage without recycle (SSWR) and double stage with permeate recycle (DSPR) as shown in Fig. 2. The configurations were simulated to investigate the performance of each configuration to varying operation parameters. The following assumptions were considered for the hollow fiber membrane module for the study. (1) Transport properties of the membrane are independent of variation in gas composition throughout the separation. (2) No flux coupling between gas components. (3) Deformation of the hollow fiber under pressure is negligible. Based on these assumptions, sensitivity analysis was carried out on both configurations and the results were discussed.

The membrane data for each module was computed based on the module specification provided by the end user into the simulator. Parameters considered in the membrane module design are as listed in the TABLE II. The process simulation for the cleaning and CH_4 enrichment process was validated by the plant upgrading process result.

TABLE II
 MEMBRANE MODULE DESIGN PARAMETERS

Properties		units
Fiber internal diameter	220.98	μm
Fiber outer diameter	370.84	μm
No. of fibers	10,000	
Active length	1.1	m
Membrane area per module	11.6293	m ²
Pot length	0.0509	mm
No. of shell	1	
CO_2 viscosity	0.0153	c_p
Packing porosity	73.8%	%
Module housing	Stainless steel	

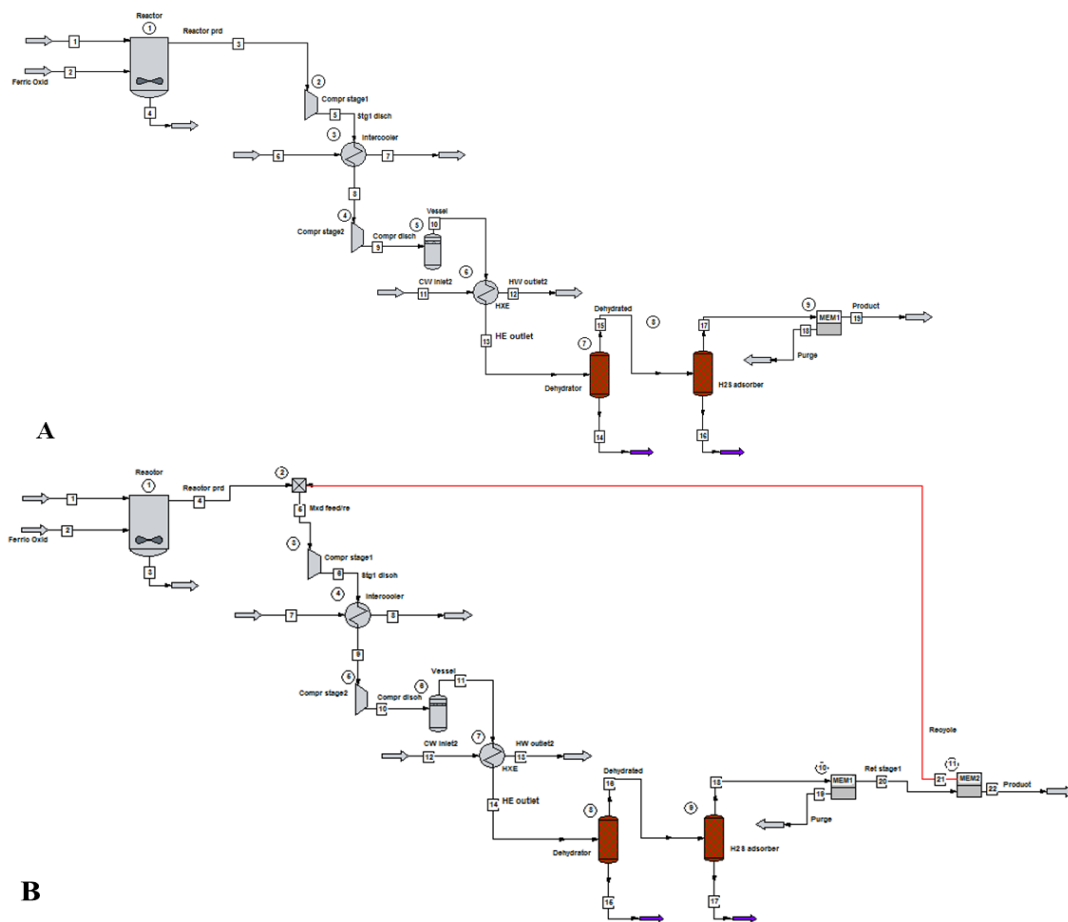


Fig. 2. Schematic diagram of the membrane configuration having (A) Single stage without recycle (SSWR), (B) Double stage with permeate recycle (DSPR).

IV. RESULT AND DISCUSSION

A. Process Simulation Validation

TABLE III shows the results of the simulation against those of plant upgrading. The plant had 91.16% CH₄ recovery with 87.2% CH₄ purity. The two proposed configuration for this analysis was simulated to determine their performance against plant result. SSWR yielded 87.47% CH₄ recovery with 77.89% CH₄ purity while DSPR

yielded 89.01% CH₄ recovery with 84.59% CH₄ purity. The simulated DSPR process suggests a good approximation to the plant data with a percentage error of 2.64%. The error in the comparison could be attributed to higher CO₂/CH₄ selectivity of the plant membrane and feed pressure drop within the membrane module. Sensitivity study was carried out on both configurations to study the effect of varying process operating condition on CH₄ recovery, purity and compressor power requirement.

TABLE III
 VALIDATION OF THE PROCESS SIMULATION WITH PLANT DATA

	Plant result		Simulated result									
	Final stage		SSWR		DSPR		Initial stage		Final stage			
	Feed	Permeate	Retentate (Sale gas)	Permeate	Retentate	Mixed feed with recycle	Compr dischar ge/cleaned Feed 1	Permeate	Retentate/feed 2	Permeate to recycle	Retentate (product)	
Flow (m ³ /hr)	80.000	23.000	46.000	30.060	49.410	98.550	98.030	33.210	64.820	18.560	46.300	
Pressure (bar)	1.000	1.082	8.280	1.080	9.400	1.000	9.900	1.500	9.230	0.800	8.769	
Av. Mol. Wt.	27.606	31.847	18.384	38.501	21.079	28.890	28.949	39.594	23.496	34.426	19.111	
CH ₄ (% mol.)	55.000	40.120	87.200	18.351	77.892	50.660	50.934	14.657	69.518	31.947	84.590	
CO ₂ (% mol.)	38.934	53.997	4.900	79.201	14.854	43.626	43.862	83.265	23.677	63.843	7.563	
N ₂ (% mol.)	5.000	4.543	7.693	2.049	6.849	4.730	4.756	1.685	6.329	3.566	7.437	
O ₂ (% mol.)	0.400	0.964	0.265	0.399	0.405	0.446	0.448	0.393	0.476	0.643	0.410	
H ₂ S (% mol.)	0.002	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	
H ₂ O (% mol.)	0.664	0.372	0.005	0.000	0.000	0.537	0.000	0.000	0.000	0.000	0.000	
Heat values (BTU/SCF)												
HHV	967.4	202.9	882.0							130.4	861.2	
LLV	870.9	182.7	794.0							117.3	775.3	
% CH₄ Recovery			91.16		87.47						89.01	

B. Methane Recovery and Product Purity

1) Effect of %CO₂ in Feed

Figs. 3a, 4a and 5a show the effect of increased CO₂ content in feed on CH₄ recovery, purity of product and membrane area required for separation on both configurations respectively. 89.8% CH₄ recovery was achieved with about 10% CO₂ in the feed stream in the SSWR configuration requiring 25.7m² of membrane area to achieve 89.85% CH₄ in the product. In DSPR configuration, CH₄ recovery was 90.48% with a product purity of 90.69% requiring 51.75m² membrane area for 10% CO₂ in the feed. As the CO₂ concentration increased, CH₄ recovery reduces as well as purity of product but with an increase in membrane area. With 60% CO₂ in the feed, CH₄ recovery has reduced to 84.19% with a 70.54% CH₄ in the product while for DSPR, 86.8% CH₄ recovery was achieved with 82% CH₄ in the product. Membrane area increased to 27m² for SSWR while DSPR required 53.58m². DSPR achieved higher CH₄ recovery and purity over SSWR because more membrane area was available for separation. Also the recycling of permeate of the second stage membrane separation ensures that less CH₄ is lost in DSPR configuration.

2) Effect of Feed Pressure

Figs. 3b, 4b and 5b show the effect of feed pressure on CH₄ recovery for both configurations, product purity and membrane area required respectively. The process condition and membrane shell data remains the same as earlier described except for the feed pressure been varied. It was

observed as shown in Fig. 3b that increasing the feed pressure increased CH₄ recovery until a limit is reached where further increase in the pressure does not improve the CH₄ recovery rate. As the feed pressure increases from 10bar to 30bar, the CH₄ recovery increases rapidly from 93.5% to 94.96% for SSWR and thereafter the increase in pressure causes a gradual reduction in CH₄ recovery to 94.84% at 50bar. DSPR also achieved an increase in CH₄ recovery from 90.4% to 92.7% as feed pressure increased from 10-30bar. Further increase in pressure only increased the CH₄ recovery by 0.2%. CH₄ recovery was higher for SSWR when compared to DSPR because after the first stage separation, the feed pressure to the second stage which is the driving force reduces and causes a reduction in separation efficiency of the second stage as compared to the first stage. But despite a lower CH₄ recovery, the product purity achieved in the DSPR, 83.75% to 86.01%, when compared to SSWR, 77.6% to 78.8%, as shown in Fig. 3b, was much higher. At lower pressure, the high packing density advantage of hollow fiber membrane becomes increasingly important due to the high membrane area available hence higher purity can still be achieved. The increase in CH₄ recovery and purity of product can also be attributed to the fact that the increased pressure creates a greater driving force across the membrane surface area while the almost constant CH₄ recovery after its peak, in this case 30bar is due to the membrane module design.

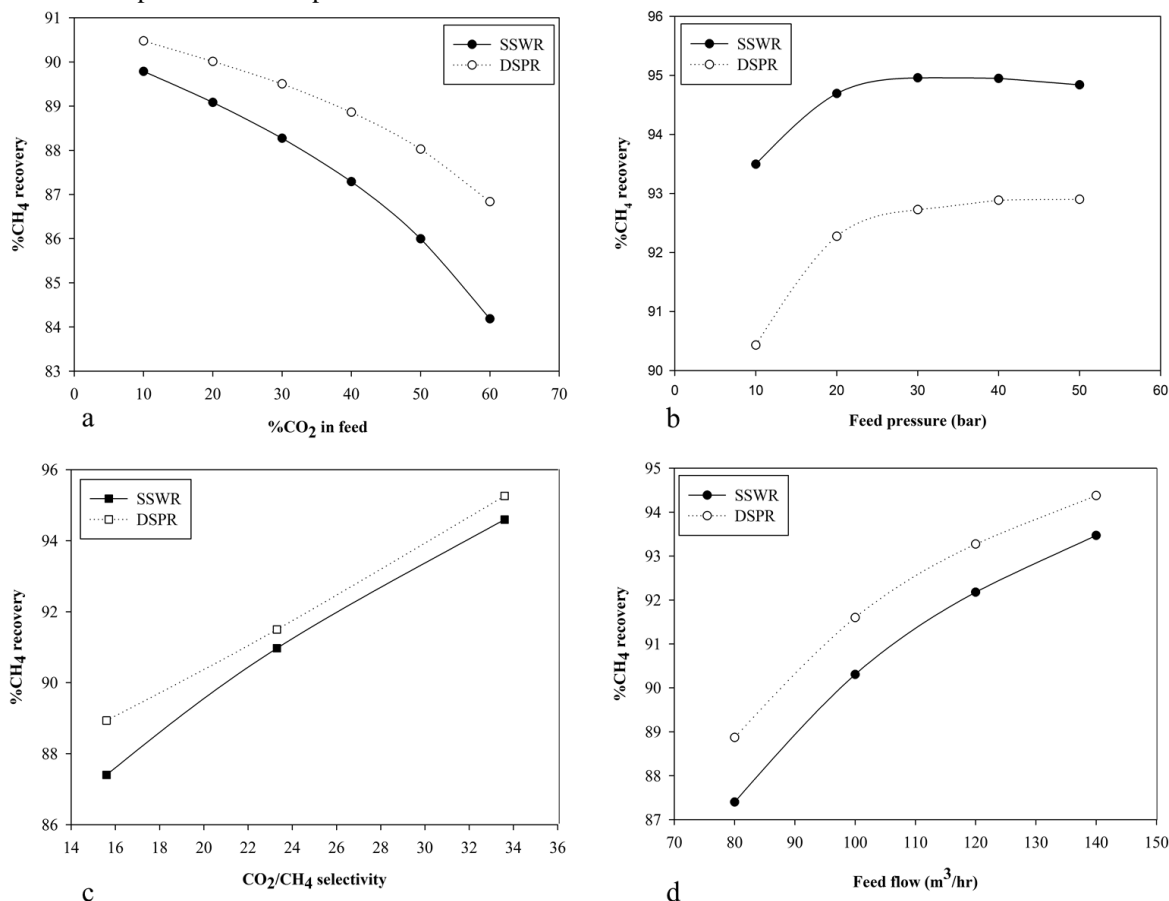


Fig. 3. Effect of variation of process parameters on CH₄ recovery

Fig. 3. (a) Effect of %CO₂ in feed, (b) Effect of feed pressure, (c) Effect of membrane selectivity, (d) Effect of feed flow rate

Improving the overall structural design will increase the membrane performance with variation in process condition [7]. This suggests that achieving high CH₄ recovery does not only depend on improving the process condition but also ensuring that an appropriate equipment design is implemented to accommodate changes in process condition. Also a higher CH₄ recovery was achieved and the required membrane area decreased as the feed pressure increases as shown in Fig. 5b. As the feed pressure increases from 10-30bar, the membrane area for SSWR decreases from 26m² to 9m² after which increasing feed pressure to 50bar only causes a reduction to 6.7m². DSPR uses higher membrane area for separation which ensures high product purity is achieved as feed pressure increases until it reaches its peak. The behaviour of DSPR is also similar to SSWR, as feed pressure increase to 30bar, there as a significant reduction in membrane area required from 57.6m² to 19.6m², after which an increase in pressure resulted in fluctuation in membrane area. This fluctuation suggests that the module has reached its pressure limit with respect to module design. It may be economically justified in some circumstances to increase the compression pressure of the feed stream to reduce membrane area cost and improve recovery but the structural design of the membrane casing to safely accommodate such pressure must be considered. Hence a feed pressure of 18bar will be considered as an optimal and safe operating pressure for the DSPR based on the module design data provided.

3) Effect of Membrane Selectivity

Membrane intrinsic properties have high influence on CH₄ recovery which increases with an increase in the selectivity of the membrane [3]. Figs. 3c, 4c and 5c shows the effect of selectivity on CH₄ recovery, product purity and membrane area required respectively. An increase in the membrane selectivity from 15.8 to 33.3, while other process conditions were held constant, resulted in a spontaneous increase in the CH₄ recovery from 87.4% to 94.6% for SSWR while DSPR increased from 88.9% to 95.3% as shown in Fig. 3c. The product purity for both configuration increased as selectivity increased from 15.6 to 23.3, SSWR increased from 78.3% to 79.5% while DSPR increased from 84.6% to 85.3%. A reduction was observed as the selectivity increases to 33.6 with SSWR CH₄ purity in product reducing to 76.6% and DSPR to 82.7 as shown in Fig. 4c. The reduction could be attributed to selectivity properties of the membrane towards CO₂. If less CO₂ permeate through the membrane, the CO₂ concentration in the product stream increases hence the product purity decreases. Also it has been reported in literature that highly permeable polymers have low selectivity and vice versa [19]. Hence a general trade off relationship between gas permeability and selectivity is required.

Aside increase in CH₄ recovery, increasing the selectivity reduces the required membrane area for separation as shown in Fig. 5c. With 15.6 CO₂/CH₄ selectivity, membrane area required to achieve 87.4% CH₄ recovery was 26.38m² for SSWR and at 33.3 selectivity, 12.63m² was required to achieve 94.6%CH₄ recovery. For DSPR, 52.40m² was required to achieve 88.9% CH₄ recovery with 84.6% product purity and at 33.6 selectivity, 27.69m² was required to

achieve 95.3% CH₄ recovery. The decrease in membrane area is due to the intrinsic property of the membrane to permeate more CO₂ over the initial contact area of the membrane and the feed gas.

4) Feed Flow Rate

Figs. 3d and 4d shows the effect of increasing feed flow rate on CH₄ recovery and product purity. The plant been studied has the capacity to produce between 20-100m³/hr biomethane. The base feed flow rate been used for this study was 80m³/hr. Increasing the feed flow rate, CH₄ recovery increased from 87.4% and 88.9% at 80m³/hr to 93.5% and 94.4% at 140m³/hr feed flow for both SSWR and DSPR respectively as shown in Fig. 3d. The product flow rate peaked at 103m³/hr for SSWR and 100.5m³/hr for DSPR. This shows that if the plant is operating at full capacity, more CH₄ will be recovered in the process and CH₄ lost will be reduced. The product purity reduces as the feed flow rate increases as shown in Fig. 4d. CH₄ in product reduces from 78.27% to 69.24% for SSWR while DSPR had a reduction from 85.02% to 72.29%. It was also observed that as the feed flow rate increases from 80m³/hr to 140m³/hr, there was a slight increase in the membrane area. The membrane area increased from 26.38m² to 28.4m² for SSWR and from 53.2m² to 54.6m² for DSPR. The reduction in product purity as feed flow increases was due to insufficient membrane area for separation hence more CO₂ is not permeated and its presence in the retentate reduces the purity of the product.

C. Compressor Power Requirement

The effect of feed composition, feed pressure and feed flow rate on compression power required for both configurations has been investigated. Fig. 6a, 6b and 6c shows the effect of increased %CO₂ in feed, feed pressure and feed flow rate on compressor power required for separation respectively. As CO₂ increases in the feed stream from 10-60%, the compression power required to facilitate the transportation of CO₂ through the membrane surface increases from 8.89hp to 34.62hp for SSWR and 17.05hp to 53.39hp for DSPR as shown in Fig. 6a. The required power per pressure increase is the minimum driving force to ensure that CO₂ permeate through the membrane surface area available and the target CH₄ recovery was achieved. Increasing feed pressure, 10-50bar, increased the required compressor power from 22.33hp to 38.82hp for SSWR and 39.78hp to 69.68hp for DSPR as shown in Fig. 6b. As earlier discussed, increasing feed pressure reduces the required membrane area but the limit of the feed pressure depends on the membrane module design pressure capacity. Increasing feed flow rate increased the compressor power required for separation as show in Fig. 6c. As the feed flow rate increases from 80m³/hr to 140m³/hr, the compressor power required for SSWR increased slightly from 22.18hp to 25.22hp because there was no recycling of the permeate stream. %CO₂ in the recycled stream of DSPR increased from 60.62 to 79.18% which increased the compressor power from 37.73hp to 46.26hp. As earlier discussed, increasing CO₂ in the suction of the compressor increases the required power for membrane separation to achieve the desired CH₄ recovery.

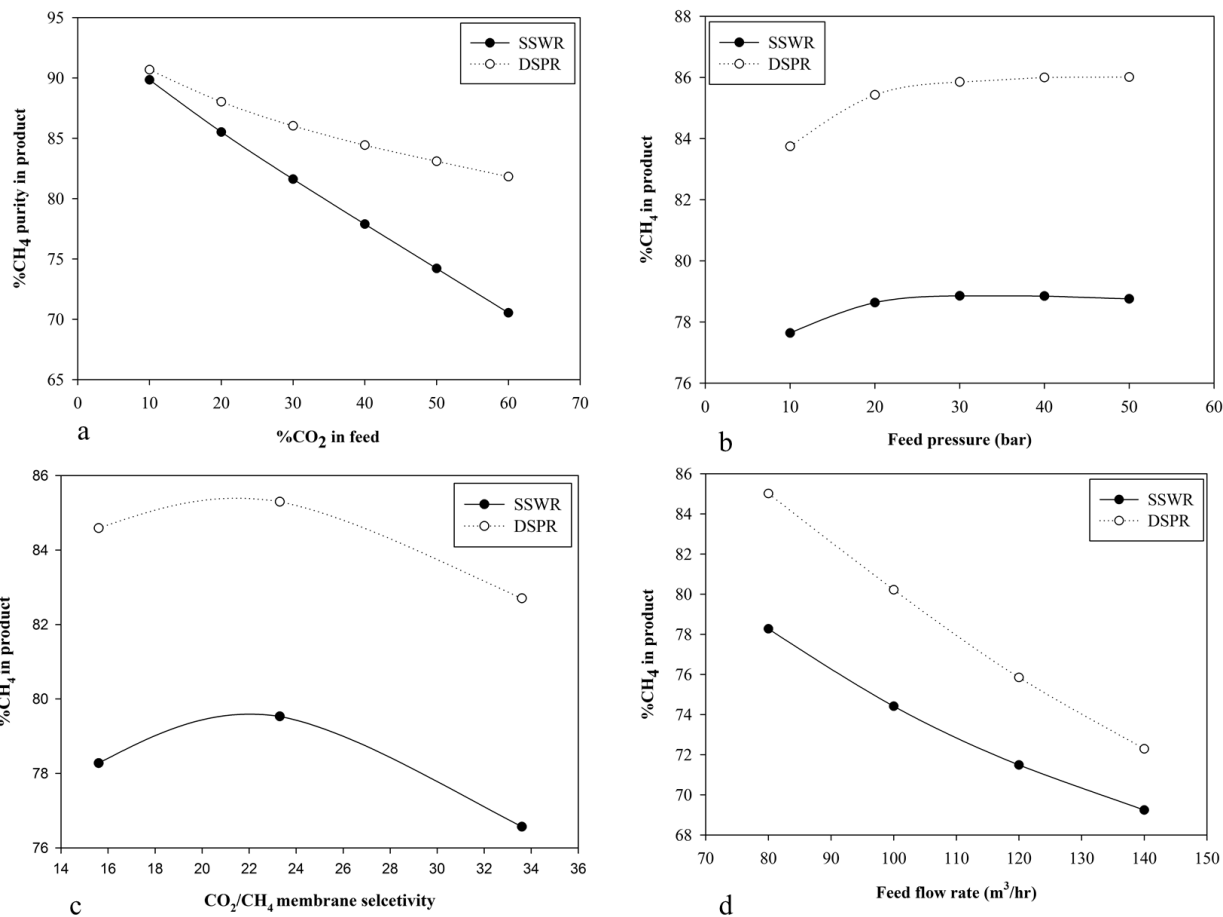


Fig. 4. Effect of varying process parameters on product purity (a) Effect of %CO₂ in feed; (b) Effect of feed pressure (c) Effect of membrane selectivity (d) Effect of feed flow

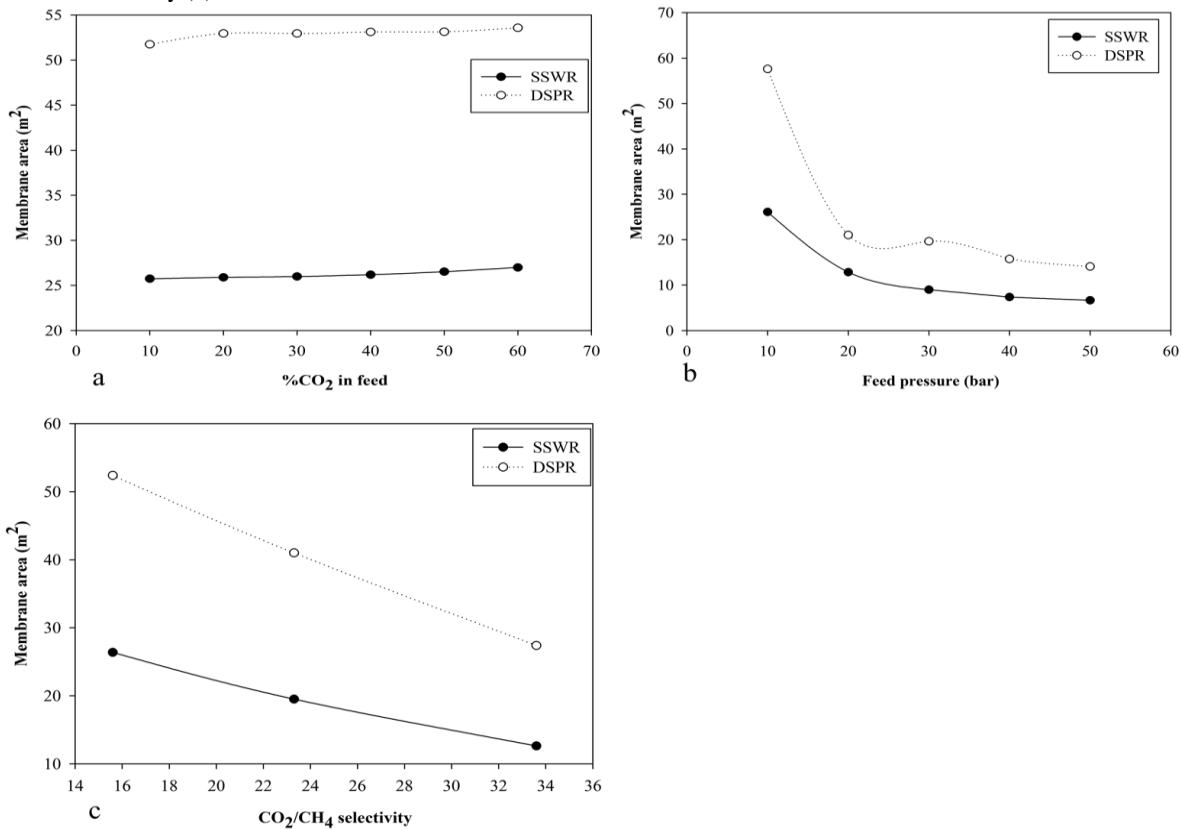


Fig. 5. Effect of variation of process variables on membrane area (a) Effect of %CO₂ in feed (b) Effect of feed pressure (c) Effect of membrane selectivity

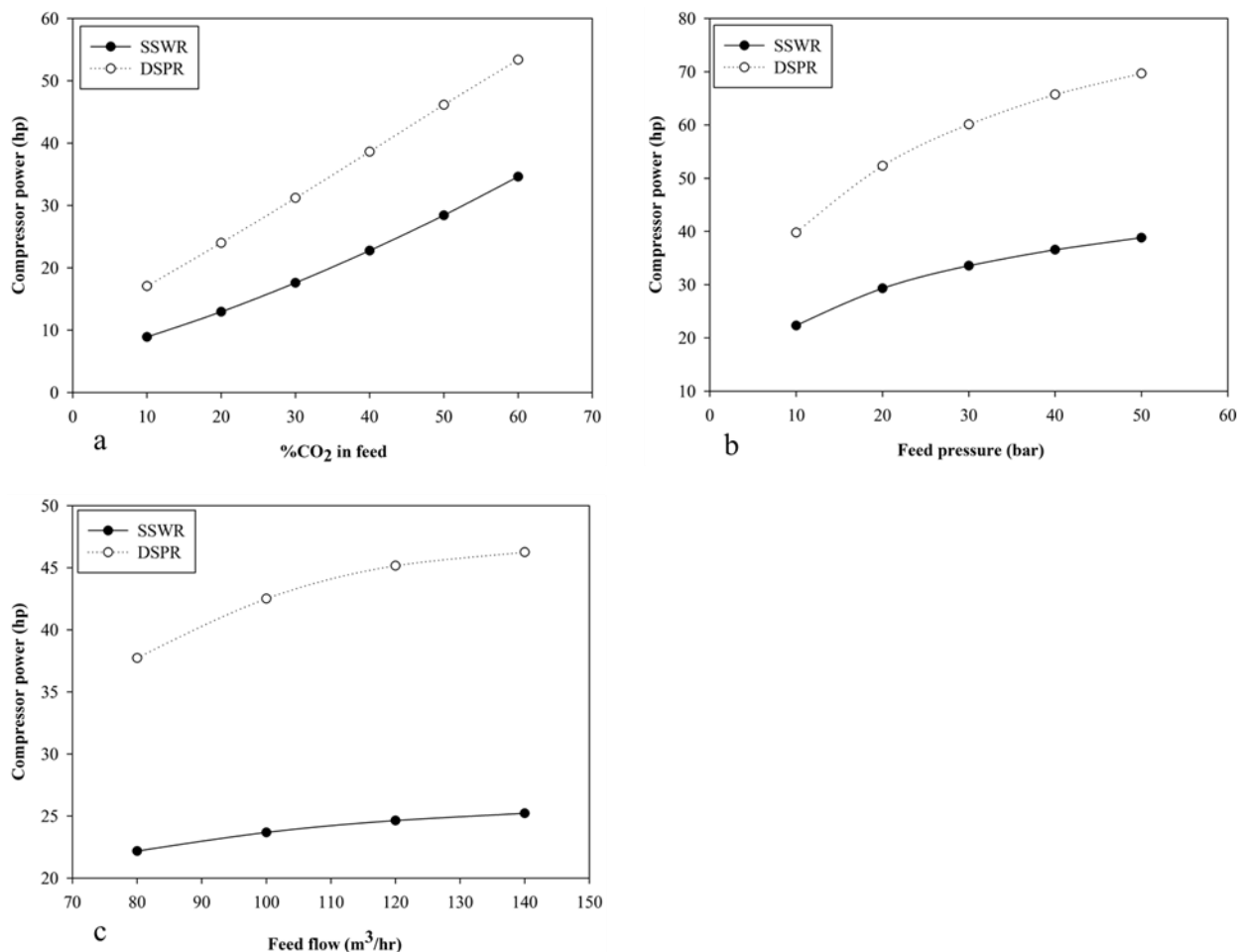


Fig. 6. Effect of variation of process parameters on compressor power (a) Effect of %CO₂ in feed (b) Effect of feed pressure (c) Effect of feed flow

The result of the parametric sensitivity analysis is consistent with previously published simulated work of Deng and Hagg [1], Ahmad, et al. [3] and Chmielewski, et al. [2] on membrane performance in methane enrichment process.

V. CONCLUSION

Parametric study of a biogas upgrading plant that considered cleaning of impurities in biogas and subsequent CH₄ enrichment using membrane has been presented. The simulated process was validated with an existing biogas upgrading plant result. The parametric study investigated the effect of varying feed composition, pressure and flow rate in enrichment process on CH₄ recovery on two configurations. Increasing CO₂ in feed reduces the CH₄ recovery by 6.23% for SSWR and 4% for DSPR. Product purity for SSWR reduced by 21.5% while DSPR product purity reduced by 9.7% as CO₂ increases in the feed. An increase in membrane selectivity from 5.68 to 33.33 increased CH₄ recovery by 8.2% for SSWR and a reduction in membrane area by 52.1%. Also, the increase in selectivity increased CH₄ recovery by 7.11% in DSPR and a reduction in membrane area by 47.73%. The product purity in DSPR was 8% higher than SSWR. Pressure increase also increased CH₄ recovery up to the module design configuration limit. Increasing feed flow rate from 80-140m³/hr increased CH₄ recovery by 6.98% for SSWR. Similarly, 6.2% increase in CH₄ recovery was achieved in DSPR with a product purity that is 4.4%

higher than SSWR. SSWR configuration requires less compression power and membrane area but CH₄ recovery and purity of the product stream was low. DSPR achieved higher CH₄ recovery and product purity at the expense of compressor power and membrane area. If the desired end use of the product requires high product purity and recovery like the plant under study, a DSPR is recommended though the investment cost may be high due to increased membrane area and compression cost. SSWR configuration will be suitable if product purity required is not high.

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