

Comparative Study of Membrane and Absorption Processes Performance and their Economic Evaluation for CO₂ Capturing from Flue Gas

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Abstract: As the main aim of this study, simulation and economic assessment of membrane technologies in comparison absorption process for CO₂ capturing from specified flue gas was conducted. For this purpose, the PRO/II v.10 software and Aspen Process Economic Analyzer v.10 were used. In this simulation, the flue gas flow rate is 8162 kmole/h and the concentration of CO₂ in flue gas is 8-22% mole. The objective function in the simulation of CO₂ capturing is to remove 85% of CO₂ from the flue gas stream. The amount of required solvent and membrane surface, as well as various costs such as equipment costs, installed costs, total capital cost, total utility cost and total operating cost for different concentrations of CO₂ in flue gas (8-22% mol.) was assessed for both membrane-based and absorption-based units. For CO₂ selectivity and permeability values of 28 and 1097 barrer, respectively, the total capital cost in the membrane-based process is very higher than the absorption process. So, the total capital cost of the membrane unit was about 2.3 times higher than the total capital cost of the absorption process. In a low concentration of CO₂, the total utility cost and total operating cost of the membrane-based process were about 2.2 times higher than the absorption process. However, by increasing the CO₂ concentration the difference between these costs in two processes decreased. By analyzing the selectivity effects on the total capital costs, it is obtained that with a selectivity value of 280 and the same permeability the costs of the membrane-based process became comparable to the costs of the absorption process.

keywords: CO₂ Capture, Membrane separation, Simulation, Economic Evaluation, Absorption method, PRO/II v.10

1. Introduction

Climate change is one of the greatest environmental, social and economic threats in the world. Europe Union has confirmed a 20% reduction of greenhouse gases released by 2020 (climate change and energy. 2018). Releasing high amounts of carbon dioxide is one of the main elements causing the greenhouse phenomenon and warming of the earth. By net production about 450 million tone carbon dioxide per year, Iran has the highest contribution for releasing this gas among Middle East countries (Mousavi, Lopez, & Blesl, 2017). About 5% of released carbon dioxide in the world is related to cement industry half of which is related to cement

production (converting limestone to refined lime) and the other half is related to the combustion process (Barker, Turner, & Davison, 2009). Sources of carbon dioxide emission are divided into combustion and non-combustion groups. Combustion sources include power production plants, refineries and also steel, cement and petrochemical industries. Non-combustion resources related to natural gas refineries and synthesis gas production sections in petrochemical industries. Table 1 suggests the volume of existing carbon dioxide in the flue gases released from different industries. Several methods are existed for separating carbon dioxide from the air. In this research, the

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economic assessment of the membrane method and the conventional absorption method are conducted. If the membrane method with existing characteristics is not cost-effective economically, it will be passed through simulator software and selectivity of the membrane is defined so that it could compete with the absorption method. If a membrane with these specifications is synthesized in the future, the membrane method will be definitely a superior method rather than the absorption method.

1.1. Various Methods for Carbon Dioxide Capturing

The separation process of CO₂ from flue gas flows is conducted through various methods. These methods are mainly categorized in 5 following groups (Ghasemzadeh, Jafari & Babalou, 2016):

1. Absorption
2. Adsorption
3. Cryogenic
4. Membrane
5. Hybrid process (membrane contactor)

Many elements are effective in choosing an appropriate process for refining acidic gases among which the most important ones include: mass or molar flow, temperature and inlet gas pressure, concentration of feed, ultimate specifications of refined gas, process economy and environmental affairs. All of these elements are effective in selecting the proper process for sweetening (Sadegh, Stenby, & Thomsen, 2013). Absorption processes could be divided into three groups: absorption with chemical reaction, absorption without chemical reaction (physical absorption) and hybrid absorption process (both physical and chemical). In low partial pressures of carbon dioxide, chemical solvents have high absorption capacity which is suitable to use in the post-combustion state for flue gas of power plants. However, in higher partial pressures, physical solvents are preferred (Wang, Lawal, Sidders & Ramshaw, 2011). Membrane technology is developing rapidly so that since 1980 has been used in some fields of gas purification on a large scale. Membrane technology has some advantages like a simple operation, suitable size and weight and space efficiency, environment and some disadvantages like membrane blocking or fouling and limited lifetime; however, in the membrane processes for gas separation, the membrane fouling is not very important (Mulder, 2012). Most studies have been conducted for separating gases through

membrane processes for single-stage systems [15]. Regarding that it is not possible to obtain the products with high purity, one of the suggested procedures is designing multi-stage membrane systems (Jafari, ghasemzadeh & Basile, 2017). Multi-stage membrane arrangements are similar to distillation columns. In order to reach an optimum enrichment in membrane processes, the following procedures could be used:

1. Redirecting a part of passed stream into feed in a single-stage membrane system.
2. Using in Series Membrane Cascade (ISMC)
3. Counter-Stream Recycle Cascade (CRC)

1.2. Conventional Membranes for Carbon Dioxide Capturing

Generally, membranes are divided into two organic (polymer) and inorganic (mineral) categories which are both used in the separation of carbon dioxide in outlet flue gases. One of the limitations of polymer membranes is that the high temperature of flue gas destroys the membrane easily. Also, the polymer membrane is not resistant to existing corrosive gases. Swelling, softness, and lack of strength against high pressures could be considered among other membrane problems (Khalilpour, Mumford & Rubin, 2015). According to their structure, inorganic membranes are divided into porous and non-porous categories. In inorganic porous membranes, ceramic or porous metal carrier is coated by an upper porous layer which must supply mechanical strength with the least resistance against mass transition. Inorganic membranes could operate under 300-800°C temperature range. Temperatures higher than 1000°C have been reported, too. Although inorganic membranes are more expensive than polymer membranes, they have significant advantages like abrasion, durability and thermal resistance of holes' structure. Today, there is a tendency towards applying and development of inorganic porous membrane, especially zeolite and silicate membranes, for separating carbon dioxide due to high selectivity and higher chemical resistance compared to polymer membranes.

Therefore, as a first approach, in this study design and simulation of multistage (7 Step) membrane unit by PRO/II software and its economic evaluation by Aspen process economics analyzer software at different operating conditions concerning the absorption process have been presented.

Table 1. volume percent of carbon dioxide in outlet gas from various industries and power plants' flue (Chen, S. 2016; Riboldi, L. 2015; Sanchez, D. 2019;Yousef, A. M. 2018;Aaron, D. 2005)

Type of plant or power plant	volume percent of CO ₂ in flue gas
Natural gas	3-5
Coal mining power plant	13-15
Cement	15-25
Iron and steel	15-20
Ammonia (flue gas)	8
biogas	25-35
Power plant boilers with natural gas or coal as fuel	8-15

2. Literature on Membrane Process

Zhao et al. (2010) evaluated the energy and economy of the multi-stage membrane process for separating carbon dioxide. The obtained results suggest that there is a correlation between permeability and selectivity of the membrane with the economy and energy demand of membrane processes. In higher permeability and selectivity, the economy and energy are more cost-effective (Zhao, Blum & Stolten, 2010).

Hassan et al. (2012) simulated and optimized separation of different concentrations of carbon dioxide from combustion gases through absorption and membrane methods. Economical evaluation was done for different concentration and this helps to decide to choose suitable technology considering different scenarios (Hasan, Elia & Floudas, 2012).

Tuinier. (2011) carried out a basic study of the economics of CO₂ capture with membrane and absorption technology. The flue gas containing 12.9 vol.% CO₂. The results show that the preferred technology highly depends on assumptions related to the availability of utilities. Also, the capital cost of the membrane unit is 1.7 times larger than the absorption unit (Tuinier et al. 2011).

Simon Roussanly et al. (2014) presented a new systematic methodology for the design and optimization of membrane systems for CO₂ capture incorporating both technical and cost models. In this work, graphical solutions to the separation problem are generated to design a cost-optimal membrane system that satisfies CO₂ capture ratio and product purity requirements. The result shows a comparison between the cost model considered and models available in order to show that the competitiveness of the membrane system designed is due to an improved design and not a possible underestimation of the membrane capture cost (Roussanly, et al. 2014).

Simon Roussanly et al. (2016) identified the membrane properties required to enable

cost-competitive post-combustion CO₂ capture from a coal power plant using membrane-based processes. This numerical model is used to assess the cost -efficiency of 1600 sets of membrane properties (selectivity and permeance) for post-combustion CO₂ capture from a coal power plant. The results show that to achieve this competitiveness with simple process configurations requires a permeance of at least $3 \frac{m^3}{m^2.h.bar}$ with high selectivity, or alternatively a selectivity of at least 65 with high permeances (Roussanly, Anantharaman & Rubin, 2016).

wang et al. (2017), evaluated CO₂ Capture Technologies from Coal-fired Power Plants. This work reviewed the basic process designs of chemical absorption and membrane-based separation processes for CO₂ capture. In addition, some energetic and economic estimates from other researchers for these two CO₂ capture technologies are summarized. The result shows that the membrane-based separation process does not possess an obvious advantage over the MEA-based chemical absorption process at the typical 90% CO₂ capture degree in terms of both energy consumption and cost. (wang et al. 2017)

Anselmi et al. (2019), Simulated the CO₂ capture separation unit. In this simulation, three technologies for CO₂ capture, absorption, adsorption, and separation using polymer membranes were considered, modeled, and compared. The results show that for a targeted CO₂ purity of 95%, the membrane process appears to be less energy -consuming. (Anselmi et al. (2019).

3. Simulation of Carbon Dioxide Capturing Unit

In this section, simulation of CO₂ captured from flue gas is studied in three various concentrations through PRO/II v.10 software. A significant characteristic of this software is its capability to connect with other important software such as MATLAB, Aspen HYSYS,

Aspen Process Economic Analyzer, Excel, etc. (Jafari, Behroozsarand & Ghasemzadeh, 2018).

In order to the economic evaluation of this unit, the conducted simulation was linked into Aspen Economic software and the economic analysis of these units is done. Other specification of PRO/II v.10 software in comparison with other commercial software, for simulating of chemical processes, is the possibility of simulation of single-stage and multi-stage arrangements of membrane units (Galli, Bozzano, Manenti & Pirola, 2018)

3.1. Simulation of Amine Absorption Unit

The PRO/II v.10 simulation program contains a mass balance method for modeling the removal operation of H_2S and CO_2 using amines. The Amine Package (AMINE) used to model the removal of H_2S and CO_2 from acid gas feeds using aqueous amine systems. Data is provided for amines MEA, DEA, DGA, DIPA, and MDEA. Results obtained for MEA and DEA are accurate enough for use in final design work. The recommended temperature, pressure, and loading range (gram-moles sour gases per gram-moles amine) for each amine system available in PRO/II v.10. Inlet flue gas (Air rich in CO_2) and amine solvent specifications for absorption tower indicated in Table 2.

A schematic diagram of the absorption unit is shown in Figure 1. In the absorption tower, carbon dioxide is absorbed by the amine solution during a calorific reaction. This absorption tower has 20 real stages. The pressure of the absorption tower is about 3 bar. The CO_2 -lean gas and CO_2 -rich amine solution leave the absorption tower (T-100) from the tower top and bottom, respectively. After

passing through a VLV-100 expansion valve, CO_2 -rich amine stream pressure reduces to 2 bar, then, enters into V-101 separator. In order to recover amine, it must be warmed until $105.1^\circ C$ and recovered in T-101 (Stripper). Heat should be imposed to break amine bonds with acidic gases. Required heat is provided by warm regenerated amine and transferred into CO_2 -rich amine in Amine-Amine Heat Exchanger, E-102. The number of stages in T-101 tower is 18 and the upper and lower pressures of the tower are 1.8 bar and 1.9 bar, respectively. Amine without carbon dioxide also exits from the bottom of the stripper. Since some water and amine are wasted, so regenerated amine enters into a control mixer (MIX-100, MIX-101) in order to measure the level of water and amine; and if be required, water or amine make-up are injected into this mixture. In order to increase the pressure up to 3 bar, regenerated amine stream enters into pump (P-100) and in order to decrease the temperature to $40^\circ C$, it enters into (E-103), and finally regenerated amine is returned into T-100 absorption tower. Since a mole percent of carbon dioxide is different in outlet gases, simulation of this unit has been done with three concentrations of 8%, 12% and 22% mole of carbon dioxide. The objective function is reaching into a specified value of carbon dioxide in treated gas.

3.2. Simulation of Membrane Unit

For symmetric membranes, this model only applies to a cross flow pattern. The unit supports between 1 and 10 feeds. Multiple feeds are combined into a single feed at the lowest pressure among all the feed streams. Permeation proceeds from the feed side across the membrane to the permeate side.

Table 2. Specification of inlet gas and solvent (Arachchige & Melaaen, 2012)

Name of stream	Inlet gas	Solvent
Temperature $^\circ C$	160	40
Pressure bar	1	3
Molar flow kmol/h	8162	60000
Component	Mole Fraction	Mole Fraction
Water	0.072	0.873
MEA	0.000	0.110
Carbon dioxide	0.220	0.017
Nitrogen	0.685	0.000
Oxygen	0.023	0.000

• Main assumptions

- ✓ A constant pressure on both permeate and residue sides.
- ✓ The driving force is partial pressure as calculated by ideal gas law.
- ✓ The residue side is well mixed.
- ✓ The permeate side is plug flow.
- ✓ The gas membrane unit is governed by the following equation:

$$F_i = Area \times K_i \times (P_{i-residu} - P_{i-surface}) \quad (1)$$

where:

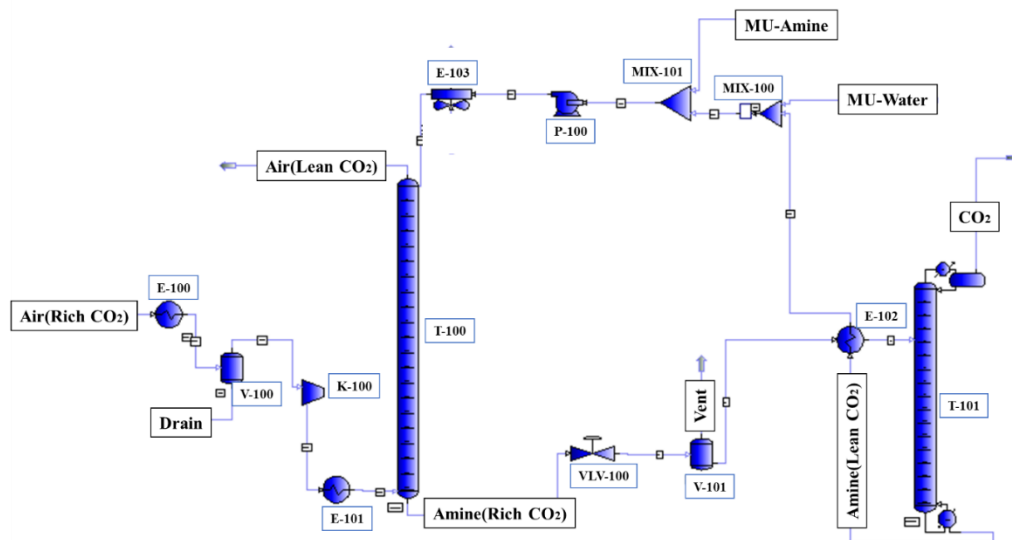
F_i = flow of component i in standard (volume/time) units, Area = area of membrane, K_i = Permeability constant of component i (volume / [area*time*pressure])

P_i = partial pressure of component i in (pressure) units

The selected arrangement in this work is a series arrangement. Inlet gas stream with the same characteristics as summarized in Table 2, enters the membrane separation stage

after reducing temperature and increasing pressure. Simulation of membrane units was also conducted with three concentrations of 8, 12, and 22% mole of carbon dioxide. Considered parameters and characteristics for simulation of membrane unit such as the thickness of the membrane, permeability of carbon-dioxide based on barrer and selectivity of carbon dioxide compared to nitrogen and oxygen are presented in Table 3.

On the other hand, the schematic diagram of the membrane unit is shown in Figure 2. In this research, a series arrangement of membrane has been used is 6 stages. The total number of the membrane is 6 and pressure enhancement in compressors is 10 bar. In the enrichment section, 6 compressors are used up to 10 bar for compressing feed and passed streams. The number of stages in enrichment and stripping parts are depended on the selectivity and desired purity level for upper and lower products.



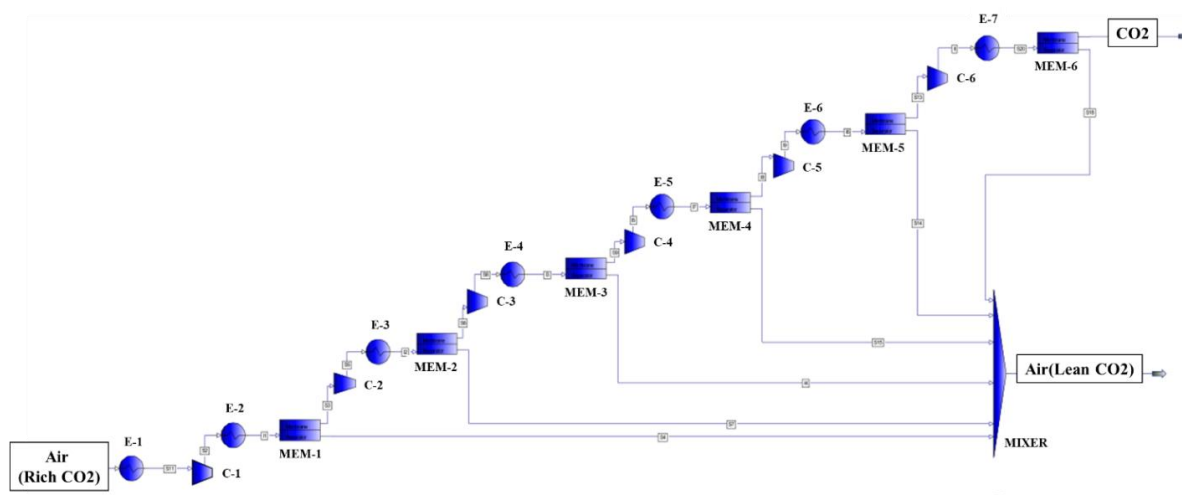


Figure 2. Schematic diagram of simulation membrane unit in PROII v.10 software for CO₂ capturing.

4. Economic Evaluation

An acceptable plant design must present a process that is capable of operating under conditions that will yield a profit. Since net profit equals total income minus total cost and taxes, knowledge of chemical engineers from very different costs in production processes is essential [24]. Now this software i.e. APEA (was known before as Aspen ICARUS in previous versions) is one of comprehensive and unique software in the field of plant design, economic evaluation of the chemical industries and developing professional reports for economic assessment [25]. APEA uses the equipment models contained in the Icarus Evaluation Engine a knowledge base of design, cost, and scheduling data, methods, and models to generate preliminary equipment designs and simulate vendor-costing procedures to develop detailed Engineering (Vozniuk, 2010).

Procurement-Construction (EPC) estimates. APEA provides tree diagrams that let you view, track, and revise information such as power distribution, process control networks, tiered contracts, areas, and their equipment specs, and installation procedures. Size of equipment is a prerequisite to costing and the results of size calculations performed during process simulation are loaded automatically by APEA.

In economic evaluation of a chemical process, some cases such as total capital cost, total operating cost, total product sales, total utility cost, equipment cost, total installed cost and desired rate of return are obtained. Capital and utility cost reduction is of vital importance in the process industry. The following is a list of some of the commonly used

terminology in economic evaluation with its description (Kallevik, 2010):

- **Installed cost** represents the total direct material and labor costs associated with the project component (including installation bulks).
- **Equipment cost** represents the bare equipment cost associated with the project component.
- **Equipment weight** represents the empty shipping weight of the equipment.
- **Total installed weight** is the equipment weight plus the weight of all bulks for installation such as piping, civil, and electrical.
- **Total utilities cost** include cooling water consumption, steam at various levels, electricity, hot oil, refrigerants, fire heat and ... annually.
- **The Operating Cost:** Indicates, by period, the total expenditure on the following items necessary to keep the facility operating: raw materials, operating labor cost, maintenance cost, utilities, operating charges, plant overhead, subtotal operating costs, and G and A costs (general and administrative costs incurred during production. This is calculated as a percentage of the subtotal operating costs.).
- **Total Capital Cost:** The capital needed to supply the necessary manufacturing and plant facilities is called the fixed-capital cost, while that necessary for the operation of the plant is termed the working capital. The sum of the fixed-capital cost and the working capital is known as the total capital cost.

Economic evaluation of a chemical process in APEA software includes the following steps [25]:

1. Obtaining simulation/process data for streams and unit operations
2. Adding the cost of feed and products streams (raw materials/feed, product, waste disposal, feed credits, etc...)
3. Specify the utility type (cooling water, HP steam, MP steam, LP steam, and power) in equipment.
4. Mapping unit operations to constituent equipment
5. Sizing equipment based on simulation process data and design standards
6. Evaluating equipment for cost based on the sizing

Since the input feed (flue gas) and outlet stream (clean gas) don't have economic value, feed and product cost is not defined to the software [25]. We must choose the type of utility for all energy streams so that the software could calculate the utility costs. The type of utility in pump and compressor is power type, in air conditioner, is air type, in condenser is cooling water type, and in the heater is low-pressure steam, and in reboiler is high-pressure steam. The next step is the determination of the equipment. Changes which could be done in unit operation of different equipment include:

a) Absorption unit:

- Condenser of distillation tower (T-100): TEMA shell and tube heat exchanger
- Coolers (E-100, E-101, and E-103): TEMA shell and tube heat exchanger
- Compressor (K-100): centrifugal compressor
- Heat Exchanger (E-102): TEMA shell and tube heat exchanger - Absorption tower: single-diameter and valve tray towers
- Distillation tower (T-100 & T-101): single-diameter and valve tray tower
- Pump (P-100): centrifuge pump
- Reboiler of distillation tower (T-101): heat exchanger of Kettle reboiler
- Separators (V-100, V-101): two-phase vertical separator

b) Membrane Unit:

- Membranes: it is not defined in Aspen software.
- Compressors (C1-C6): centrifugal compressor
- Coolers (E1-E7): TEMA shell and tube heat exchanger

After specifying the equipment, the sizing of equipment is done. Then, by specifying the geometric dimensions, the economic evaluation could be conducted. Equipment sizing is done according to simulation data and standard

design. Evaluation of equipment cost is also done based on the obtained geometric dimensions.

Since, membranes not defined for APEA software, related prices and costs for the economic evaluation of membrane processes are adopted from related tables or charts in literature. However, software is used for estimating the cost of compressors and conditioners. Finally, these costs will be summed [26]. For economic evaluation of the membrane process, first, the selectivity of the membrane should be calculated and then the price of carbon dioxide capture is calculated from Fig. 3 and then the price of the unit surface area of membrane is determined according to Fig.4 and is multiplied by the required area which is obtained by PRO/II v.10. A significant point in an economic evaluation of the membrane system is that the presented charts for membrane cost estimating is related to the vacuum mode, therefore, it must be corrected for non-vacuum mode. Selectivity is about 28. The cost of the membrane based on the mentioned charts is about 82 \$ per m² (in 2005). The cost index of 2005 is 500 and it is 576 for 2016. Therefore, the membrane cost is 95\$ per m² (Chemical Engineering Plant Cost Index, 2018). Eventually, using the related tables, fixed capital and working capital costs of equipment and then total capital costs are obtained.

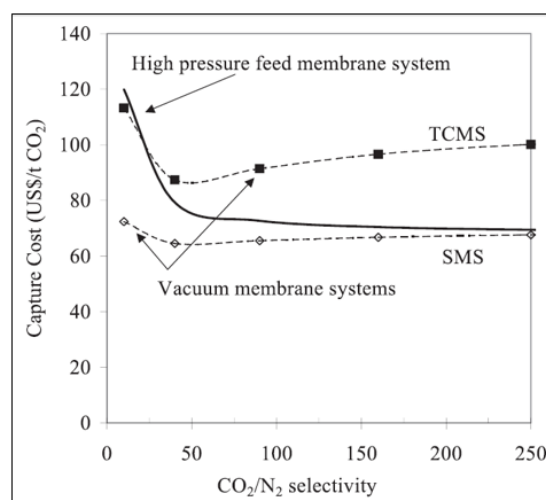


Figure 3. Changes of expenses of carbon dioxide separation based on CO₂/N₂ selectivity in different systems (SMS: single-stage membrane system, TCMS, and TCMS-RR: two-stage cascade membrane system with and without retentate recycle) (Abanades, J. C., et al. (2015).)

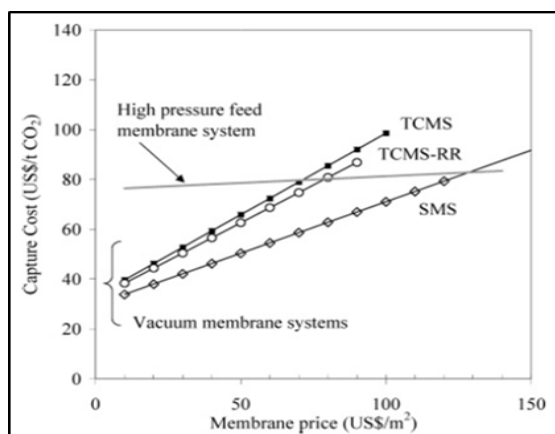


Figure 4. Cost changes of carbon dioxide separation based on membrane purchase cost in different systems (Abanades, J. C., et al. (2015).)

5. Results and Discussion

For simulation and economic assessment, the inlet flue gas specification was considered as presented in Table 2.

Before the sensitivity analysis of CO₂ capture by the absorption and membrane process, it is necessary to validate the overall simulation procedure and results. As

simulation results validation, in Table 4, our simulation results have been compared with theoretical results presented literature (Hassan. S.M 2005; Jakobsen. J et al 2017), showed that the simulation is well validated. Two important results in the absorption process are the rate of CO₂ recovery and circulating amine.

Since the innovation of this article is to simulate a multi-stage membrane unit in PRO/II software, it has not been done in previous researches. Therefore, there is no validation of the simulation of the membrane unit.

By unit simulation, different equipment and streams specification, as well as required utilities, were obtained, then, based on the simulation results, determination of the different costs and economic assessment was conducted. In table 5, a summary of equipment and utility cost for all equipment, in table 6, a summary of usage and cost of utility and table 7, the result of overall cost for absorption unit are given.

Table 4. Comparison of simulated CO₂ capture unit and published CO₂ capture study.

Properties	Simulation	Ref. (Hassan. S.M 2005)	Error, %
Air lean CO ₂ kmol/h	6800	6835	0.91
CO ₂ Recovery, %	85	85	0.00
Molar flow of solvent, kmol/h	60000	60175	0.28

Table 5. Summary of equipment and utility costs for absorption unit (mole fraction of CO₂, 22%)

Name	Equipment Cost [M USD]	Installed Cost [M USD]	Equipment Weight [k LBS]	Installed Weight [k LBS]	Utility Cost [USD/HR]
E-100	0.06	0.17	18.00	42.60	9.52
K-100	19.21	20.10	34.30	524.10	558.60
V-100	0.22	0.44	78.30	116.00	0.00
E-101	0.04	0.12	10.80	29.40	8.02
T-100	0.48	0.95	14.70	241.42	0.00
V-101	0.13	0.37	44.50	87.80	0.00
E-102	0.97	1.61	39.25	562.70	0.00
T-101	8.25	16.50	3378.00	5131.00	3263.60
P-100	0.04	0.19	2.80	37.20	4.30
E-103	0.60	0.85	234.00	294.00	47.62

Table 6. Overall usage and cost of each utility for absorption unit (mole fraction of CO₂, 22%)

Utilities					
Name	Fluid	Rate	Rate Units	Cost per Hour	Cost Units
Electricity		9926.60	kW	572.00	USD/H
Cooling Water	Water	844.00	M BTU/H	190.00	USD/H
HP Steam	Steam	994.00	M BTU/H	3140.00	USD/H

Similarly, simulation and economic assessment of the membrane unit was conducted. In this case, inlet gas and membrane specifications are considered as presented in Tables 2 and 3. The obtained results as different required costs are showed in tables 8-10. In table 8, a summary of equipment and utility cost for all equipment, in table 9, a summary of usage and cost of utility and table 10, the result of overall cost for membrane unit is given.

5.1. Effect of CO₂ Concentration

In this section, the economic assessment for both the absorption unit and membrane unit was conducted in different CO₂ concentrations. CO₂ concentration in the inlet flue gas was changed in the range of 8-22%, and its effect on the different required costs was evaluated.

According to Table 1, in most industries, the CO₂ concentration is in the range of 8-25%. Therefore, in this study, three concentrations of CO₂ in this range including 8, 12 and 22%

are considered to evaluate the performance of both membrane and adsorption processes.

For the absorption unit, the required solvent rate is plotted against the CO₂ concentration in flue gas in Fig. 5. This diagram presented the rate of required solvent for CO₂ concentration range from 8% to 22%. In the adsorption unit, 70% of capital is related to use solvent and 20% is also related to energy. By increasing the CO₂ concentration from 8 to 22%, the rate of applied solvent is increased from 1070 to 60000 kmol/h. The amount of solvent used depends on CO₂ concentration, absorber pressure, and temperature. Low temperature and high pressure are better for the absorption process. Since it causes problems in the absorber at temperatures below 35-40 °C, the amount of amine consumed depends more on the CO₂ concentration and the pressure of the adsorption tower. The simulation results show that at constant temperature and pressure, the amount of amine consumption is linearly proportional to the concentration.

Table 7. Overall cost figures for absorption unit (mole fraction of CO₂, 22%)

Summary	
Total Capital Cost [M USD]	63.28
Total Operating Cost [M USD/Year]	40.22
Total Utilities Cost [M USD/Year]	34.18
Equipment Cost [M USD]	30.00
Total Installed Cost [M USD]	41.30

Table 8. Summary of equipment and utility costs for membrane unit (CO₂, 22%, S: 28, P:1097 barrer)

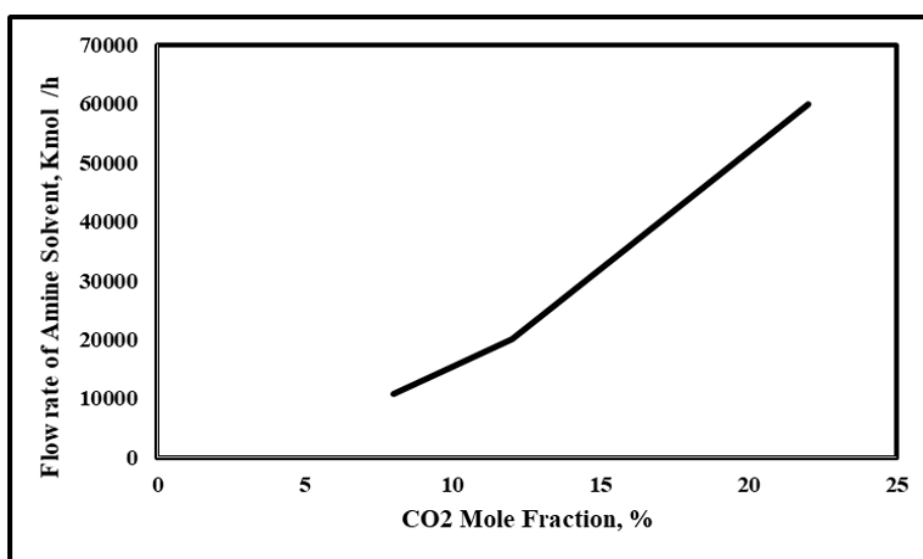
Name	Equipment Cost [M USD]	Installed Cost [M USD]	Equipment Weight [k LBS]	Installed Weight [k LBS]	Utility Cost [USD/HR]
E-1	0.08	0.18	27.40	53.00	3.96
C-1	47.80	49.00	826.30	1108.00	1718.80
E-2	0.16	0.37	30.00	52.00	21.98
MEM-1	0.13	0.19	-	-	0.00
C-2	14.59	15.2	267.00	377.00	515.64
E-3	0.04	0.21	10.20	31.00	6.52
MEM-2	0.05	0.08	-	-	0.00
C-3	8.72	9.15	166.00	248.00	386.73
E-4	0.03	0.21	8.30	28.00	5.11
MEM-3	0.03	0.04	-	-	0.00
C-4	8.65	9.10	162.00	240.00	343.76
E-5	0.03	0.21	8.00	22.10	4.55
MEM-4	0.02	0.03	-	-	0.00
C-5	8.65	9.10	162.00	240.00	343.76
E-6	0.03	0.17	8.00	22.10	4.31
MEM-5	0.01	0.02	-	-	0.00
C-6	8.65	9.10	162.00	240.00	343.76
E-7	0.03	0.16	8.00	22.10	4.18
MEM-6	0.01	0.01	-	-	0.00

Table 9. Overall usage and cost of each utility for membrane unit (CO₂, 22%, S: 28, P: 1097 Barrer)

Utilities					
Name	Fluid	Rate	Rate Units	Cost per Hour	Cost Units
Electricity		63865	kW	3678.60	USD/H
Cooling Water	Water	225.80	M BTU/H	50.60	USD/H

Table 10. Overall cost figures for membrane unit (CO₂, 22%, S: 28, P: 1097 barrer)

Summary	
Total Capital Cost [USD]	142.22
Total Operating Cost [USD/Year]	42.12
Total Utilities Cost [M USD/Year]	32.66
Equipment Cost [USD]	97.62
Total Installed Cost [USD]	102.68

**Figure 5.** Flow rate of amine changes of used solvent in different concentrations of CO₂.

In Figs. 6 and 7, Equipment cost, total installed cost and total capital for absorption and membrane unit are plotted against the CO₂ concentration in the inlet flue gas. These diagrams suggest that by increasing CO₂ concentration from 8% (mole fraction) to 22% (mole fraction), equipment cost, total installed cost and total capital cost in absorption process increased more than 20%, 47% and 51%, respectively, while these costs for membrane process were increased 53%, 53% and 55%.

In the membrane process, by increasing CO₂ concentration, the permeate stream through the membrane has been increased. On the other hand, by this increase, the area of the membrane, heat exchanger and capacity of compressors will be increased. As a result, total installed costs and equipment

costs will increase. Also, by increasing CO₂ concentration, the circulating amine solvent will be increased in the absorption unit. Moreover, by this increase, the equipment capacity will be increased and consequently, the cost of equipment and total installed costs will be increased.

Fig 8 shows the required surface area of the membrane by increasing the CO₂ concentration from 8% mole to 22% mole. It seems that, by increasing CO₂ concentration from 8% to 22%, the permeate stream was increased. As the permeate stream in the membrane is increased, the area of membrane will be increased. These results suggest that by increasing CO₂ concentration from 8% (mole fraction) to 22% (mole fraction), surface area of the applied membrane increased more than 53%.

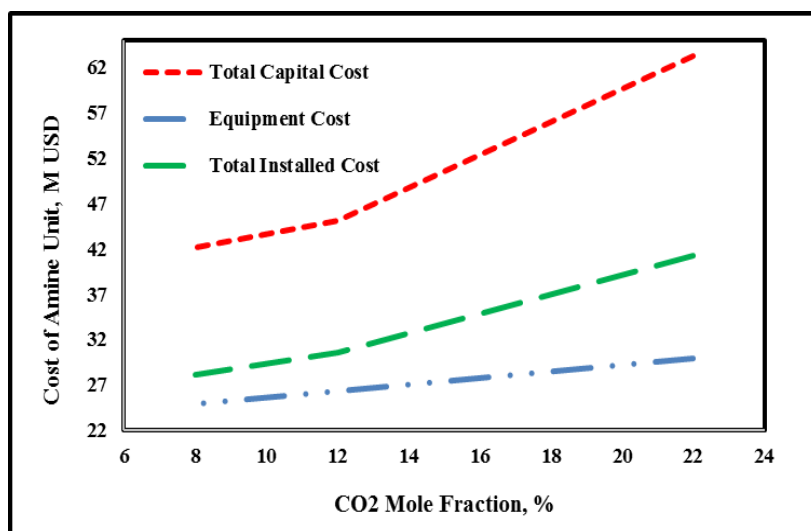


Figure 6. Equipment cost, total installed cost and total capital costs in different concentrations of CO₂ for the absorption unit.

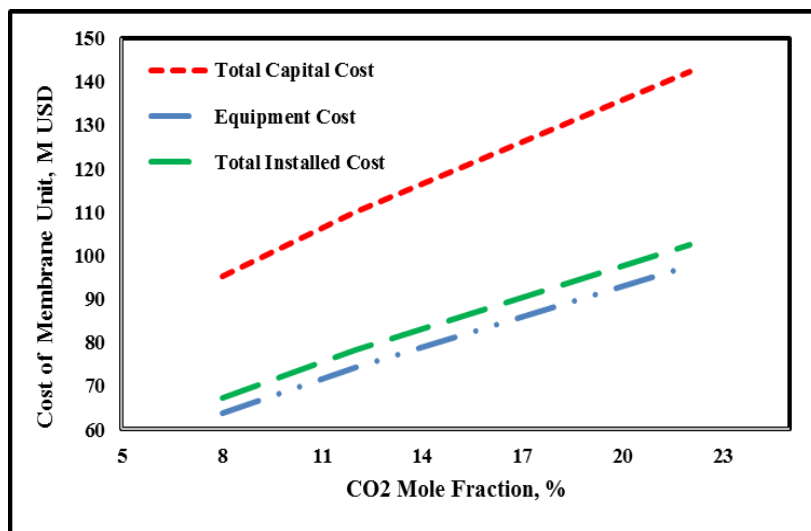


Figure 7. Equipment cost, the total installed cost and total capital costs in different concentrations of CO₂ for membrane unit (S:28, P: 1097barrer).

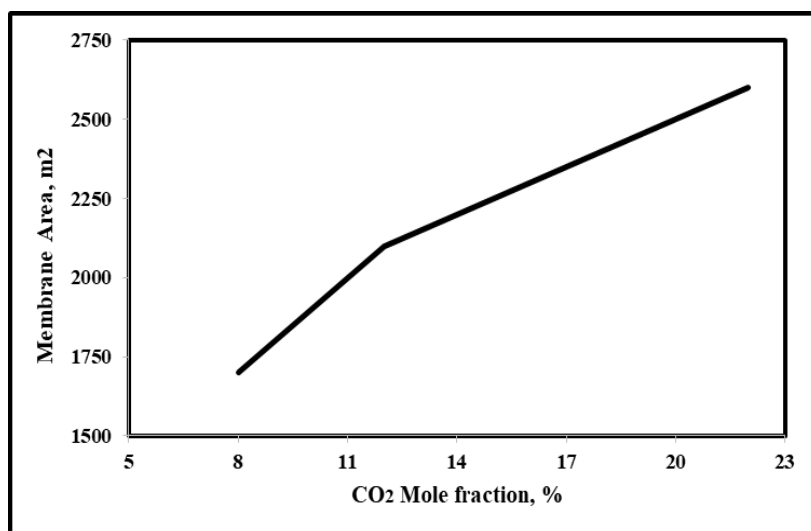


Figure 8. Required surface area of the membrane as a function of CO₂ concentration in inlet flue gas (S:28, P:1097barrer).

In Figs. 9 and 10, total operating cost and total utility cost for absorption and membrane unit are plotted against the CO₂ concentration in the inlet flue gas. These diagrams suggest that by increasing CO₂ concentration from 8% (mole fraction) to 22% (mole fraction), total operating cost and total utilities cost in absorption process increased more than 203% and 257%, respectively, while these costs for membrane process were increased 46% and 50%.

It is indicated that by increasing CO₂ concentration, the permeate stream was increased. By increasing this parameter, the power of compressors and pumps, the consumption of steam and cooling water of heat exchangers and ... will be increased. As a result, the total utilities and operating costs were increased.

5.2. Effect of Membrane Selectivity

The results so far show that the membrane process at different concentrations of CO₂ is

not economically feasible. In the membrane process, to separate CO₂ from the flue gas stream, the selectivity of this component against N₂ is very important.

A membrane unit has equipment such as a membrane, compressor, and cooler. The permeability of the membrane depends on the area of the membrane and the purchase price of the membrane and selectivity of CO₂ also depends on the purchase of compressors and coolers. As reported in Table 7, more than 99% of the cost of purchasing equipment depends on the compressor purchase. Therefore, in the process of CO₂ capture in the range of these concentrations, the selectivity of CO₂ is more important than the permeability of CO₂. Thus, by increasing CO₂ selectivity, the number of membrane stages, the number of compressors and the total capital cost of the membrane unit will decrease.

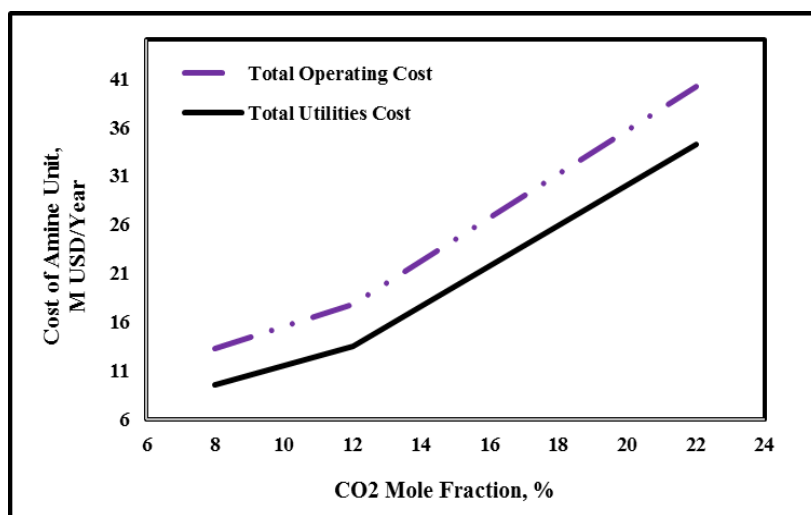


Figure 9. Total operating cost and total utilities cost in different concentrations of CO₂ for the absorption unit.

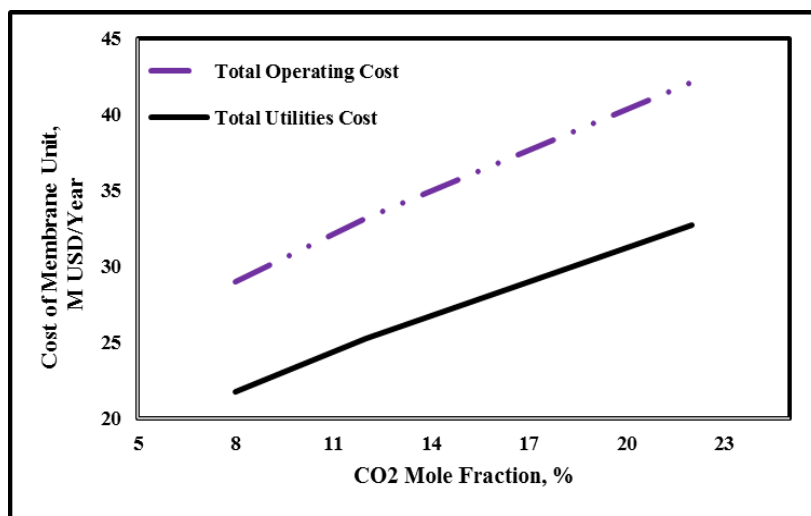


Figure 10. Total operating cost and total utilities cost in different concentrations of CO₂ for membrane unit (S: 28, P:1097barrer).

At constant permeability, 1097 barrer, the selectivity of the membrane was changed up to 700 and its effect on the different costs was evaluated. in fig. 11 and 12 equipment costs, total installed costs, total capital cost, total operating cost, and total utilities cost in membrane unit are shown against the selectivity of CO_2/N_2 in membranes. They suggest that by increasing the selectivity of CO_2/N_2 from 28 to 280, equipment cost, total installed cost and total capital cost in membrane process decreased more than 85%, 93%, and 84%, also, total operating cost and total utilities cost in membrane process decreased more than 79% and 87%. These results suggest, if we can reach CO_2 permeability of 1097 and the selectivity of 280,

the membrane separation method could compete with the absorption method economically and technologically to CO_2 capturing. Also, if we can reach CO_2 permeability of 1097 and the selectivity 280, the total capital cost in the membrane unit becomes 1.22 times higher than the absorption unit, but total operating cost and total utility cost of membrane units becomes about half of these costs in absorption unit.

By increasing the selectivity from 280 to 700, no change in the results will occur. Because, in CO_2 selectivity above 280, the number of membrane compressors and membrane stages remains constant, thus not affect the investment cost.

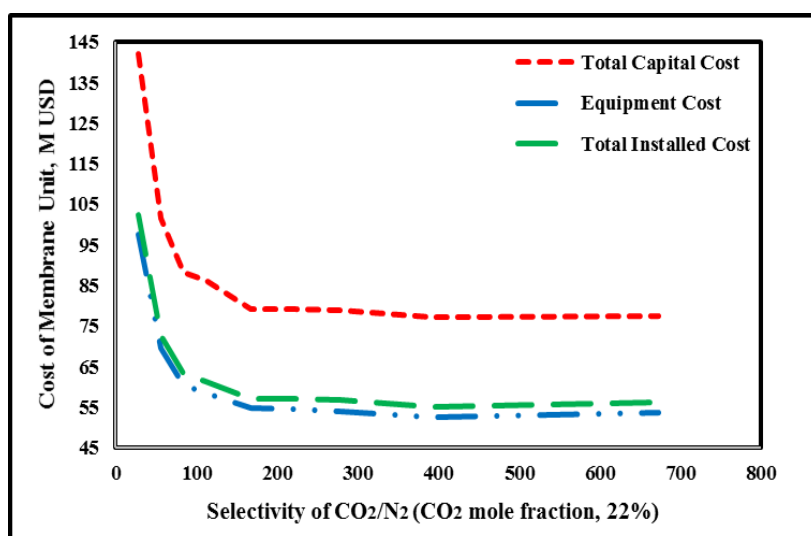


Figure 11. Equipment cost, total installed cost and total capital costs in different selectivity of CO_2/N_2 for membrane unit, ($P=1097$ barrer).

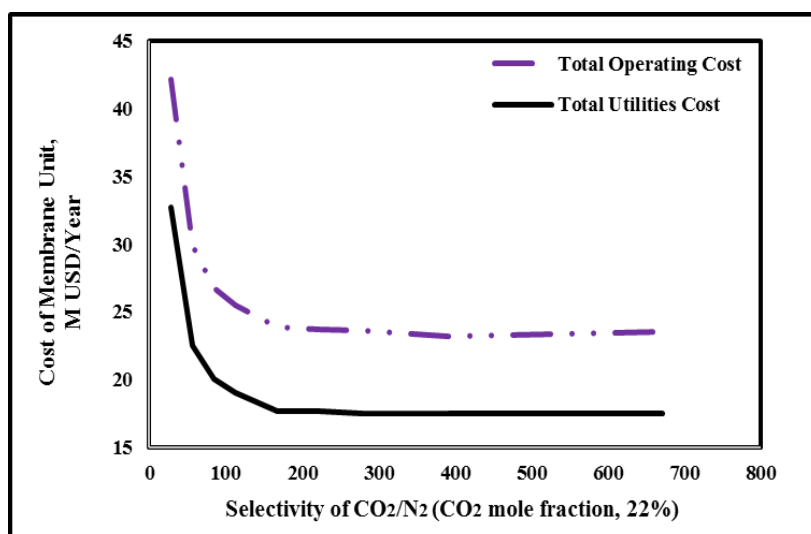


Figure 12. Total operating cost and total utilities cost in different selectivity of CO_2/N_2 for membrane unit, ($P=1097$ barrer).

6. Conclusion

Chemical absorption and membrane processes were introduced as methods for separating carbon dioxide from combustion gas; and the arrangements, constituents, and effective parameters in each method were briefly explained. Simulation and economic evaluation of both processes to CO₂ capturing from a typical combustion gas were conducted. Based on these simulations, the amount of applied solvent in the absorption process and the required membrane surface area in the membrane process were measured. Also, equipment purchase and installation costs, total capital cost, operation cost and utility cost for both processes were calculated for different concentrations of carbon dioxide in the inlet combustion gas. For applied selectivity and permeability in all concentration ranges of CO₂, the required capital costs for the membrane process are higher (over 2.2 times) than the absorption process. Therefore, the membrane process is more expensive. In - terms of total operating and utility costs, membrane process costs were about 1.01- 2.3 times higher than these costs for the absorption process. In a higher concentration of carbon dioxide, the total operating cost of the two processes is closer to each other. In other words, the sensitivity of operational cost and utility cost to CO₂ concentration in the absorption process is higher than this sensitivity in the membrane process. For example, by increasing carbon dioxide concentration from 8% to 22%, total operating cost and utilities cost in the absorption process increased more than 204% and 265%, respectively, while these costs for the membrane process were increases 45% and 50%, respectively. It is concluded that the most effective cost in the membrane unit is the compressor cost so that the compressor purchase cost alone is more than the sum of equipment purchase costs in the absorption unit. Effective factors on the cost of the membrane were the rate of selectivity and permeability of the membrane. For considered operation, if the selectivity of carbon dioxide to nitrogen in combustion gas was 280 and the membrane permeability was 1097 barrer, the membrane process could be economically competing with the chemical absorption process and applied instead of the absorption process. Even, operating and utility costs in the membrane unit become lower than these costs in the absorption unit. Therefore, for commercialization and industrial application of the membrane process must be focus on the enhancement of membrane performance

(improving the selectivity) and /or improving the membrane fabrication technology and reducing the membrane purchase cost.

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