

Assessment of a membrane contactor process for pre-combustion CO₂ capture by modelling and integrated process simulation

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Abstract

A membrane contactor process for pre-combustion CO₂ capture from shifted synthesis gas originated from IGCC power plant is assessed from the technical and economical point of views. The process is designed as pressure swing absorption and desorption in a closed loop. The design basis for process simulation were synthesis gas containing CO₂ and H₂ only, and the CO₂ capture efficiency was fixed to 90%. The CO₂ gas was absorbed in ionic liquid [bmim][TCM] inside a hydrophobic, porous hollow fibre membrane contactor. One-dimensional mathematical model of membrane contactor developed in MATLAB was integrated to the process simulation software (HYSYS) through Cape-Open simulation compiler. The energy evaluation of this process revealed that compressors are the most energy demanding process equipment. The specific energy requirement for this process is estimated 0.75 MJ/kg CO₂. A parametric study was also performed to analyse the effect of CO₂ concentration in feed gas and liquid to gas ratio. The capital cost investment and total operating costs of CO₂ capture unit were also evaluated. The capital investment required for capturing 0.14 M ton CO₂/year including CO₂ compression is 47.4 M \$, and the operating cost per year is 9.04 M \$. The membrane absorber contributed about 39 % to total investment cost. The specific cost of this capture unit is calculated to be 87 \$/ton CO₂.

1. Introduction

Greenhouse gas emissions are imposing great threat to increase the climate temperature. Carbon capture and storage (CCS) is addressing this challenge in order to mitigate the CO₂ emissions. The widely known strategies for reducing CO₂ emissions are pre-combustion or post-combustion CO₂ capture, or through oxy-fuel combustion. Post-combustion has been investigated the most among the above-mentioned three approaches due to retrofitting to existing power plants. However, post-combustion CO₂ capture offers many challenges, equipment corrosion, low CO₂ partial pressures, and pressurization of gas stream to storage site are a few to mention here. Pre-combustion CO₂ capture involves the mitigation of CO₂ before the fossil fuel is burned to produce power [1] and can be applied to both coal/natural gas fired power plants. The benefits of pre-combustion CO₂ capture encompass the less energy exhaustive process, high gas volumes, high gas pressure as driving force, less regeneration energy requirement, and more prominently the generation of another important fuel gas, hydrogen (H₂). The energy generation demand is typically 10–16%, which is approximately half of that in the case of post-combustion CO₂ capture [2]. Nevertheless, this technology is still under scrutiny for different physical and chemical absorbents and a CCS plant based on it is yet on launching pad.

The pre-combustion CO₂ capture from an Integrated Gasification Combined Cycle (IGCC) plant is comparatively an efficient technology to reduce emissions, but the overall cost of CO₂ capturing plant is potentially higher. There are some other challenges such as operational cost of the CO₂ capture plant, lack of sufficient experimental data and operational activities. Moreover, the IGCC power plant with CO₂ capture also tends to reduce the overall efficiency of the power plant. In order to combat this challenge, more energy efficient solvents and processes should be investigated and the absorption equipment be optimized.

There are, in general, two types of absorbents namely chemical absorbents and physical absorbents. Chemical absorbents react with CO₂ gas and enrich the mass transfer. Aqueous amine, carbonate-bicarbonate, hydroxide solutions are preferably employed on pilot and industrial scale, and in membrane contactor processes [3], but high regeneration energy associated with it increases the Capex of the industrial plant. On the other hand, physical solvents absorb CO₂ as pressure of the gas increases. Generally, high pressures are favourable for physical solvents [4]. For CO₂ absorption, physical solvents are favoured in case of high pressure and low temperatures and when large gas volumes are available at high pressures [5,

6]. The commercially available physical absorbents include polypropylene carbonate[7], Methanol (Rectisol)[8, 9], N-methyl-2-pyrrolidone (Purisol), Dimethylether polyethylene glycol (Selexol)[10, 11], and water. The physical absorbents exhibit stripping of acid gas by pressure swing, which reduces the energy penalty as in case of chemical absorbents, where the regeneration of absorbents demands high energy. The selection of an ideal physical absorbent could be based on CO₂ solubility and viscosity of the absorbent. The vapour pressure of the selected absorbent should also be as low as possible in order to circumvent the entrainment and solvent losses in regeneration step. The commercial physical solvents as mentioned above grab a few disadvantages such as reduced mass transfer (Selexol), validity upto low temperature operation (Rectisol), volatile solvent (Purisol), and economically not adequate to achieve high product gas purity. Ionic liquids are investigated as physical absorbent in recent years for CO₂ capture due to their task specific nature. The ionic liquids could be alternative promising physical absorbents due to less energy demand during stripping process as these exhibit insignificant low vapour pressures. Ionic liquids are also reported to have significant CO₂ solubility. The surface tension of ionic liquid with respect to membrane material should be tested first before its final selection. The only challenge in its implementation arises due to high viscosity. In this work ionic liquid Butyl-3-methylimidazolium tricyanomethanide ([Bmim][TCM]) was used as a CO₂ absorbent due to its high thermal stability, moderate viscosity and high CO₂ absorption capacity. More detailed information on the selection of the ionic liquid and the validation of the mathematical models by experimental data of the same ionic liquid can be found in our prior publications [12-15].

The process under assessment in this work is for coal fired power plants. According to literature, a coal fired power plant costs 27-39 \$/ton of CO₂ while natural gas combined cycle plant costs 48-102\$/ton [16]. The cost of CO₂ capture plant integrated to IGCC power plant depend on different factors including place, utilities, choice of separation method. The thermal regeneration of absorbent, absorbent pumping, heating/cooling of liquid and gas streams contribute to added auxiliary cost. Power, heating and cooling energy penalties are comparatively lower in case of Selexol unit for CO₂ absorption than Rectisol and MDEA processes [17]. Dave et al.[5] described the process design of CO₂ capture from syngas using DMEPEG as solvent in a packed tower and resulted in 90.4% CO₂ absorption and 89% solvent saturation. The also studied the co-absorbed hydrogen recovery from the solvent and reported to be 55.7% based on process design. The process design of IGCC power plant and CO₂ recovery using Selexol solvent was reported in [18]. The CO₂ recovery process helped to

control CO₂ emissions by 90.9% at the cost of reduction in net electric power by 145MW. In another study, the 90% CO₂ removal by Selexol solvent accounted for 5-7% reduction in LHV (Thermal efficiency) [19].

Pressure-swing membrane absorption and desorption process in a closed loop is employed in the present scenario for process simulation studies. The general flow scheme of a pre-combustion CO₂ capture process by means of pressure swing absorption and stripping is represented in Figure1.

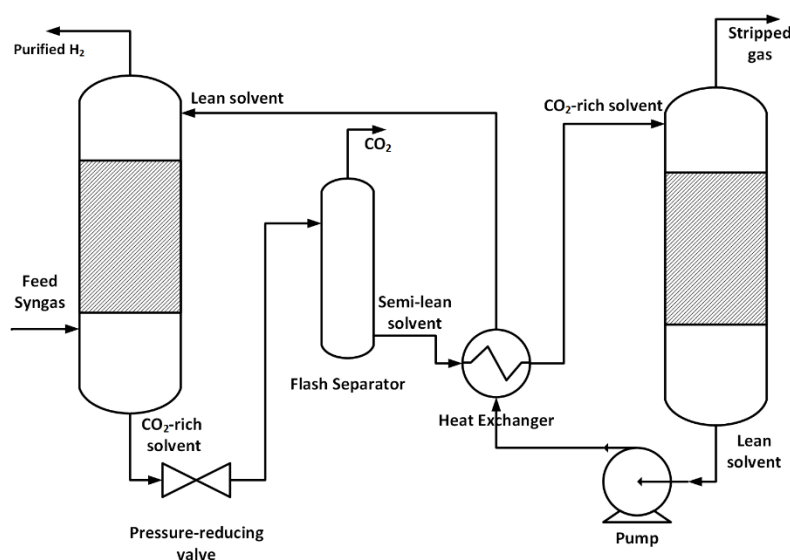


Figure1. General layout of CO₂ capture process by absorption and stripping

The absorption of CO₂ in solvent is carried out in an absorber for pre-combustion process. The CO₂-rich solvent is passed through the pressure reduction valve and is fed to the flash tank to separate CO₂ from the rich solvent. After heat exchange with the lean solvent, the rich solvent is fed to the top of the stripper. The depressurization in stripper causes the stripping of CO₂ from loaded solvent. The lean solvent is pumped again to the absorber to complete the continuous process.

Here, a membrane contactor based pressure swing absorption and desorption process for pre-combustion CO₂ capture is designed by employing an ionic liquid as absorbent. Energy and economic evaluation has been performed for this proposed process. Compared with the process presented in Figure 1, in the membrane absorption process the conventional packed column is replaced with hydrophobic hollow membrane contactor as absorber unit. The desorption of CO₂ is carried out by flash separators and pressure reduction valves. The ionic liquid [bmim][TCM] absorbs the CO₂ in the absorber which is quite promising solvent. The energy

and cost analyses are estimated in order to predict the overall performance of this process and are compared with other CO₂ capture processes that involved physical absorbent in conventional packed column.

2. The proposed process concept

2.1 CO₂ capture in IGCC power plant

Figure 2 represents the generic layout of an IGCC power plant with CO₂ capture. The coal is converted into carbon monoxide and hydrogen after the gasification process. The hot syngas exiting the gasifier at a temperature of 1100-1500°C [20] is cooled down and high pressure or low-pressure saturated steam is produced because water is being used as cooling agent. The solid particles contained in the syngas after gasification are removed by water scrubber. The syngas is cooled down to ambient temperature after water wash scrubber.

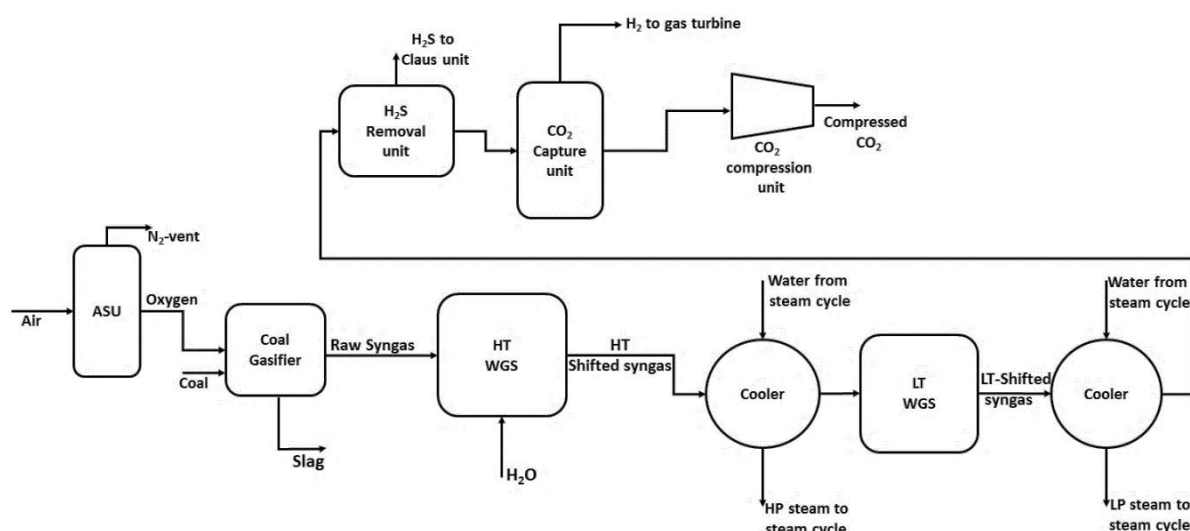


Figure 2. Generic layout of an IGCC power plant with CO₂ capture

The sulphur contents which are present in syngas as H₂S, are removed from the syngas in order to avoid the corrosion of the gas turbine and pollutant emissions. The removed hydrogen sulphide is sent to Claus unit to get elemental sulphur. After H₂S removal, the syngas is fed to the CO₂ capture unit. The IGCC power generation efficiency ranges between 40-46% (thermal energy efficiency).

2.2. Process flow diagram

Figure 3a demonstrates the pressure swing absorption-desorption cyclic membrane contactor process for CO₂ separation from shifted syngas. The shifted syngas after particulate and H₂S

removal is being considered as feed gas for this process simulation. The feed gas, containing 45% CO₂ and 55% H₂, is passed through the membrane absorber where CO₂ is absorbed in ionic liquid [bmim][TCM] at 20bar and 50°C. It was assumed that only CO₂ will be absorbed in ionic liquid [bmim][TCM] and whole amount of H₂ gas leaves the absorber as the solubility of H₂ in the ionic liquid are reported [21, 22] remarkably low. The CO₂-rich absorbent leaves the membrane absorber at 20bar and the pressure of this stream was reduced by introducing it to the pressure-reducing valve. The pressure drop of 11.8 bar occurs in the valve and CO₂-rich absorbent is sent to the flash separator 1 where part of CO₂ is stripped off. The flash separator 1 operates at 7.5 bar pressure. The semi-lean absorbent from flash separator 1 is further introduced to the pressure reducing valve followed by flash separator 2. The flash separator 2 is operated at atmospheric pressure (1bar). The lean absorbent is at low pressure after flash separator 2 and pressure is increased to 20bar by booster pump. Heat exchanger is installed after the pump to bring the temperature of the solvent to 50°C. The stripped CO₂ from both the flash separators is mixed and compressed to 75bar as shown in Figure 3b. Membrane absorber and flash separators accomplish the pressure swing in this process.

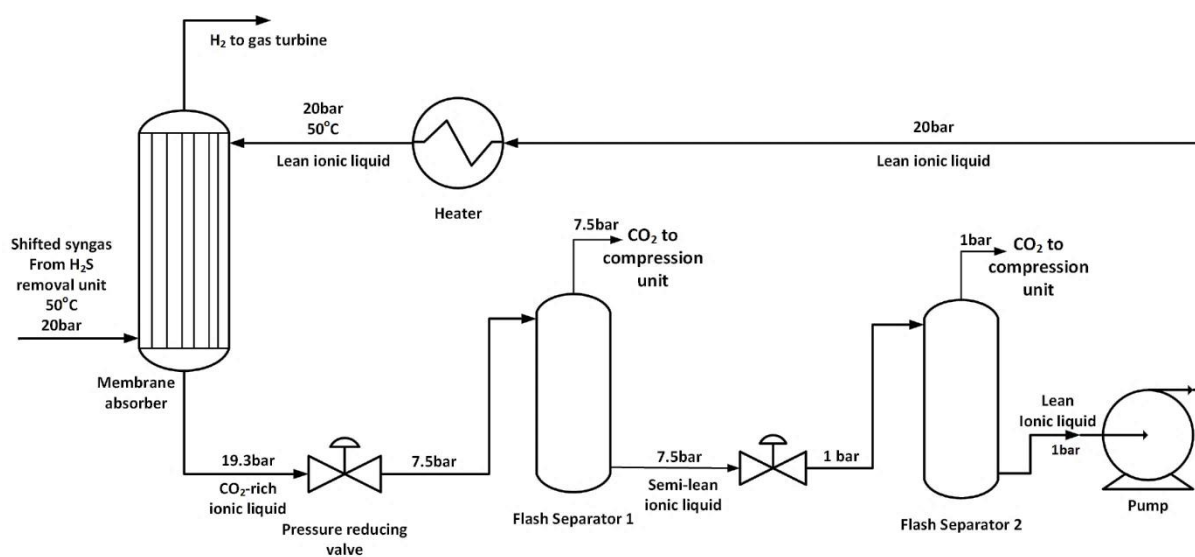


Figure3a. Process flow sheet diagram of pressure-swing membrane absorption and desorption process

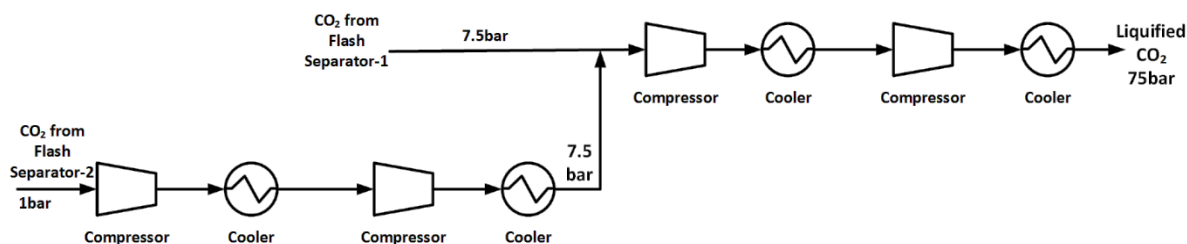


Figure3b. Process flow sheet for CO₂ compression

2.3. Property models of liquid, gas and membrane

2.3.1. Property model of the gas

The shifted syngas is a mixture of carbon dioxide and hydrogen along with some other gases in traces. The gas stream assumed here is after the removal of H₂S and other impurities. To simplify the process simulation, only carbon dioxide and hydrogen are considered in the shifted syngas. The fluid package used in HYSYS is Peng-Robinson EOS, which predicts the gas properties with quite good accuracy [23, 24].

2.3.2. Property models of liquid

As ionic liquids are a relatively new kind of absorbents investigated for CO₂ absorption and are not pre-defined in the HYSYS, they are introduced as hypothetical component. The physical and thermal properties of this component are evaluated by defining molecular weight and density. Moreover, membrane absorber unit utilizes the physical properties of [bmim][TCM] (as shown in Fig 4) as inherited in MATLAB program. Liquid density, viscosity, and CO₂ solubility were experimentally tested and empirical correlations are developed as function of temperature [15]. These correlations estimate the properties with quite good accuracy (1-5%). The liquid physical properties for rest of the units were evaluated by UNIQUAC model. It is found to predict the ionic liquid properties with quite good agreement [25, 26].

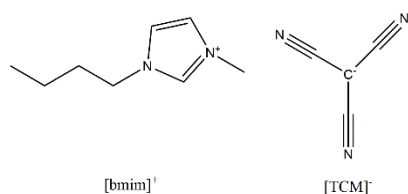


Figure 4. Molecular structure of [bmim][TCM]

2.3.3. Membrane property model

Users can formulate the membrane properties in the membrane property model. For current study, only porous membrane characteristic properties are plugged into the model. Membrane length, pore diameter, inner and outer diameters, porosity, thickness, and tortuosity are framed as input parameters. The model calculates the shell side and tube side surface areas and specific surface area of the module. The model will also approximate hydraulic diameter of the shell side of the membrane contactor. Both wetted and non-wetted modes of operations are included to the model to better estimate the performance of a membrane contactor. Membrane mass transfer coefficient is also a function of wetting and contributes to predict the overall

performance of the membrane contactor. Membrane wetting ratio can also be given as input according to the membrane properties and morphology.

3. Process design

3.1. Assumptions for the process design

The process design for pressure and temperature swing absorption is based on a few assumptions as described below, as the membrane contactor model is still preliminary: no pilot plant data are available to validate the simulation data. However, the mathematical model of membrane absorber for pre-combustion CO₂ capture was validated for experimental data in our previous study [15]

- i. Adiabatic rate based modelling is performed for membrane contactor.
- ii. Only laminar flow regime is considered for liquid side.
- iii. Hydrogen is assumed inert gas and do not pass through the membrane pores.
- iv. Membrane is considered porous and non-wetted. For the ionic liquid absorbent, the minimum pressure at which liquid can get into the pores is estimated by Young-Laplace equation, which is 0.96bar. The liquid pressure drop in this case is less than the above mentioned value.
- v. Only CO₂ and H₂ are contained in the shifted syngas as H₂S and other impurities are removed prior to the process.

3.2. Process design basis

The process designed in this work is based on the shifted syngas after waster gas shift reactor in an IGCC power plant. The process design basis acquired in this study are presented in Table 1 and the specifications of membrane contactor are provided in Table 2.

Table 1: Design basis of the process

Property	Syngas	Absorbent
	CO ₂ +H ₂	[bmim][TCM]
Pressure	20 bar	20 bar
Temperature	50 °C	50 °C
Flow rate	1000 kmol.hr ⁻¹	6500 kmol.hr ⁻¹
Molar masses	20.916 g mol ⁻¹	229.281 g mol ⁻¹
Concentration	45% CO ₂ /55% H ₂	
CO ₂ capture rate	90%	

Table 2. Specifications of membrane contactor [27]

Item	Value
Inner diameter of fibre	0.43mm
Outer diameter of fibre	0.87mm
Porosity of membrane	0.336
Length of membrane	1500mm
Number of fibres	116874000
Specific membrane area	3508 m ² .m ⁻³

3.3. Process equipment

3.3.1. Membrane absorber

To the best of our knowledge, membrane contactors have never been investigated for process design of CO₂ capture by either post or pre-combustion scenarios. Membrane absorber module is not available in the simulation tools, e.g., ASPEN-HYSYS, and thus a one-dimensional mathematical model of membrane absorption had been developed in the MATLAB by incorporating ionic liquid ([bmim][TCM]) as absorbent. The liquid is assumed to flow through the tube side, while the gases pass through the shell side of the membrane contactor in parallel and counter current direction to the liquid stream. The material and energy balances are applied on the membrane absorber to get the molar flow rates of CO₂ and H₂ both in gas and liquid phases, and temperature profiles along the length of membrane contactor. Differential equations emerge due to mass and heat balances are solved by method of orthogonal collocation method and f-solve in the MATLAB. The developed mathematical model was also validated with experimental data and details about the model can be found in our prior work [15]. The mathematical model of membrane absorber is then communicated to HYSYS via Cape-Open extension tool provided by Amsterchem [28]. This tool provides some function codes to communicate with the HYSYS. The single-phase properties, physical constants, and equilibrium data can also be predicted and retrieved using the codes. It gives the freedom to choose either the properties in the user model or from HYSYS. This tool is relatively easy as compared to other computational languages. The feed liquid properties are estimated in the MATLAB code. This model gives the molar fractions of all the components in both gas and liquid inlet and outlet streams. The temperatures and pressures are also evaluated at the outlet streams of gas and liquid.

3.3.2. Flash Separator

The flash separators are utilized in this process to strip CO₂ from the CO₂-rich solvent stream as only physical interaction is there between [bmim][TCM] and CO₂. The first flash separator operates at 7.5 bar and second flash tank at 1bar to achieve the 99% saturated ionic liquid to recycle to membrane absorber. These pressures are optimized to attain the less compression energy and recirculation rate of the solvent. The flash separator is chosen as two phase-separator and comprises of two outlet streams. One contains mainly gas in vapour stream and some CO₂ and ionic liquid in the CO₂-rich solvent stream. As vapour pressure of ionic liquid is significantly low [29] and chances of entrainment of the liquid with CO₂ in vapour stream are negligible. The sizing of flash separators is estimated by ASPEN-HYSYS.

3.3.3. Heat Exchanger

The lean solvent after stripping off all the carbon dioxide is pumped first to 20bar pressure before entering the absorber. The temperature of the lean solvent has to bring back to absorber temperature (50°C) and a heat exchanger is installed for it. The designed pressure drop in the heat exchanger is assumed negligible. Energy calculations are driven by the HYSYS.

3.3.4. Pressure reducing valve

The pre-combustion gas operates at high-pressure conditions and the CO₂ is absorbed in the ionic liquid by physical means. The ionic liquid can be recycled to the membrane absorber after depressurization of CO₂. It can be achieved by gradually reducing the pressure either by pressure reducing valves or flash tanks. The pressure reducing valves are cost effective, that is why two pressure reducing valves are incorporated into the flow sheet. A pressure drop of 11.8 bar and 6.5 bar are assumed for the two valves considering the operating pressure of flash separators. The high pressure drop considered in the first valve as the feed stream containing higher amount of liquid is being throttled.

3.3.5. Lean solvent pump

The pressure of inlet feed gas stream to membrane absorber is at high pressure. The liquid phase must be at a similar pressure as that of gas to avoid bubbling or wetting of the membrane. The feed gas pressure is approximately 20 bar and ionic liquid must be pumped at this pressure to the membrane absorber. The ionic liquid after stripping off all the CO₂ is at 1bar and is pressurized by centrifugal pump. The viscosity of ionic liquid is comparably higher than other physical absorbents, which would add surplus of energy to pump this liquid up to 20bar.

3.3.6. Compressor

CO₂ gas is stripped off from both the flash separators at different pressures (7.5 and 1bar). The CO₂ flashed in the first separator is at 7.5 bar pressure while CO₂ from flash separator is released at 1bar. The CO₂ of flash separator-2 is first compressed to 7.5 bar by installing compressors with intercooling. This stream was then mixed with CO₂ of flash separator-1. This mixed stream is further compressed to 75 bar in series of compressors with intercooling.

4. Results and discussion

4.1. Parametric analysis

4.1.1. Effect of feed gas composition

The composition of carbon dioxide in the feed shifted syngas can vary between 35- 55 %. Therefore, the influence of various feed gas CO₂ concentrations on the required membrane contactor area and total energy requirement was studied as shown in Figure 5. The CO₂ capture rate is assumed 90% for this study and liquid to gas ratio is fixed to 6.5. The results in Figure 4 indicate that membrane contactor area is reduced with concentrated CO₂ feed gas stream. There is 54% lessen membrane contactor area by varying the feed composition of CO₂ from 35 to 55%. The higher is the composition of CO₂ in feed gas, higher will be the partial pressure of the gas and hence enhanced absorption flux in the membrane contactor. Thus, the feed gas concentrated with CO₂ improves the driving force to achieve required separation.

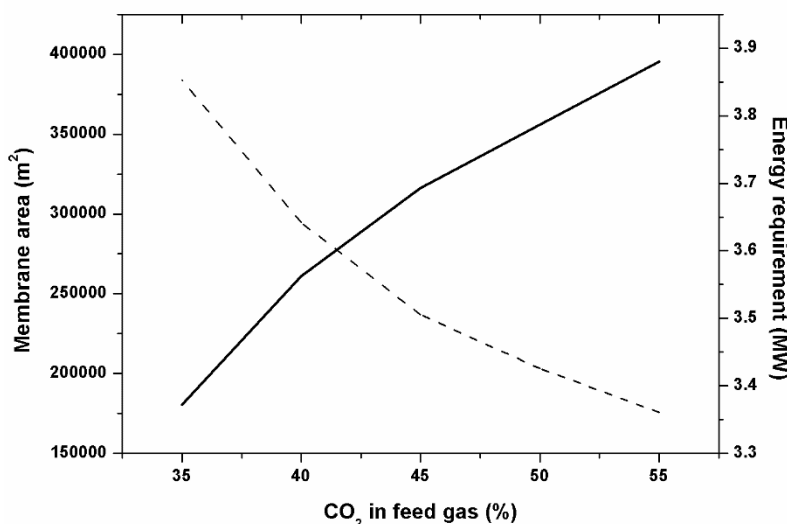


Figure 5. The influence of CO₂ composition in feed gas on membrane contactor for 90% CO₂ capture rate; membrane area (dash line) and energy requirement (solid line)

Besides the membrane area, the energy consumption has close relationship with initial composition of CO₂ in feed gas. The energy requirement increased 1.15 times by altering the

CO₂ composition in feed gas from 35- 55% to achieve 90% CO₂ capture rate. It could have been resulted from compression energy penalty due to handling of large gas flow rates.

4.1.2. Effect of L/G ratio

The liquid to gas ratio (L/G) is another prime parameter in investigating the performance of CO₂ capture process. As ionic liquids are in development phase for industrial scale production, this parameter will help in assessing the energy requirement for IL-based membrane contactor process for pre-combustion CO₂ capture. The results obtained by varying the liquid to gas ratio from one to seven are presented in Figure 5. The liquid to gas ratio is defined here in terms of molar flow rate of both gas and liquid. The CO₂ capture rate improves from 25 to 91% with increasing the L/G ratio from 1-7. The ionic liquid [bmim][TCM] is the physical absorbent and increasing the amount of ionic liquid improved the CO₂ absorption and hence the capture rate. A 90% CO₂ capture rate can be reached at an L/G ratio of 6.5. Valencia-Marquez et al.[30] performed the techno-economic study of CO₂ capture pilot plant using ionic liquid 1-decyl-3-methyl imidazolium trifluoromethane sulfonate([C10mim][TfO]). Their results displayed that a removal rate of 90% can be achieved at an L/G ratio of 3.5. In another study by Dave et al. [5], to capture 90% CO₂ from synthesis gas using physical solvent DMEPEG, was acquired at L/G ratio of about one. This difference in L/G ratio could have arose due to the gas solubility in absorption liquid. The CO₂ capture efficiency in case of physical absorbent is strongly dependent on the Henry's law constant of the absorbent and the values for [bmim][TCM] and DMEPEG at 33°C are 5.73 and 2.76 MPa respectively. The CO₂ solubility in DMEPEG is noticeably higher than ionic liquid [bmim][TCM].

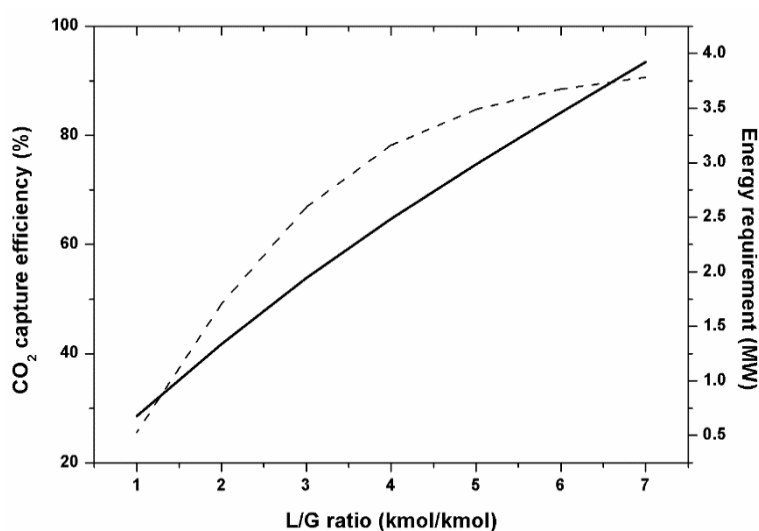


Figure 6. The CO₂ capture efficiency as function of liquid to gas (L/G) ratio; CO₂ capture efficiency (dash line), Total energy requirement (solid line)

On the other hand, the L/G ratio can also influence the energy requirement of the process as pumping and compression energies are of great concern for such kind of processes. It is also evaluated for different liquid to gas ratio. Results presented in Figure 6 indicates a substantial energy requirement by varying the L/G ratio. High energy requirement is firstly due to the pumping of large flow rates of ionic liquid, and more energy will be required by heat exchanger to heat the absorption liquid to achieve the membrane absorber temperature. In addition, large amount of CO₂ gas separated during this process need to be compressed to 75 bar.

To compare the energy demand of this process, the effect of L/G ratio on specific energy requirement is being assessed. Considering this, L/G ratio were altered from 4 to 7 to achieve 90% CO₂ capture rate and it was attained by varying the membrane area while other parameters were kept constant. The results derived from this are illustrated in Figure 7. The specific energy is defined as the ratio of total energy consumption to the mass flow rate of CO₂ captured. An increasing trend in specific energy with L/G ratio can be observed in Figure 6. The specific energy consumption per kg of CO₂ of this process is 0.75 MJ, much lower than the reported values for other capturing methods and comparable to membrane processes[31]. It reflects that membrane contactor based pre-combustion process could be a potential alternative to conventional packed column process.

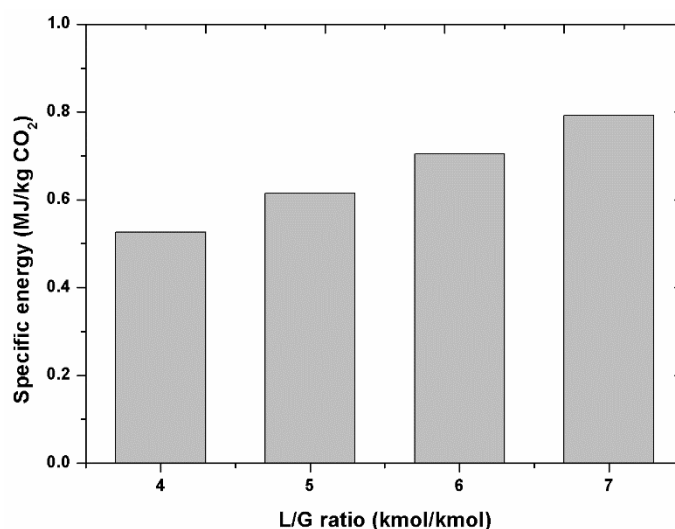


Figure 7. Specific energy demand as a function of liquid to gas ratio

Liu et al.[32] performed a process simulation study of decarbonisation of shale gas using ionic liquids. Pressure swing based absorption-desorption process was designed for this study. They designed and simulated a two-single stage and multistage flash tanks system for

decarbonisation at high-pressure conditions (60bar), and reported that the specific energy in terms of CO₂ was approximately 0.5MJ/kg CO₂ using multistage flash tank system with ionic liquids.

Another study by Shiflett et al. [33] for CO₂ capture using ionic liquid [bmim][Ac] was performed for post-combustion pressurized flue gas (6bar) and specific energy requirement was 6.01 MJ/kg CO₂. The high-energy consumption is due to the pre-heating the flash tank stream and steam requirement in the flash separator. The energy associated with steam requirement is 73% of total energy demand, thus specific energy could have been less than 6.01 MJ/kg CO₂.

4.2. Energy Performance

The immense energy penalty of the CO₂ capture unit is an obstacle in its implementation and has always been a key parameter in order to compare different technologies for CO₂ capture. The high energy associated with aqueous amines makes the CO₂ capture process energy intensive. In this section, breakdown of energy requirement with respect to process equipment is being analysed. The proposed pre-combustion capture process has mainly heat exchanger, lean solvent pump and compressors as energy consuming process equipment. The energy breakdown of each equipment is demonstrated in the Table 3. Heat exchanger is installed to heat the lean solvent to absorber temperature of 50°C and contributed 24% to total energy penalty of this proposed design process. Heat exchanger is probably exerting more energy to heat the large molar flow of absorption liquid stream to achieve the set temperature difference.

Table 3. Energy penalty of each process equipment

Process equipment	Energy (MW)
Heat exchanger	0.89
Lean solvent pump	1.03
Compressors	1.78
Total energy penalty	3.69

Lean solvent pump only adding about 28% of its part in total energy demand. The energy of lean solvent pump is consumed in pressurizing the absorption liquid stream to membrane absorber pressure of 20 bar. The lean solvent after the second flash separator is at atmospheric pressure (1 bar), and it is necessary to boost the pressure of the lean solvent to absorber pressure

(20 bar) to ensure efficient CO₂ absorption. The feed gas entering the membrane absorber is at 20 bar, thus the transmembrane pressure difference must be within a reasonable range to avoid membrane wetting and bubbling. Membrane wetting is the phenomena that occurs as a result of the pressure difference between gas and liquid streams over the liquid penetrating pressure of the porous membrane. It blocks the membrane pores and deteriorates the process performance, resulting in low CO₂ capture efficiency. The other factor that contributes to energy demand of the pump is the relatively high viscosity of ionic liquids. The viscosity of ionic liquid is substantial at ambient temperature due to which pump has to exert more energy in order to boost the solvent pressure to 20 bar. In order to reduce the pump energy penalty and achieve the membrane absorber temperature, the lean solvent is set to be 50 °C. At this temperature, the viscosity of ionic liquid is relatively low and becomes comparable to conventional physical solvents at their operating temperatures, such as Selexol; The viscosity of selexol solvent is 5.8 mPa.s at 25 °C [34] and viscosity becomes much higher at its operational temperature (-18 °C). In our case, the viscosity of IL [bmim][TCM] is 27 mPa.s at 25 °C but it decreases to 5.8 mPa.s at 80 °C. In this sense, it is reasonable to say that the viscosity of IL in operation is comparable to Selexol.

The greatest contribution can be found from the compressors, which accounts for 48% of the total energy requirement. The CO₂ gas stripped off from the flash separators 1 and 2 is at 7.5bar and 1bar pressure respectively. An immense amount of energy is required to compress the gas from these pressures to 75 bar which is the storage pressure of liquefied CO₂. Additionally, the large gas flow rates of CO₂ are removed in the process and required high compression duty as discussed earlier. The multistage compressors are installed with intercooling and large auxiliary load will be required.

It must be noted that the total energy requirement for membrane contactor process for pre-combustion CO₂ capture using ionic liquids is significantly smaller than a process using similar absorbents based on packed column. The specific energy requirement (as discussed earlier) for this process is 0.74MJ per kg of CO₂ [35]. The designed process is based on pressure swing using ionic liquids, in which ionic liquids physically absorbs CO₂ at high pressure, and then releases it by reduction in pressure via pressure reducing valves and/or flash separators. It also indicates that it is beneficial to use ionic liquid here as absorbent due to their negligible vapour pressure and low energy for regeneration compared with using aqueous amine solutions in packed column. The chances of absorbent entrainment and evaporation losses are also significantly lower.

4.3. Economic evaluation

The economics of any processing unit determines its potential to be applicable on large industrial scale. The present section describes the methodology used to estimate the total capital investment and operational cost of pre-combustion CO₂ capture unit. The basis of economic evaluation are provided in Table 4. The plant life is fixed for 30 years while membrane replacement is set to set to 10 years. There are various cost estimation methodologies available to predict the economics of a processing unit. Feng [36] compared five capital cost estimating methodologies and found installation, material and pressure factors to be varying for each method.

In the present work, total capital investment is estimated by summing up the fixed capital cost and working capital. The working capital is assumed 15% of total fixed capital cost in this study. The total fixed capital cost is calculated by inside battery limit, offsite, engineering and construction, and contingency costs. The inside battery limit cost accounts for the installation and procuring the process equipment.

Table 4. General economic parameters involved in cost evaluation

Parameter	Value
Base year	2017
Plant Life	30 years
Operating rate	8000 hrs
Discount rate	6 %
Electricity price	0.05 \$/kwh
Material of construction	304 Stainless steel
Membrane life	10 years

The PTFE hollow fibre membranes had only been tested for pilot scale [37] and the commercial price was not available, thus an estimated price of 20\$ per square meter is adapted for the cost evaluation. The installation cost of all the equipment is determined based on some typical cost factors as described by Towler and Sinnott [38]. They have also summarized the factors for estimation of offsite, engineering and construction and contingency costs. The ISBL (inside battery limit) cost is evaluated on the basis of equipment erection, piping, instrumentation, electrical and civil structure factors. The care must be taken when selecting equipment material

for high-pressure applications and 304-stainless steel is chosen based on the temperature and pressure conditions.

Table 5. ISBL cost distribution

	Cost M \$
Membrane absorber	4.86
Flash separators	0.51
Heat exchanger	0.11
Lean solvent pump	0.07
Compressors	6.75
ISBL cost (including installation cost)	22.7

The cost of ionic liquid is an important parameter in cost evaluation. As [bmim][TCM] is the physical absorbent and no reaction is involved with CO₂, thus ionic liquid is completely regenerated (99%). Secondly, the vapour pressure of ionic liquid is negligible and thus evaporation losses are very low. Considering this, ionic liquid replacement will be carried out every 8th year. The ionic liquids are yet not commercially produced at high capacity and their cost cannot be correctly predicted. The cost range for ionic liquid at commercial level may fall between 2.5- 50\$ per kg at later stage [39]. A price of 10\$ per kg of ionic liquid is chosen in this study.

Besides the total investment required for building the processing unit, operational cost of the processing unit is also of prime consideration for process optimization. Fixed production cost includes the operating and labour, maintenance, insurance, taxes and plant overhead charges and variable production cost covers the cost for raw material, utilities and consumables. Fixed and variable production costs are also assessed and cost factors used are presented in Table 5 [40].

The cost estimation methodology described above is applied to assess the economic potential of this process. The total capital investment required to build this CO₂ capture unit is summarized in Table 6 while Figure 7 illustrates the breakdown of the purchased equipment cost. Fixed capital cost is calculated by taking into account offsite batter limit cost (OSBL), engineering and construction cost and contingency factors as described in Towler and Sinnott [34].

Table 6. Total capital investment

	Factor	Cost M US\$
	Inside battery limit cost (ISBL)	22.7
	Fixed capital cost (FCC)	41.3
	Working capital (WC)	6.19
	Total capital Investment (TCI)	47.4

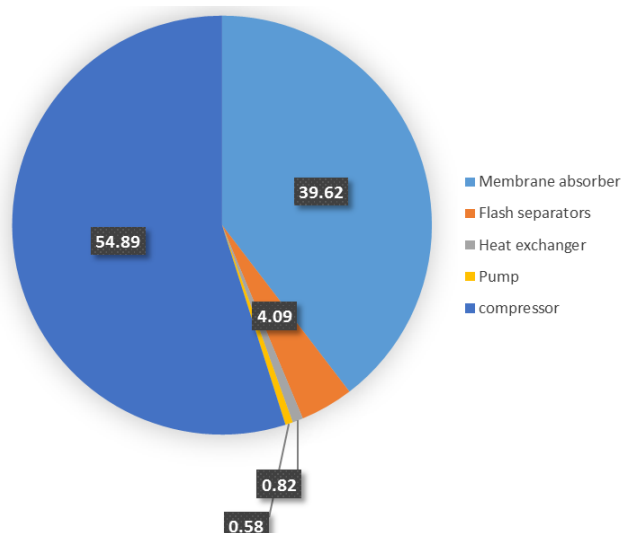


Figure 7. Breakdown of each equipment cost to total investment

Clearly, the compression cost is dominating (54%) among all the equipment cost. This is due to the immense energy requirement to compress the gas to 75 bar pressure and auxiliary load associated with it.

Table 7. Total plant operating cost

	Cost M US\$
Labour cost	0.23
Maintenance	0.68
Insurance	0.23
Taxes	0.82
Plant overhead	0.17
Fixed cost of production (FCP)	2.13
Raw materials	3.79
Utilities	1.52
Consumables	0.32
Variable production cost	5.63
Total operating cost (TOC)	9.04

The important point here is that the cost of membrane absorber only accounts for 39 % of the total capital cost. This high cost is due to the assumed cost of membrane (20 \$ per square meter of membrane). The price of membrane would have significantly reduced when prepared on large industrial scales. The total investment needed to capture 90% CO₂ (0.14M ton/year) by proposed process in this work is 47.4M \$ and the operating cost of this process is 9.04M \$. The cost of ionic liquid is significant contributor in estimated total operating cost. Total plant operating cost is presented in Table 7.

It is important to estimate the specific cost of this process with respect to CO₂ captured. The expression used to calculate the specific cost is given by [41]:

$$Cost_{specific} \left(\frac{\$}{\text{ton CO}_2} \right) = \frac{[TOC + 0.07 TCI]}{CO_2 \text{ captured}}$$

The specific cost calculated by the above expression for this process is 87\$. The capture cost per ton of CO₂ for pressure swing adsorption processes is 60\$ and is comparable to the cost calculated in this process [42]. Although, the specific cost is simulated based on laboratory scale data, yet the process is competitive with existing technologies. The CO₂ separation with membrane contactor is in developing phase and it is expected that specific cost will further reduce by optimizing the process at pilot and industrial scales.

Conclusions

The first of its kind pressure swing membrane absorption and desorption process for pre-combustion CO₂ capture in ionic liquid is studied in the present work. The membrane contactor model is developed in MATLAB and integrated to ASEPN-HYSYS simulation software by Cape-Open. The capacity of this capture process is 0.14M ton of CO₂ per year at 90% CO₂ capture rate. The required specific energy for the process is calculated 0.74 MJ/kgCO₂, which is much lower than other methods being used for pre-combustion CO₂ capture. The economic evaluation is also performed by capital cost and total operating cost estimation. The total capital investment for this process is ~47.4M \$, while operating cost per year is 9.04M \$. The membrane absorber only contributed 39% of the total capital cost. The specific cost of this process is estimated as 87 \$/ton CO₂ which defines this technology to be competitive with applied technologies for pre-combustion CO₂ capture.

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