



A novel tool for the modeling, simulation and costing of membrane based gas separation processes using Aspen HYSYS: Optimization of the CO₂/CH₄ separation process

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ABSTRACT

A key tool called “MemCal” is developed to support the simulation of membrane-based gas separation processes using Aspen HYSYS. By integrating “MemCal” with the simulator, users can simulate complex multi-component/multi-stage processes, perform sensitivity studies, and utilize the default HYSYS features to cost and optimize processes. Industrially, “MemCal” can be used to simulate the separation of air, biogas, natural gas, olefins-paraffins and other gases. The CO₂/CH₄ binary mixture separation was demonstrated in this manuscript as the first and the simplest application of “MemCal”. The simulation of various multi-stage processes suggested that the treatment cost can be dropped by optimizing the separation load distribution among the stages. For CO₂/CH₄ separation, the simulations suggested that the optimized two-stage process plus recycle resulted in the least separation cost. The installation of a third stage for boosting hydrocarbons recovery was found unjustified when membranes equivalent to or better than cellulose acetate are adopted.

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1. Introduction

Natural gas purification from its associated acid gases, particularly CO₂ is one of the major operations encountered in a typical natural gas treatment facility, whether onshore or offshore (Mazyan et al., 2016). The presence of CO₂ adversely impacts the produced gas quality, its market competency and thus its price; over 30% of the crude gas produced in US suffer from high levels of CO₂ and H₂S (Hao et al., 2002). The lack of market interest in a CO₂-rich gas is a consequence of the gas depreciated heating (calorific) value and its potential in forming corrosive solutions (Rufford et al., 2012). The CO₂ specification of the sales gas is often imposed by the gas customer depending on its end-usage. The gas utilized for power generation, heating applications, and/or water desalination is typically ~2–3 % CO₂ v/v (Foss, 2004), whereas the CO₂ content of the gas utilized as a feedstock for gas liquefaction (LNG) processes and petrochemical industries should be reduced to

ppm level for miscellaneous technical justifications as elaborated in Klinkenbijn et al. (1999).

Multiple technologies such as physical/chemical solvent absorption, solid adsorption, batch units and membranes were proven effective in reducing the CO₂ content of the crude gas (Tennyson and Schaaf, 1977). However, each of these technologies have a unique set of advantages and drawbacks in terms of their capital and operating costs, purification capability, reliability, operation/maintenance simplicity, units' weight/footprint, environmental impact, tolerance to contaminants and many other factors. Among the above-mentioned technologies and from the application point of view, the amine (chemical solvent) absorption units have dominated the acid gas removal industry. Beside their undisputable technical capabilities, the preference of chemical absorption is also regarded to industry's convenience in utilizing amine technology throughout the past few decades of natural gas industry. The reasons for the reluctance of natural gas industry in adopting alternative CO₂ removal technologies under the conventional or non-conventional applications are not different from the general reluctance observed in the oil and gas industry; these reasons have been described elsewhere (Alcheikhhamdon and Hoorfar, 2016). In this paper, the “conventional application window” is defined as (a) onshore applications, (b) crude of 5–15% v/v CO₂ and (c) moderate crude gas pressure levels (30–80 bar).

Abbreviations: CA, cellulose acetate; CAPEX, capital expenditure; LNG, liquefied natural gas; MMBTU, million British thermal unit; Mscfd, thousand standard cubic feet per day; MMscfd, million standard cubic feet per day; OPEX, operating expenditure; Ppm, part per million.

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The advantages offered by the membrane units such as the low utilities demand (e.g. power), the simplified operation, and the limited footprint have not been enough to convince gas fields' developers to replace the well-established amine units by membrane units, particularly when operating in the conventional application window. In fact, the membrane units were appreciated more outside the "conventional application window" (e.g. offshore applications). In view of the expected growth in the investments pertaining to the offshore and the high CO₂ gas fields, it was earlier projected that the membrane gas separation market size in 2020 would be five times that in 2002 (Baker, 2001). Nevertheless, the 2017 estimates suggest that membrane's contribution to the total CO₂ removal market is limited to 8%, whereas the amine processes still dominate the market at 87%. The residual market (5%) is left to the cryogenic and adsorption technologies (Sorschak, 2017).

In fact, the interest of operating companies in membrane units was a consequence of the failure in accommodating amine units onboard the offshore platforms at reasonable designs/cost, noting the associated footprint and weight constrains (Araújo et al., 2017). Hence, the consideration of the membrane process for the offshore gas treatment was not a process choice, but rather the last remedy option to accomplish CO₂ removal onboard the offshore platform.

2. Background

The CO₂ removal membrane is a selective yet imperfect separation medium, which upon its placement in service under a differential pressure it preferentially permeates CO₂ over the other natural gas components. Once permeated across the membrane thickness, the CO₂ loses most of its pressure, whereas the residual gas (mostly methane) is retained at a pressure almost equal to that of the feed. The retention of the bulk natural gas at a high pressure is a crucial process target as it minimizes the extent of subsequent gas recompression, which is known to be a costly process. The fact that the CO₂ is more permeable than the other natural gas components, particularly CH₄ is aligned with the latter indicated process target. The membrane units were commercialized more than two decades ago. However, their utilization was limited due to the following:

- 1) The membranes units could demand extensive crude gas pretreatment to prevent frequent unit shutdowns; the pretreatment could be costly and footprint demanding (Koch et al., 2005).
- 2) The membrane units' application window is relatively narrow compared to the widely-adopted amine units. In onshore facilities, the membrane units are mostly employed in low gas capacity fields, and/or in fields of high CO₂ content. In offshore, units of remarkable capacities (>500 MMscfd) were installed (Isa and Azhar, 2009). However, the offshore scope is relatively limited.
- 3) The membrane units' suppliers are less available as compared to the amine units. Further, the reliable commercial membranes which are offered in the market are limited.

The membrane technology regains attention whenever operating companies show interest in investing in offshore gas fields. Several commercial polymeric CO₂ removal membranes are available in the market with comparable selectivity figures. Each of these membranes is featured with a set of distinguishing properties including permeability, resistance to fouling/plasticization, lifetime, compactness, and other distinguishing features.

Unfortunately, the polymeric membranes suffer from a trade-off between their permeability and selectivity (Robeson Limit), i.e., the development of a highly selective CO₂ removal membrane is possible, yet at the cost of depreciated permeability. The reduced permeability of the separation membrane reflects in the de-

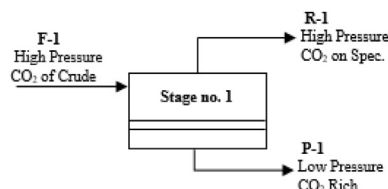


Fig. 1. Single-stage configuration.

mand for a larger membrane separation area, which in-turn reduces the market competency of the membrane process, particularly for the offshore applications. While designing the membranes, the developers do balance the trade-off such that reasonable permeability/selectivity figures are obtained. Apart from the separation capabilities, membrane suppliers also target robustness which is referred to as resilience against crudes contaminants and fouling/plasticizing constituents.

The polymeric membranes are much widely commercialized and spread compared to their non-polymeric counterparts. This is due to their satisfactory performance, low material cost and their successful history in water desalination applications (White, 2010). Table 1 lists the major polymeric CO₂ removal membranes and their properties.

3. Process design configurations

The slip of hydrocarbons along with the separated CO₂ is one of the major drawbacks resulting from the imperfect selectivity of commercial membranes (Peters et al., 2011). Unfortunately, the quantity of the methane escaping to the permeate side is significant to the extent that the process loses its economic competency. Membranes with both high permeability and selectivity (if available) would be fabricated from exotic materials; these materials may complicate the fabrication process and increase the cost. Process designers were successful in overcoming the concern of hydrocarbons slip by manipulating units' design rather than altering the membrane itself. The adjustment of the process conditions (*P,T*), and/or the installation of recovery stages were proven to be effective in controlling hydrocarbons slip (Datta and Sen, 2006).

Several membrane process configurations were elaborated in literature and implemented in industry; all these configurations aim at meeting the product gas CO₂ content, maximizing methane recovery, and minimizing the treatment cost. Despite the existence of numerous process configurations (Xu and Agrawal, 1996), those listed in Table 2 are the most encountered (Datta and Sen, 2006; Hao et al., 2008; Bhide and Stern, 1993; Ahmad et al., 2012; Qi and Henson, 1998). The configurations in Table 2 forms the outcome of the logical combination that could be potentially formed from one, two and three membrane stages.

Compared to the multi-stage processes, the single-stage process (see Fig. 1) is distinguished by its small weight and footprint, which are both crucial for offshore applications; these features are due to the lack of the recovery stage(s) and its associated inter-stage compressor(s). When conventional CO₂ removal membranes are employed, the hydrocarbons (mostly methane) losses in the single stage process (and hence the associated loss in revenues) could be significant to the extent that it is unacceptable for operating companies. It should be noted that a membrane stage does not imply single membrane unit, but rather a bank of membrane units installed in parallel, sharing common feed, product and permeate stream manifolds (Hao et al., 2008).

The two-stage process reduces the extent of the methane loss in the process, yet at the cost of increased unit footprints, weight and capital expenditure (Echt, 2017). Fig. 2 demonstrates various design configurations of the two-stage process (Datta and

Table 1Details of commercial CO₂ removal membranes.

Base Polymer	Cellulose Acetate		Polyimides	Perfluorinated
Product supplier	UOP		Schlumberger	MTR Inc.
Commercial name	Separex®		CYNARA	Z-Top
Module	Spiral wound		Hollow fibers	Spiral wound
Pure CO ₂ permeability (Barrer)	6 (Scholes et al., 2012)		11.5 (Scholes et al., 2012)	150 (Scholes et al., 2012)
Field test CO ₂ /CH ₄ selectivity	Moderate: 12–15 (Peters et al., 2011)		High: 20–25 (Baker, 2001)	Low: 6.5 (Abetz et al., 2011)

Table 2

Summary of process configurations.

Parameter	Config. tag	Process overall product	Process overall permeate	Recycle
Single stage	I	R-1	P-1	-
Two stages	II.A	R-1	P-2	R-2 to F-1
	II.B	R-2	P-1	P-2 to F-1
	II.C	R-1 + R-2	P-2	-
Three stages	III.A	R-2	P-1 + P-3	R-3 to F-1
	III.B	R-2	P-1 + P-3	R-3 to F-2
	III.C	R-2 + R-3	P-1 + P-3	-

I: Single stage
 II: Two stages
 III: Three stages
 F-X: Membrane stage no. X feed gas
 P-X: Membrane stage no. X permeate gas
 R-X: Membrane stage no. X residual gas

Sen, 2006). The main differences distinguishing the two-stage designs (illustrated in Fig. 2) are the membrane stream concentration scheme and the recycle stream allocation. For example, configurations II.A/II.C adopt “permeate concentration” whereas configuration II.B adopt “residual gas concentration”. In terms of stream recycle, configuration II.C involves no recycle stream.

The three-stage process is a further design development adopted by process designers to increase process methane recovery, yet at the cost of a further increase in the unit footprint, weight and capital expenditure. Fig. 3 demonstrates the various configurations of the three-stage process, each involving a single compression station. Configurations other than those listed in Fig. 3 exist, but were excluded from this study as they involve sev-

eral compression stations (Agrawal and Xu, 1996). The presence of multiple compressor stations is not desired industrially, noting the associated increase in unit's footprint, control scheme complexity and shutdown frequency (due to compressors failure).

In all of the configurations shown in Fig. 3, the first stage is responsible for bulk CO₂ removal, i.e. preliminary “residual gas concentration”. The permeate stream resulting from this stage is highly enriched with CO₂, and hence disposed directly with no attempt for its associated hydrocarbon recovery. The second stage is often responsible for final product quality control whereas the third stage is responsible for “permeate concentration”. The process of permeate concentration intends to maximize unit's overall hydrocarbon recovery. It should be noted that the compressors indicated in these schematics could represent more than one compressor installed in series to achieve the targeted pressure. In terms of stream recycle, configuration III.C involves no recycle stream.

Membrane units with more than three stages are rarely encountered in industry (Ohs et al., 2016; Favre, 2010). This is regarded to the footprint, extensive capital expenditure (more recycle compressors) and the complexity of such units.

4. Study objective and workflow

Several studies were conducted to evaluate the economics of the process configurations listed in Table 2 (Figs. 1–3) (Datta and Sen, 2006; Lock et al., 2015; Hao et al., 2008; Bhide and Stern, 1993; Koch et al. 2005; Rezakazemi et al., 2017). Each of these studies evaluated one or a limited number of process configurations under a narrow range of variations in crude gas CO₂ content, flowrate, crude gas pressure and natural gas market price.

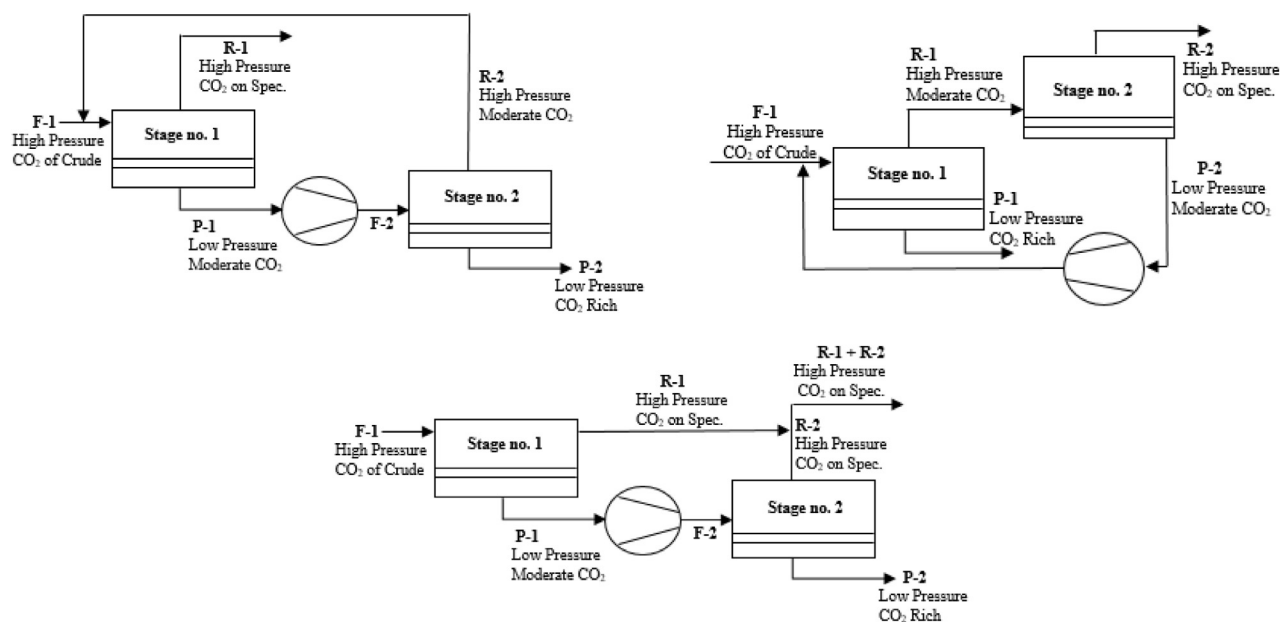


Fig. 2. Various two-stage configurations (II.A, II.B, II.C) subject to membrane function, sequencing and recycling scheme.

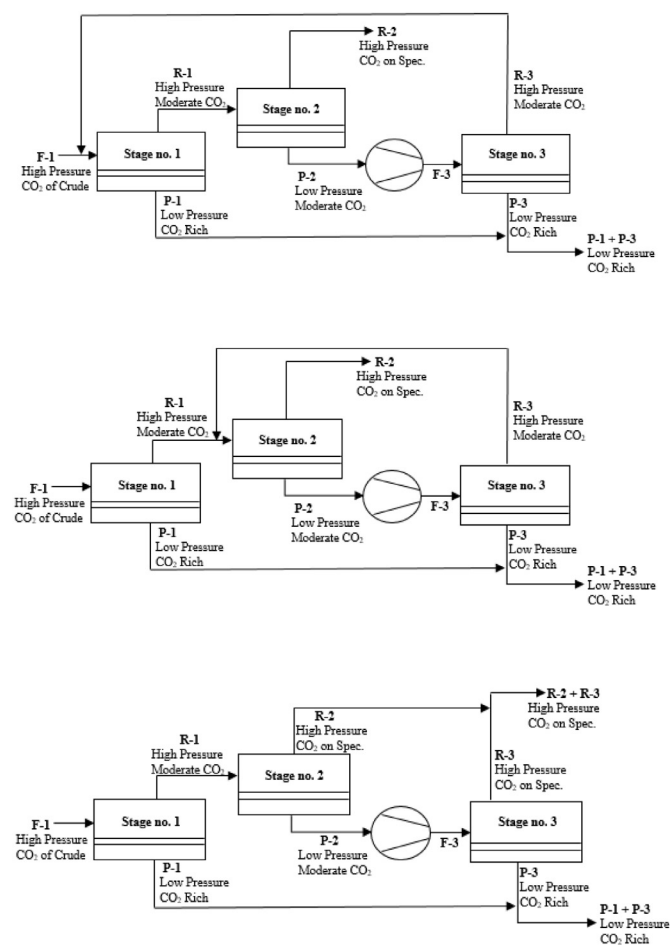


Fig. 3. Three-stage configurations (III.A, III.B, III.C) subject to membrane function, sequencing and recycling scheme.

The computational difficulties in the analyses, particularly when handling ‘multi-component’ crude gas in configurations involving recycle streams, are one of the main reasons for the lack of detailed studies in the past. This manuscript bridges the gap observed in the literature by evaluating the economics of the various process configurations listed in Table 2 (Figs. 1–3) under a wide range of process parameters. In this study, the computational limitations were overcome with the aid of a simulator which resolves the recycle operation computations, and hence the associated material/energy balances smoothly.

On the high level, the work presented in this manuscript targets the development of a guide which directs process designers towards selecting the cost optimum process configuration. The selection is performed post matching the following variables:

- Crude gas capacity
- CO₂ content
- Feed gas pressure
- Natural gas market price.

Each combination formed from the above-mentioned indicated parameters creates a “study set” that has an associated optimum process configuration. Apart from that, the model, utilized to produce the results corresponding to CO₂ removal process, must be reliable and easily tunable to model other natural gas processes (e.g. N₂ removal) in future studies.

The workflow adopted to produce the results in this manuscript is as follows:

- All the “study sets” of crude gas capacity, CO₂ content, pressure and natural gas market price are imposed on the developed models for simulation.
- An optimization process is conducted by manipulating membrane area distribution between the various stages (applicable for all multi-stage configurations), such that the targeted product CO₂ content is met at the least cost. The single-stage process is exempted from this optimization process.
- Tables are produced to consolidate the outcome of the optimization and costing analyses, illustrating the optimum process design configurations for all the “study sets”.
- The data will be analyzed to investigate the sensitivity of the treatment cost with respect to the variation in the “manipulated variables”, i.e. crude gas flowrate, CO₂ content, feed gas pressure and natural gas market price.

5. Modeling

The numerous material balances that need to be resolved simultaneously within a multi-stage membrane unit causes the prediction of unit’s performance using conventional mathematical solvers (e.g. spreadsheet) to be challenging. Further, the struggle to solve the indicated balances obstructs any intended process optimization. Hence, the development of a flexible, efficient, and user-friendly model is crucial to simulate, evaluate and optimize such processes. Multiple thermodynamic simulators (e.g. UniSim®, HYSYS, CHEMCAD, aspenONE, ProMax®, etc.) are available to model, optimize and troubleshoot natural gas conditioning operations. Unfortunately, none of these simulators (including Aspen HYSYS) offer “membranes” among their standard “process functions”. Despite this limitation, a prodigious number of researchers (Lock et al., 2015; Rezakazemi et al., 2017) were able to manage membrane units modeling. The selection of “Aspen HYSYS” is regarded to the convenience of users with the simulator, particularly those in the oil and gas industry.

The following two options are available to overcome the indicated limitation:

- Create a spreadsheet outside the simulation environment; the “external spreadsheet” has an in-built coding for membrane calculations, and exchanges input/output with the simulation. If recycle streams are involved, the exchange of data between the simulator and the external spreadsheet continues until the simulator converges.
- Create a membrane unit operation using the HYSYS “Extension” and HYSYS “Custom” features; these features allow users to develop a tailored made unit operations “within” the HYSYS environment, i.e. no exchange of data occurs outside the simulation environment.

Out of the two listed options, authors adopted option “b” due to its faster convergence, ease of data management, reduced demand for programming alteration, and the ability to utilize the in-built simulator utilities such as HYSYS “Optimizer” and “Case Study” tools to accomplish the study objectives (Lock et al., 2015). MemCal (V.0.95beta) which is a HYSYS extension was developed for this purpose; this extension is the core tool which allowed authors to model membrane units in Aspen HYSYS.

5.1. Model supporting theory

The membrane separation process is modeled based on the solution-diffusion mechanism, which is governed by the following mass transfer equation (Ahmad et al., 2012):

$$Q_i = J_i A (P_f x_i - P_p y_i) \quad (1)$$

where,

Q_i	Permeation rate of component i (mol s ⁻¹)
J_i	Component i membrane permeance (mol m ⁻² Pa ⁻¹ s ⁻¹)
A	Membrane area (m ²)
P_f	Feed side pressure (Pa)
P_p	Permeate side pressure (Pa)
x_i	Feed side component i molar fraction
y_i	Permeate side component i molar fraction
i	Component index.

In a membrane module section, the mass balance across a membrane is presented by the following:

$$F_{pi} = F_{fi} - F_{ri} \quad (2)$$

where,

F_{pi}	Permeation rate of component i (mol s ⁻¹)
F_{fi}	Feed rate of component i (mol s ⁻¹)
F_{ri}	Residual rate of component i (mol s ⁻¹).

The flow and composition of permeate and residual gas changes across the membrane area. Using Eqs. (1) and (2), the local area mass balance is presented by the following:

$$dF_{pi} = 0 - dF_{ri} = J_i dA (P_f x_i - P_p y_i) \quad (3)$$

Using Eqs. (1)–(3), the local flow rate of the individual component and thus the local composition trend can be predicted.

5.2. Solving methodology

In most membrane applications, the simultaneous solving of multi-component material balances is often regarded as a major challenge preventing units' designers from predicting membrane units' performance precisely. In fact, solving these equations analytically is not possible, which implies that these equations are to be solved numerically. Pan (1986) proposed a differential algorithm to solve the above equations using the 'shooting method'. However, this method fails to converge when the stage cut (the permeate to feed ratio) is high. This drawback prevented the adoption of Pan's method in the membrane-related process simulators. It should be noted that convergence failure does not suggest inaccuracy. In fact, results from Pan's method were found consistent with an alternative method proposed by Coker et al. (1998). The series solution-based method developed by Coker et al. (1998) is efficient in solving Boundary Value Problems (BVPs), and in overcoming the convergence obstacles experienced in Pan's model; the series solution method was adopted in MemCal.

In brief, the individual components differential equations are first discretized through upwind finite difference method (FDM). Then, the Gaussian elimination algorithm (Strang, 2016) is used to solve equations and thus map the flowrate across the membrane module (Coker et al., 1998). The adopted algorithm is robust, efficient, and converges at the stage cuts as high as 99.9%. The convergence was determined by an error function with a default composition calculation tolerance of 0.001%.

5.3. Coding and integration

The algorithms were first coded in Visual Basic (VB.NET), then compiled into a Dynamic Link Library (DLL) file. The DLL file communicates with HYSYS through a Component Object Model (COM), which allows the exchange of data from and to HYSYS. Further, an External Definition File (EDF) was established to define the data and user-interface. The interface was designed such that it is user-friendly, enabling easy and quick modeling of membrane processes. The extension was denoted "MemCal"; the module icon and user interface are shown in Figs. 4 and 5, respectively.

As demonstrated in Fig. 5, the interface allows the users to manipulate module's configuration (hollow fiber/spiral wound), flow

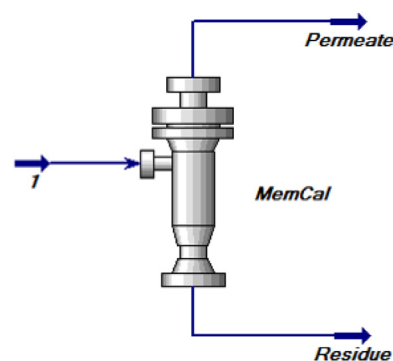


Fig. 4. Membrane stage representation in HYSYS flow diagram.

pattern (co-current/counter-current), membrane area, membrane gas permeance values and permeate pressure. The algorithms running in the background vary according to the user's selections. It is worth highlighting that in the current version (V.0.95beta) of MemCal, module's configuration selection (hollow fiber/spiral wound) does not impose a certain flow pattern as this is left to the user's selection. In the future MemCal version, it is planned to link the co-current/counter-current options to the hollow fiber configuration, and link the cross-flow mode (not in current version) to the spiral-wound configuration; the described changes causes the tool to be more realistic. In the current MemCal version, the selection of module's configuration only alters the method followed in calculating unit's pressure drop.

Unlike the other developed models which support binary mixtures modeling only, Fig. 5 illustrates the ability of MemCal to support the simulation of an endless number of components. Given that most of the industrial applications including natural gas treatment are of multi-component (not binary) natures, MemCal can support their modeling, including but not limited to acid gas removal applications. The substitution of the approximated binary models with the multi-component models definitely results in more realistic designs. This improvement reduces the extent of design margins which are often adopted by designers to cover the uncertainties arising from inaccurate simulation predictions. Furthermore, an accurate simulation moderates the requirement for establishing a pilot unit to verify the success of a separation process prior to its scale up (Lock et al., 2015). It is also worth indicating that apart from the constant numerical figures, MemCal allows users to enter components' permeance in the form of pressure/temperature functions. The permeance function is to be configured in a standard spreadsheet within the HYSYS simulation environment, and the numerical outcome of it is to be exported to MemCal. This feature is particularly important for simulating the effect of membrane's plasticization on unit's overall performance.

5.4. MemCal model validation

MemCal was validated by comparing its predictions with the outcome of an experimental setup that utilizes asymmetric cellulose acetate membrane for the recovery of hydrogen from acid gases. The series solution (Coker et al., 1998) adopted in MemCal showed an almost perfect overlap with the experimental data across the entire stage cut range. Alternatively, Davis's "Approximate Method" (Davis, 2002) which replaces the complex differential equations with logarithmic mean difference resulted in poor prediction of unit's performance, particularly at high stage cuts. It is worth highlighting that the selection of 30–60% as the stage cut range for the comparison in Fig. 6 is regarded to the experimental data's availability. However, the accuracy of MemCal holds over the entire stage cut range (0–99%).

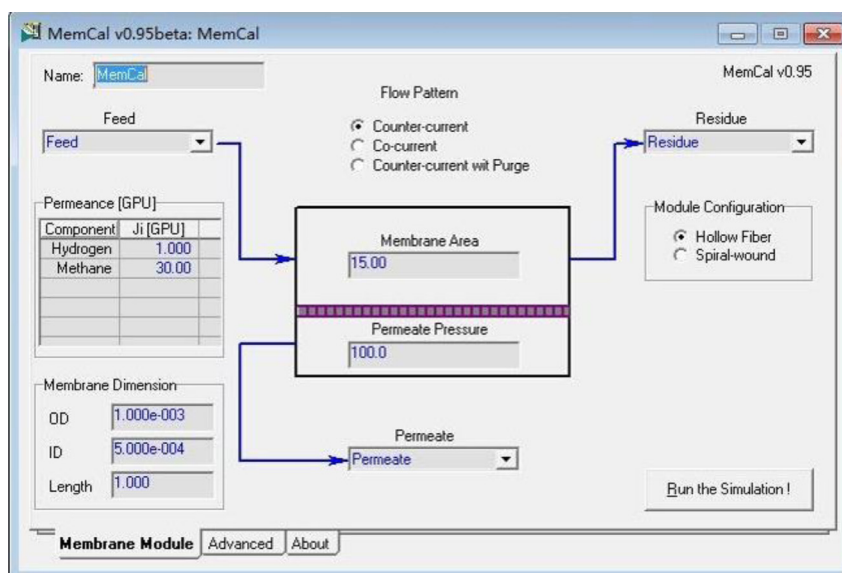


Fig. 5. Membrane unit interface.

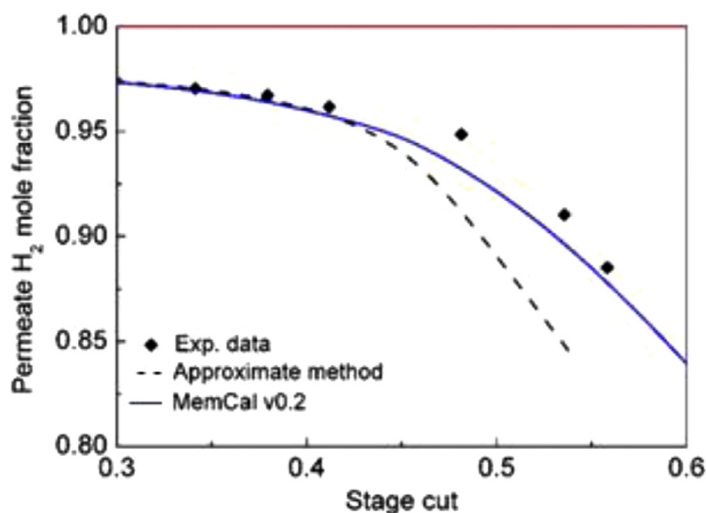


Fig. 6. MemCal model data validation.

To verify the correct implementation of the series solution equations into MemCal, the results presented by Coker *et al.* for the “ternary separation” case (Coker *et al.*, 1998) were compared with the outcome of MemCal. Fig. 7 demonstrates the composition profile across the membrane fiber length as predicted by MemCal. The same profile was predicted by Coker *et al.* (1998), i.e. the results were reproduced successfully. The slight offset between the results is justified by the calculation tolerance adopted in MemCal, which can be tuned by the MemCal user at the cost of an increased convergence time.

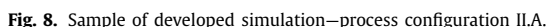
5.5. Simulation development

Apart from MemCal, the Aspen HYSYS simulation model was enhanced by utilizing the conventional unit operations available in Aspen HYSYS (e.g. compressors, coolers, mixers, etc.) and readily-available recycle blocks (where applicable). The Peng–Robinson fluid package was selected to map the thermodynamics of the CO₂ removal process, noting its suitability for the selected gases. Fig. 8 demonstrates sample of the performed simulations.

For comparing the configurations and reducing the errors arising from different changes applied to different process configuration files, all the configurations indicated in Table 2 where modeled in a single Aspen HYSYS file (case). Furthermore, an input/output interface dashboard (spreadsheet) was created to retrieve all the critical input/output parameters from the simulation, allowing for inputs manipulation and outputs monitoring without the need for entering the individual configuration's simulation environment. This interface is crucial as it allows users to compare and analyze the results, particularly during the sensitivity analyses exercise. Also, a costing model was imposed on the simulation; the costing spreadsheet imports the required simulation outputs from the respective configuration simulation to estimate its associated gas treatment cost. The costing equations are described in the following section.

6. Costing scope, methodology and envelope

The reduction of the gas treatment cost is a primary objective for all natural gas producers. The subject gains more attention as the natural gas market price drops, noting that the operating com-



The product loss contribution to the treatment cost depreciates as the number of recovery stages increases. Unfortunately, the increase in the recovery stages is accompanied with an unavoidable CAPEX/OPEX increase and an increase in unit footprint. The treatment cost in this manuscript is expressed in terms of the cost spent to produce a thousand standard cubic feet of methane gas (US\$/MSCF of methane product), upon dropping the crude gas CO₂ content from its arrival value down to the required specifications. As an observation, some of the previous manuscripts (Bhide and

Table 3
Treatment cost computation basis.

Cost description	Tag	Value
<i>Capital expenditure (CAPEX)</i>		
First membrane module installation	MC	Unit area cost (\$/m ²)
Compression system	CC	$8650 \times (\text{horse power}/\mu)^{0.82}$
Fixed cost	FC	$MC + CC + PTC$
Base unit cost	BPC	$1.12 \times FC$
Project contingency	PC	$0.20 \times BPC$
Total fixed investment	TFI	$BPC + PC$
Start-up cost	SC	$0.1 \times VOM$
Total unit investment	TUI	$TFI + SC$
Annual capital related cost	CRC	$A = A/P$ (TUI, interest%, plant life)
<i>Operational expenditure (OPEX)</i>		
Annual material maintenance	CMC	$0.05 \times TFI$
Annual local tax and insurance	LTI	$0.015 \times TFI$
Annual direct labor	DL	No. of operators \times operator annual rate No. of operators = 1 per 25 MMscfd
Annual labor overhead	LOC	$1.15 \times DL$
Annual membrane replacement	MRC	Unit area cost (\$/m ²)
Annual variable operating and maintenance	VOM	$CMC + LTI + DL + LOC + MRC + UC$
<i>Product loss</i>		
Annual value of energy lost in permeate	CH ₄ LS	(Annual energy lost in permeate) \times (gas energy market price)
<i>Treatment cost</i>		
Gas processing cost (\$/MSCF)	GPC	$(CRC + VOM + CH_4LS)/(\text{annual gas production})$

heating value is the most appropriate and fair measure for evaluation of the transacted gas quality. In fact, most of the natural gas contracts and financial transactions are performed based on the gas gross heating value.

The “treatment cost” quoted in this manuscript accounts for the differential cost arising from the presence of the CO₂ removal unit; this includes the membrane unit and its supporting systems (e.g. recovery compressors). No value was considered for the resulting unit permeate as it is assumed to be flared. The pre-treatment unit cost is excluded from the cost calculations, noting that the uncertainties in the pre-treatment unit design is subject to the contaminants present in the crude. Further, the main infrastructure and the supporting utilities which are installed irrespective of the CO₂ removal unit presence will not be accounted for in the analysis. Apart from the spare membrane area, the costing considers no further sparing within the unit; this is particularly important for the costly items (e.g. recycle compressors) which influence units’ CAPEX extensively.

6.3. Costing study envelope

The factors influencing the treatment cost are numerous. Table 4 summarizes these factors and specifies the values or the ranges adopted to generate the 96 “study sets” (combinations) considered in this work. Each “study set” will be imposed on the simulation corresponding to the configurations described in Table 2. The simulation outcome is then exported to the costing spreadsheet configured within the simulator to calculate the treatment costs tabulated in Table 5.

Subject to the treated gas ultimate usage, the treatment process design targets producing gas of either 2–3% CO₂ v/v (US pipeline standard), or 50–100 ppmv CO₂ (Foss, 2004; Klinkenbijn et al., 1999). Given that natural gas is mostly used for power generation or heating applications, and in view of the failure of membrane units in achieving the “ppm” levels of CO₂ at reasonable areas, the costing analyses in this manuscript will be conducted based on a fixed targeted CO₂ content which is 2% CO₂ v/v. As for the crude gas composition, the study is conducted considering a binary CO₂/CH₄ mixture with CO₂ concentrations ranging from 5% to 40%

Stern, 1993; Rezakazemi et al., 2017) expressed the treatment cost per unit volume of crude rather than product; such expression cannot be used for cost comparison purposes, particularly when the compared crudes possess different composition, and the extent of CO₂ removal is high. For the cost comparison to be fair, the analysis must be conducted on an equal basis, i.e. must be product based (Hao et al., 2008).

When multiple hydrocarbons are present in the crude, it is more sensible to normalize the treatment cost to the overall product gas gross heating value (US\$/ MMBTU), noting that the gross

Table 4
Treatment cost impacting factors.

Parameter	Value/range	Increment	No. of analyzed data points	Unit
<i>Manipulated variables forming a “study set”</i>				
Crude flow	50–500	225	3	MMscfd
Crude CO ₂ content	5–40 (residual is CH ₄)	5	8	v/v %
Crude pressure	30–90	60	2	Barg
Energy market price	3–5	2	2	US\$/MMBTU
<i>Fixed variables used in the “study sets”</i>				
Location	Onshore	Not applicable	Not applicable	–
Crude temperature	50			degC
Membrane material/configuration	CTA/hollow fiber			–
Flow configuration	Counter-current			
Membrane CO ₂ permeance	50 (Baker, 2001)			GPU
Membrane CH ₄ permeance	3.33 (Baker, 2001)			GPU
Membrane selectivity	15 (Baker, 2001)			–
Permeate pressure	1.0			Barg
Sales gas CO ₂	2			v/v%
Plant life time	20			Years
Plant availability	95			%
Labor cost	40			\$/h
Investment interest rate	10			%
Membrane life time	4			Years
First membrane module cost	500			\$/m ²
Membrane replacement cost	100			\$/m ²
Recycle comp. efficiency	75			%
Recycle comp. turbine efficiency	30			%
Spare membrane modules	20%			% of calculated area
Membrane disposal cost	0			\$/m ²

Table 5
Optimum configuration/treatment costs of study sets.

Crude gas pressure	% CO ₂ in crude (v/v)	Crude flow 50 MMscfd	Crude flow 275 MMscfd	Crude flow 500 MMscfd
30 barg	<i>Treatment cost (US\$/MSCF of methane) at 3 US\$/MMBTU</i>			
	<i>Treatment cost (US\$/MSCF of methane) at 5 US\$/MMBTU</i>			
	5%	Optimum configurations II.A, minor advantage over II.C		
		0.48	0.46	0.45
		0.53	0.52	0.51
	10%	Optimum configurations II.A, minor advantage over II.C		
		0.84	0.81	0.80
		0.95	0.92	0.91
	15%	Optimum configurations II.A, minor advantage over II.C		
		1.00	0.95	0.94
		1.07	1.02	1.00
	20%	Optimum configurations II.A, minor advantage over II.B		
		1.21	1.16	1.14
		1.33	1.27	1.26
	25%	Optimum configurations II.A, minor advantage over II.B		
		1.37	1.30	1.28
		1.46	1.38	1.36
	30%	Optimum configurations II.A, minor advantage over II.B		
		1.53	1.45	1.43
		1.60	1.55	1.52
	35%	3 \$/MMBTU: II.B offers minor cost advantage over II.A 5 \$/MMBTU: II.A offers minor cost advantage over II.B		
		1.69	1.64	1.63
		1.93	1.84	1.82
	40%	3 \$/MMBTU: II.B offers minor cost advantage over II.A 5 \$/MMBTU: II.A offers minor cost advantage over II.B		
		1.80	1.75	1.74
		2.07	1.94	1.91
90 barg	<i>Treatment cost (US\$/MSCF of methane) at 3 US\$/MMBTU</i>			
	<i>Treatment cost (US\$/MSCF of methane) at 5 US\$/MMBTU</i>			
	5%	Optimum configurations II.A, minor advantage over II.C		
		0.23	0.22	0.21
		0.27	0.25	0.24
	10%	Optimum configurations II.A, minor advantage over II.C		
		0.41	0.39	0.37
		0.48	0.46	0.44
	15%	Optimum configurations II.A, minor advantage over II.C		
		0.50	0.48	0.44
		0.55	0.52	0.50
	20%	Optimum configurations II.A, minor advantage over II.B		
		0.61	0.53	0.51
		0.66	0.58	0.56
	25%	Optimum configurations II.A, minor advantage over II.B		
		0.71	0.63	0.61
		0.78	0.70	0.68
	30%	Optimum configurations II.A, minor advantage over II.B		
		0.85	0.75	0.73
		0.96	0.86	0.84
	35%	Optimum configurations II.A, minor advantage over II.B		
		0.95	0.84	0.81
		1.06	0.95	0.92
	40%	3 \$/MMBTU: II.B offers minor cost advantage over II.A 5 \$/MMBTU: II.A offers minor cost advantage over II.B		
		1.01	0.91	0.88
		1.12	1.04	0.99

v/v; this range is within the typical crude gas CO₂ content encountered in natural gas fields worldwide (Qi and Henson, 1998). The industrial records for the treatment of crudes with the CO₂ content higher than 45% v/v is limited, noting that the production of such a gas would be accompanied with technical and possibly commercial concerns (Isa and Azhar, 2009; Finn and Brien, 2014). The presence of other natural gas components (e.g. N₂, H₂S, heavy hydrocarbons) is often associated with design complications and an increased CO₂ separation cost (Rezazakemi et al., 2017). The simulation and costing of multi-component (more than two) systems are planned in future studies; it is worth indicating that “MemCal” supports such simulation work.

Unlike the amine absorption process during which the crude gas pressure has limited influence on unit's performance, the membrane system's performance is highly influenced by its oper-

ating pressure (Bhide and Voskericyan, 1998). The pressure difference across the membrane dictates unit's footprint, and influences units' methane losses. The crude gas pressure window selected for the costing analyses is 30–90 barg; the latter indicated range covers the typical pressures (20–70 barg) encountered at the treatment facilities' inlet (Qi and Henson, 1998). In this study, and irrespective of the simulated configuration, the permeate pressure is uniformly set at 1 barg; such low setting ensures maximizing the CO₂ permeation potential and minimizing the membrane area, which in turn reflects positively on unit's methane recovery. It is understood that the drop in the permeate pressure increases the compression system's stage count and power. However, the long term (say 20 years) benefits illustrated by the boost of methane recovery often overcomes the compression unit's CAPEX/OPEX escalation. The study assumes no footprint limitation for fitting the

compression system which could constitute of two or three stages. The latter indicated assumption does not hold in offshore applications where the footprint is very restricted. When the permeate side is operated at 1 barg (2 bar absolute), the pressure ratios corresponding to crude gas pressures of 30 and 90 barg would be ~ 15 and 45, respectively. The latter indicated ratios are calculated based on the absolute pressure values across the membrane. A pressure ratio of 15 is well within industry's norms, whereas a pressure ratio of 45 is above that. Nevertheless, it is assumed that the mechanical strength of the used fibers supports the indicated pressure ratios.

The simulations are conducted based on the separation properties (permeability and selectivity) pertaining to the conventional Cellulose Acetate (CA) membranes (see Table 4); it is assumed that neither CO₂ concentration nor operating pressure impacts these properties. The CA membranes dominate the current acid gas removal market despite their imperfect separation capability. Coupled with an appropriate pretreatment and an optimized process design, the performance of CA units could be made very satisfactory. The estimates (Tennyson and Schaaf, 1977; Scholes et al., 2012) suggest that 80% of the natural gas treatment membranes installed worldwide are CA based. Given that this research intends to support the current and the near future market with a process configuration selection guide, it is logical to perform the analyses considering the properties associated with the currently used CA membranes.

Apart from the process parameters, market parameters also do influence the treatment cost of a gas. The annual contribution of the product loss in the treatment cost is quantified based on the value of the energy dissipated with the hydrocarbons in membranes' non-recoverable permeate stream(s). The costing analyses in this manuscript consider 3–5 US\$/MMBTU as the gas energy market price; the latter indicated price range was recorded when the cost of a crude oil barrel ranged from 40 to 100 US\$ (between 2017 and 2013, respectively) (NASDAQ, 2017).

7. Treatment cost analyses outcome

This section demonstrates the outcome of the analysis performed to cost the operation of purifying the natural gas from its CO₂ content. To cover all the process configurations, a total of 672 simulations were performed. These simulation runs correspond to the imposition of the 96 “study sets” generated from Table 4 on the 7 process configurations described in Table 2. For each of the study sets, an optimization process was conducted to control the distribution of the areas between the various membrane stages in a particular process configuration. The latter indicated optimization aims at attaining the targeted product quality at the least treatment cost. Post the completion of the optimization process (single stage exempted), the costing tool built within the simulation environment selects the optimum process configuration for the imposed “study set”. Table 5 demonstrates the outcome of the above-mentioned analysis. For each of the study sets, the treatment cost is expressed in US\$/MSCF of the produced methane; the optimum process configuration for the “study set” is also tabulated.

It should be noted that the cost figures quoted in Table 5 vary subject to the case-specific design variables (e.g. targeted product gas CO₂ contents). Apart from the cost figures, it is crucial for process designers to understand the impact and the sensitivity of altering various design parameters on the treatment cost of a gas, particularly those listed under the “manipulated variables” in Table 4 (i.e. crude gas flow, CO₂ content, pressure and market gas price). Section 8 elaborates the obtained results and demonstrates the indicated sensitivity.

Table 6

Example illustrating the importance of area optimization.

Parameter	Area set 1	Area set 2	Area set 3
Configuration	III.A		
Crude gas flow	50 MMscfd		
Crude pressure	30 barg		
Crude CO ₂ content	20% v/v		
Product CO ₂ content	2.0% v/v		
Gas market price	3.0 US\$/MMBTU		
Stage 1 area	11,000 m ²	11,000 m ²	1,000 m ²
Stage 2 area	62,000 m ²	41,000 m ²	51,000 m ²
Stage 3 area	1000 m ²	11,000 m ²	26,000 m ²
Compression power	3.8 MW	2.1 MW	2.9 MW
Methane recovery	94.6%	88.9%	86.0%
Treatment cost	1.29 \$/Mscf	1.34 \$/Mscf	1.72 \$/Mscf

8. Results and discussion

8.1. Optimum configuration analyses

The figures in Table 5 suggest that when conventional commercial membranes are utilized, the single stage configuration offers absolutely no economic advantages over multi-stage configurations, irrespective of the evaluated “study set”. This is due to the significant and irrecoverable methane losses. In such systems, methane losses contribute most to the total treatment cost (>50%). Hence, the adoption of the single-stage design is not recommended, unless the footprint or weight limitation dictates that. In addition, it is observed that the competency of the single-stage configuration (against multi-stage configurations) deteriorates as the crude gas CO₂ content increases (see Fig. 9).

The CO₂ rich crudes demand a larger separation area to achieve the targeted gas quality, which in turn increases the likelihood of methane slip, and subsequently amplifies the gas treatment cost. In fact, this is applicable to both single and multi-stage configurations. However, the methane slipped in multi-stage configurations is partially recoverable; whereas that slipped in single-stage processes is not at all recoverable. The latter justifies the steep cost trend observed for single-stage units compared to that of multi-stage (see Fig. 9). Oppositely, when the crude gas CO₂ content (say 3% v/v) is close to the targeted gas quality, the membrane area demand will be minimal, and thus the lost methane quantity will be insignificant to justify the installation of a recovery stages. This makes the single-stage process more appealing to unit's designers when the crude is of low CO₂ content; noting the simplicity and the low CAPEX/OPEX of such units.

As for the multi-stages units, and prior to the conclusion of a particular configuration being the “optimum”, process designers should subject the unit to an optimization process to control the area assigned to each stage. In a certain configuration the targeted separation can be achieved using various stages' area sets, each resulting in a different treatment cost. To further clarify the above statement an example is illustrated in Table 6.

In fact, the above-mentioned optimization is one of the highlights of this manuscript since none of the earlier published papers addressed this subject thoroughly; the published literature granted more attention to the effect of individual stage area variation rather than the distribution of separation load among the installed stages (Colling et al., 2004; Narayan and Patton, 2004). The control of area distribution across the various stages within a membrane unit is analogous to the “Pinch Analysis” conducted in systems containing multiple heat exchangers, where the distribution of heat transfer areas across the exchangers is optimized to maximize unit's heat integration. Unfortunately, no analytical solution is available so far to predict the optimum distribution of areas across the multiple membrane stages, which in turn implies

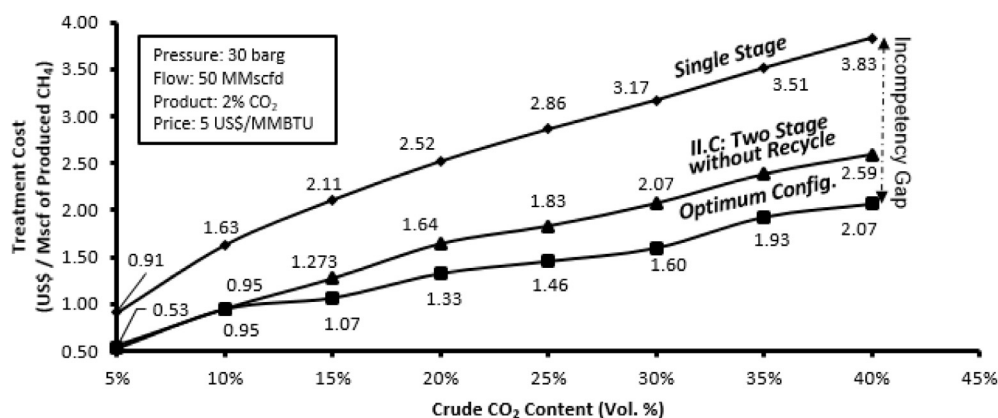


Fig. 9. Competency gap analysis demonstrating the cost differences between various configurations.

that such an optimization is best performed by “trial and error”. The “Case Study” feature available in HYSYS along with the developed tool “MemCal” supports this exercise. Post defining the lower/upper bound for each stage area (manipulated variable), the “Case Study” tool generates all the possible area combinations and subsequently apply them to the respective process configuration model. The upper bound is often selected to be 3 times the area required to achieve the targeted separation using a single-stage unit, whereas the step size is defined such that no more than 1000 area combinations per configuration are obtained. The generated results are then exported to a spreadsheet where the non-converged runs or the combinations failing to achieve the targeted separation (CO_2 content = 2% v/v) are disqualified. The succeeding area combinations are then analyzed for judging the optimum among them. This process is followed for all six multi-stage configurations prior to identifying the “optimum” configuration. To optimize the treatment cost, the “Case Study” function (sensitivity-based) was selected over the “Optimizer” function (gradient-based) since the second was associated with some convergence issues; both functions are part of HYSYS in-built features.

Upon subjecting the multi-stage units to the above-mentioned optimization, the configurations with no recycle (II.C/III.C) were found economically incompetent to their counterparts with recycle (II.A/II.B/III.A/III.B); the incompetency gap widens as the crude gas CO_2 content increases (see Fig. 9). The minor benefits resulting from the absence of the recycle stream in units II.C/III.C (e.g. ease of simulation) do not often offset their noted economic incompetency, particularly when the crude is rich in CO_2 . Among the residual multi-stage configurations (II.A/II.B/III.A/III.B), the three-stage configurations (III.A/III.B) were found to offer no or at the best insignificant cost advantages over the two-stage configurations (II.A/II.B). This observation implies that when conventional CO_2 removal membranes are utilized, no more than two stages are required to achieve cost effective separation, noting the installation of a third stage overwhelms the unit with additional footprint and unjustified CAPEX. The differences in the treatment cost between configurations II.A and II.B were found insignificant.

8.2. Variations in process design inputs

The following sections illustrate the impact of the variations made in the main process design inputs on the gas treatment cost.

8.2.1. Crude gas flow

The chemical processes’ costing basics suggest that the “capacity normalized CAPEX” of an equipment depreciates as its capacity increases (Turton et al., 2009). As a rule of thumb, most of the

equipment encountered in chemical processes conform to the sixteenth rule: $\text{CAPEX} \propto (\text{Capacity})^{0.6}$ (Turton et al., 2009). Unfortunately, the capacity advantage is not applicable to the membrane units, wherein the membrane area/flow relation is linear: $\text{CAPEX} \propto (\text{Capacity})^1$. The latter indicated linearity is demonstrated best in the single-stage configuration trends (see Fig. 10), where the scale-up advantages offered by multi-stage configurations’ equipment (e.g. recycle compressors) are isolated. The linear membrane area/flow demand (Eq. (1) in Section 5.1), and the linear membrane procurement unit rate (\$/m²) both justify the independence of the treatment cost from the crude gas flow. On the other hand, the multi-stage configurations experience a minor cost advantage at higher throughputs. In fact, this advantage is not regarded to the membrane modules, but rather to the other equipment present in the unit (mainly recycle compressors). This observation is consistent with other relevant studies’ outcome (Hao et al., 2008; Bhide and Stern, 1993).

Among the multi-stage processes listed in Table 2, process configuration II.A was selected to demonstrate the treatment cost/crude gas flow relation, noting that this configuration is proven to be the cost optimum for most of the “study sets”; similar trends are obtained upon simulating other “study sets”.

8.2.2. Crude gas CO_2 content

The values in Table 5 suggest that the treatment cost of a gas is highly influenced by its CO_2 content. To interpret the relation, units’ designers should first understand the correlation between the separation area requirement and the crude gas CO_2 content. The following equation (rearrangement of Eq. (1)) suggests that

$$A = Q_{\text{CO}_2} / [J_{\text{CO}_2} (P_f x_{\text{CO}_2} - P_p y_{\text{CO}_2})] \quad (4)$$

- The area required for separation is directly proportional to the amount of CO_2 that must be permeated to achieve the targeted gas quality. At a particular crude gas rate, this amount increases as the CO_2 content in the crude gas intensifies.
- The area required for separation is inversely proportional to the CO_2 partial pressure difference experienced across the membrane (Echt, 2017). At a fixed crude gas pressure, this difference intensifies as the crude gas CO_2 content increases.

The above points suggest that the enrichment of the crude gas with CO_2 can drive the membrane area to a higher or lower value depending on the resultant balance. To isolate the other factors influencing the quantification of the membrane area requirement, this analysis is demonstrated best in single-stage configuration, see Fig. 11.

The membrane area trend in Fig. 11 can be used as an indirect mean to qualitatively predict the trend of methane losses (see

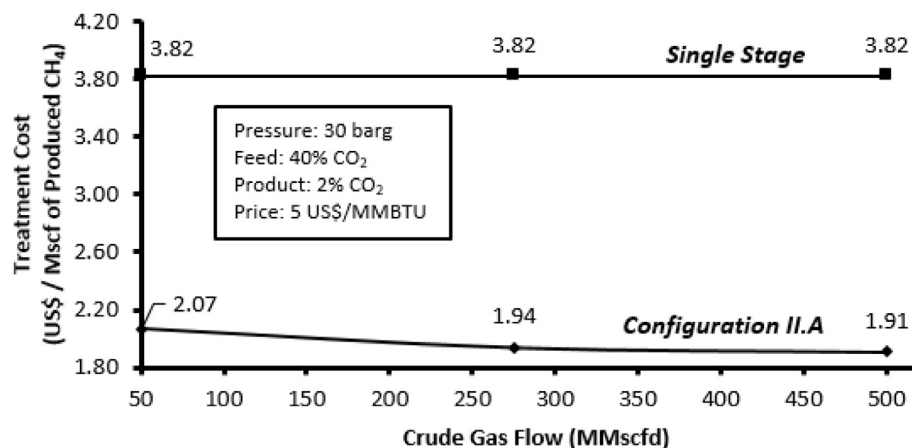
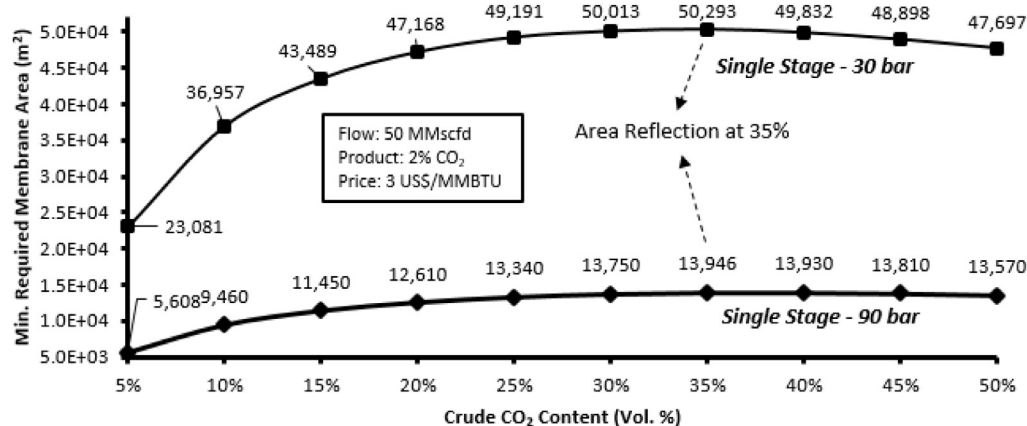
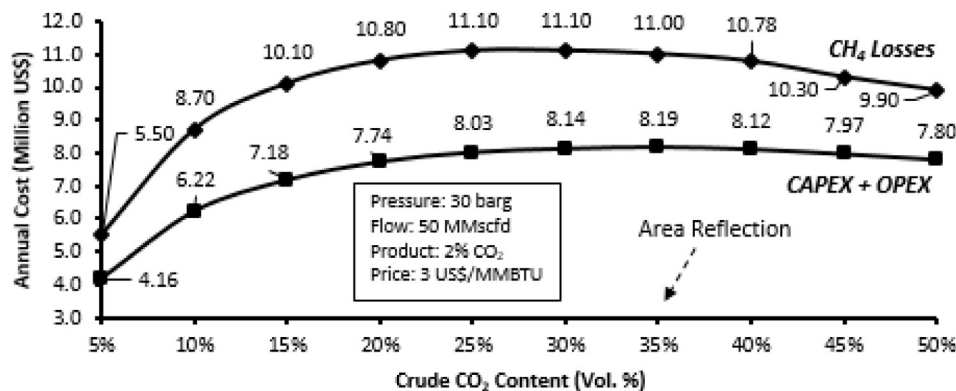


Fig. 10. Impact of crude gas flow on gas treatment cost.

Fig. 11. Impact of crude gas CO₂ content on membrane area.Fig. 12. Impact of crude gas CO₂ content on treatment cost contributors.

the elaboration in Section 8.1). The membrane area trend can also infer unit's CAPEX/OPEX trend, noting that the single-stage unit's CAPEX is dominated by the area dependent membrane cost. Fig. 12 demonstrates the latter indicated correlations.

The observed drop in unit's CAPEX, OPEX and methane losses post the area flip point (35% in Fig. 12) is not reflected in the flow normalized treatment cost which sustains its proportionality (see Fig. 9). At a fixed crude gas flow (e.g. 50 MMscfd), the net methane in the crude gas reduces as the crude gas CO₂% increases, causing less amount of net methane to be available for later recovery as a product. The latter implies that the reduced CAPEX, OPEX and methane losses (post area reflection) will distribute over

lesser amount of net produced methane, causing the normalized cost to be in the favor of increasing the treatment cost (US\$/Mscf of produced CH₄). As for the multi-stage configurations, the trend in Fig. 13 suggest that as the crude gas CO₂ containment increases, the treatment cost increases yet at a decreasing rate.

8.2.3. Crude gas pressure

It is important for process designers to understand the dependency of membrane unit's CAPEX, OPEX and methane losses on the variations in the crude gas pressure. The impact of the crude gas pressure on unit's CAPEX in membrane systems is opposite to the trend observed in conventional process systems, where the high

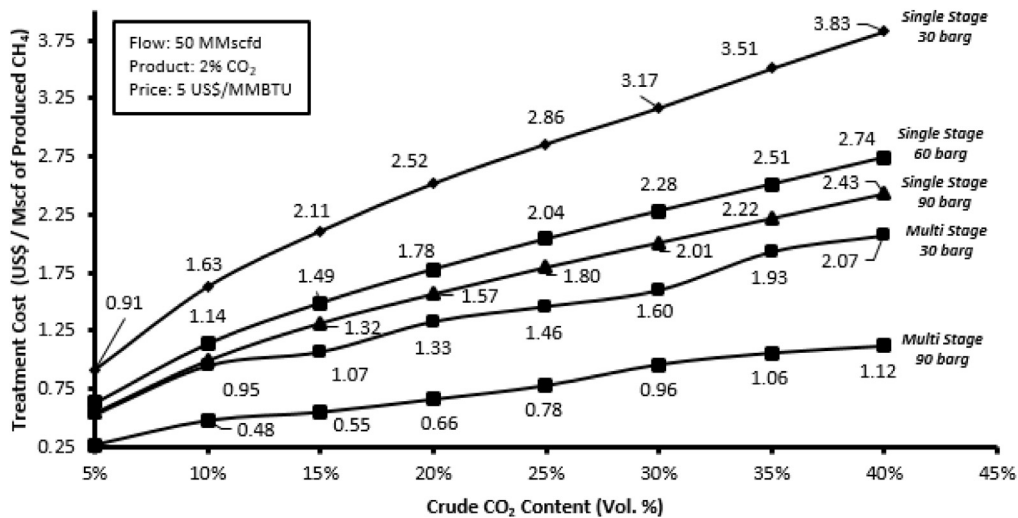


Fig. 13. Effect of crude gas pressure/crude gas CO₂ content on treatment cost.

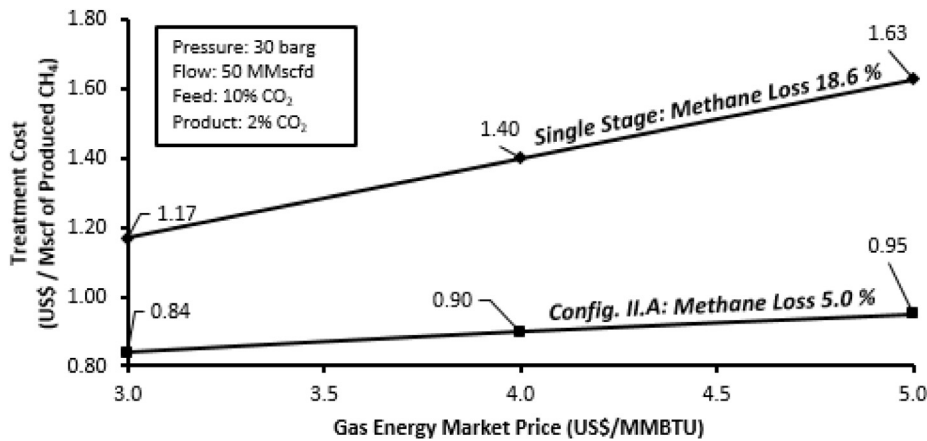


Fig. 14. Impact of energy market price on gas treatment cost.

pressure is often associated with thicker vessels and thus costlier units. In membrane units, the CAPEX which is dominated by the membrane cost drops as the crude gas pressure increases (see Fig. 13); this is regarded to the associated increase in CO₂ permeation potential (partial pressure boost), which in turn confines the membrane area required for separation. The OPEX is often associated with the CAPEX, and thus similar trends apply. The positive effect of the pressure on unit's CAPEX/OPEX is subject to the crude gas being readily available from the producing wells at a high pressure, implying the absence of the feed gas compression requirement. Compared to the low capacity compressors installed within the membrane unit, the high capacity crude gas compressors are costly and footprint demanding (Bounaceur et al., 2006). As for the methane losses, it is worth stating that the boost in the permeation potential upon the pressure increase is not limited to CO₂, but rather extended to the methane gas which its permeation is undesired. Luckily, the drop in the required separation area often overcomes the boost in the methane permeation potential, causing the net permeation rate of methane (losses) to drop as the crude gas pressure increases. The latter elaboration endorses the outcome of the previous simulation/costing studies in (Hao et al., 2008).

In a multi-stage unit, the treatment cost gains (when the unit is operated at a higher pressure) are mainly regarded to the savings made in unit's CAPEX/OPEX. This is unlike the single-stage units in

which the cost gains are primarily due to the noted drop in unit's methane losses. Irrespective of the configuration, the positive impact of the crude gas pressure on the treatment cost is observed to be the most when the unit is originally operating at a low pressure.

The performed analysis assumes that membrane performance is sustained at high crude gas pressures. Unfortunately, the increase in the crude gas pressure can result in the deterioration of membrane performance, subject to the presence of plasticization promoting components (e.g. aromatics); membrane plasticization intensifies methane losses and subsequently results in increasing the treatment cost. It is worth highlighting that MemCal supports the simulation of plasticization effects, noting that components' permeance values can be tabulated in the form of pressure equations; the latter indicated feature promotes accurate modeling of membrane units.

8.2.4. Gas energy market price

Beside CAPEX and OPEX, Section 6.2 lists "Product Loss" among the main treatment cost contributors. The non-intended permeation of methane along with CO₂ (due to membranes' imperfect selectivity) could be significant. The value of the dissipated energy is equivalent to the revenues which would have been acquired by the operating companies if the slipped methane was preserved in the product gas. Thus, the value of the lost methane and subse-

quently the crude treatment cost are directly proportional to the gas energy market price.

The trends in Fig. 14 suggest that the treatment cost in single-stage units is highly dependent on and sensitive towards the variations in the gas energy market price; this is regarded to the fact that methane losses in such units are significant and irrecoverable, forming >50% of the total treatment cost. The latter indicated relation is not observed in multi-stage units where the “product loss” contribution to the total treatment cost is minor (<10%). In fact, the stable treatment cost observed in the multi-stage units is advantageous, as it boosts investor’s confidence in the economic viability of a project, noting that the risks associated with the future treatment cost increase and thus profit loss are diluted.

8.2.5. Other sensitivities

Membrane unit price (expressed in \$/m²) is another treatment cost influencing factor, providing that membrane first installation and periodic replacements impact units’ CAPEX and OPEX, respectively. Despite the proven impact on the treatment cost, the analysis showed that the membrane unit price has no influence on the conclusion of the “optimum” process configuration.

9. Conclusions

“MemCal” is a vital HYSYS extension that allows process designers to simulate, optimize and evaluate the cost of complex multi-stage membrane processes. In this manuscript, “MemCal” was utilized to simulate and subsequently evaluate the cost of the CO₂ removal process, where the crude gas CO₂ content was dropped from 5–40% v/v down to 2% v/v. When the conventional CA membranes are used, configuration II.A offered the least treatment cost among the other evaluated configurations for most of the evaluated “study sets”. Thus, proposals of adopting three-stage design to boost unit’s methane loss were found economically unjustified. The control of membrane area distribution between the stages was proven to be a critical factor in optimizing units’ treatment cost. “MemCal” empowers such optimization processes, particularly for the processes involving recycle streams, noting that their convergence is proven difficult.

Furthermore, it was demonstrated that unit’s capacity scale-up has an insignificant impact on the gas treatment cost. Oppositely, the increase of the crude gas pressure has a notable positive impact. Also, it was observed that the treatment cost in multi-stage units are not much sensitive to the variations in the gas energy market price; this is unlike the single-stage units for which the corresponding treatment cost is very sensitive to such variations. Finally, the analysis of the optimum configuration was found to be independent of the membrane unit price.

Considering the potential of “MemCal”, this tool can be utilized to model and optimize other natural gas separation applications; the standalone or simultaneous removal of H₂S, N₂, aromatics and CO₂ from multi-component natural gas mixtures are examples of the intended future work.

References

- Abetz, V., Brinkmann, T., Shishatskiy, S., Wind, J., 2011. Polymer membranes for separation of CO₂—an overview. 2nd International Conference on Energy Process Engineering—Efficient Carbon Capture for Coal Power Plants.
- Agrawal, R., Xu, J., 1996. Gas separation membrane cascades II. Two-compressor cascades. *J. Membr. Sci.* 112 (2), 129–146.
- Ahmad, F., Lau, K., Shariff, A., Murshid, G., 2012. Process simulation and optimal design of membrane separation system for CO₂ capture from natural gas. *Comput. Chem. Eng.* 36, 119–128.
- Alcheikhhamdon, Y., Hoorfar, M., 2016. Natural gas quality enhancement: A review of the conventional treatment processes, and the industrial challenges facing emerging technologies. *J. Nat. Gas Sci. Eng.* 34, 689–701.
- Araújo, OCADQF, Reis, ADC, Medeiros, JLD, Nascimento, JFD, Grava, WM, Musse, APS, 2017. Comparative analysis of separation technologies for processing carbon dioxide rich natural gas in ultra-deep-water oil fields. *J. Cleaner Prod.* 155, 12–22.
- Baker, R., 2001. Future directions of membrane gas-separation technology. *Membr. Technol.* 2001 (138), 5–10.
- Bhide, B., Stern, S., 1993. Membrane processes for the removal of acid gases from natural gas. II. Effects of operating conditions, economic parameters, and membrane properties. *J. Membr. Sci.* 81 (3), 239–252.
- Bhide, B., Voskericyan, A., 1998. Hybrid processes for the removal of acid gases from natural gas. *Fuel Energy Abstr.* 39 (5), 347.
- Bounaceur, R., Lape, N., Roizard, D., Vallieres, C., Favre, E., 2006. Membrane processes for post-combustion carbon dioxide capture: a parametric study. *Energy* 31 (14), 2556–2570.
- Coker, D., Freeman, B., Fleming, G., 1998. Modeling multicomponent gas separation using hollow fiber membrane contactors. *AIChE J.* 44 (6), 1289–1302.
- Colling, C., Huff, G., Bartels, J., 2004. Processes using solid perm-selective membranes in multiple groups for simultaneous recovery of specified products from a fluid mixture (US20040004040 A1). 2004.
- Datta, AK, Sen, PK, 2006. Optimization of membrane unit for removing carbon dioxide from natural gas. *J. Membr. Sci.* 283 (1–2), 291–300.
- Davis, R., 2002 Sep. Simple Gas Permeation and Pervaporation Membrane Unit Operation Models for Process Simulators. *Chem. Eng. Technol.* 25 (7), 717.
- Echt, W., 2017. Hybrid Systems: Combining Technologies Leads to More Efficient Gas Conditioning. UOP LLC.
- Favre, E., 2010. Polymeric membranes for gas separation. *Compr. Membr. Sci. Eng.* 155–212.
- Finn, AJ, O'Brien, J., 2014. Processing of carbon dioxide rich gas. GPA Conference.
- Foss MM, Interstate natural gas quality specifications and interchangeability.CEE. 2004.
- Hao, J., Rice, P., Stern, S., 2002. Upgrading low-quality natural gas with H₂S- and CO₂-selective polymer membranes. *J. Membr. Sci.* 209 (1), 177–206.
- Hao, J., Rice, P., Stern, S., 2008. Upgrading low-quality natural gas with H₂S- and CO₂-selective polymer membranes. *J. Membr. Sci.* 320 (1–2), 108–122.
- Isa, M., Azhar, M., 2009. Meeting technical challenges in developing high CO₂ gas field offshore. 24th World Gas Conference (WGC).
- Klinkenbijl, J., Dillon, M., Heyman, E., 1999. Gas pre-treatment and their impact on liquefaction processes. Gas Processors Association.
- Koch, D., Buchan, W., Cnop, T., 2005. Proper Pretreatment Systems Reduce Membrane Replacement Element Costs and Improve Reliability. UOP LLC., Des Plaines, IL.
- Latest U.S National Average Natural Gas Price & Chart [Internet]. NASDAQ, 2017 Available from <https://www.nasdaq.com/markets/natural-gas.aspx>.
- Lock, S., Lau, K., Ahmad, F., Shariff, A., 2015. Modeling, simulation and economic analysis of CO₂ capture from natural gas using cocurrent, countercurrent and radial crossflow hollow fiber membrane. *Int. J. Greenhouse Gas Control* 36, 114–134.
- Mazyan, W., Ahmadi, A., Ahmed, H., Hoorfar, M., 2016. Market and technology assessment of natural gas processing: a review. *J. Nat. Gas Sci. Eng.* 30, 487–514.
- Narayan, R., Patton, C., 2004. Membrane gas separation process.
- Ohs, B., Lohaus, J., Wessling, M., 2016. Optimization of membrane based nitrogen removal from natural gas. *J. Membr. Sci.* 498, 291–301.
- Pan, C., 1986. Gas separation by high flux, asymmetric hollow fiber membrane. *AIChE J.* 32 (12), 2020–2027.
- Peters, L., Hussain, A., Follmann, M., Melin, T., Hägg, M-B., 2011. CO₂ removal from natural gas by employing amine absorption and membrane technology—a technical and economical analysis. *Chem. Eng. J.* 172 (2–3), 952–960.
- Qi, R., Henson, M., 1998. Optimization-based design of spiral-wound membrane systems for CO₂/CH₄ separations. *Sep. Purif. Technol.* 13 (3), 209–225.
- Rezakazemi, M., Heydari, I., Zhang, Z., 2017. Hybrid systems: combining membrane and absorption technologies leads to more efficient acid gases (CO₂ and H₂S) removal from natural gas. *J. CO₂ Util.* 18, 362–369.
- Rufford, T., Smart, S., Watson, G., Graham, B., Boxall, J., Costa, JDD, et al., 2012. The removal of CO₂ and N₂ from natural gas: A review of conventional and emerging process technologies. *J. Pet. Sci. Eng.* 94–95, 123–154.
- Scholes, CA, Stevens, GW, Kentish, SE, 2012. Membrane gas separation applications in natural gas processing. *Fuel* 96, 15–28.
- Sorschak, K., 2017. XEBEC Investor Presentation First Quarter 2017.
- Strang, G., 2016. Introduction to Linear Algebra. Cambridge Press, Wellesley, MA: Wellesley.
- Tennyson, R., Schaaf, R., 1977. Guidelines can help choose proper process for gas treating plants. *Oil Gas J.* 10 (75).
- Turton, R., Bailie, R., Whiting, W., Shaeiwitz, J., 2009. Estimation of capital costs. Analysis, Synthesis and Design of Chemical Processes, 3rd edn Pearson, Upper Saddle River, NJ.
- White, LS, 2010. Evolution of natural gas treatment with membrane systems. *Membr. Gas Sep.* 313–332.
- Xu, J., Agrawal, R., 1996. Gas separation membrane cascades I. One-compressor cascades with minimal exergy losses due to mixing. *J. Membr. Sci.* 112 (2), 115–128.