Subsea natural gas dehydration with membrane processes: Simulation and process optimization

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Abstract

Subsea processing enables broader exploration of oil and gas reservoir, giving an increased focus on developing alternative processes for subsea oil and gas treatment. This work provides a first evaluation of a new proposed subsea natural gas dehydration process with the use of a membrane contactor with triethylene glycol (TEG) for dehydration of the natural gas in combination with thermopervaporation for regeneration of the TEG. Simulation models are developed in Aspen HYSYS V8.6 and process optimization is performed on three different process designs with respect to staging of the regeneration. By introducing two thermopervaporation units in series the TEG flow rate is reduced by 55%, the membrane volume by 14.6% and the energy demands by 37.8%, compared to a design with one thermopervaporation unit. However, increasing the number of regeneration stages increases the complexity as additional heaters are introduced.

Keywords: Subsea natural gas dehydration, Membrane Contactor, Thermopevaporation, Process Optimization, Triethylene glycol

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1. Introduction

Natural gas from the reservoir is normally saturated with water, which may condense during transportation causing flow assurance problems such as hydrate formation, corrosion and erosion [1]. Removal of the water can prevent these challenges and dehydration of natural gas is therefore one of the main processing steps in natural gas treatment. Several methods can be used for dehydration of natural gas, such as absorption into a solvent or adsorption onto a solid [2]. The most common technology used in the oil and gas industry today is the absorption process with the use of glycols, typically triethylene glycol (TEG), as illustrated in Fig. 1. The inlet scrubber is applied to remove hydrocarbon liquids and/or free water from the wet gas before it enters the glycol contactor. In the glycol contactor the natural gas meets the glycol in a counter-current flow and the dry gas leaves from the top of the column. The rich TEG (TEG with the absorbed water) is regenerated before it is reused as lean TEG in the glycol contactor. In addition to the topside dehydration facility, injection of hydrate inhibitor such as monoethylene glycol (MEG) or methanol is commonly used to prevent hydrate formation from the reservoir to the topside treatment facility [3].

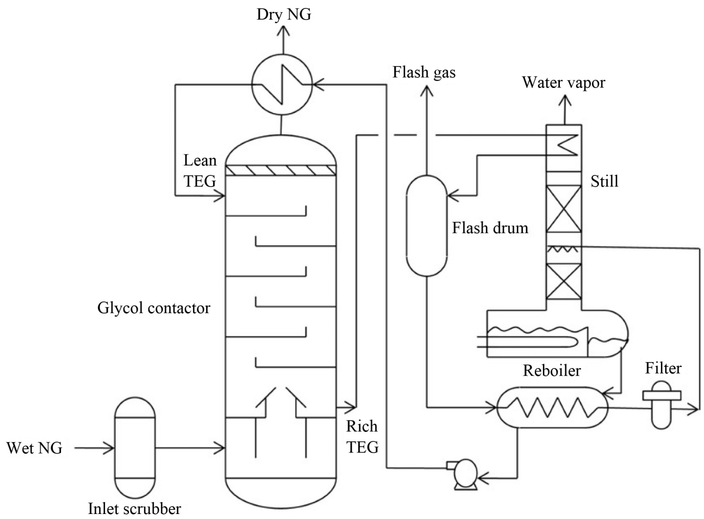


Figure 1: Process design for commonly used natural gas dehydration process with glycol [4]

With increasing focus on subsea processing and the ultimate vision of directly exporting of the produced hydrocarbons from the reservoir to the market [5], alternative technologies for subsea dehydration of natural gas should be investigated. Subsea dehydration can reduce the water content in downstream processing steps to acceptable levels, and eliminate the need for continuously injection of hydrate inhibitor. In addition, subsea dehydration can provide subsea to shore production, as the gas will be processed subsea and directly exported [6]. When considering a technology for subsea installation several design criteria should be considered. Unmanned operation is required as the process is placed at a remote location and limiting moving parts is favorable to reduce the need for maintenance [7]. With respect to high accessibility and retrievability, high modularity is preferred. Depending on the water depth of the installation, compact design is favorable due to the limitations in size and weight for the installation cranes [8, 9]. The complexity of the system should be kept low and the number of process equipment to a minimum. To avoid leakage and minimize the potential of failure, the number of subsea connection points should be low, which makes it important to consider how the process design should be installed in subsea modules.

Membrane technology being compact with no moving parts, in addition to flexible operation due to high modularity make membranes and membrane contactors to interesting technologies for subsea operation [10]. Membrane contactor is a hybrid technology combining the advantages of both absorption and membrane technology. One of the advantages compared to conventional adsorption columns is the higher surface area per unit contactor volume, providing increased contact area in a smaller module. During the last decade, membrane contactors have been intensively studied, especially for CO_2 capture [11–33]. Promising result have been reported from one pilot test of natural gas dehydration with membrane contactor, showing dehydration to pipeline specification and stable operation [34–36].

Regeneration of TEG is commonly achieved with distillation technology, but the system complexity and energy demand make the process less feasible for subsea operation. This encourage the research for alternative technologies for subsea regeneration of TEG. Wijmans et al. [37] patented a natural gas dehydration process where the regeneration step is replaced with vacuum pervaporation. Pervaporation combines permeation through a membrane with evaporation, leading to a possibly higher recovery of the TEG. In vacuum pervaporation, a vacuum pump is applied on the permeate side to achieve a low vapor pressure and improve the separation performance. An alternative to the vacuum pervaporation; i.e. driving force by vacuum pump, is thermopervaporation where the low vapor pressure is maintained by condensing the permeate; i.e. driving force created by temperature differences. The access to large amount of cooling water on the seabed, makes thermopervaporation an interesting technology for subsea operation. Several researchers [38–56] have investigated dehydration of monoethylene glycol (MEG) with vacuum pervaporation. Little information is found on research for dehydration of TEG with vacuum pervaporation [57–59], and no data for thermopervaporation. Due to the limited investigation of membrane contactor for dehydration of natural gas and TEG regeneration with thermopervaporation, more research is needed to evaluate the potential for subsea operation.

In this work, a natural gas dehydration process with the use of membrane technology is evaluated through modelling and simulation. Compared to the commonly used absorption process in Fig. 1, the conventional packed absorption column is replaced with a hollow fiber membrane contactor unit. In addition, the regeneration part is replaced with a thermopervaporation unit. The models of the membrane contactor [60] and thermopervaporation unit [59] are already published and will not be described here. They are written in MATLAB and implemented in Aspen HYSYS V8.6 with the use of MATLAB CAPE-OPEN (version 1) [61]. Here, process design with different number of regeneration stages are evaluated, in addition to optimization of the operation conditions for the system with respect to TEG flow rate and membrane module sizes. Moreover, the different designs are compared based on subsea feasibility with respect to sizes, number of stages and energy demands. This is the first and an important step in the feasibility evaluation of the new proposed dehydration process.

2. The proposed process concept

The proposed subsea natural gas dehydration process is illustrated in Fig. 2 and the selected capacity is a natural gas feed of 25 MSm^2/d, saturated with H_2O at reservoir conditions. Åsgard transport condition is selected as the dehydration requirements, with a water dew point of -18°C at 69 barg [5, 62]. The pressure and the temperature of the natural gas from the reservoir are reduced to 80 bar and 25°C and a scrubber is applied to remove the liquid hydrocarbon and/or free water. The natural gas from the scrubber is saturated with water, which may condense if the temperature in the pipeline is reduced. Therefore, to prevent condensing of the water before the membrane contactor inlet, the natural gas is heated to 30°C. The properties of the wet natural gas feed to the membrane contactor are as given in Table 1. The wet natural gas enters the membrane contactor on the shell side and the triethylene glycol (TEG) flows counter-current inside the fibers. The H_2O is transported over the membrane and absorbed into the TEG, providing a dry natural gas ready to be exported. The rich TEG is reduced in pressure to 1 bar and heated before it enters the thermopervaporation unit for regeneration by removal of the water. The regenerated TEG is reused in the membrane contactor for absorption of H_2O. Prior to the membrane contactor inlet the temperature and pressure of the TEG is adjusted to the same conditions as the membrane contactor gas inlet.

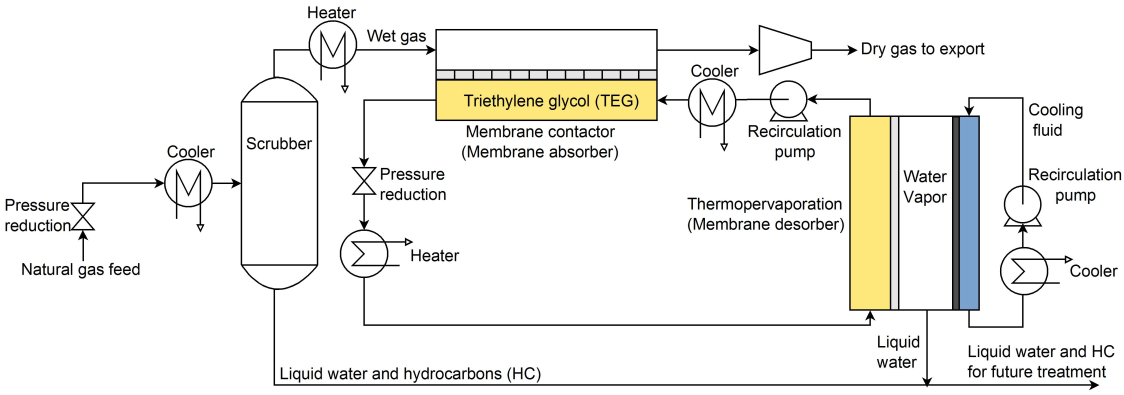


Figure 2: Simplified illustration of the proposed subsea natural gas dehydration system

Table 1: Properties of the wet natural gas for the membrane contactor inlet

Parameter	Value
Temperature [°C]	30
Pressure [bar]	80
Flow [Sm^2/d]	24.9
Molar flow [kmol/h]	$4.4{\cdot}10^4$
H_2O content [ppm]	600

2.1. Simulation assumptions

Simulations are carried out using Aspen HYSYS V8.6 process simulator. Modelling of the membrane contactor and thermopervaporation unit are performed in MATLAB and implemented into Aspen HYSYS with the use of MATLAB CAPE-OPEN (version 1). The other units in the design (scrubber, heaters, coolers, pumps and pressure relief valves) are standard units in Aspen HYSYS. The design pressure drop in the scrubber, heaters and coolers are assumed to be negligible. The thermodynamic models used to calculated the thermodynamic properties in the simulations are Peng-Robins (PR) for the natural gas and the non-random two-liquid model (NRTL) for the liquid phase. With the use of MATLAB CAPE-OPEN it is possible to choose if the physical properties of the gas and liquid should be retrieved from Aspen HYSYS properties or from defined correlations inside the model. In the developed models, all the physical properties required for the TEG and the natural gas are calculated based on defined correlation inside the model. The HYSYS flow sheet gives the feed conditions to the membrane modules such as the temperature, the pressure, the composition and the total flow rate, and the outlet conditions are provided by the model. Only water is assumed to be transported over the membrane in the membrane units.

2.2. MATLAB CAPE-OPEN unit operation in Aspen HYSYS

CAPE-OPEN is a standard for communication between chemical engineering software components, facilitating interoperability between process simulators. Most simulation software are CAPE-OPEN compliant, including Aspen HYSYS, which means that a CAPE-OPEN unit operation and property package can be included in the simulation [63]. CO-LaN is a non-profit organization responsible for the management of the CAPE-OPEN standard [64]. "MATLAB CAPE-OPEN unit operation" developed by AmsterChem [61] is a software for developing unit operation models to be implemented in CAPE-OPEN compliant software. This give the advantage that all the function in the MATLAB library can be used within the simulation software. In addition to simple implementation of MATLAB models as unit operations.

2.3. Membrane contactor

The membrane contactor model is based on a hollow fiber module configuration with counter-current flow. On the shell side a one-dimensional model is used for the gas flow. For the liquid in the lumen side, a two-dimensional model is developed to describe the temperature and the concentration profiles in axial and radial directions. The partial differential equations describing mass, temperature and pressure changes are discritized by orthogonal collocation. Consequently this results in nonlinear algebraic equations solved in MATLAB by Newton-Raphson type iteration. In a previous work, the model was validated against high pressure experimental data and a sensitivity study with respect to membrane and module properties was performed [60]. The result from this evaluation provides the selection of membrane and module properties used for the membrane contactor in this simulation, as listed in Table 2. To meet the subsea requirements of long-term stable operation, a thin composite membrane has been selected to avoid pore wetting, which significantly reduces the separation performance. The selected membrane for this process evaluation consist of a porous polypropylene (PP) support coated with a dense layer of Teflon®AF2400.

Table 2: Specifications of membrane contactor

Parameter	Value	Unit
Fiber inner diameter	600	μm
Membrane length	1	m
Packing density	1500	m_m^2/m_t^2
Membrane thickness (porous, PP)	200	μm
Membrane porosity	0.71	
Membrane thickness (dense, AF2400)	1	μm
Membrane permeability (dense, AF2400)	3000	Barrer
Operation pressure	80	bar
Operation temperature	30	°C

2.4. Thermopervaporation unit

The thermopervaporation model is based on the plate-and-frame module configuration, with alternated channels for the TEG, air gap and the cooling water. The TEG and the cooling water are modelled as two-dimensional laminar flow, while the air gap is considered as a stagnant phase. The model consist of a temperature dependent permeability correlation for the dense Teflon®AF2400 layer, developed based on results from vacuum pervaporation experiments [59]. The same solving methods is used for the thermopervaportion model as for the membrane contactor. In a previous work a sensitivity study was performed for the thermopervaporation unit with respect to membrane and module properties [59]. The selected specifications used for this evaluation are based on the results from the sensitivity study, as given in Table 3.

Table 3: Specifications of thermopervaporation

Parameter	Value	Unit
Feed channel thickness	4	mm
Air gap	1	cm
Membrane length	1	m
Membrane width	1	m
Membrane thickness (porous, PP)	25	μm
Membrane porosity	0.41	
Membrane thickness (dense, AF2400)	1	μm
Cooling water channel thickness	5	mm
Cooling water operation temperature	4	°C
Cooling water velocity	0.05	m/s
Operation pressure	1	bar

3. System optimization

There are several design variables that can be adjusted to obtain the optimum process design. The goal for the process is to meet the dehydration specifications for the dry natural gas, which is a water dew point of -18°C at 69 barg. In all the cases the dehydration criteria must be fulfilled. As mentioned above, based on the sensitivity analysis of the membrane units the membrane and module parameters are given specific values. In addition, a given natural gas feed is used with the conditions as given in Table 1. This leaves us with five variable for the optimization of the process design, including TEG flow rate, TEG regeneration temperature, membrane contactor size (number of fibers), size of the thermopervaporation unit(s) (number of feed channels) and the number of regeneration stages in series.

The liquid temperature in the thermopervaporation unit will decrease along the module due to heat of evaporation and heat transfer between hot and cold fluid. A reduction in the liquid temperature results in reduced separation performance. It is possible to improve the separation performance by using several stages of the regeneration in series with heating between the stages. Three different process designs with respect to number of stages for the regeneration were investigated, including one (Design 1), two (Design 2) and three (Design 3) stages. A simplified illustration of the three process designs are given in Fig. 3. The remaining variables are then used in the optimization of each process design.

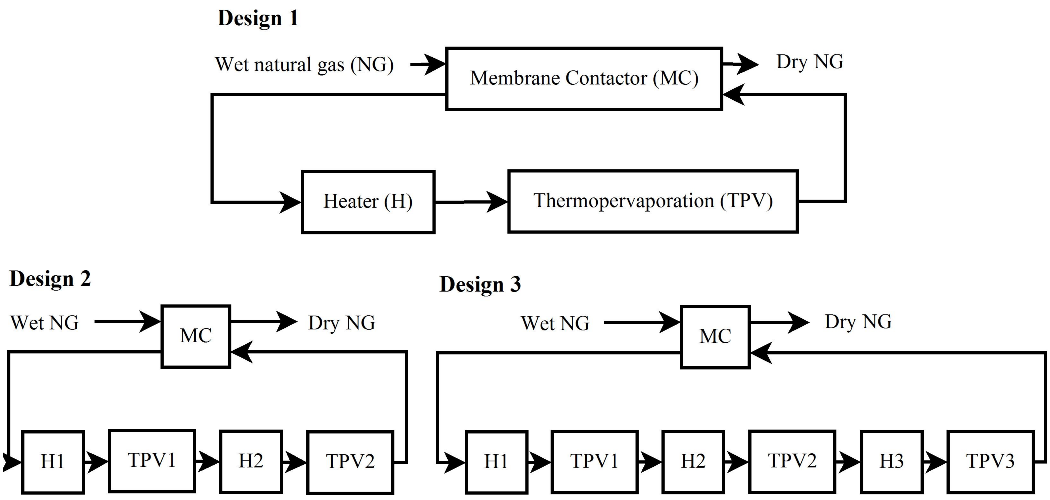


Figure 3: Simplified illustration of the three different design configurations of the dehydration process with one, two or three stages of regeneration. The illustration only shows the main differences between the designs, as the rest of the process is the same. MC indicates the membrane contactor, TPV is a thermopervaporation unit and H indicates an electrical heater.

The optimization is performed in MATLAB. Aspen HYSYS is accessed from MATLAB via activeX/COM interface. Aspen HYSYS is used for the process simulation as different process configurations can easily be simulated, in addition standard unit operations, flash calculations and thermodynamic properties can be used. The optimization problem solved in MATLAB with the fmincon function can be written as:

$$\min_{\mathbf{x}} f(\mathbf{x}) \tag{1}$$

$$s.t. \quad \mathbf{c}(\mathbf{x}) \leq 0 \tag{2}$$

$$\mathbf{c_{eq}}(\mathbf{x}) = 0 \tag{3}$$

where f is the objective function to be minimized, $\mathbf{c}$ is the inequality constraints, $\mathbf{c_{eq}}$ is the equality constraints and $\mathbf{x}$ is the optimization variables. The main equality constraint is the dehydration pipeline specification for the natural gas. The other constraints are mainly for simulation purposes to avoid infeasible operation conditions. Every time Aspen HYSYS is called by the optimization routine it return values with a given accuracy based on the solver tolerance for the membrane modules. To assure accurate results from the optimization routine, the calculation tolerance of the membrane modules are given a much stricter requirements than for the optimization. In addition, to avoid iterations in Aspen HYSYS the recycle block is removed and equality constraints are added on the component molar flow of TEG and H_2O to ensure that the recycle loop is closed. The optimization variables ($\mathbf{x}$), which are equal to the design variables are: molar flow of TEG (fTEG) and H_2O (fH_2O) in the regeneration loop at the membrane contactor inlet, number of fibers in the membrane contactor (Nf) and number of feed channels in the thermopervaporation unit(s) (Nc). The regeneration temperature of the TEG is also evaluated as an optimization variable. However, during the first evaluation it is found that the temperature always reached the upper limit for the variable which is decided based on material limitations. Therefore, the temperature is fixed to 100°C in the optimization to reduce the computational time.

3.1. *Objective function*

The objective function to be minimized is a function based on investment and operating costs. The investment cost includes the main equipment in the regeneration loop, such as the membrane contactor, the thermopervaporation unit(s), the heater(s) and the TEG recirculation pump. The operating cost includes the electrical power required for the heater(s) and the pump in the TEG regeneration loop. The size of the membrane unit(s) and the energy demand are parameters of interest. These are included in the objective function as the investment cost of the membrane unit is proportional to the membrane area and the energy consumption is included in the operating cost. The objective function is given in Eq. 4.

$$f = \mathrm{C_I} \cdot \mathrm{ACCR} + \mathrm{Q}_{EL} \cdot \mathrm{C_{EL}} \tag{4}$$

where $\mathrm{C_I}$ is the total capital cost, ACCR is the annual capital charge ratio, Q_{EL} is the annual electrical power demands and $\mathrm{C_{EL}}$ is the electricity price.

A typical installation factor of six is used to estimate the total capital cost for fluid processes [65], but to adjust for the subsea conditions this factor is increased with 50%. Based on this, the total capital cost can be calculated with Eq. 5. However, this does not include the cost of the ships for the subsea installation, which is not considered in this evaluation.

$$C_I = 9\left(C_{e,MC} + C_{e,TPV} + C_{e,P} + C_{e,H}\right) \tag{5}$$

$C_{e,MC}$ is the cost of the membrane contactor, $C_{e,TPV}$ is the cost of the thermopervaporation unit(s), $C_{e,P}$ is the cost of the TEG recirculation pump and $C_{e,H}$ is the cost of the electrical heater(s).

To include the capital cost in the objective function, it should be transformed to annual cost, which is achieved with the use of the annual capital charge ratio (ACCR).

$$ACCR = \frac{\left[i\left(1+i\right)^n\right]}{\left[\left(1+i\right)^n - 1\right]} \tag{6}$$

where i is the interest rate and n is the number of years (lifetime of investment).

One uncertainty in this evaluation is the limitation in cost estimations for subsea processing equipment. Therefore, onshore cost values are used with adjustment to include for subsea operation conditions. The cost for the membrane contactor and thermopervaporation unit are assumed based on reported onshore prices for membrane modules [66–68], with an increase by 50% to provide the more robust module shell for subsea operation. The price of the thermopervaporation unit is assumed to be higher than the membrane contactor based on more complexed module preparation with a lower packing density of the module. An overview of the cost values and parameters in the objective function are given in Table 4.

Table 4: Cost values and parameters used in the objective function

Category	Parameter	Value
Capital Expenditure (CAPEX)	Membrane contactor cost ($C_{e,MC}$)	45 \$/m^2
	Thermopervapoartion unit cost ($C_{e,TPV}$)	75 \$/m^2
	Recirculation pump cost($C_{e,P}$)	Eq. 7
	Electrical heater cost ($C_{e,H}$)	Eq. 8
	Total fixed capital cost (C_I)	Eq.5
	Annual capital charge ratio (ACCR)	Eq.6
Annual Operating Expenditure (OPEX)	Electricity cost (C_{EL})	0.0627 \$/kWh
Other assumptions	Equipment lifetime (n)	5 years
	Interest rate (i)	5%

The equipment cost of the TEG circulation pump ($C_{e,P}$) and the electrical heater(s) ($C_{e,H}$) are found based on a linear approximation from cost estimation in Aspen HYSYS V9 given in Eq. 7 and Eq. 8 respectively. A factor of 50% is added to the equipment cost to adjust for the subsea conditions.

$$C_{\mathrm{e,P}} = 1.5\,(0.3872 \cdot S + 153,691) \tag{7}$$

$$C_{\mathrm{e,H}} = 1.5\,(2.0993 \cdot S + 8240.7) \tag{8}$$

Here, S is the size parameter, which for the recirculation pump is the liquid flow rate [L/s] and the for the electrical heater(s) is the energy demands [kW].

4. Results and discussion

With the use of the fmincon function in MATLAB the optimum point with respect to the selected variables are found. Three different process designs are evaluated with different numbers of regeneration stages in the range from one to three, as illustrated in Fig. 3. The results from the optimization for the different process designs, including the values of the optimization variables are given in Table 5.

The temperature drop in the thermopervaporation unit affects the separation performance, due to the reduced driving force over the membrane. By introducing staging of the regeneration with heating between each thermopervaporation unit it is expected that the separation performance would be increased and provide a higher purity of the lean TEG. When the purity of the lean TEG is increased the required membrane area in the membrane contactor and the TEG flow rate is reduced. Reducing the TEG flow rate will also have an effect on the energy demand from the heaters and the pump. It is therefore expected that staging of the regeneration will provide a better design with respect to membrane sizes, TEG flow rate and energy demands. However, it is important to remember that increasing the number of stages introduces additional heaters, which increase the complexity. As expected, increasing the number of regeneration stages gives a decreased TEG flow rate as the purity of the lean TEG is increased. By increasing from one stage (Design 1) to two stages (Design 2) a larger benefit is provided compared to further increaseing from two to three stages (Design 3). The TEG flow rate is reduced by 55.0% (353.6 kmol/h) for Design 2 compared to Design 1. Going from Design 2 to Design 3, a further reduction of 40.4% (116.7 kmol/h) is obtained. In conventional absorption dehydration processes a TEG circulation rate of 15-40 $\mathrm{L_{TEG}/kg_{H_2O removed}}$ is commonly reported

[1, 69, 70]. This value is based on optimization of the conventional dehydration process, and it is important to note that when new technologies are used for both the absorption and the regeneration step, this value might not be an optimum for the new process. The results in Table 5 show that the TEG circulation rate for all the evaluated designs are higher than the reported TEG circulation rate for the conventional absorption process, but the value is reduced as the number of regeneration stages are increased.

Table 5: Values of the optimization variables and the result for the optimum point of the different process designs with respect to membrane sizes and energy demands. The volume calculation is based on only the active membrane area and a cooling wall thickness of 1 mm.

Parameter	Design 1	Design 2	Design 3	Unit
Optimization variables				
Molar flow TEG (fTEG)	590.6	267.1	160.0	[kmol/h]
Molar flow H_2O (fH_2O)	52.2	22.0	12.4	[kmol/h]
Number of MC fibers (Nf)	21.271	19.941	19.560	[$x10^6$]
Number of TPV feed channels				
1 (Nc1)	8.206	3.374	1.915	[$x10^3$]
2 (Nc2)		3.651	1.830	[$x10^3$]
3 (Nc3)			2.105	[$x10^3$]
Results				
Lean TEG flow rate	642.8	289.1	172.4	[kmol/h]
TEG circulation rate	197.8	89.4	53.5	[$L_{TEG}/kg_{H_2Oremoved}$]
Lean TEG purity	98.95	99.02	99.08	[wt%TEG]
MC area	40,095	37,588	36,869	[m^2]
MC volum	27	25	24.6	[m^3]
TPV area				
1	16,412	6784	3829	[m^2]
2		7123	3660	[m^2]
3			4209	[m^2]
TPV volum				
1	255	105	59	[m^3]
2		111	57	[m^3]
3			65	[m^3]
Total membrane volume	282	240	206	[m^3]
Heater duty				
1	4509	2048	1234	[kW]
2		797	471	[kW]
3			437	[kW]
Recirculation pump duty	243	110	66	[kW]
Total energy demand	4752	2955	2208	[kW]
Objective function value	9.45	7.85	7.06	[mill$]

Increasing the number of regeneration stages also reduces the value of the objective function. Fig. 4 illustrates the contribution from the different parts of the objective function. From these results it can be seen that the main contribution to the objective function is the cost of the membrane units and the electricity. This explain why the value of the objective function is reduced with increased staging, even though additional heaters are added. However, the contribution from the heater(s) to the objective function is increased with the number of regeneration stages. Increasing the number of regeneration stages reduces the contribution from the regeneration part (thermopervaporation unit(s), heater(s), recirculation pump and electricity) of the objective function, while the membrane contactor percentage is increased.

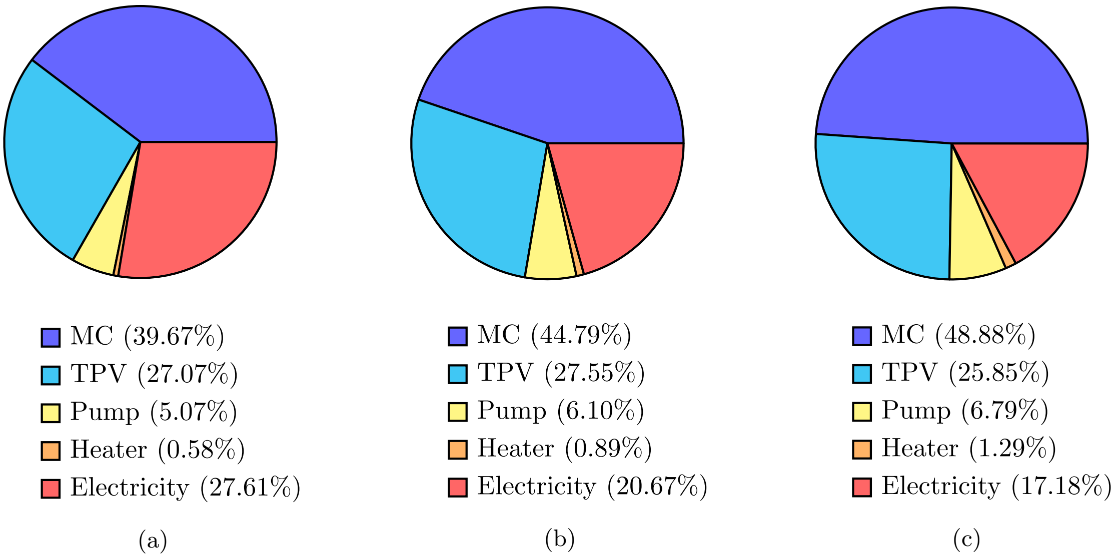


Figure 4: Contribution of the different parts on the cost function for the different process designs with respect to stages of regeneration; (a) Design 1 (one stage), (b) Design 2 (two stages) and (c) Design 3 (three stages). MC is an abbreviation for membrane contactor and TPV for thermopervaporation.

Important factors for subsea operation are the size of the system and the energy demand. As for the TEG flow rate, increasing the number of stages reduces the total membrane volume and the energy demand. Comparing the result of Design 1 and Design 2 shows that the total membrane volume is reduced by 14.6% (41 m^3) and the total energy demand is reduced by 37.8% (1796 kW). Going from Design 2 to Design 3 gives a further decrease of the membrane volume by 14.3% (34 m^2) and 25.3% (747 kW) in the energy requirements. The complexity of the system is increased with the additional heaters when the number of regeneration stages is increased. In addition, the largest benefits are given when going from Design 1 to Design 2. Therefore, a further increase in number of regeneration stages might not be favorable. Compact design is important due to installation and retrieval cost, in addition to limitations in weight for the installation cranes. It is found that increasing the number of thermopervaporation stages is preferred from a separation point of view. However, a next step for the practical evaluation of subsea installation is to evaluate how the system should be designed in subsea installation modules. Big subsea modules would require a large ship for the installation, while smaller modules would lead to more subsea connection points which have a larger potential for leakages and failures. All these factors make it favorable to find the most compact design of the process.

The results reveals that the low packing density of the thermopervaporation membrane module gives large volumes for this units. Even thought the membrane area in the membrane contactor is more than the double of the thermopervaporation unit, the volume is much larger. The low packing density of the thermopervaporation unit (64.4 m^2/m^3) is based on the selected plate-and-frame module configuration, in addition to the air gap and the cooling water channel which increases the size of the unit. From a practical point of view, it could be favorable to reduce the size of the thermopervaporation unit and increase the membrane contactor size to reduce the total membrane volume. An analysis is performed for Design 1, where the number of fibers in the membrane contactor is changed simultaneously as the number of feed channels in the thermopervaporation unit is adjusted to meet the dehydration criteria for the natural gas (Fig. 5a). In this analysis the TEG flow rate is kept constant at the value found at the optimum point for Design 1 (Table 5). From the result given in Fig. 5b, it can be seen that the objective function has a minimum point as reported above, however with respect to the total membrane volume, another minimum point can be found (MC fibers 32x10^6) with a higher value of the objective function.

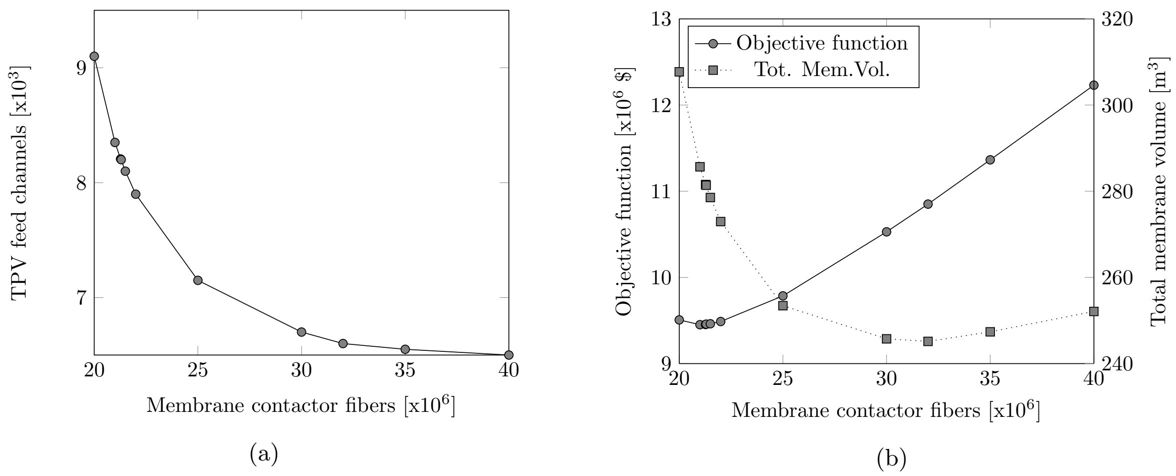


Figure 5: a) The number of feed channels in the thermopervaporation unit as a function of numbers of fibers in the membrane contactor, when the dehydration criteria for the natural gas is received. b)The value of the objective function and the total membrane volume (membrane contactor and thermopervaporation unit) as function of changing the number of fibers in the membrane contactor simultaneously as the number of feed channels in the thermopervaoration is adjusted.

The low packing density of the thermopervaporation module is related to the selection of membrane module, another module configuration, for instance tubular membrane modules, could provide a higher packing density and reduce the volume. The need for the air gap and the cooling water would still limit the packing density, but it is expected that the packing density would be higher. However, when changing the module configuration new models and evaluations of the module parameters are needed to evaluate the exact benefit on the packing density.

The temperature drop and the low packing density of the thermopervaporation unit are two limiting factors for the proposed process. Adding spacers in the feed channels could increase the separation performance, as the introduction of mixing or turbulence might prevent or reduce the temperature and concentration polarization. This was proven by Krish et al. [71], for the butanol-water mixture and it is expected that the same effect might be seen for TEG-water mixtures. Another alternative that could be interesting to analyse is to introduce a heating channel or an electrical wire inside the feed channel to maintain the liquid temperature along the membrane module. However, if heating is placed close to the membrane material it is important to remember the temperature limitations for the membrane material. Avoiding the temperature drop could increase the separation performance and maybe a system as suggested in Design 1 without staging of the regeneration could be sufficient. This alternative moves the heating previously used between the regeneration stages to inside the membrane module, which means that the complexity of the system is moved to the membrane module and it is therefore a trade-off between system complexity or membrane module complexity.

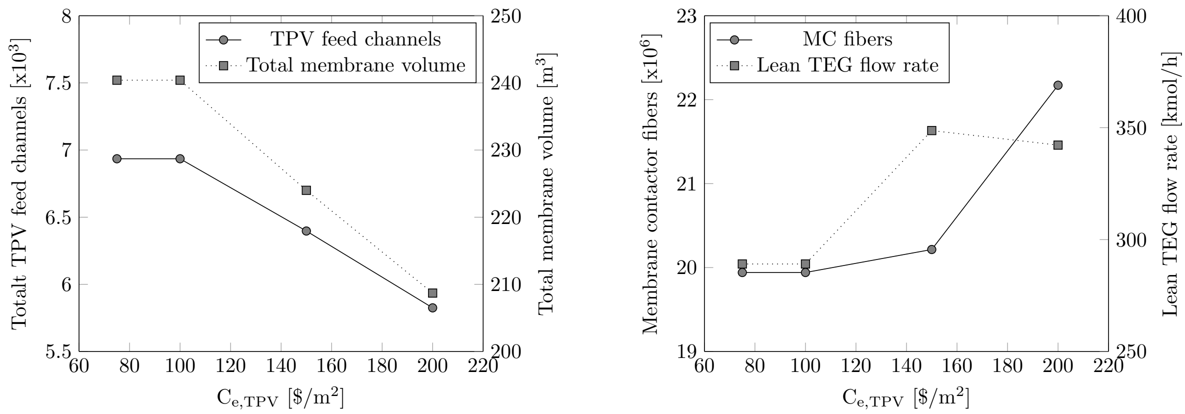
4.1. Price sensitivity study

As the cost values are based on assumptions and are uncertain values a sensitivity study for some of the parameters in the objective function was performed to evaluate how the optimum point is changed for Design 2.

4.1.1. Price of thermopervaporation

The price of the membrane units are related to the membrane area and therefore also to the size of the units. However, the thermopervaporation unit has a much lower packing density compared to the membrane contactor resulting in a much higher volume, even if the membrane area is much lower. In the objective function there is no penalty on the volume of the units, as the price is only related to the membrane area. As shown from the investigation for Design 1 (Fig. 5b) it was found that with respect to the total membrane volume another minimum could be found. It is therefore of interest to evaluate how the optimum conditions for the system are changing when the price of the thermopervaporation unit is increased, while keeping all other cost values constant. Increasing the price difference between the membrane units will introduce a penalty on the volume. In this sensitivity study four different prices were used for the thermopervaporation unit; 75 (base case), 100 , 150 and 200 $\$/m^2$.

The results given in Fig. 6 shows that increasing the price from 75 to 100$\$/m^2$ do not change the optimum point. But, as expected, with a further increase of the price the size of the thermopervaporation unit is reduced. As the size of the thermopervaporation units are decreased, the total membrane volume is reduced as the largest contribution is given by the thermopervaporation units (Fig. 6a). A decrease in the regeneration size, results in a decreased purity of the lean TEG and hence an increase in the membrane area in the membrane contactor, as illustrated in Fig. 6b. The lean TEG flow rate is increased when the price is increased to 150 $\$/m^2$. However, when the thermopervaporation price is further increased to 200 $\$/m^2$ a small reduction in the TEG flow rate can be seen, with a larger increase of the membrane area in the membrane contactor.



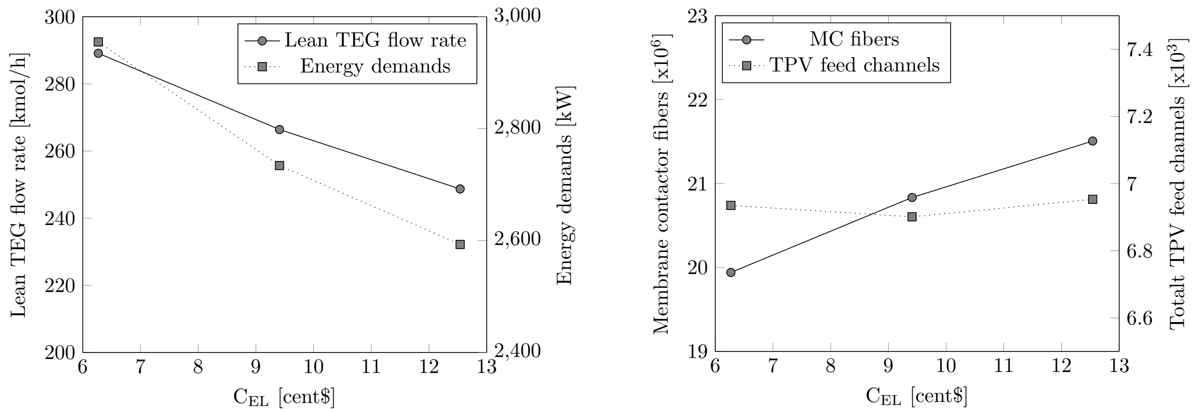
(a) (b)

Figure 6: (a) The effect of the thermopervaporation cost on the total number of feed channels in the thermopervaporation units (TPV) and the total membrane volume (thermopervaporation units and membrane contactor). (b) The sizes of the membrane units with number of fibers in the membrane contactor (MC) and the lean TEG flow rate as function of the thermopervaporation cost.

4.1.2. Electricity price

In the base case, the electricity price is set to 0.0627 $/kWh, for the sensitivity evaluation the price are increased by 50% (0.0941$/kWh) and 100% (0.1254$/kWh). The results (Fig. 7) show that changing the electricity price give some changes in the membrane sizes and TEG flow rate with respect to optimum design.

An increase in the electricity price results in a larger value of the objective function as the operating cost is increased, and the values for the three cases are $7.85\text{x}10^6$\$ (0.0627 \$/kWh), $8.63\text{x}10^6$\$ (0.0941\$/kWh) and $9.36\text{x}10^6$\$ (0.1254\$/kWh). As the electricity price is increased a reduction of the TEG flow rate can be seen of 7.9% and 14% compared to the base case when the electricity price is increased by 50 and 100% respectively. This is expected as a decrease in the flow rate is reducing the energy demands for the heater(s) and the recirculation pump (Fig. 7a). A decrease in the TEG flow rate, requires an increase in the membrane area for the membrane contactor (Fig. 7b) to be able to dehydrate the natural gas to the pipeline specification.



(a) (b)

Figure 7: (a) The effect of the electricity cost on the lean TEG flow rate and the total energy demands. (b) The sizes of the membrane units with number of fibers in the membrane contactor (MC) and the total number of feed channels in the thermopervaporation units (TPV) as function of electricity cost.

5. Conclusion

A new process concept for subsea natural gas dehydration with membrane processes is evaluated, indicating a promising potential for subsea operation. The proposed process is a regenerative design where a membrane contactor is used for dehydration of natural gas with triethylene glycol (TEG) in combination with thermopervaporation for regeneration of TEG. Three different process designs with staging of the regeneration with heating between the stages are considered. Process optimization are performed and the result reveals the following conclusions:

- Staging of the regeneration with heating between the stages are preferred as it reduces the size of the membrane units, the energy demands and the TEG flow rate. Increasing the number of regeneration stages from one to two gives a reduction of the TEG flow rate of 55% (353.6 kmol/h), in the total membrane volume of 14.6% (41 m^3) and in the energy requirements of 37.8% (1796 kW).

- The liquid temperature drop in the thermopervaporation unit is found to be a limiting factor for the system. The driving force over the membrane and the separation performance is reduced as the temperature drops. Therefore, staging of the regeneration would be preferred.

- Plate-and-frame module configuration for the thermopervaporation module provides a low packing density, which gives a large volume of the membrane unit.

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