Feasibility Study for the Treatment of Kuwait Sour Gas by Membranes

Yousef Alqaheem* and Abdulaziz Alomair

Petroleum Research Center, Kuwait Institute for Scientific Research, Kuwait

* Corresponding author. Tel.: +965 24956929. Fax: +965 23980445.
Email address: yqaheem@kisr.edu.kw

Abstract

Natural gas is mainly used in Kuwait for power generation and hydroprocessing. Yet, the gas needs to be treated as it contains amounts of carbon dioxide and hydrogen sulfide. Amine process is utilized but the unit is energy intensive and suffers from flooding. Alternatively, the membrane can treat natural gas with no flooding issue and lower energy input. In this work, a commercial membrane system was simulated in UniSim® for the removal of Kuwait sour gas having 12 mol% carbon dioxide and 4 mol% hydrogen sulfide. The system was compared to the amine unit in terms of technical and economic points of view. Results show that the membrane system was not capable of reaching the performance of the amine process due to the tradeoff between product purity and gas recovery. The membrane-amine process integration reduced the energy consumption by 15%, yet, the membrane's high capital cost makes the integration uneconomical.

Keywords: sour gas, gas-separation membrane, amine process, UniSim.

1. Introduction

Kuwait produces about 3 million barrels of crude oil daily. During the oil extraction, associated gas is released which contains mainly methane. The gas is then sent for treatment as it contains significant amounts of carbon dioxide and hydrogen sulfide. These impurities are corrosive and can damage pipelines (*1*). Therefore, the contents should be reduced to 2 mol% carbon dioxide and 4 ppm hydrogen sulfide (*2*). After the treatment by the acid gas removal (AGR) unit, the gas is moved to a liquefied petroleum gas (LPG) plant to produce commercial methane (natural gas).

In Kuwait, most of the natural gas is used for electricity generation and hydrogen production. Figure 1 shows the process of gas production in Kuwait.

The AGR unit is based on an amine process where carbon dioxide and hydrogen sulfide are chemically absorbed using methyl diethanolamine (MDEA) solution. The rich amine solution is then sent to a regenerator in which the amine solution is restored and used again.

Despite the excellent separation performance of the amine unit, the process has some issues related to foaming, flooding, and waste disposal (*3*). Also, the amine process has high capital and operating costs due to the regeneration step (*4*). On the other hand, membranes can provide an alternative solution for acid gas removal with minimum energy (*5*). The technology was commercialized 40 years ago and it was installed in many refineries worldwide. The membranes are compact, easy to integrate, and reported to have a continuous operating life of five years (*6*).

The original membrane was made from cellulose acetate due to its excellent chemical stability and good permeation (*7*). The transport mechanism is based on the pressure gradient across a dense membrane (*8*). First, the gas is absorbed on the membrane surface then it diffuses through flexible pores called free volume (*8*). In recent years, the membrane performance was enhanced by incorporating newly engineered materials such as polyphenylene oxide, polyimide and Pebax® (*9*).

In this project, an amine process was developed in Honeywell UniSim® to simulate the unit for the treatment of 70,000 $m^3 h^{-1}$ of sour gas. This gas is produced from west fields of Kuwait with carbon dioxide and hydrogen sulfide concentreations of 12 mol% and 4 mol%, respectively (*10*). An economic assessment was also carried out to calculate the capital and operating costs of the amine process to produce natural gas with 2 mol% carbon dioxide and 4 ppm hydrogen sulfide.

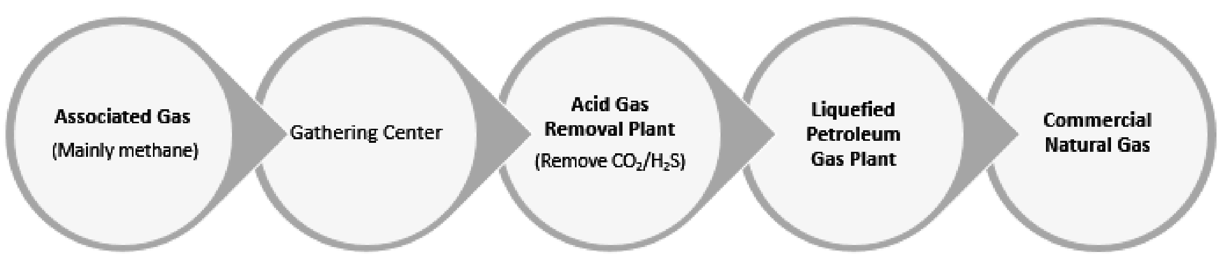

Figure 1. Steps for gas production in Kuwait.

After that, a membrane system was developed in UniSim® for the treatment of Kuwait sour gas. The system was compared with the amine process in terms of technical and economical points of view. This includes the quality and quantity of the produced gas along with the capital and operating expenses. The option of membrane–amine integration was also assessed.

2. Methodology

The chemical composition of the sour gas of the west fields is given in Table 1. The gas contains mainly methane (65 mol%) with impurities such as carbon dioxide (12 mol%) and hydrogen sulfide (4 mol%). The gas is fed at a rate of 70,000 m^3 h^{-1} with a pressure of 52 bar and a temperature of 38°C. Methyl diethanolamine (MDEA) has been selected as the amine solution as the compound is known to have less energy requirement compared to other amine solutions (*11*). MDEA enters the absorber at 45°C in which carbon dioxide and hydrogen sulfide are dissolved in the amine by chemical absorption (*12*):

$$CO_2 + R_3N + H_2O \leftrightarrow R_3NH^+ + HCO_3^- \quad (1)$$

$$H_2S + R_3N \leftrightarrow R_3NH^+ + HS^- \quad (2)$$

The product (sweet gas) leaves the absorber with a carbon dioxide concentration of 2 mol% and hydrogen sulfide content of 4 ppm to meet the requirement of pipeline specifications. After that, the rich amine solution is sent to a regenerator to recover back the amine solution by removing carbon dioxide and hydrogen sulfide using heat. The regenerated amine solution is then pumped and recycled back to the absorber. Table 2 shows the operating conditions of the amine process.

The membrane process was developed in UniSim® using a component splitter, spreadsheet, and adjust functions. Details of the system is described elsewhere (*13*). The sour gas enters the membrane at 52 bar and the product (retentate) leaves at 51 bar. The permeate (concentrated sour gas) leaves the unit at 2 bar and it is then compressed back to 52 bar using a compressor. Hydrocarbons recovery is calculated based on the methane content in the feed and the product using the following:

$$\text{Hydrocarbons Recovery (\%)} = \frac{\left(n_{C_1}\right)_{\text{Feed}}}{\left(n_{C_1}\right)_{\text{Product}}} \times 100 \quad (13)$$

Table 1. Composition of the sour gas of west fields of Kuwait.

Component	**Mole (%)**
Methane (CH_4)	65
Carbon dioxide (CO_2)	12
Hydrogen sulfide (H_2S)	4
Ethane (C_2H_6)	10
Propane (C_3H_8)	5
Butane (C_4H_{10})	2.5
Water (H_2O)	1
Nitrogen (N_2)	0.5

Table 2. Operating Conditions of the Amine Process for Sour Gas Treatment.

Property	**Condition**
Feed Gas	
Flowrate	70,000 m^3/h (60 MMSCFD)
Pressure	52 bar
Temperature	38°C
Amine Solution	
Pressure	52 bar
Temperature	45°C
Purity	45 wt%
Absorber Trays	22
Regenerator Trays	23
Product Gas Target	
CO_2 (mol%)	2
H_2S (ppm)	4

The simulation was performed in three steps. First, the amine process was developed in UniSim® for the treatment of Kuwait sour gas to produce sweet gas containing 2 mol% carbon dioxide and 4 ppm hydrogen sulfide. The second step was to develop an independent membrane module in UniSim® for the treatment of Kuwait sour gas. This was achieved by assuming the membrane area and monitoring product gas specifications. The third step was to integrate the membrane unit before the amine process and study the advantages of adding the membrane in terms of energy reduction and product quality. An economic assessment was carried out as well.

3. Results and Discussion

Amine Process Simulation: The first objective of this study is to simulate the amine process in UniSim® to treat the gas of west fields of Kuwait. This unit is already installed and the aim is to determine the performance and the cost of the unit. Figure 2 shows the solved process flow diagram of the amine unit and the product (sweet gas) with carbon dioxide content less than 2 mol% and less than 4 ppm of hydrogen sulfide. Hydrocarbons recovery is 99.6% which means that only 0.4% of hydrocarbons were dissolved in the MDEA solution. The required MDEA flow rate is 243 m^3 h^{-1}. The regenerator consumes about 66% of the total energy due to high power input for steam generation to heat the rich-amine solution.

Capital expenditure (CAPEX) of the amine process was carried out using econometric software for chemical plants (CAPCOST). The capital cost is based on the price of the units in addition to the cost of the MDEA solution. Based on Figure 2, the units are the absorber, regenerator (with reboiler and cooler), fired heater (for steam generation for regenerator), heat-exchanger (for amine lean/rich streams), amine cooler, amine pump, cooling towers (for coolers), mixers, flash vessels, and filters. CAPCOST uses the following equation to calculate the purchase cost of the equipment (C_p^o):

$$\log_{10} C_p^o = K_1 + K_2 \log_{10}(X) + K_3[\log_{10}(X)]^2 \tag{14}$$

where K_1, K_2, K_3, and X are parameters that differ from one unit to another. For example, to calculate the cost of the absorber, the X values represent the unit volume. It should be noted that the purchase cost is the starting price tag and does not cover special materials, delivery, or

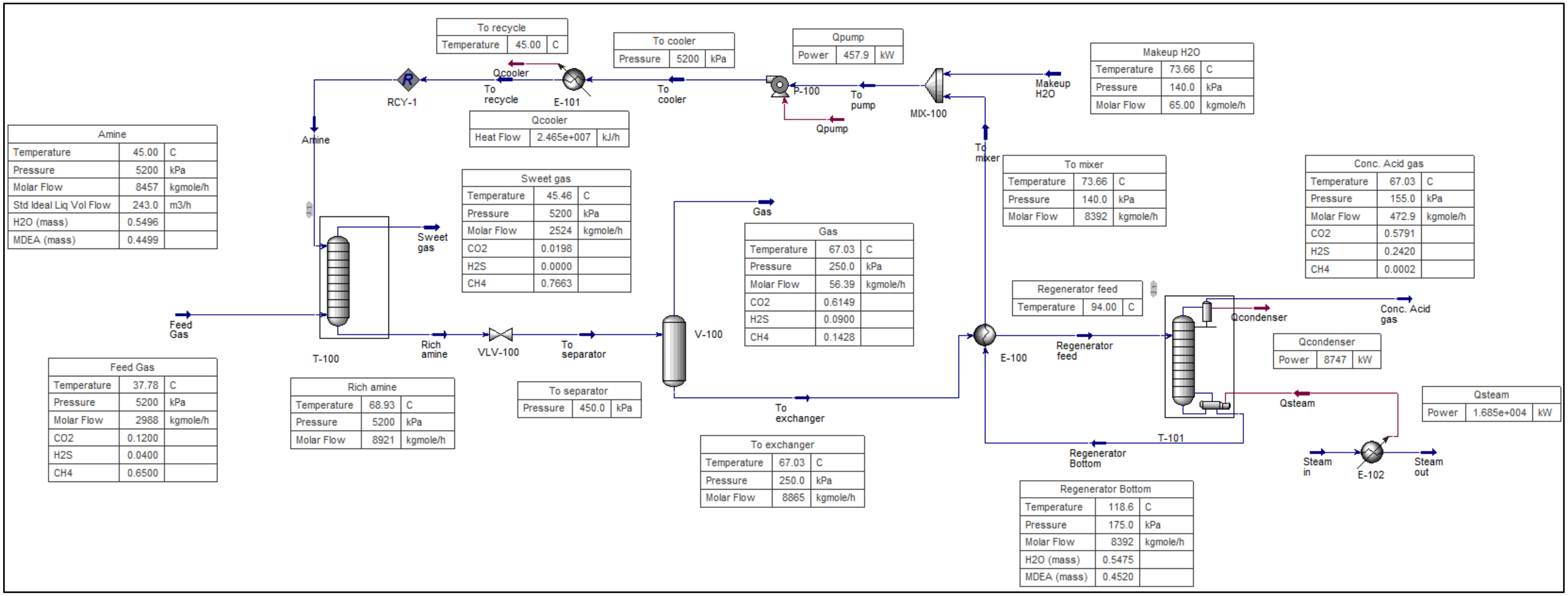

Figure 2. Solved flowsheet of the amine process to treat sour gas of west fields of Kuwait.

installation. The bare module (C_{BM}) equation covers the previous criteria:

$$C_{BM} = C_p^o(B_1 + B_2 F_M F_P) \quad (15)$$

where B_1 and B_2 are constants for each unit. F_m is related to the material of construction and F_p corresponds to unit operating pressure. The constants and equations for various units are given in the Appendix.

It is worth mentioning that CAPCOST was developed in 2001 and the prices are different today. So the following correction was used based on Chemical Engineering Plant Cost Index (CEPCI) to calculate the recent equipment cost:

$$\text{Updated Equipment Cost} = \frac{CEPCI_{2023}}{CEPCI_{2001}} \times C_{BM} \quad (16)$$

where $CEPCI_{2023}$ is 806 and $CEPCI_{2001}$ is 397. This means that the cost of equipment doubled from 2001 to 2023.

After the calculation of the main equipment, other units such as mixers, vessels, valves, etc, are assumed to be 10% of the equipment cost. Therefore, the capital fixed investment (CFI) is the total cost of all equipment. Following this, the cost of MDEA solvent is assumed to be 10% of CFI. Now CAPEX can be calculated by:

$$\text{Capital Expenditure (CAPEX)} = \text{CFI} + \text{Solvent Cost} \quad (17)$$

Based on the data in the Appendix, the price of equipment for the amine process to treat 70,000 $m^3\ h^{-1}$ of sour gas containing 12 mol% carbon dioxide and 4 mol% hydrogen sulfide is 12.9 million \$. The most expensive unit is the regenerator which includes the reboiler, condenser, and fired heater for steam generation. The regenerator material was assumed to be made from stainless steel with a design pressure of 7 bar. The tower consists of 23 stages of valve trays. The regenerator accounts for 48% of the total equipment cost. The price of the MDEA solution is 1.2 million \$ and this gives CAPEX of the amine process of 12.9 million \$. Calculation details of the economic assessment of the amine process are shown in Table 3.

The next step was to estimate the annual operating expenses (OPEX) based on the maintenance cost and utility bills as follows:

$$\mathrm{OPEX} = \mathrm{Maintenance\ Cost} + \mathrm{Utility\ Bills} \tag{18}$$

Table 3. CAPEX of the Amine Process for the Treatment of Kuwait Sour Gas.

Equipment	**Specifications**	**Cost ($)**
Absorber with Trays	SS / Valve Tray / 2m D / 12m H / 22 Stages / 62 bar	4,796,702
Regenerator with Trays	SS / Valve Tray/ 23 Stages / 7 bar	1,165,554
Regenerator Reboiler	CS / Kettle / 193 m^2/ LP Steam / 7 bar	1,264,695
Regenerator Condenser	CS / Fixed Tube / 83 m^2	164,664
Heat Exchanger (L/R Amine)	CS / Floating Head / 355 m^2 / 7 bar	361,874
Cooler (Amine Lean)	CS / 162 m^2 / Fixed Tube	189,739
Pumps	CS / Centrifugal / 458 kW / 60 bar	552,133
Steam Boiler	CS / Fired Heater / 16.8 MW	1,348,837
Cooling Towers	CS / Water / 8.7 MW	886,872
Others	Valves / Drums / Pipes / Mixers / Filters	1,073,107
Fixed Capital Investment (FCI)		11,804,176
MDEA Solvent	45 wt% MDEA	1,180,418
Total Capital Expenditure (CAPEX)		12,984,594

SS: Stainless steel. CS: Carbon Steel. D: Diameter. H: Height. LP: Low pressure. L/R: Lean/Rich. m: meter

The maintenance cost is assumed to be 10% of the fixed capital investment (CFI) while the utility bills are from the steam generation, pump, and cooling towers. The Steam production (S_c) cost was calculated based on the combustion of natural gas in a fired heater:

$$S_c(\$/ton) = \frac{a_F(H_g - h_f)}{450\,\eta_B} \tag{19}$$

where a_F is price of natural gas ($ per MMBtu), H_g is enthalpy of steam at 159°C (Btu), h_f is the enthalpy of the condensed steam at 159°C (Btu), and η_B is the boiler efficiency (assumed 0.86). Using the data from the simulation and a gas price of 2.5 $ per MMBtu, the steam production cost is 5.8 $ per ton. The required tons of steam is 232 per year and if we assume an annual operation duration of 8000 h, the steam generation cost is 1.3 million $ per year.

The utility bill of the amine pump is based on the power input and the electricity tariff as follows:

$$\text{Pump Utility Bill (\$/year)} = P \times RT \times t \quad (20)$$

where P is the power input of 458 kWh, RT is the electricity tariff of 0.0167 per kWh, and t is the annual operation time of 8000 h. This gives a bill of 60,822 $ per year.

The last bill is from the cooling towers. Mainly, the towers consist of tanks, pumps, and fans. The calculation of electricity input is based on the tons of cooling (1 ton = 18,000 Btu). For instance, the fan consumes 0.042 kWh per ton of cooling while the pump requires 0.00714 kWh per ton of cooling (*15*). Therefore, the following equations are used to calculate the cooling tower bills:

$$\text{Cooling Tower Utility Bill} = \text{Fan}_{\text{Utility Bill}} + \text{Pumps}_{\text{Utility Bill}} \quad (20)$$

$$\text{Cooling Tower Fans Utility Bill (\$/year)} = 0.042 \times TC \times RT \times t \quad (21)$$

$$\text{Cooling Tower Pumps Utility Bill (\$/year)} = 0.00714 \times TC \times RT \times t \quad (22)$$

where TC is the tons of cooling. The simulation states that the amine cooler needs 1,947 tons of cooling while the condenser (from the regenerator) requires 2,487 tons of cooling. This gives a bill of 28,938 $. Consequently, the total annual operating cost is 2.6 million $ as presented in Table 4.

Table 4. OPEX of the Amine Process for the Treatment of Kuwait Sour Gas

Bill	**Cost ($)**
Maintenance	1,180,418
Steam Generation	1,348,837
Pumps	60,822
Cooling Towers	28,938
Total Operating Cost	2,619,015

Membrane Process Simulation: The previous results tell that the amine process has a capital cost (CAPEX) of 12.9 million $ and an annual operating fee (OPEX) of 2.6 million $ to treat 70,000 m^3 h^{-1} of Kuwait sour gas. The process reduced carbon dioxide content from 12 to less than 4 mol% and hydrogen sulfide concentration from 4 mol% to less than 4 ppm. The second objective of this study is to use a commercial membrane unit made from Pebax® to treat Kuwait sour gas. The developed case in UniSim® is given in Figure 3. The simulation was run based on changing the membrane area and observing carbon dioxide and hydrogen sulfide content in the product gas. Hydrocarbons recovery was monitored as well and the data is presented in Table 5. Unfortunately, the membrane system did not reach the performance of the amine process due to the excessive loss of hydrocarbons. For example, the amine process recovers 99.6% of the hydrocarbons to produce less than 2 mol% carbon dioxide and 4 ppm of hydrogen sulfide. At this recovery rate, the membrane almost did not separate the impurities because the product gas still contains 11.9 mol% of carbon dioxide and 3.7 mol% of hydrogen sulfide. Increasing the membrane area caused a significant loss in hydrocarbons recovery. For example, using an area of 10,000 m^2 reduced carbon dioxide to 6.7 mol% and hydrogen sulfide to 0.6 mol% but hydrocarbons recovery was only 89.4%. This loss is unacceptable for the refinery as the lost hydrocarbons can be sold for profit. The third objective of this study is to integrate the membrane with the amine process to monitor the reduction in energy consumption.

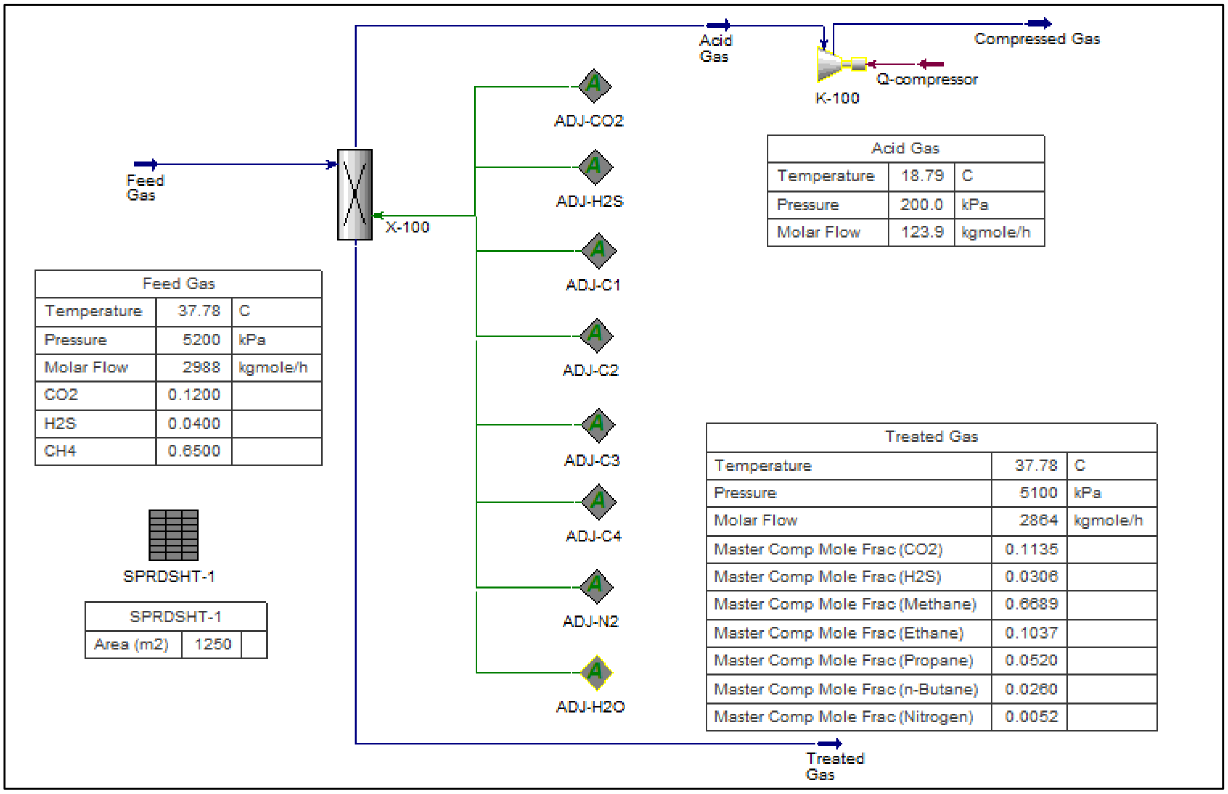

Figure 3. Developed membrane process in UniSim® for sour gas treatment.

Table 5. Performance of Pebax® Membrane for Sour Gas Treatment.

Membrane Area (m^2)	**Product CO_2 (mol%)**	**Product H_2S (mol%)**	**Hydrocarbons Recovery (%)**
10,000	6.7	0.6	89.4
5,000	9.1	1.5	94.6
2,500	10.6	2.4	97.0
1,250	11.3	3.1	98.6
750	11.7	3.4	99.2
350	11.9	3.7	99.6

Amine–Membrane Integrated Process: It was found before that the amine unit is an energy-intensive process due to the regeneration step. The membrane system could not reach the amine performance due to the high loss of hydrocarbons. In this section, the membrane was integrated to study the reduction in energy in the amine process. First, the sour gas enters the membrane where hydrocarbons recovery of 98.6% was achieved. This gives a product containing 11.3 mol% carbon dioxide and 3.1 mol% hydrogen sulfide as given previously in Table 5. This stream then enters the amine unit to produce the sweet gas. The solved simulation case is given in Figure 4 and the use of the membrane integration decreased the overall energy of the amine process by 16%. This reduction in operating energy was noticed in the steam regeneration, heat exchanger, cooler, and pump. Furthermore, the amine flowrate was dropped by 11%. It should be noted that the membrane system also consumes energy but it is only 1%. This gives a total reduction in energy of 15%.

The next step was to calculate the capital cost of the membrane system and estimate the new operating fee of the membrane and the modified amine process. After that, the data will be compared with the previous results where only the amine unit was used. From Figure 4, the required membrane area is 1,250 m^2 and it is estimated that the membrane skid (including pipes and valves) costs 500 $ per m^2 (*13*). So, the price of the membrane skid is 625,000 $. The membrane system requires also a compressor as the concentrated sour gas leaves the system at 2 bar. The compressor will increase the pressure back to 52 bar and this needs a power input of 329 kW. The cost of the compressor is about 1,000 $ per kW and therefore the compressor costs 329,000 $. To include the delivery and installation fee, 60% is added to the cost of the membrane skid and compressor. This gives a CAPEX of 1,526,400 $.

The operating cost of the membrane system is based on the maintenance fee and the compressor bill. The maintenance fee is assumed to be 10% of CAPEX and the compressor bill is calculated from the power input and electricity tariff similar to Equation 20. The annual operating cost is 196,330 $. Details of the economic assessment of the membrane system are shown in Table 6.

As stated before, the use of the membrane system reduced the overall energy input of the amine process by 15%. Table 7 shows a comparison between the amine unit and the membrane-amine integrated process. The amine standalone process recovered 99.6% of hydrocarbons while the integrated system recover 98.6%. The integrated process resulted in reduction in MDEA flowrate from 243 to 216 $m^3 h^{-1}$. The power input for the steam generation was dropped from 16.8 to 14.1 MW. Similarly, a decrease in energy consumption was noticed in the cooling towers and amine pump. This resulted in annual savings of 84,705 $ but the membrane system costs 1,526,400 $. Therefore, 18 years are needed to payback the membrane process which is not reasonable since the membrane lifetime is only five years. Furthermore, the membrane-amine integrated process caused a loss of 1% of hydrocarbons compared to the independent amine process. This stream accounts for 455 $m^3 h^{-1}$ of natural gas which gives a yearly flow of 3,640,000 m^3 (equivalent to 129,870 MMBtu). Using a natural gas price of 2.5 $ per MMBtu, the annual hydrocarbons loss is 324,675 $. This gives annual net savings of -239,970 $ in addition to the needed membrane capital cost of 1,526,400 $. This implies that the hydrocarbons loss is severe and the integrated process cannot compensate it. Thus, at the moment, the standalone amine process is more technically and economically feasible than the membrane or the integrated membrane-amine process. This was explained by the high capital cost of the membrane system and the tradeoff in hydrocarbons recovery.

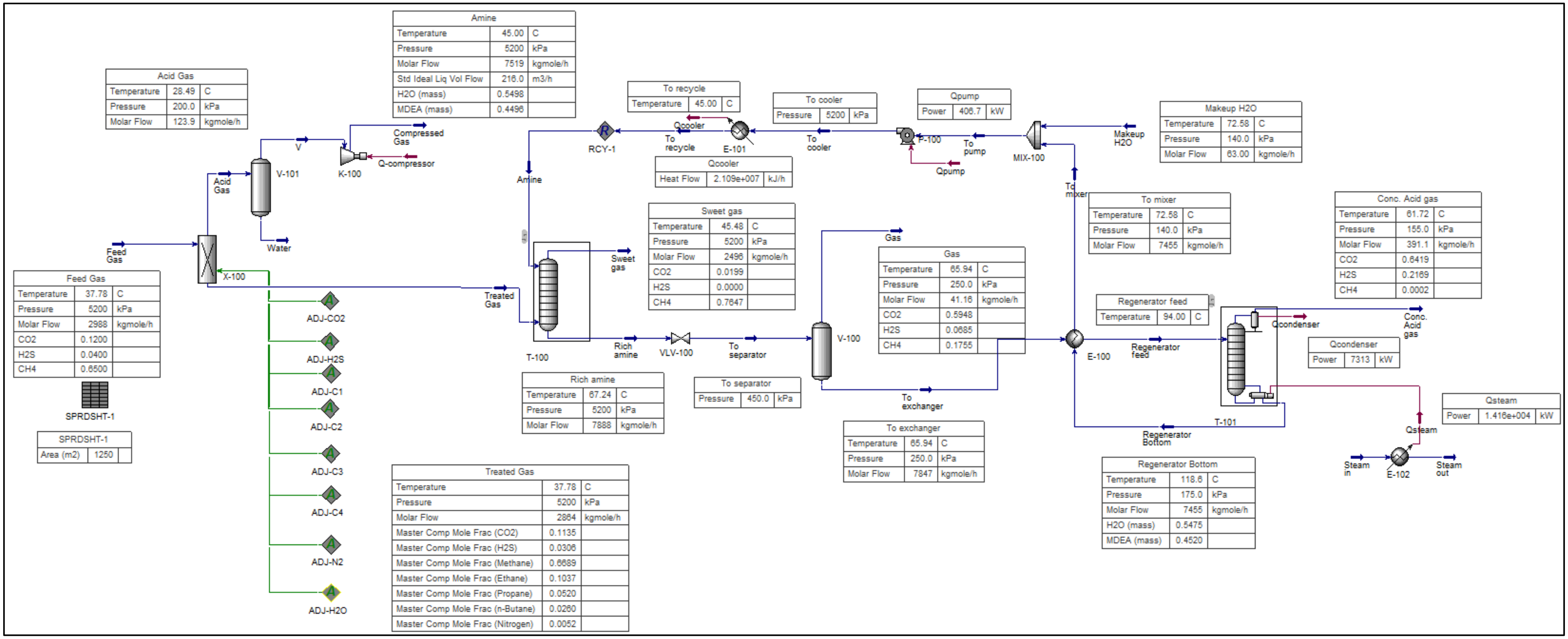

Figure 4. Process flowsheet of membrane–amine integration for the treatment of Kuwait sour gas.

Table 6. Economic Assessment of the Membrane System for Amine Process Integration.

Parameter	**Value**
Capital Cost	
Membrane skid	500 \$ per m^2
Membrane area	1,250 m^2
Membrane cost (M)	625,000 \$
Compressor price	1000 \$ per kW
Compressor power	329 kW
Compressor cost (C)	329,000 \$
Installation cost factor (F)	60%
Total capital cost (M+C)×1.6	1,526,400 \$
Annual Operating Cost[1,2]	
Maintenance Cost (MT)	10% Capital Cost
Electricity Tariff (TF)	0.0167 \$ per kWh
Compressor Bill (CB)	43,691 \$ per year
Total Operating Cost (MT+ CB×TF)	196,330 \$ per year

1) Labor cost is not covered. 2) Yearly operation of 8000h

Table 7. Comparison between the Amine Unit and Membrane-Amine Integrated Process.

Property	**Amine Process**	**Membrane–Amine Integrated Process**
Hydrocarbons Recovery	99.6%	98.6%
MDEA Flowrate	243 m^3 h^{-1}	216 m^3 h^{-1}
Steam Generation	16.8 MW	14.1 MW
Cooling Towers	15.5 MW	13.1 MW
Amine Pump	458 kW	407 kW
Membrane Compressor	–	329 kW
Membrane Capital Cost	–	1,526,400 $
Annual Operating Fee[1]	2,619,015 $	2,534,310 $
Annual Savings	–	84,705 $
Hydrocarbons Loss	–	1%
Annual Hydrocarbons Loss[2]	–	-324,675 $
Annual Net Membrane Savings	–	-239,970 $

1) Operation of 8000h. 2) Based on natural gas price of 2.5 $ per MMBtu.

4. Conclusions

In Kuwait, natural gas is produced during oil extraction and it is used mostly for power generation and hydrogen production. Still, the gas needs treatment before utilization as it contains significant amounts of carbon dioxide and hydrogen sulfide. Currently, the amine process is employed to produce a sweet gas containing 2 mol% carbon dioxide and 4 ppm hydrogen sulfide with hydrocarbons recovery of 99.6%. Despite the high efficiency of the amine unit, the process suffer many issues such as flooding, foaming, waste disposal, and high operating costs. Alternatively, polymeric membranes can provide a solution for gas separation with minimum energy with no liquid waste. In this project, a commercial membrane was developed and simulated in UniSim® for the treatment of the sour gas of west fields of Kuwait. Results showed that the membrane, unfortunately, was not capable of reaching the amine process performance in terms of product quality and quantity. The second step was to integrate the membrane unit to reduce the energy consumption of the amine process. Indeed, the membrane decreased the power input to the amine process by 15% with annual savings of 84,705 $. However, the membrane pay pack duration is 18 years which is unacceptable. Furthermore, the membrane caused a loss of 1% of hydrocarbons due to the tradeoff limitation in the polymeric membranes and cost 324,675 $ per year. Therefore, at the present, the amine process is favorable technically and economically compared to the membrane system.

Funding: The project was funded by Kuwait Institute for Scientific Research (KISR) under the code PF112K.

Conflict of Interest: The author declares no conflict of interest.

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Appendix

The following data was used to calculate the capital investment of the amine process for sour gas treatment.

<table>
<tr><th>Equipment</th><th>Equations</th><th>Input data</th><th colspan="6">Parameters</th></tr>
<tr><td rowspan="2">Tower
(Vertical Vessel)</td><td rowspan="2">$$C_{BM} = C_p^o(B_1 + B_2F_MF_P)$$
$$\log_{10} C_p^o = K_1 + K_2\log_{10}(V) + K_3[\log_{10}(V)]^2$$
$$V = \frac{\pi d^2}{4}l$$
$$F_P = \frac{\frac{(P+1)D}{2[850-0.6(P+1)]} + 0.00315}{0.0063}$$</td><td rowspan="2">Tower diameter (d)
Tower length (l)
Vessel pressure (P)</td><td>K_1</td><td>K_2</td><td>K_3</td><td>B_1</td><td>B_2</td><td>F_M</td></tr>
<tr><td>3.4974</td><td>0.4485</td><td>0.1074</td><td>2.25</td><td>1.82</td><td>3.1 (SS)</td></tr>
<tr><td rowspan="2">Tower
(Valve Trays)</td><td rowspan="2">$$C_{BM} = C_p^oNF_{BM}F_q$$
$$\log_{10} C_p^o = K_1 + K_2\log_{10}(A) + K_3[\log_{10}(A)]^2$$
$$A = \pi dl$$</td><td rowspan="2">Tray diameter (d)
Tray spacing (l)</td><td>K_1</td><td>K_2</td><td>K_3</td><td>F_{BM}</td><td colspan="2">F_q</td></tr>
<tr><td>3.3322</td><td>0.4838</td><td>0.3434</td><td>1.8 (SS)</td><td colspan="2">1 (N ≥ 20)</td></tr>
<tr><td rowspan="4">Regenerator
Reboiler (Kettle)</td><td rowspan="4">$$C_{BM} = C_p^o(B_1 + B_2F_MF_P)$$
$$\log_{10} C_p^o = K_1 + K_2\log_{10}(A) + K_3[\log_{10}(A)]^2$$
$$A = \frac{Q}{U\Delta T_{LMTD}}$$
$$\Delta T_{LMTD} = \frac{\Delta T_1 - \Delta T_2}{\ln\left(\frac{\Delta T_1}{\Delta T_2}\right)}$$
$$\Delta T_1 = T_{h,\text{in}} - T_{c,\text{in}}$$
$$\Delta T_2 = T_{h,\text{out}} - T_{c,\text{out}}$$</td><td rowspan="4">Reboiler duty (Q)
Overall heat coeff. (U)
Hot Stream temperatures
Cold stream temperatures
Reboiler area (A)</td><td>K_1</td><td>K_2</td><td>K_3</td><td>B_1</td><td>B_2</td><td>F_M</td></tr>
<tr><td>4.4646</td><td>-0.5277</td><td>0.3955</td><td>1.63</td><td>1.66</td><td>1 (CS)</td></tr>
<tr><td>C_1</td><td>C_2</td><td>C_3</td><td colspan="3" rowspan="2"></td></tr>
<tr><td>0.03881</td><td>-0.11272</td><td>0.08183</td></tr>
</table>

	$\log_{10} F_P = C_1 + C_2 \log_{10}(P) + C_3[\log_{10}(P)]^2$							
Regenerator Condenser (Fixed tube)	Same as the above equations	Condenser duty (Q) Overall heat coeff. (U) Hot Stream temperatures Cold stream temperatures	K_1	K_2	K_3	B_1	B_2	F_M
			4.3247	-0.303	0.1634	1.63	1.66	1 (CS)
			C_1	C_2	C_3			
			0.03881	-0.11272	0.08183			
Steam Generator (Fired heater)	$C_{BM} = C_p^o F_M F_P$ $\log_{10} C_p^o = K_1 + K_2 \log_{10}(Q) + K_3[\log_{10}(Q)]^2$	Fired heater duty (Q)	K_1	K_2	K_3	F_M	F_P	
			7.3488	-1.1666	0.2028	2.13 (CS)	1 (P < 20 barg)	
Rich/Lean Amine Heat Exchanger (Floating Head)	$C_{BM} = C_p^o(B_1 + B_2 F_M F_P)$ $\log_{10} C_p^o = K_1 + K_2 \log_{10}(A) + K_3[\log_{10}(A)]^2$ $A = \frac{Q}{U \Delta T_{LMTD}}$ $\Delta T_{LMTD} = \frac{\Delta T_1 + \Delta T_2}{2}$ $\Delta T_1 = T_{h,\text{in}} - T_{c,\text{in}}$ $\Delta T_2 = T_{h,\text{out}} - T_{c,\text{out}}$ $\log_{10} F_P = C_1 + C_2 \log_{10}(P) + C_3[\log_{10}(P)]^2$	Heat exchanger duty (Q) Overall heat coeff. (U) Hot Stream temperatures Cold stream temperatures	K_1	K_2	K_3	B_1	B_2	F_M
			4.8306	-0.8509	0.3187	1.63	1.66	1.0 (CS)
			C_1	C_2	C_3			
			0.03881	-0.11272	0.08183			

<table>
<tr><td>Amine Cooler
(Fixed Tube)</td><td colspan="8">Same as Regenerator Condenser but the LMTD equation is changed to:
$\Delta T_{LMTD} = \dfrac{\Delta T_1 + \Delta T_2}{2}$</td></tr>
<tr><td rowspan="4">Pump
(Centrifugal)</td><td rowspan="4">$C_{BM} = C_p^o(B_1 + B_2 F_M F_P)$
$\log_{10} C_p^o = K_1 + K_2 \log_{10}(P) + K_3[\log_{10}(P)]^2$
$\log_{10} F_P = C_1 + C_2 \log_{10}(P) + C_3[\log_{10}(P)]^2$</td><td rowspan="4">Pump power input (P)</td><td>K_1</td><td>K_2</td><td>K_3</td><td>B_1</td><td>B_2</td><td>F_M</td></tr>
<tr><td>3.3892</td><td>0.0536</td><td>0.1538</td><td>1.89</td><td>1.35</td><td>1.60 (CS)</td></tr>
<tr><td>C_1</td><td>C_2</td><td>C_3</td><td colspan="3" rowspan="2"></td></tr>
<tr><td>-0.3935</td><td>0.3957</td><td>-0.00226</td></tr>
<tr><td>Cooling Towers
(CT)</td><td>CT Capital Cost = 200$ per ton of cooling</td><td colspan="7">Tons of cooling (Q)
Assume water temperature from 30–45°C</td></tr>
<tr><td>Other equipment
(Vessels, mixers,
filters, …)</td><td colspan="8">10% of equipment cost (total of the above)</td></tr>
<tr><td>Fixed Capital
Investment (CFI)</td><td colspan="8">The total cost of the above (all equipment)</td></tr>
<tr><td>Solvent Cost</td><td colspan="8">10% of the fixed capital investment (CFI)</td></tr>
<tr><td>Capital
Expenditure
(CAPEX)</td><td colspan="8">Capital Expenditure (CAPEX) = CFI + Solvent Cost</td></tr>
</table>