Membrane-Distillation Hybrid Process to Reduce the Energy Footprints in Abu-Attifel Gas Plant

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Abstract

This study investigates the possibility of reducing the heat load required in the refrigeration cycle used in the NGL plant at Abu-Attifel filed. Gas membrane filters were installed pre to the propane refrigeration cycle. These membranes will recover most of light hydrocarbons from the stream entering to the refrigeration cycle, and thus an increase in the boiling point of the natural gas that is being processed occurs. The addition of the membrane filters was found to decrease the amount of heat load in both the refrigeration cycle and the heat duty in the condenser in the de-ethanizer column. Material and energy balances were made on the modified process. Two cases of light hydrocarbon recoveries were examined (85%, and 90%). A polymath code was developed to solve the design transport equations for the membrane units to determine the required membrane area and flow rates and compositions around the membrane. The existing gas plant process were also simulated and the results were compared with the existing operational data for validation purposes and reasonable agreements were obtained. The preliminary results indicate that there is considerable savings that can be attained when using gas membrane elements before the natural gas is fed to the cooling cycle section.

Keywords: Gas membrane; refrigeration cycle; propane recovery.

1. Introduction

Abu-Attifel field is one of Mellitah oil and gas company concession for production of oil and gas. It is located about “450 km” south-East of Benghazi in the great sand sea. Abu-Attifel reservoir is in production since 1970. Water is injected at pressure equal to 2133 psia at a volumetric ratio of 2:1 with the oil. The crude oil streams from the production wells are collected into arrival in manifolds and conveyed to the gas-oil separation plant, where the oil, gas, and water are separated using the necessary facilities in Abu-Attifel oil field and gas production:

- Gas oil separation.
- Oil stabilization.
- Gas compression.

1.1. Gas Processing

Abu-Attifel field produces about 100,000 BopD of crude oil. Oil produced by three separation stages is collected and stabilized by gas boots and then stored in atmospheric oil storage tanks. The stabilized oil is transported north a distance of “132 km” through a 30 inch diameter pipeline to concession (103A) which belongs to Zueitina oil company and then north through 40 inch diameter pipeline a distance (222 km) long to Zueitina terminal. The compression facilities are designed to compress all the medium pressure gas; 2nd stage and 3rd stage gas, as well as the boot gas to the first stage pressure of 700 psia- The atmospheric gas from the gas boots is compressed by turbo compressor to about 86 psia and sent to second turbo compressor where it is combined with the third stage gas Before being compressed by the second stage gas compressor. This compressor raises the pressure to about 256 psia. The gas is then combined with the 2nd stage gas and compressed by turbo compressors. C_3 to the first stage pressure each compression stage is equipped with inlet cooler and scrubbers and condensate removal. The liquid hydrocarbon recovered by gas plant were combined with Zueitina condensate and stabilized at concession (103A) and then sent to Zueitina terminal fractionation facilities through a (20 inch) pipeline, to be separated into C_3, C_4, C_5^+ products. The dry gas produced from Zueitina gas plant is compressed and re injected into the Inlisor field, at concession (103 D) to improve the recovery of Zueitina oil company field at (103A) and (103D) most at the associated gas was then continuously flared. As a result, the National Oil Corporation (NOC) and Mellitah Company decided to construct as gas plant at Abu-Attifel field to treat about (384 MM ScfD) of associated gas. The gas plant is designed to recover NGL mixture from the associated gas stream produced at Abu-Attifel.

1.2. NGL Recovery Plant

The NGL recovery plant recovers the liquids from the associated gases coming from Abu-Attifel oil center. In addition, natural gas treatment, and condensate and heavy liquids extraction are carried out in the NGL recovery plant.

1.3. The NGL Plant Products

- NGL stream (i.e. C_3^+ cut) to be sent into (10 inch) new pipeline liquid down by others.
- The dry gas stream (C_1, C_2) to be sent into the existing 34″ gas line.
- Liquid recovery process is based on straight refrigeration technology by the use of an external mechanical refrigeration cycle where the refrigerating medium is propane.

1.4. Product Stream Characteristics

- Dry gas (C_1, C_2).
- C_3^+ cut (NGL) or condensate.

1.5. Dry Gas Steam

- Water dew point less than -50 °F at 615 psia.
- Hydrocarbon dew point less than -50 °F at 615 psia.

1.6. Delivery Conditions (Upstream Pressure Control Valve)

- $P \geq 595$ psia
- Temperature ≤ 85 °F

1.7. Specification of NGL Stream (C_3^+)

- Ethane mole fraction not higher than 0.002
- Delivery conditions (at pipe line pump discharge upstream flow control valve)
- $P \geq 800$ psia and $T \leq 85$ °F

1.8. NGL Recovery Plant Stages

1. Raw gas cooling and liquid separation system.
2. Condensate stabilization system.
3. Hot oil system.
4. Gas drying system.
5. Chilling train and dry gas metering system.
6. De-Ethanizer system.
7. Condensate delivering and metering system.
8. Propane Refrigeration system.
9. Process water degassing system.

1.9. Objectives

The overall goal of this study is to investigate the possibility of adding membrane elements to recover the light components from natural gas before entering into the refrigeration cycle to minimize the head load needed in both refrigeration cycle and deethanizer column are follows:

1. Developing the process flow diagram for the new proposal.
2. Overall mass and energy balances for the proposal in three different cases.
3. Cost evaluation of the new proposal.

2. Material and Methods

2.1. New Proposal

In this study an attempt was made to utilize membrane technology for gas separation to reduce the energy load requirement in both the refrigeration cycle and condenser in the de-ethanizer column. The removal of light hydrocarbons from the natural gas stream using membrane would increase the dew point of the gas stream, and as a result would minimize the heat removal needed to partially condensate the natural gas stream before introduced to the de-ethanizer column. The heat removal needed in the fractionation tower in the condenser will also be lower when natural gas stream contains less light hydrocarbons. The specifications of (C_3^+) produced were maintained as in the original design. The commercial software polymath was used to solve the material balance and design equations for the membrane section. The results obtained for recovered gas (i.e. flow rate and composition) were used as feed input to the commercial software HYSYS to examine the effect of feed flow rate and composition and the head load.

2.2. Gas Separation Using Membrane

Targets of membrane separation through a thin film where feed gas, at high pressure P_1, contains some low molecular weight species (MW < 50) to be separated from small amount of higher-molecular-weight species usually a sweep gas is not needed, but the other side of the membrane is maintained at much lower pressure P_2; often near-ambient to provide an adequate driving force. The membrane is deuce; these species are absorbed at the surface and then transported through the membrane by one or more mechanisms. Then, perm selectivity depends on both membrane absorption and transport rate mechanisms are formulated in terms a partial pressure or fugacity driving force using the solution-diffusion model of the products are permeate enriched in A and a retentive enriched B. If the membrane is micro porous, pore size is extremely important because it's necessary to block the passage of species B otherwise, unless molecular weights of A and B differ appreciably, only a very modest separation a achievable as was discussed in connection with Knudsen diffusion.

2.3. Design Equation for Gas Separation Membranes

$$nr_j = nf_j - nf_j - nr_j \tag{2.1}$$

Table 2.1: composition and average relative volatilities

Stream	(205) Feed	(134) Top	(24) Bottom	(α_i) Average volatilities
CH_4	0.36000	0.56137	-	10.0430
C_2H_6	0.23899	0.36129	0.002262	2.97800
C_3H_8	0.19001	0.02496	0.484720	1.0000
i-C_4H_{10}	0.04860	0.00017	0.135198	0.4125
n-C_4H_{10}	0.05858	0.00065	0.162270	0.34880
i-C_5H_{12}	0.01850	-	0.051519	0.15667
n-C_5H_{12}	0.01852	-	0.051602	0.12224
n-C_6H_{14}	0.01033	-	0.028790	0.04490
CO_2	0.05344	0.06020	-	3.81470
N_2	0.0016	0.00307	-	100.317

$$Am_j = np_j - Pm_j\left(\frac{Pf_j - 2 \times Pp_j + Pr_j}{2}\right) \tag{2.2}$$

$$np_j = np_j - Pm_j\left(\frac{Pf_j - 2\times Pp_j + Pr_j}{2}\right) = 0 \tag{2.3}$$

$$Pf_j = Pf_j - \frac{Pf_j}{\sum_{j=1}^{n} nf_j} \tag{2.4}$$

2.4. Design Equation for Fractionation Tower

A complete fraction nation de-ethanizer column (307 - VE-01), was designed to fulfill the following requirements. The height, diameter and number of valve trays required valve trays selected for the column design which gives more effective separation with high efficiency over a wide range of operating conditions that permits reduction of the reflux. The compositions in mole fraction of fed, over head and bottom products and average relative volatilities are shown in Table 1 and 2. The minimum number of theoretical plates (Ntheo) was computed using Fensk's equation:

$$N_{min} = \frac{\log\left(Y_{LKD}/Y_{hKD}\right) \times \left(X_{hKW}/X_{LKW}\right)}{\log\left(\alpha_{LK}/\alpha_{hK}\right)_{av}} \tag{2.5}$$

L.K: The light key component. Taken as ethane.

h.K: The heavy key component. Taken as propane.

Calculations lead to $N_{min} = 8.028$

Since the feed is a mixture of saturated vapor and liquid with liquid percentage of 64%.

Using Underwood equation to calculate R_{min}:

$$\sum \frac{\alpha_i X_{fj}}{\alpha_i - \theta_j} = 1 - q \tag{2.6}$$

$$\sum \frac{\alpha_i X_{fj}}{\alpha_i - \theta_j} = R_{min} + 1 \tag{2.7}$$

Trial and error solution was executed to find $\theta = 1.292957 \cong 13$

Substituting (θ) in equation (2.7), the minimum reflux ratio was found to be $R_{min} = 0.3036$

The reflux ratio was taken as $1.5\,R_{min}$, which is 0.4554.

The number of theoretical trays were found using Gilliland correlation:

$$\frac{R_t - R_{min}}{R_t + 1} = 0.083 \tag{2.8}$$

$$\frac{N - N_{min}}{N + 1} = 0.65 \tag{2.9}$$

By substituting, $N_{Theo} = 28$ trays.

Overall efficiency is estimated using O'Connell equation:

$$\eta E_{\circ} = 51 - 32.5 \log\left(\mu\, \alpha_{av}\right) \tag{2.10}$$

Viscosity calculations shown in Appendix (A)

$\mu_1 = 0.0785$ cp, $\alpha_{av} = 2.97$, and $\eta E_{\circ} = 70\%$

Actual number of plates:

$$N_{actual} = \frac{N_{theo}}{E_{\circ}} = \frac{28}{0.7} = 40 \text{ trays}$$

Table 2.2: Specification of the de-ethanizer

Item	
Hight of the column	111.54 ft
Distance between trays at the top	23 ft
Distance between trays at the middle	20 ft
Distance between trays at the bottom	19 ft
Diameter at the top	78 ft
Diameter at the bottom	101.68 ft
Diameter at the middle	91.84 ft
Tray type	Valve tray

2.5. Validation

The purpose of this section is to compare simulation results with original operational data to verify the viability of the package used in the study.

Table 2.3: Final results of the de-Ethanizer

Item	
Temperature at the top	-33 °F
Pressure at the top	366.9 psia
Vapor density at the top	2.412 lb/ft^3
Gas flow rate	4516 lb/hr
Temperature at the bottom	210 °F
Pressure at the bottom	372.7 psia
Pressure drop in the column	6 psia
Liquid density at the bottom	26.73 lb/ft^3
Liquid flow rate	2.267×10 bbl / day

Table 2.4: Comparing between existing data and simulation results for the de-ethanizer column

Item	Existing data	Simulation results
No. of stages	40	40
Q_c, (Hp)	18520.29	16176.492
Q_R, (Hp)	26530.43	24162.5
C_2 mole fraction at top	0.451	0.3723
C_2 mole fraction at bottom	0.00207	0.0019
C_3 mole fraction at top	0.0003	0.0001
C_3 mole fraction at bottom	0.5109	0.4912
Condenser Temperature, °F	-27.4	-32.91
Reboiler temperature, °F	208.58	210.9

2.6. Effect of Membrane on Refrigeration Load

Table 2.5: De-ethanizer simulation results using gas membrane with recovery (85-90%) of methane

Item	85% Recovery	90% Recovery
Refrigeration cycle load, (Hp)	11827.32	10469.13
Temp. at top, °F	-30.77	-30.51
C_2 mole fraction at top	0.3871	0.3954
C_2 mole fraction at bottom	0.0015	0.0012
C_3 mole fraction at top	0.0001	0.0002
C_3 mole fraction at bottom	0.4549	0.5607
Q_c, (Hp)	10458.126	15936.753
Q_R, (Hp)	171356.91	30816.296

Table 2.6: Summary results of cost estimations

Item	Original operation	New proposal 85%	New proposal 90%
Cost of refrigeration cycle load, ($ / day)	2214.0854	1683.1354	1489.85
Cost of refrigeration cycle load, ($ / 5 years)	3941072.012	2995981.0	2651933
Net profit, ($ / 5 years)			
Gas membrane cost, ($)	-	447134.6	562527.91
Revenues, ($ / 5 years)	-	527956.412	726611.102
Revenues at end of project, ($)	-	10559128.24	14532222.04
ROR (%)	-	16.04%	21%

2.7. Membrane Area

Table 2.7: The results from gas membrane separations using polymath

Item	Area, (ft^2)	Membrane cost, ($)
Recovery 85%	4.49×10^4	417134.6
Recovery 90%	6.055×10^4	562527.91
Material selection	Rubber	-
Cost	100 ($/m^2$)	-

2.8. Cost Estimation

Cost is summarized in Table 2.6

3. Conclusion and Recommendation

1. Gas membrane separation technology was found to be capable of reducing the content of light hydrocarbons in the natural gas fed to the gas processing.

2. The preliminary results indicate that the amount of heat load in the refrigeration cycles was decreased which consequently minimizes the cost associated with the cooling process to produce the required two phase stream just after the refrigeration cycle.

3. The cost estimation results show that there is a considerable saving that can be attained when using gas membrane elements before the natural gas is fed to the cooling cycle section. This amount is proportional to the amount of light hydrocarbons that are recovered in the membrane section.

4. The rate of return for this new proposal was found to be 26.211 % when the designed recovery was 85 % and was found to be 28.35 % when the designed recovery was set to 90 %.

5. Further and more detailed and accurate optimization to determine the optimum membrane recovery is recommended before this proposal can considered for implementation.

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