Petroleum and Coal

Article **Open Access**

Enhancement and Optimization of an Egyptian Natural Gas Processing Plant

Abeer M. Shoaib[1], Ahmed A. Bhran[1,2]*

[1] *Department of Petroleum Refining and Petrochemical Engineering, Faculty of Petroleum and Mining Engineering, Suez University, Suez, Egypt*

[2] *Chemical Engineering Department, College of Engineering, Al Imam Mohammad Ibn Saud Islamic University, Al Riyadh, Saudi Arabia*

Received April 10, 2019; Accepted July 21, 2020

Abstract

The objective of the present study is to develop and optimize an existing Egyptian natural gas processing plant. The existing plant uses a methanol injection system for the purpose of hydrate prevention, which has many disadvantages. In this work, two alternative solutions are introduced to overcome drawbacks issued by using methanol. The first alternative is to use a tri-ethylene glycol (TEG) contactor for dehydration, while the second alternative is to use mono ethylene glycol (MEG). The suggested alternatives are comparatively studied to select the best alternative for the investigated plant. Although both of the two studied alternatives can achieve product within the required specifications, it is found that the second alternative is better from the viewpoint of economics. Additionally, operating conditions of the natural gas plant using MEG injection have been studied and optimized to achieve the maximum annual profit for the considered plant. Two correlations relating the gas conditions at the injection point to the productivity of sales gas and condensate are introduced.

Keywords: *Natural gas; gas dehydration; Condensate production; NGL recovery.*

1. Introduction

Natural gas is composed mainly of methane besides some heavier hydrocarbons such as ethane, propane, etc. It also contains some non-hydrocarbon as, for example, hydrogen sulfide and carbon dioxide. The standard specifications of sales gas should be met for the natural gas to be sold. Consequently, the wellhead natural gas, which is generally saturated with water vapor, will require a dehydration process to remove water from it [1-3]. The dehydration process is necessary since the presence of water vapor in natural gas causes safety issues during its transportation. As an example, the presence of water can lead to gas hydrate formation, which is the most significant restrictions when treating, storing, and transporting natural gas. Removing water from natural gas through dehydration not only adjusts the gas dew point but also avoids hydrate formation and pipeline blockage as well as reduces the possibility of equipment corrosion [4-6].

Many research work in the literature discussed the various methods used for gas dehydration, which include solid and liquid desiccant and cooling/refrigeration with glycols/methanol systems [7–11]. From the comparison done between various approaches used for natural gas dehydration, it is concluded that solid desiccant adsorption technique using silica gel, alumina or molecular sieves as adsorbent is favored for achieving very low dew point [7, 12]. However, gas dehydration by liquid desiccant absorption using glycol is economically attractive because it needs lower energy compared to the adsorption process. This can be attributed to the existence of hydroxyl groups in glycols, which makes hydrogen bonds with water molecules. This consequently makes glycol a good absorber for water. The most widely used glycol is Tri-Ethylene Glycol (TEG) due to its high water affinity, high hygroscopicity, regeneration capability, high chemical stability, low evaporation loss rate, low vapor pressure, and low thermal degradation rates during regeneration [13-18]. Recently, Shoaib et al. introduced an optimization program aiming to improve the natural gas dew point through studying the operating conditions of the dew point control unit and their influence on the dew point and condensation production rate [19].

One of the most effective and reliable methods to avoid hydrate formation is hydrate inhibitors injection because they reduce the temperature at which hydrate can be created or/and delay hydrate formation. Alcohols such as methanol, diethylene glycol (DEG) and monoethylene glycol (MEG), are commonly used as thermodynamic hydrate inhibitors in the oil and gas industries. Methanol changes hydrate formation conditions by decreasing water activity. However, a large quantity of methanol is lost in the gas phase in comparison with glycols [20-24]. Additionally, MEG is preferred over DEG for cases where the temperature is -10°C or lower because the viscosity of MEG is increased significantly at low temperatures [20].

The gas processing plant considered in the present study was established to separate condensates from associated natural gas and prevent hydrate formation using methanol injection before the sweetening process. However, the use of methanol has many disadvantages; the daily use of methanol is about 4000 liters, which costs $4,000 /day, which is very expensive. Additionally, disposal of methanol is achieved through a specialized company with extra cost. Moreover, methanol is very dangerous to the environment with the possibility of ignition.

The purpose of this research work is to introduce alternatives for methanol as a hydrate inhibitor to reduce the operating cost as well as harmful environmental impact. Two alternative solutions are introduced and discussed from the viewpoint of quality and cost. The first alternative considers using the tri-ethylene glycol (TEG) contactor for the dehydration process, while the second alternative uses mono-ethylene glycol (MEG). Moreover, the impact of operating conditions of the gas at the injection point on both sales gas and condensate production is studied. Correlations relating these operating conditions to sales gas production as well as condensate production are introduced. Optimization of the operating conditions aiming to maximize plant revenue is also studied in this work.

2. Case study

The case study presented in this work is for an Egyptian Petroleum company located in the Western Desert with a gas plant capacity up to 20 MMSCFD. Figure 1 shows the process flow diagram of the investigated gas processing unit. The feed gas composition is shown in Table 1.

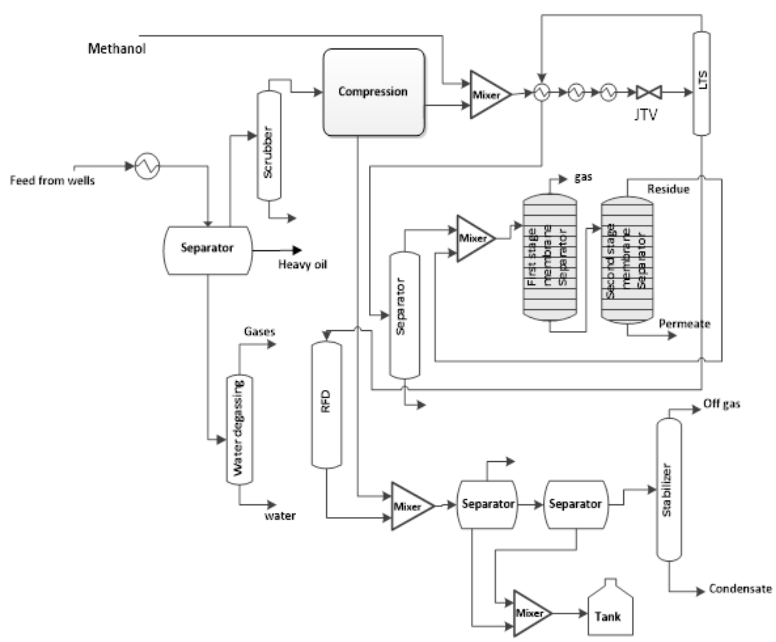

Figure 1. Process flow diagram for an Egyptian natural gas processing unit

Table 1. Feed gas composition of the considered natural gas processing unit

Component	Mole fraction	Component	Mole fraction
H_2O	0.2705884519	n-Decane	0.00589790699
H_2S	0.00000882288	Undecanes*	0.00489826174
CO_2	0.0280	Dodecanes*	0.00439843911
Nitrogen	0.00719744582	Tridecanes*	0.00409854554
Methane	0.03588172928	Tetradecanes*	0.00369868744
Ethane	0.09373063935	Pentadecanes*	0.00309890028
Propane	0.07782544015	Hexadecanes*	0.00249911313
i-Butane	0.01149592041	Heptadecanes*	0.00219921956
n-Butane	0.02479120228	Octadecanes*	0.00179936146
i-Pentane	0.01079616873	Nonadecanes*	0.00159943241
n-Pentane	0.01119602684	Eicosanes*	0.00129953883
n-Hexane	0.01969301149	Heneicosanes*	0.00089968073
M-cyclopentane	0.00259907766	Docosanes*	0.00069975168
Benzene	0.00109960978	Tricosanes*	0.00049981815
Cyclohexane	0.00439843911	Tetracosanes*	0.00039985452
n-Heptane	0.00959659443	Pentacosanes*	0.00029989089
M-cyclohexane	0.00279900671	Hexacosanes*	0.00029989358
Toluene	0.00409854554	Heptacosanes*	0.00019992726
n-Octane	0.0082970556	Octacosanes*	0.00019992905
Benzene	0.0011995743	Nonacosanes*	0.00009996363
m-Xylene	0.00249911313	Triacontanes*	0.00009996363
p-Xylene	0.00059978715	Hetriacontanes*	0.00009996363
o-Xylene	0.00139950335	Dotriacontanes*	0.00009996363
n-Nonane	0.00689755225	Tritriacontanes*	0.00009996363
124-MBenzene	0.00089968073		

The associated gas feed comes from Jasmine and Melihah west deep wells at temperature and pressure of 3°C and 5.5 barg, respectively. The inlet gas is heated to 30°C in a gas/gas heat exchanger for emulsion breaking enhancement and water separation. Then, it is sent to a three-phase separator to be separated into three main streams; gas, heavy oil, and water. The upper gas stream from the three-phase separator is then sent to a scrubber to remove any associated liquids. The resulted gas stream is compressed in a four-stage compressor to increase the pressure from 2 barg to 139.7 barg; in each stage, the compressed gas is cooled and sent to a scrubber to remove any produced liquids. Methanol, which is injected into the gas stream, comes out of a four-stage compressor to inhibit hydrate formation. In order to remove C_1^+ (hydrocarbons heavier than methane) and adjust the dew point, the gas is cooled by passing through three heat exchangers, Joule Thomson valve (JTV) and low-temperature separator (LTS), where the temperature is decreased gradually in the exchangers to reach 18 °C and fed to the JTV. The gas in the JTV is exposed to a reasonable pressure drop resulting in a sudden decrease in temperature (from 18°C to -2.58°C). The outlet gas is fed to the LTS to separate gas (mainly methane) from the produced condensate.

The top product from the LTS is heated to 30°C in a heat exchanger, then fed to a separator to remove any remaining liquids. The resulted gas, which is free of condensates, is now passed through a two-stage membrane for the purpose of acid gas removal. The obtained sales gas composition is shown in Table 2.

Table 2. Composition of the produced sales gas

Component	Mole fraction	Component	Mole fraction
H_2O	< 0.0001	Ethane	0.1561
Methanol	0.0003	Propane	0.0702
H_2S	< 0.0001	i-Butane	0.0049
CO2	0.0199	n-Butane	0.0073
Nitrogen	0.0148	i-Pentane	0.0011
Methane	0.7241		

The separated condensate is sent to a recycle flash drum to separate a top product (gas), which is sent to the third stage of compression, and then heated and sent to a filter coaleser to separate water-methanol mixture from condensate. The condensate is sent to a stabilizer to reduce the Reid vapor pressure (RVP); then, it is passed through a heat exchanger for cooling before storage or sale.

3. Results and discussion

As indicated before, two alternative solutions are suggested to overcome the disadvantages of using methanol as a hydrate inhibitor. The following subsections discuss the two considered alternatives. HYSYS Version 10, with the fluid package; Peng-Robinson EOS [25-26] was applied as the simulation tool in the current study. This simulation software considers also the economic calculations required to assess and make a comparison between the original plant and the two proposed modified plants.

3.1. Replacement of methanol injection system with a glycol tower (contactor)

The first suggested solution to improve the investigated plant is to use a TEG as an absorbent for water. The original plant, as well as the plant modified with TEG, were simulated as presented in Figures 2 and 3, respectively.

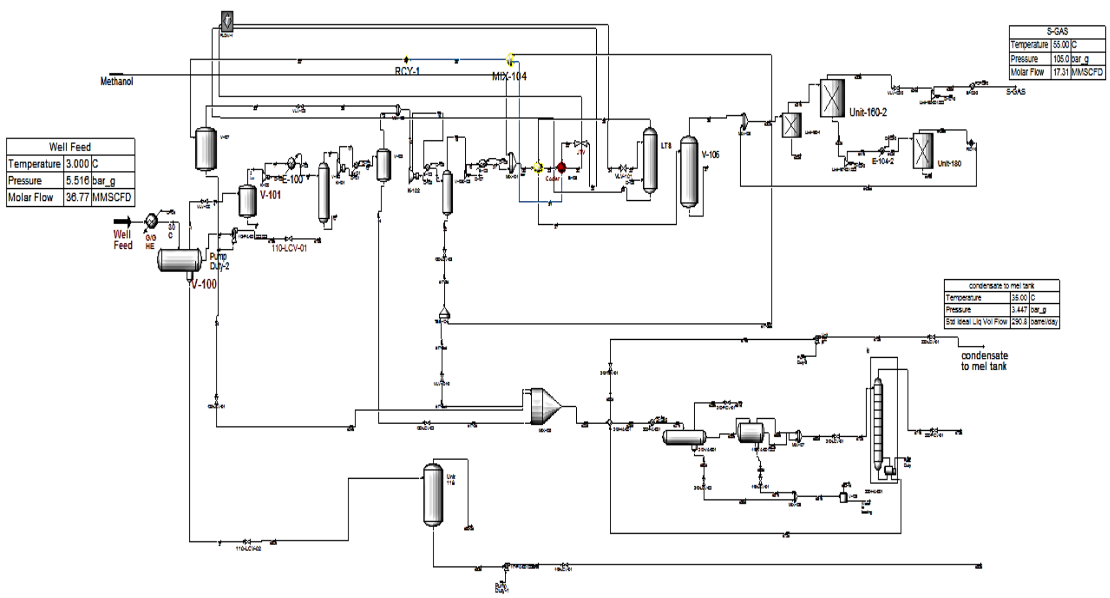

Figure 2. The simulated process flow diagram of the original plant

TEG contactor has many advantages compared to the present methanol hydrate inhibition technique. From these advantages, TEG is recoverable [27], and the daily operating cost in the case of using TEG is much lower compared to the cost in case of using the methanol approach. It was found that the monthly consumption of TEG that can effectively achieve the desired degree of hydrate inhibition is only 4 barrels. The needed operating pressure of the proposed glycol tower is 60 bar. Thus, the present four-stage compressor should be replaced with a two-stage compressor. The pressure of 60 bar is suitable for the sweetening process, so there is no need to use the JTV. Therefore, the JTV should be replaced with a propane chiller for cooling at the same pressure. It was also noticed that replacing the methanol injection system with TEG tower works efficiently, and the same sales gas specification is obtained. However, it was found that condensate production from the investigated gas processing plant would be reduced by about 70 bbl/day. This reduction can be attributed to the miscibility of the condensate constituents with TEG.

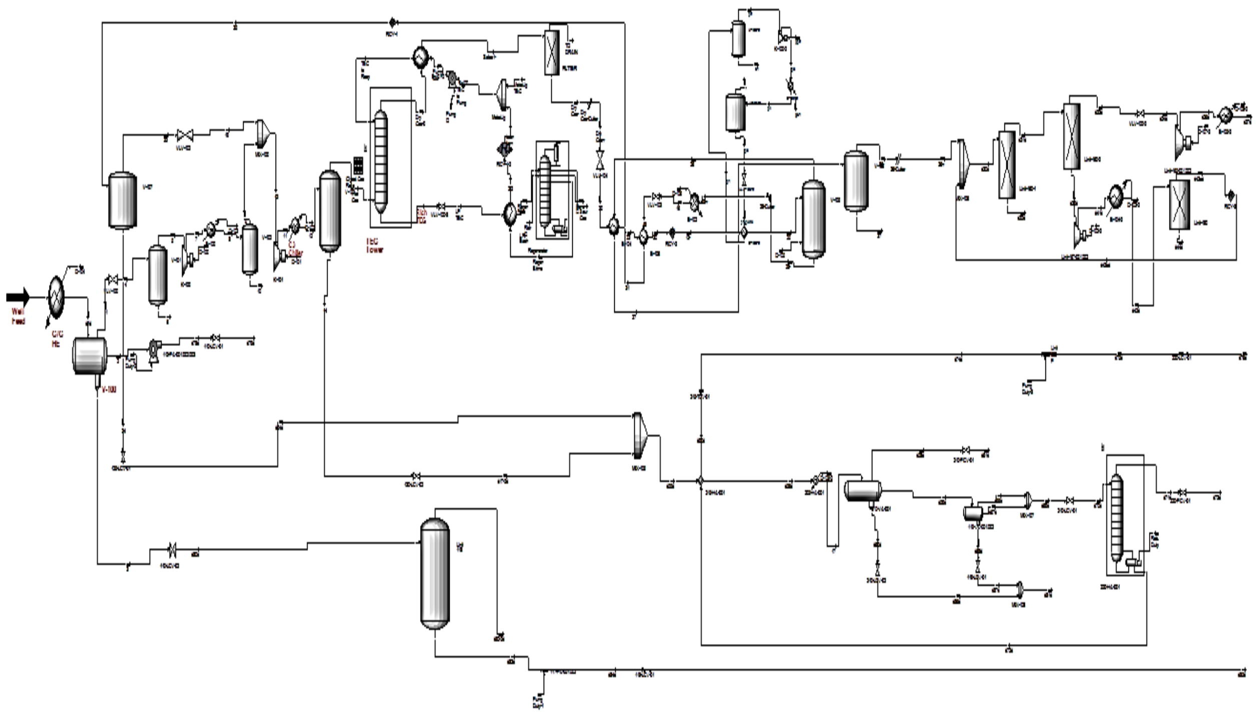

Figure 3. The simulated process flow diagram of the plant modified for the use of TEG contactor for the dehydration process

The simulated economic results calculated by Aspen Process Economic Analyzer showed that the annual cost of the consumed methanol is $1,460,000 in addition to a disposal cost of $11,500. Accordingly, the total annualized operating cost related to methanol usage is about $1,471,500. On the other hand, the annual consumption of TEG would be 8,271 L with a cost of $10,339. The proposed modification of using TEG instead of methanol for hydrate prevention requires the installation of a contactor and a new two-stage compressor, which costs $ 4,300,000. In conclusion, using TEG contactor instead of a methanol injection technique leads to a yearly saving in operating cost of about $1,461,161. The estimated return on investment (ROI) for applying the TEG contactor route is 0.34, which is very reasonable and indicates the effectiveness of this modification from the economic viewpoint.

3.2. Using MEG instead of methanol as a hydrate inhibitor

The second suggested an alternative for improving the original gas plant is directed to replace methanol by MEG as a hydrate inhibitor in order to overcome the above-mentioned problems come from using methanol as well as to increase the plant profitability. The simulated plant, modified according to the second alternative, is presented in Figure 4. It should be noticed that the use of MEG needs the installation of a regeneration system for MEG recovery, which costs $245,800/year. The annual consumption of MEG is 52.56 tons, which costs $58,000. The estimated saving in the annual operating cost in case of using MEG instead of methanol is $1,413,500. Therefore, it is clear that the proposed modification has many benefits; this is because MEG is recoverable, does not need the high capital cost for installation of the MEG regeneration system, environmentally safe and applied with the lower operating cost. The ROI of this suggested modification is only 0.167, which is very low and indicates the validity of the introduced modification route. From the above discussion and calculations, it is clear that replacing methanol by MEG is the best alternative for improving the present plant due to its lower ROI, in addition to its other benefits compared to methanol. Thus, this alternative is recommended economically and environmentally to be applied for enhancing the existing natural gas processing plant.

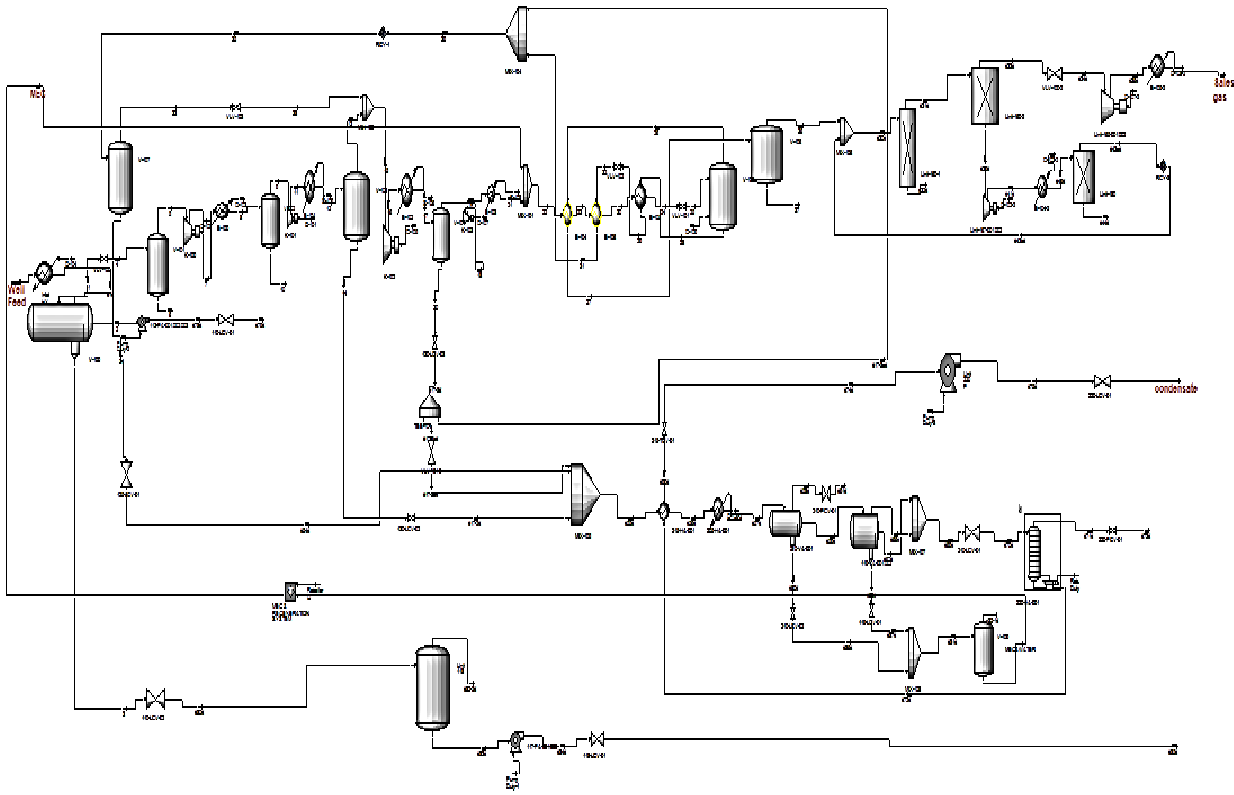

Figure 4. The simulated process flow diagram of the plant modified for the use of MEG instead of methanol as a hydrate inhibitor

3.3. Optimization of the operating condition to maximize profit

The current research work considers also studying the effect of operational variables, especially temperature and pressure of the natural gas at the injection point of MEG on the sales gas and condensate productions. The results illustrated in Table 3 show that the increase of pressure leads to a decrease in sales gas and condensate productions while increasing the temperature, the sales gas and condensate productions are increased. Two correlations are introduced in this study to relate the sales gas as well as condensate production to the temperature and pressure of the gas at the injection point. Equations 1 and 2 are derived using regression analysis to correlate the sales gas and condensate productions with the gas conditions (temperature and pressure), respectively. The obtained R square values of equations 1 and 2 are 0.98 and 0.96, respectively. This consequently confirms the validity of the proposed correlations.

$$\text{Sales Gas production} = 19.61-0.035\,P+1.46E\text{-}5\,P^2-0.00045\,T^2+0.000494\,PT \quad (1)$$

$$\text{Condensate production} = 0.40458-0.00142\,P+1.5E\text{-}6\,P^2-1.1\,E\text{-}5\,T^2+1.54\,E\text{-}5\,PT \quad (2)$$

Where sales gas and condensate productions are in MMSCFD, pressure, and temperature are in barg and ^oC, respectively.

The introduced correlations are used for building up an optimization program aiming to obtain the gas optimum operating conditions (at the injection point) at which the maximum plant profit can be achieved. LINGO software version 17 is used as the optimization tool to determine these optimum conditions. For optimization calculations, the price of sales gas and condensate is taken as \$2.34/m^3 and \$66.5/bbl, respectively. The optimization results listed in Table 4 show that the maximum annual profit of \$182.7x10^7 could be achieved at pressure and temperature of 100 barg and 67.2 ^oC, respectively. This corresponds to sales gas and condensate production of 17.54 and 0.3314 MMSCFD, respectively.

Table 3. Effect of pressure and temperature at injection point on the sales gas and condensate production

Pressure, bar-g	Temperature, oC	Sales gas flowrate, MMSCFD	Condensate flowrate, MMSCFD
100	59	17.54671208	0.327631286
102	59	17.609061	0.327930209
104	59	17.60259275	0.32595482
106	59	17.57188451	0.327002874
108	59	17.58993954	0.325593595
110	59	17.58981083	0.334248778
112	59	17.58005509	0.325337919
114	59	17.57445594	0.329613433
116	59	17.57058233	0.324849658
118	59	17.56866016	0.324503265
120	59	17.55702442	0.326330496
122	59	17.55598089	0.324241505
124	59	17.55333558	0.323979478
126	59	17.54903304	0.323797996
128	59	17.54432084	0.323632228
130	59	17.53958886	0.323473049
132	59	17.53494968	0.323318205
134	59	17.53049122	0.323167408
136	59	17.5260406	0.323019773
138	59	17.52178672	0.32287495
140	59	17.51765715	0.322733776
142	59	17.51364916	0.322595857
144	59	17.50975987	0.322461102
146	59	17.50598579	0.322329452
148	59	17.50232364	0.322200859
150	59	17.49876883	0.322075252
152	59	17.49531818	0.321952571
154	59	17.49196815	0.32183276
156	59	17.48871523	0.321715767
158	59	17.48555603	0.321601526
160	59	17.50155742	0.321852961
140	40	17.03831587	0.303247264
140	42	17.1069864	0.305209078
140	44	17.16209846	0.307723777
140	46	17.21203763	0.309621361
140	48	17.27588028	0.312773461
140	50	17.3254083	0.31476853
140	52	17.37115701	0.316670698
140	54	17.38523415	0.317793096
140	56	17.43571553	0.320467238
140	58	17.47027651	0.32199002
140	60	17.50360089	0.323308926
140	62	17.54469368	0.323913569
140	64	17.55315968	0.326299378
140	66	17.57247058	0.327758034
140	68	17.59346655	0.328902205
140	70	17.61270695	0.329994505
140	72	17.63777489	0.330878209
140	74	17.63928786	0.332220764
140	76	17.65143647	0.333271042
140	78	17.66277339	0.334295589
140	80	17.67592328	0.335328336

Table 4. Comparison between the current and optimized gas processing plants

Parameter	Current plant	Optimized plant
Sales gas production (MMSCFD)	17.51	17.54
Condensate production (MMSCFD)	0.3241	0.3314
Temperature (oC)	59	67
Pressure (bar-g)	140	100
Annual profit ($)	179.54×10^7	182.7×10^7

4. Conclusion

Egypt has begun to realize the importance of the gas industry and works to develop this industry, which becomes very important all over the world. Therefore, the present work is directed to improving and optimizing an existing Egyptian gas processing plant to increase its profitability. The present plant uses methanol injection techniques to avoid hydrate formation, which can cause problems when treating, storing, and transporting natural gas. Therefore, two suggested alternatives consider the hydrate inhibition process is introduced in this study. The first introduced alternative considers replacing the original methanol injection system with a dehydrator tower applying TEG as an absorbent of water where it is responsible for hydrate formation. The second alternative considers the replacement of the existent methanol injection system with the MEG injection system. The original plant, as well as the modified plants according to the two proposed modification routes, are simulated via HYSYS Version 10, with Peng-Robinson EOS. The results, including the economic study of the original as well as the two proposed suggestions, showed that the best modification route for improving the investigated gas plant is the second alternative solution. This can be attributed to the lower ROI value (0.167) in addition to other benefits in case of using the MEG injection technique compared to the original plant, and the plant modified by the first suggested an alternative approach.

The present study also introduces two correlations that relate the sales gas and condensate production with the operational conditions (pressure and temperature) of the gas at the injection point of MEG. Additionally, LINGO software version 17 is used in this research work to determine the optimum gas conditions at which the maximum plant profit can be obtained. The results indicate that the maximum annual profit of $\$182.7 \times 10^7$ could be achieved at the temperature and pressure of 67.2°C and 100 barg, respectively.

References

[1] Sayed A, Ashour I, Gadalla M. Integrated process development for an optimum gas processing plant. Chemical Engineering Research and Design, 2017; 124: 114–123.

[2] ES: 1606/2008, 2008. Natural Sales Gases Egyptian Standards. Egyptian Organization for Standardizations & Quality, Egypt.

[3] Younger AH. Natural Gas Processing Principles and Technology- Part I, vol. 1–5. University of Calgary, Canada, 366 (2004).

[4] Rahimpour MR, Saidi M, Seif M. Improvement of natural gas dehydration performance by optimization of operating conditions: A case study in Sarkhun gas processing plant. Journal of Natural Gas Science and Engineering, 2013; 15: 118-126.

[5] Mokhatab S, Poe WA. Handbook of Natural Gas Transmission and Processing. Gulf Professional Publishing (2006).

[6] Carroll J. Dehydration of Natural Gas. 2014: 175–195. http://dx.doi.org/10.1016/b978-0-12-800074-8.00006-5.

[7] Petryk MR, Khimich A, Petryk MM, Fraissard J. Experimental and computer simulation studies of dehydration on microporous adsorbent of natural gas used as motor fuel. Fuel, 2019; 239: 1324-1330.

[8] Gandhidasan P, Al-Farayedhi AA, Al-Mubarak AA. Dehydration of natural gas using solid desiccants. Energy, 2001; 26: 855–68.

[9] Karimi A, Abdi MA. Selective dehydration of high-pressure natural gas using supersonic nozzles. Chem. Eng. Process, 2009; 48: 560–8.

[10] Gas Processors Suppliers Association. GPSA Engineering Data Book.12th ed. Tulsa, OK: Gas Processors Suppliers Association; (2004). Section16, 1–13.
[11] Bahadori A, Mokhatab S. Simple correlation accurately predicts densities of glycol solutions. Pet. Sci. Technol. 2009; 27: 325–30.
[12] Netusil M, Ditl P. Comparison of three methods for natural gas dehydration. J Nat Gas Chem., 2011; 20(5): 471–476.
[13] Kong ZY, Mahmoud A, Liu S, Sunarso J. Revamping existing glycol technologies in natural gas dehydration to improve the purity and absorption efficiency: Available methods and recent developments. Journal of Natural Gas Science and Engineering, 2018; 56: 486–503.
[14] Bahadori A. Natural Gas Processing, Technology and Engineering Design, Chapter 9 - Natural Gas Dehydration. 2014: 441-481. http://dx.doi.org/10.1016/b978-0-08-099971-5.00009-x.
[15] Paymooni K, Rahimpour MR, Raeissi S, Abbasi M, Baktash MS. Enhancement in triethylene glycol (TEG) purity via hydrocarbon solvent injection to a TEG+water system in a batch distillation column. Energy Fuels, 2011; 25: 5126–5137.
[16] Bahadori A, Vuthaluru HB, Mokhatab S. Analyzing solubility of acid gas and light alkanes in triethylene glycol. J. Nat. Gas Chem., 2008; 17: 51–58.
[17] Rincón MD, Jiménez-Junca C, Duarte CR. A novel absorption process for small-scale natural gas dew point control and dehydration. J. Nat. Gas Sci. Eng., 2016 ; 29 : 264–274.
[18] Chebbi R, Qasim M, Jabbar NA. Optimization of triethylene glycol dehydration of natural gas, Energy Reports, 2019; 5: 723–732.
[19] Shoaib AM, Bhran AA, Awad ME, El-Sayed NA, Fathy T. Optimum operating conditions for improving natural gas dew point and condensate throughput. Journal of Natural Gas Science and Engineering, 2018; 49: 324–330
[20] Bahadori A. A simple mathematical predictive tool for estimation of a hydrate inhibitor injection rate. Nafta, 2011; 62: 213-223.
[21] Bahadori A, Vuthaluru HB. Prediction of methanol loss in vapor phase during gas hydrate inhibition using Arrhenius-type functions. J. Loss Prev. Process Ind.,2010; 23: 379-384.
[22] Elgibaly AA, Elkamel AM. A new correlation for predicting hydrate formation conditions for various gas mixtures and inhibitors. Fluid Phase Equilibria, 1998; 152: 23-42.
[23] Ghiasi MM, Bahadori A, Zendehboudi S, Jamili A, Rezaei-Gomari S. Novel methods predict equilibrium vapor methanol content during gas hydrate inhibition. J. Nat. Gas Sci. Eng., 2013; 15: 69-75.
[24] Kamari A, Bahadori A, Mohammadi AH, Zendehboudi S. New tools predict monoethylene glycol injection rate for natural gas hydrate inhibition. Journal of Loss Prevention in the Process Industries, 2015; 33: 222-231.
[25] AEA technology PLC, Thermodynamic and HYSYS, 2000
[26] HYSYS. Users guide. Tech. rep., Aspen Technology Inc.; 2009.
[27] Wu M, Wang S, Liu H. A study on inhibitors for the prevention of hydrate formation in gas transmission pipeline. J. Nat. Gas Chem.,2007; 16: 81-85.

To whom correspondence should be addressed: professor Abeer M. Shoaib, Department of Petroleum Refining and Petrochemical Engineering, Faculty of Petroleum and Mining Engineering, Suez University, Suez, Egypt, E-mail: abeer.shoaib@suezuniv.edu.eg