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Research paper

Optimization of triethylene glycol dehydration of natural gas

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ABSTRACT

Dehydration using triethylene glycol (TEG) is widely used in natural gas processing to avoid corrosion and plugging of the flow lines. Optimization using Aspen HYSYS simulator and optimizer tool was performed to minimize the processing cost considering different sets of parameters: TEG circulation rate, numbers of theoretical trays (in the absorber and the stripping gas column), feed gas pressure and temperature, gas flow rate, gas price level and stripping gas rate. In addition to typical dew point depression simulation results reported in other publications, the present investigation also provides the minimum processing cost including both utilities and capital cost. The results based on the parametric optimization study yield criteria for design and optimum operating conditions for the dehydration process.

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1. Introduction

Natural gas is a major source of energy and a source of feedstocks used for the production of a large number of petrochemicals; therefore, its production is vital for the world economy. The gas needs to be dehydrated in order to avoid corrosion following water condensation especially in the presence of acid gases, and plugging that may occur if the natural gas temperature reaches the hydrate formation temperature (Manning and Thompson, 1991, p. 139).

Every plant (whether simple or complex) typically has a dehydration unit. There are different alternatives for natural gas dehydration. Reviews can be found in Neagu and Cursaru (2017) and the references therein. The most widely used techniques are: (i) liquid desiccant absorption techniques using TEG typically, (ii) solid desiccant adsorption with silica gel, alumina or molecular sieves as adsorbent, and (iii) cooling/refrigeration with glycols/methanol used to prevent dehydration (Manning and Thompson, 1991; Kidnay et al., 2011; GPSA, 2012). With TEG dehydration, a dew point depression of 60 °F–120 °F is typically reached (GPSA, 2012). Molecular sieves units are more costly from both capital and operation sides, but a very low dew point of –150 °F can be attained (GPSA, 2012). Netusil and Dittl (2011) compared the energy requirements for the three above alternatives.

Usually a triethylene glycol dehydration unit, circulating a highly concentrated solution of TEG, is used to absorb H₂O in a contactor (absorber). The regeneration of TEG by removal of

water gained from natural gas is performed in a regeneration unit where heat is provided to remove the absorbed water. More removal of H₂O requires higher concentrations of TEG, which may necessitate the use of stripping gas or other techniques including Drizo, Coldfinger and vacuum processes (GPSA, 2012). The Drizo process can achieve a dew point depression of 180–220 °F, and for the stripping gas, vacuum, and Coldfinger processes, the water dew point depression is typically 100–150 °F. BTEX and VOCs in natural gas are partially absorbed by TEG in the contactor. Environmental aspects addressing BTEX and VOCs emissions to the ambient air at the top of the regenerator are considered for TEG units in (Braek et al., 2001; Darwish et al., 2004; Darwish and Hilal, 2008).

Gupta et al. (1996) performed a sensitivity analysis, along with a chemical analysis, to investigate and reduce excessive TEG losses from a gas dehydration unit in Bombay offshore. Gandhidasan (2003) did a parametric study to find correlations for the effects of saturated feed gas pressure and temperature on the water content of the gas, the dry gas dew point including a temperature approach to equilibrium taken as 8 °C, the TEG circulation rate, the minimum TEG concentration and the design of the contactor. Jokar et al. (2014) considered the replacement of the existing TEG absorber plate column in Farashband plant by a structured packing one and found the change to be economic. Chukwuma and Jacob (2014) investigated the effects of parameters including the TEG circulation rate and the number of plates in the absorber on the water content of the dry gas in a TEG unit in Niger Delta. Increasing the stripping gas rate was found to have more impact than higher reboiler temperature level. Ranjbar et al. (2015) followed a relative sensitivity approach to investigate the effects of TEG circulation rate and the inlet

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List of symbols

B	Objective function in Eq. (2)
B_r	B/B_{ref}
B_{ref}	B for the reference case
BTEX	Benzene, toluene, ethylbenzene and xylenes
C1	methane
C2	ethane
C_{Bm}	Equipment bare module cost
C_{OL}	Cost of labor
C_{RM}	Cost of raw materials
C_{UT}	Cost of utilities
C_{WT}	Cost of water treatment
COM_d	Cost of processing
CEPCI	Chemical engineering plant cost index
DP	Dew point
DPD	Dew point depression = reduction in water dew point after TEG absorption
EOS	Equation of state
FCI	Fixed capital investment (total module cost)
GP	Gas price
HYSYS	Flowsheet simulator
LTS	Low temperature separator
M	Thousand
MM	Million
N	Theoretical number of trays in the contactor
N_A	Theoretical number of trays in the regenerator
N_B	Theoretical number of trays in the stripper
P	Pressure
S	Standard conditions: 1 atm and 288.7 K (60 °F)
SGR	Stripping gas rate
T	Temperature
TEG	Triethylene glycol
VOC	Volatile organic compound
\$	US dollar

gas temperature to the absorber of a feed gas at 115 bar and 50 °C. The optimum values found were lower TEG rate and inlet gas temperature, drier gas and lower reboiler duty. Kamin et al. (2017) used Design Expert software along with HYSYS simulator, and data for a feed gas at 115 bar and 50 °C to Farashband plant in order to keep the glycol loss, the reboiler duty, the dry gas water content and the hydrate formation temperature within specific ranges by optimizing the TEG circulation rate, the number of plates in the absorber and the reboiler temperature. Parametric study results by Olbrich (Olbrich, 1988; Manning and Thompson, 1991) include dew point depressions at 600 psia (41.4 bars) and two temperatures 80 °F (299.8 K) and 100 °F (310.9 K) for different TEG circulation rates and concentrations. Curves from GPSA (2012) provide water removal fraction versus circulation ratio for different TEG concentrations at 1000 psia (68.9 bar) and 140 °F (333.2 K). A technical and economic analysis was performed by Neagu and Cursaru (2017) by considering the effect of the reboiler temperature and stripping gas rate (injected into the reboiler) for a feed gas at 30 °C and 4100 kPa to a TEG dehydration unit.

They found that stripping gas has little impact on the fixed capital cost, and concluded that enhancing the stripping gas rate is an appropriate way to enhance TEG concentration and decrease the dry gas dew point.

In contrast to the previous investigations mentioned above, the objective of the present investigation is to minimize the processing cost of TEG units including capital cost, and utility requirements for the reboiler, TEG pumping, and stripping gas. Due to the fact that the required dew point of the dry gas depends on the country (Neagu and Cursaru, 2017), optimization is performed for a number of design parameters including TEG circulation rate to provide more general results. The water dew point depression caused by TEG dehydration can be increased by enhancing TEG concentration (a key factor), and to some extent, the TEG circulation rate and the number of trays in the contactor - (Figs. 8 – 11 – 8 – 20 in Manning and Thompson, 1991). Enhancing the TEG concentration requires higher reboiler temperature. However, above 404 °F (206.5 °C) TEG degradation occurs, in which case stripping gas is required. For TEG concentrations higher than 99 wt% stripping gas can be used (Fig. 8 in Parrish et al., 1986 in which N_B is the number of equilibrium stages below the reboiler).

Process simulation along with a costing model and optimization are required. In the present work, design variables are varied to obtain the minimum processing cost. In addition to typical dew point depression simulation results reported in other publications (Olbrich, 1988; Manning and Thompson, 1991; GPSA, 2012), the present investigation also provides the minimum processing cost including both utilities and capital cost. The process simulation and optimization are presented first, followed by a discussion of the results and conclusions regarding the selection of the absorber pressure and temperature levels, the stripping gas rate, the TEG circulation rate and the numbers of theoretical trays in the absorber and stripper.

A typical dehydration process is shown in Fig. 1. The feed gas enters the contactor from the bottom and, through countercurrent contact with the TEG entering the contactor from the top, loses part of its water content picked up by the TEG leaving from the bottom. The rich glycol is sent to a flash tank where it loses light hydrocarbons, then to the top of the stripper to provide a reflux, minimize TEG vapor losses, and gain temperature increase. Heat integration is achieved through heat exchange with the lean TEG leaving the reboiler. In order to reuse the lean TEG in the contactor operating at high pressure, a circulation pump is required as the regenerator column pressure (about atmospheric) is significantly lower than the contactor pressure. Before entering the contactor, heat exchange between the lean glycol and the dehydrated gas leaving the absorber from the top achieves further cooling of the lean glycol following prior cooling by heat transfer between the lean glycol and the rich glycol. Fig. 1 shows a dehydration unit equipped with a stripping column used to remove H₂O from the TEG leaving the reboiler with stripping gas picking up part of the water from the TEG leaving the regenerator. The stripping gas is sent to the bottom of the regenerator after leaving the stripper. Another option is to discard the stripping column and to send the stripping gas into the reboiler.

2. Process simulation

Optimization of TEG units first requires simulation of a base case. Aspen HYSYS simulation was performed using the TST (Twu–Sim–Tassone) EOS glycol package. The model is recommended for TEG dehydration modeling as it provides accurate estimates for water content, activity coefficients, and dew point temperature (Aspen HYSYS Simulation Basis, 2004). A TEG dehydration unit typically has a contactor, a regenerator column, a

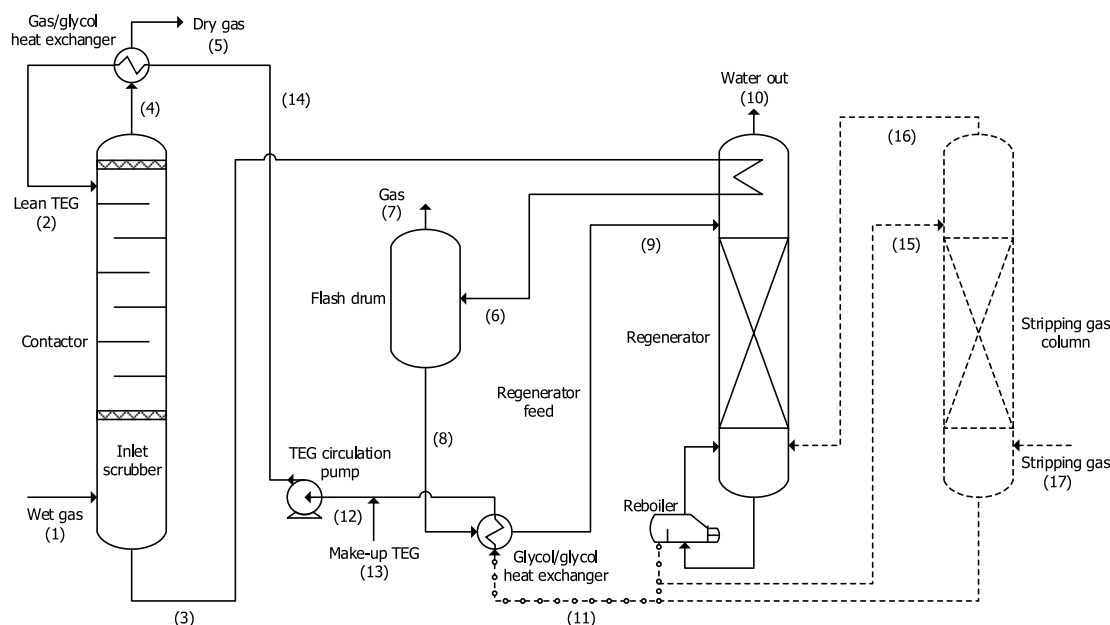


Fig. 1. Process for TEG dehydration stream (○) without stripping gas and stream (- -) with stripping gas.

stripping column (for high TEG concentrations), a glycol-gas heat exchanger, a cooling coil, a reboiler, a flash tank, a lean glycol-rich glycol heat exchanger, pumps and filters. The decrease in dew point following dehydration is called water dew point depression. The objective is to meet the required dew point for the dried gas, or similarly to reach a specific dew point depression, given the dew point - of the inlet gas to the absorber. The TEG circulation rate is typically in the range of 2–5 gal TEG/lb H₂O removed (GPSA, 2012).

Higher dew point depressions require higher TEG concentrations. Above 99 wt% TEG, stripping gas is widely used as an option to regenerate TEG at atmospheric pressure as the reboiler temperature should not exceed 404 °F (479.8 K) to avoid degradation of the glycol. As in Olbrich (1988), the feed composition selected consists of 91 mole % of methane (C1) and 9 mole % of ethane (C2). The different levels for the feed pressure, temperature and flow rate are shown in Table 1.

3. Process optimization

A cost model was embedded in the simulation study. The original multi-variable steady state Optimizer in Aspen HYSYS Version 8.8 was used to minimize the cost. We distinguish three cases: (i) case a: no stripping gas used for TEG wt% of 99 and less, (ii) case b: $N_B = 0$ (direct injection into the reboiler with no stripping column), and (iii) case c: N_B equal to one or more (stripping column used with injection above the reboiler).

3.1. Design parameters

The design parameters are fixed in each optimization case. They include the feed condition given by any of the cases in Table 1, the circulation rate m , gal TEG/lb H₂O in feed gas, the number of theoretical trays N in the contactor, and the number of theoretical trays N_B in the stripping gas column. In addition, two gas prices are considered: low and high.

3.2. Design variables

In contrast to the design parameters fixed in each optimization run, the design variables are changed by the optimizer to reach the lowest dehydration cost.

The process variables selected for optimization are:

- Case a: (i) temperature of rich glycol into the regenerator
- Case b: (i) temperature of rich glycol entering the regenerator and (ii) rate of stripping gas sent into the reboiler
- Case c: (i) temperature of rich glycol entering the regenerator, (ii) rate of stripping gas (from the stripping column) sent above the reboiler

The other variables are either fixed as discussed in the methodology section or varied as parameters as discussed in the design parameters section.

3.3. Costing and objective function

The dehydration requirement is to meet the maximum water content allowed for the sales gas; this is directly related to the minimum water dew point depression to be achieved in the absorber. As the target is gas dehydration to meet a specified water removal, optimizing the process consists of minimizing the cost while meeting the dew point depression requirement. The cost of processing (excluding the cost of depreciation) is given by (Turton et al., 2009):

$$COM_d = 0.180 FCI + 2.73 C_{OL} + 1.23 (C_{UT} + C_{WT} + C_{RM}) \quad (1)$$

where FCI is the fixed capital investment, C_{UT} is the utility cost, C_{OL} is the cost of labor and C_{RM} is the cost of raw materials. The cost of water treatment C_{WT} is discarded, and the cost of labor does not change. As we need to minimize the cost of processing, we define the objective function as

$$B = 0.180 FCI + 1.23 (C_{UT} + C_{RM}) \quad (2)$$

The cost of raw materials is calculated as basically the cost of make-up TEG to compensate for the losses. The losses of hydrocarbons in the regenerator are neglected, but the losses of stripping gas (if applicable) are included as utilities.

FCI is taken as the total module cost using

$$FCI = 1.18 \sum_{i=1}^n C_{Bm,i} \quad (3)$$

The correlations of the bare module costs, $C_{Bm,i}$, are given in Turton et al. (2009). The bare module costs are updated according

Table 1
Feed condition.

Feed composition (dry basis)	Mole %
C1	91
C2	9
Feed pressure levels	300 psia (2068 kPa), 600 psia (4137 kPa) and 1400 psia (9653 kPa)
Feed temperature levels	80 °F (299.8 K) and 100 °F (310.9 K)
Gas flow rate (dry gas)	80 MMscfd (26.2 Sm ³ /s) and 160 MMscfd (52.4 Sm ³ /s)

Table 2
Natural gas prices and costs of steam and electricity used.

Cost	Gas price \$/MSm ³	Steam \$/1000 kg	Electricity \$/1000 kWh
Low	105.6	8.29	29.9
High	429.4	33.75	121.6

to

$$C_{\text{Bm, year B}} = C_{\text{Bm, year A}} \frac{\text{CEPCI}_{\text{year B}}}{\text{CEPCI}_{\text{year A}}} \quad (4)$$

The details related to the cost correlations used are given in the [Appendix](#).

The electricity cost for the pump was calculated based on the gas price and the heating value of the gas. Two limiting cases for the gas price, \$2.99/Mscf (May 2016) and \$12.16/Mscf (July 2008) (EIA, 2016) were considered in the analysis. The unit costs used for steam and electricity are given in [Table 2](#). The TEG cost used is \$0.7/lb, and the FCI cost was estimated as of July 2018 using a CEPCI cost index of 607.3 ([ChemEngOnline, 2018](#)).

The total module cost FCI requires the bare module costs of the process vessels (absorber, vertical flash separator, regenerator and gas stripping column if applicable), heat exchangers (glycol–glycol, gas–glycol and regenerator reboiler), absorber trays and demister pads, packing in regenerator and gas stripping column, and pumps (the only moving part in the TEG dehydration unit). One spare pump was added to the total module cost.

4. Methodology

In order to determine the water content at saturation, the dry gas at specific temperature and pressure values is mixed with a stream of water at the same pressure but at a slightly higher temperature. The slight temperature difference is set to allow for a slight cooling effect due to the change of phase of water without changing the temperature of the gas after saturation. Mixing is followed by separation. The flow rate of the water mixed with the dry gas is adjusted to ensure that the liquid flow rate from the separator is very small but not zero to guarantee saturation, and the temperature of the water stream is such that the gas temperature remains the same after saturation.

The optimum range for the feed gas temperature to the absorber is typically 80 °F–100 °F (299.8 K–310.9 K) as lower temperatures result in higher TEG viscosity causing lower overall absorber efficiency, while higher temperatures yield more TEG vaporization losses ([Manning and Thompson, 1991](#)). The flash drum pressure is taken as the mid-range pressure 62.5 psig (430.8 kPag) and the lean TEG temperature is set higher than the gas temperature leaving the contactor by 15 °F (8.3 K) ([Manning and Thompson, 1991](#)).

TEG is added to make up for losses, occurring mainly from the top of the absorber and the regenerator along with minor losses occurring from the flash tank. The reboiler temperature was set at 400 °F (477.6 K) in case stripping gas is used (cases b and c). The stripping gas used is dry gas.

5. Results and discussion

We define a reference cost B_{ref} as the objective function calculated for the no-stripping gas case, $N = 2$, $T = 100$ °F (310.9 K), $P = 600$ psia (4137 kPa), $m = 2$ gal TEG/lb (0.0167 m³/kg) water in feed gas, gas flow rate = 80 MMscfd (26.2 Sm³/s) and TEG concentration = 99.0 wt%. B_{ref} was found equal to 237,350 \$/yr using CEPCI for July 2018 (607.3). The choice of the reference is not unique and is simply used to make results of order 1 for the same gas flow rate. The cost results are presented in terms of the relative objective function B_r defined as $B_r = B/B_{\text{ref}}$.

Flow summaries for key streams following optimization to minimize the cost B are shown in [Tables 3](#) and [4](#) for both the selected base case, and a second case with gas stripping column.

The no-stripping (case a) and stripping cases (b and c) correspond to different regenerated TEG concentrations: (i) 99.0 wt% (no-stripping gas case), and (ii) regenerated TEG concentration: 99.5 wt%, $N_B = 1$ and optimum stripping gas flow rate = 1.744×10^{-4} mol/s. The lower gas price, \$105.6/MSm³ ([Table 2](#)), was used in both optimization cases. A small part of the water picked up by the glycol solution in the contactor leaves with the vapor exiting the flash tank, and the remaining part is removed in the regenerator to bring back the lean glycol concentration to 99.0 wt% ([Table 3](#)) and 99.5 wt% ([Table 4](#)). The composition of the stream leaving the regenerator (Vapor out) mainly consists of water vapor and a small fraction of hydrocarbons as seen from [Table 3](#) (about 0.27 wt%). The higher fraction of hydrocarbons (about 8.325 wt%) in [Table 4](#) is essentially due to the presence of stripping gas leaving the regenerator as part of the overhead stream. TEG glycol losses (TEG make up stream) are less by about 28% in the case of stripping gas ([Table 4](#)) compared to the non-stripping gas case ([Table 3](#)). The TEG losses mainly occur in the regenerator where temperature is high and pressure is relatively low (about 1 atmosphere). The use of stripping gas would lead to more glycol losses in the case of an ideal-liquid mixture. In the present case, the decrease in the glycol loss when stripping gas is used is attributed to the difference in glycol mole fraction in the reflux stream (0.04097 versus 0.04874 in the no-stripping gas case) along with non-ideality of the glycol–water liquid mixture resulting in dependency of the equilibrium ratios of glycol and water upon the reflux composition. The corresponding duties are shown in [Table 5](#).

The detailed costings after optimization are given in [Tables 6–8](#) for the bare module costs and FCI, the utility costs, and B , respectively. The listed equipment in [Table 6](#) includes, in particular, the bare module costs and the fixed capital investment due to the stripping gas column and packing for the stripping gas case, and the additional utility cost associated with the use of stripping gas is also included in [Table 7](#). The increase in the FCI for the stripping gas case in [Table 6](#) is small as compared to the no-stripping case. A similar conclusion is revealed in [Neagu and Cursaru \(2017\)](#). The utility cost associated with the use of stripping gas is significant ([Tables 7](#) and [8](#)).

5.1. Comparison with published dew point and dew point depression results

The water content values of the saturated gas were determined at different temperatures for 300 psia (20.7 bar) and 1000

Table 3

Material stream results for optimized TEG dehydration unit without stripping gas (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, regenerated TEG concentration: 99.0 wt%, low gas price). Feed dew point = 310.9 K, sales gas dew point = 273.7 K.

Stream name	1	2	3	4	5	6	7
Vapor fraction	1.000	0.000	0.000	1.000	1.000	0.004	1.000
Temperature, K	310.9	319.3	312.3	312.8	313.9	322.6	322.6
Pressure, kPa*	4137	4137	4137	4137	4137	532	532
Mass flow, kg/s	19.1904	0.6733	0.7071	19.1566	19.1566	0.7071	4.6576×10^{-4}
Component mass fraction							
Total hydrocarbons	0.9981030	–	0.0007841	0.9998331	0.9998331	0.0007841	0.9944687
Water	0.0018970	0.0100009	0.0565523	0.0001645	0.0001645	0.0565523	0.0055237
TEG	–	0.9899991	0.9426637	0.0000024	0.0000024	0.9426637	0.0000076
Stream name	8	9	10	11	12	13	14
Vapor fraction	0.000	0.136	1.000	0.000	0.000	0.000	0.000
Temperature, K	322.6	429.1	374.8	475.2	345.0	345.0	345.4
Pressure, kPa*	532	101	101	101	101	101	4137
Mass flow, kg/s	0.7066	0.7066	0.0333	0.6732	0.6732	3.7036×10^{-6}	0.6732
Component mass fraction							
Total hydrocarbons	0.0001291	0.0001291	0.0027346	–	–	–	–
Water	0.0565859	0.0565859	0.9971554	0.0099991	0.0099991	0.0099991	0.0099991
TEG	0.9432850	0.9432850	0.0001100	0.9900009	0.9900009	0.9900009	0.9900009

*Only significant pressure drops were included for optimization.

Table 4

Material stream results for optimized TEG dehydration unit with stripping gas (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, $N_B = 1$, regenerated TEG concentration: 99.5 wt%, low gas price). Stripping gas flow rate: 1.744×10^{-4} mol/s. Feed dew point = 310.9 K, sales gas dew point = 269.8 K.

Stream name	1	2	3	4	5	6	7	8	9
Vapor fraction	1.000	0.000	0.000	1.000	1.000	0.004	1.000	0.000	0.113
Temperature, K	310.9	319.3	312.3	312.8	314.1	324.7	324.7	324.7	427.1
Pressure, kPa*	4137	4137	4137	4137	4137	532	532	532	101
Mass flow, kg/s	19.1904	0.6434	0.6778	19.1560	19.1560	0.6778	4.6018×10^{-4}	0.6773	0.6773
Component mass fraction									
Total hydrocarbons	0.9981030	–	0.0008080	0.9998673	0.9998673	0.0008080	0.9940188	0.0001332	0.0001332
Water	0.0018970	0.0049959	0.0547299	0.0001302	0.0001302	0.0547299	0.0059717	0.0547630	0.0547630
TEG	–	0.9950041	0.9444621	0.0000025	0.0000025	0.9444621	0.0000095	0.9451038	0.9451038
Stream name	10	11	12	13	14	15	16	17	
Vapor fraction	1.000	0.000	0.000	0.000	0.000	0.000	1.000	1.000	
Temperature, K	372.0	473.1	349.8	349.8	350.2	477.6	473.1	310.9	
Pressure, kPa*	101	101	101	101	4137	101	101	101	
Mass flow, kg/s	0.0370	0.6434	0.6434	2.6538×10^{-6}	0.6434	0.6501	0.0098	0.0030	
Component mass fraction									
Total hydrocarbons	0.0832478	0.0000428	0.0000428	–	0.0000428	–	0.3057008	0.9998550	
Water	0.9166808	0.0049542	0.0049542	0.0050000	0.0049542	0.0094333	0.3016478	0.0001437	
TEG	0.0000714	0.9950031	0.9950031	0.9950000	0.9950031	0.9905667	0.3926514	0.0000014	

*Only significant pressure drops were included for optimization.

Table 5

Duties for optimized TEG dehydration without stripping gas (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, regenerated TEG concentration: 99.0 wt%) and with gas-stripping column (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, $N_B = 1$, regenerated TEG concentration: 99.5 wt%).

Duty in kW	No stripping gas	Stripping gas
TEG recirculation pump	3.143	3.012
Reboiler	151.6	159.5
Glycol/glycol heat exchanger	266.4	240.5
Gas/glycol heat exchanger	52.58	59.34
Reflux coil	20.36	23.79

psia (68.9 bar), and found in very good agreement with those obtained from the charts of McKetta and Wehe (Fig. 4–6 in Manning and Thompson (1991)) and Robinson et al. (Fig. 4–7a in Manning and Thompson (1991)), respectively as seen from Fig. 2. It is noted that reliability of Fig. 4–6 requires the pressure to be less than 500 psia (3447 kPa) while Fig. 4–7a can be used for pressure up to 2000 psia (13,786 kPa) (Manning and Thompson, 1991). The difference in deviations of the present results for the 6895 kPa

case compared to the 2068 kPa one in Fig. 2 are attributed to the fact that Fig. 4–6 (McKetta and Wehe chart) and Fig. 4–7a (Robinson et al. chart) are from different investigators.

For the purpose of the current study, the water content was calculated for different water dew points and three pressure levels: 300, 600 and 1400 psia (20.7, 41.4 and 96.5 bars). The relationship provides the dew point corresponding to a specific water content. The dew point of the gas decreases after dehydration, and the corresponding dew point drop is termed dew point depression. Dew point depression results are compared with the simulation results in Figs. 8–11, 8–15, 8–16, and 8–20 in Manning and Thompson (1991) for a feed gas at a pressure of 4137 kPa, and at two temperatures: 310.9 K (Fig. 3a) and 299.8 K (Fig. 3b). The deviations vary from about 2 to 10%.

5.2. Effect of parameters

5.2.1. No-stripping gas case

Fig. 4 and Fig. 5 (a–c) show the dew point depression (DPD) and relative processing cost (B_r) results for a feed flow rate of 80 MMscfd (26.2 Sm³/s), feed gas temperatures of 80 °F (299.8

Table 6

Cost estimates compared for TEG dehydration without stripping gas (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, regenerated TEG concentration: 99.0 wt%) and with gas-stripping column (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, $N_B = 1$, regenerated TEG concentration: 99.5 wt%). Low gas price considered in both cases.

Equipment (number of units if more than 1)	No stripping gas		Stripping gas	
	Size per unit	C_{BM} (July 2018, \$)	Size per unit	C_{BM} (July 2018, \$)
Contactor	15.07 m ³	329,260	15.07 m ³	329,150
Contactor tray (8)	1.889 m ²	12,050	1.888 m ²	12,050
Mist eliminator (2)	1.889 m ²	11,250	1.888 m ²	11,250
Flash drum	0.363 m ³	13,050	0.354 m ³	12,900
Glycol/glycol heat exchanger	44.29 m ²	103,100	33.83 m ²	93,100
Gas/glycol heat exchanger	58.91 m ²	119,100	61.04 m ²	121,000
Regenerator	2.170 m ³	28,500	2.07 m ³	27,808
Regenerator packing	0.964 m ³	1,700	0.925 m ³	1,600
Reboiler	16.03 m ²	139,700	20.23 m ²	143,300
Glycol circulation pump (2), 1 stand by	3.143 kW	78,000	3.012 kW	76,600
Stripping gas column	-	-	0.460 m ³	14,200
Stripping gas column packing	-	-	0.460 m ³	810
Total C_{BM} (July 2018, \$)	-	1,009,200	-	1,015,960
FCI (\$)	-	1,190,900	-	1,198,830

Table 7

Utility costs compared for TEG dehydration without stripping gas (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, regenerated TEG concentration: 99.0 wt%) and with gas-stripping column (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, $N_B = 1$, regenerated TEG concentration: 99.5 wt%). Low gas price considered in both cases.

Utility	No stripping gas		Stripping gas	
	Requirement	Cost (\$/yr)	Requirement	Cost (\$/yr)
Electricity (pump)	3.492 kW	910	3.347 kW	880
TEG make-up	3.704 × 10 ⁻⁶ kg/s	180	2.654 × 10 ⁻⁶ kg/s	130
Steam (reboiler)	0.0671 kg/s	17,600	0.0706 lb/hr	18,500
Stripping gas	-	-	0.0042 Sm ³ /s	13,700
Total	-	18,690	-	33,210

Table 8

Annual cost compared for TEG dehydration without stripping gas (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, regenerated TEG concentration: 99.0 wt%) and with gas-stripping column (feed gas: 310.9 K and 4137 kPa, gas flow rate 26.2 Sm³/s, $N = 2$, $m = 0.0167$ m³/kg, $N_B = 1$, regenerated TEG concentration: 99.5 wt%).

Cost	No stripping gas (Low gas price)	No stripping gas (High gas price)	Stripping gas (Low gas price)	Stripping gas (High gas price)
$B = 0.18 FCI + 1.23 C_{UT}$ (\$/yr)	237,350	305,450	256,600	370,800

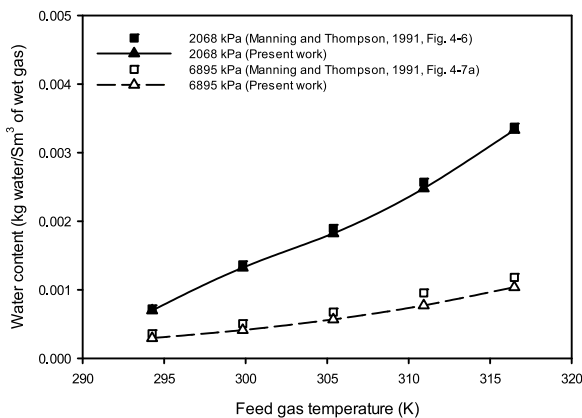


Fig. 2. Comparison of water content prediction in this work with the values obtained from McKetta and Wehe and Robinson et al. charts (Fig. 4–6 and 4–7a in Manning and Thompson (1991)).

K), $N = 1$ (Fig. 4) and 100 °F (310.9 K) (Fig. 5) with three different numbers of theoretical trays in the absorber ($N = 1$ –3) corresponding to cases a–c in Fig. 5.

- Effect of TEG rate

Increasing the circulation rate in the range $m = 2$ –6 gal/lb (0.0167–0.0501 m³/kg) water in gas, for TEG concentrations of 98.5 and 99.0 wt% at $P = 600$ psia (4137 kPa) enhances the dew point depression as seen from Fig. 4 ($T = 299.8$ K) and Fig. 5a–c ($T = 310.9$ K).

At high values of the circulation rate m , the effect of increasing the glycol circulation rate m on the dew point depression becomes less significant especially for $N = 2$ and 3 as seen from Fig. 5. However, the corresponding increase in the cost is significant.

- Effect of lean TEG concentration

For a given circulation rate, higher values of the dew point depression and the dehydration cost are obtained at the higher lean TEG concentration, 99.0 wt% TEG (Figs. 4 and 5). However, for a fair comparison, the required value of the minimum dew point depression needs to be fixed (if the feed gas temperature is the same) as a typical requirement for gas dehydration is to achieve a maximum specific dew point value for the dried gas, i.e. a minimum dew point depression needs to be achieved. The comparison is performed using the graphical procedure illustrated in Fig. 4 for a selected (minimum dew point depression) DPD of 26.6 K as an example. The processing cost is found lower at higher TEG concentration with a lower value of the relative cost B_r approximately equal to 0.83 for $N = 1$ and a lean TEG concentration of

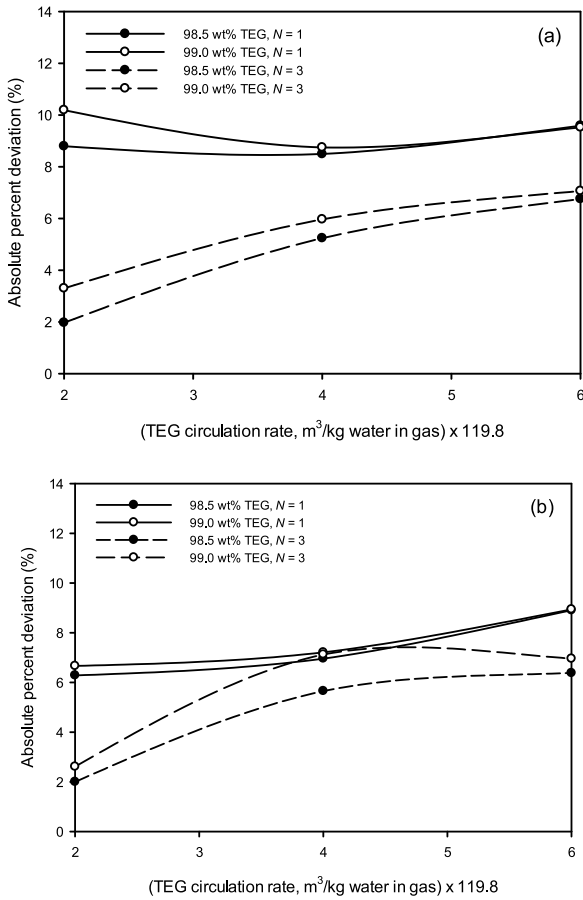


Fig. 3. Comparison of dew point depressions in this work with the simulation results in Fig. 8–11, 8–15, 8–16, and 8–20 in Manning and Thompson (1991) for (a) feed gas at 310.9 K and 4137 kPa (b) feed gas at 299.8 K and 4137 kPa.

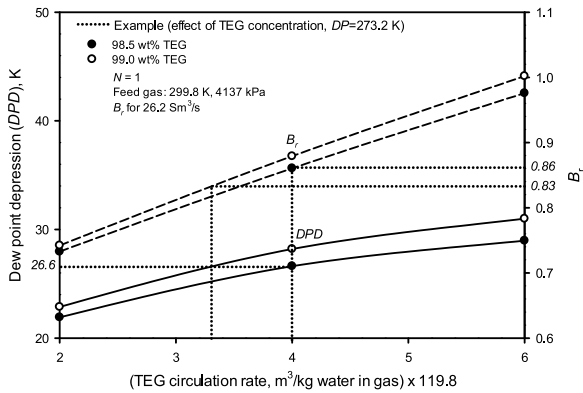


Fig. 4. Dew point depression (solid line) and cost ratio B_r (low gas price, dashed line) at 299.8 K and 4137 kPa for $N = 1$.

99.0 wt% versus B_r equal to about 0.86 for 98.5 wt% TEG with the same number of theoretical trays in the absorber.

• Effect of the number of theoretical trays in the absorber N

Increasing N enhances not only the dew point depression, but also the cost (Fig. 5a–c). The cost is seen to increase more than the dew point depression and this becomes more visible for larger N . Operating at a low circulation rate of 2 gal/lb (0.0167 m³/kg) water in gas yields the following depressions and lowest costs obtained at the lower TEG concentration of 98.5 wt%: $DPD = 22.6$ K and $B_r = 0.88$ for $N = 1$, $DPD = 34.4$ K and $B_r = 0.98$ for

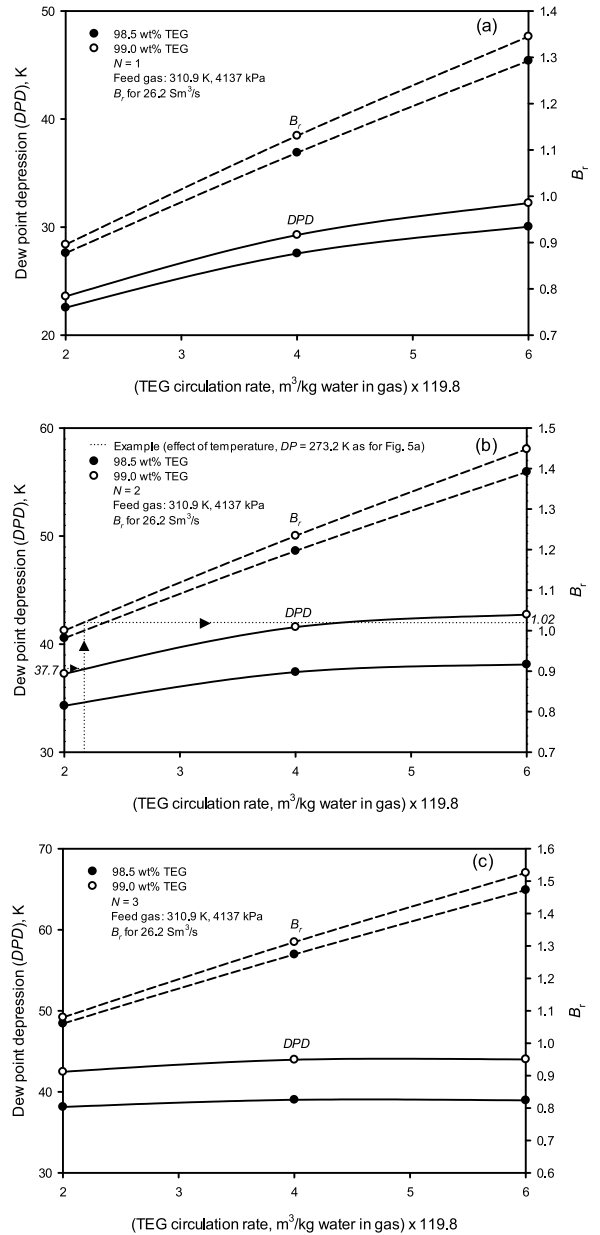


Fig. 5. Dew point depression (solid line) and cost ratio B_r (low gas price, dashed line) at 310.9 K and 4137 kPa for (a) $N = 1$, (b) $N = 2$, (c) $N = 3$, and a gas flow rate of 26.2 Sm³/s.

$N = 2$, and $DPD = 38.2$ K and $B_r = 1.06$ for $N = 3$. Therefore, the recommended number of trays is $N = 1$ for low dew point depressions (DPD), $N = 2$ for intermediate values of DPD and $N = 3$ for high values of DPD.

• Effect of feed gas rate and gas price

The simulations were performed for $N = 2$, $T = 80$ °F (299.8 K), $P = 600$ psia (4137 kPa), two TEG concentrations, 98.5 and 99.0 wt%, and a feed gas rate of 160 MMscfd (52.4 Sm³/s) (Fig. 6).

The economy of scale is not substantial as shown in the following comparison. For a circulation rate of 4 gal/lb (0.0334 m³/kg) water and a TEG concentration of 98.5 wt% as an example, the relative cost B_r is found to be 1.72 (Fig. 6) for a feed gas rate of 52.4 Sm³/s and low gas price (GP), whereas B_r is $2 \times 0.96 = 1.92$ (same conditions as in Fig. 4 with $N = 2$ instead of $N = 1$) for two similar trains processing 26.2 Sm³/s of feed gas each. The reduction of the cost is about 11.6%. Therefore, the effect of

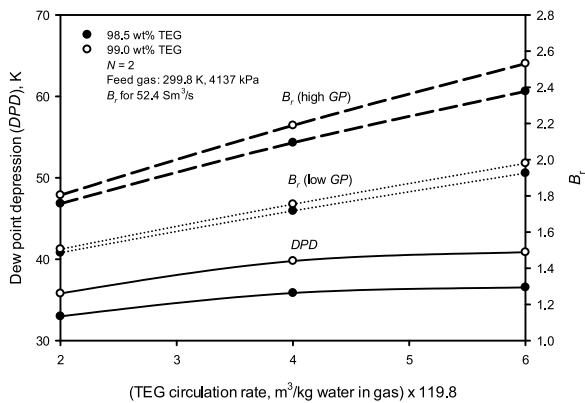


Fig. 6. Dew point depression (solid line) and cost ratio B_r (high gas price: dashed line, low gas price: dotted line) for $N = 2$ at 299.8 K and 4137 kPa and a gas flow rate of 52.4 Sm^3/s .

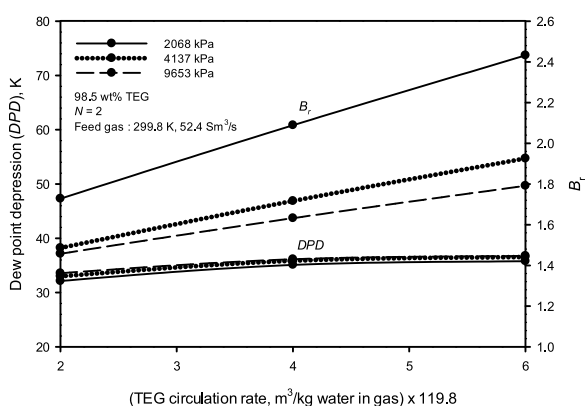


Fig. 7. Effect of absorber pressure on cost ratio B_r (low gas price) for $N = 2$ at 299.8 K. Gas flow rate: 52.4 Sm^3/s , regenerated TEG concentration = 98.5 wt%.

economy of scale is not substantial. At higher gas prices, the cost of processing is higher (Fig. 6). This finding is expected because of the fact that the corresponding increased cost of utilities, is basically not affected by the economy of scale. The differences between the processing cost for the two lean TEG concentrations (98.5 and 99.0 wt%) are higher at a high gas price (Fig. 6) for the same reason.

• Effect of feed gas pressure

The effect of feed gas pressure on the cost is shown in Fig. 7 for $N = 2$, $T = 80^\circ\text{F}$ (299.8 K), gas rate = 160 MMscfd (52.4 Sm^3/s), $P = 300$ (2068), 600 (4137) and 1400 (9653) psia (kPa), and TEG concentration = 98.5 wt%. The cost is significantly higher at 2068 kPa due to significantly higher water content in the feed gas. At higher pressures, 4137 and 9653 kPa, the effect of pressure is small as two adverse effects become comparable: higher capital cost due to higher pressures, and lower utilities due to less water content. At high circulation rate, m , the effect of utilities is predominant, and at lower values of m , the effect of enhanced capital cost is larger. At all pressures, increasing m yields higher costs. On the other hand, dew point depressions are slightly higher at high pressures as seen from Fig. 7. Therefore, operating at high pressure is recommended if the feed gas is at high pressure. Otherwise, considering the additional compression and cooling cost before absorption needs to be compared with the dehydration cost saving prior to a decision to operate at high pressure.

• Effect of feed gas temperature

Lower gas dew point is reached after dehydration at 80°F (Fig. 4) despite lower dew point depression at this temperature as compared to dehydration at 100°F (Fig. 5a). Operating at the lower temperature range value, 80°F (299.8 K), yields significantly less dehydration cost as indicated below using Figs. 4 and 5a. A dew point requirement of 273.2 (299.8–26.6) K is selected as an example. For a feed gas at 310.9 K, the dew point depression required is of 37.7 K, and the corresponding reduced cost B_r is 1.02 for $N = 2$ and 99.0 wt% TEG (Fig. 5b). For a feed gas at 299.8 K the DPD required is lower, 26.6 K. For $N = 1$ and 99.0 wt% TEG, the relative cost (0.83 from Fig. 4) is significantly lower. However, this may require additional cost due to cooling if the feed gas is originally at a higher temperature than 299.8 K. In such a case, the decision regarding the operating temperature requires a comparison of the additional cost to the saving in operating at a lower temperature.

• Optimum temperature of rich glycol feed to the regenerator

The optimum temperature of rich glycol feed to the regenerator is determined as a function of m for $N = 1-3$, $P = 600$ psia (4137 kPa), gas flow rate = 80 MMscfd (26.2 Sm^3/s), two TEG concentrations: 98.5 and 99.0 wt%, and two temperatures 80°F (299.8 K) and 100°F (310.9 K). The results show that higher optimum rich glycol feed temperatures to the regenerator are obtained at lower N and higher TEG wt%. At higher TEG concentration (99.0 wt%), the optimum temperature of the rich glycol feed to the regenerator is roughly 10 K higher than the corresponding value for the lower TEG concentration (89.5 wt%). The increase in temperature is expected as higher temperatures are expected in the regenerator due to the need to reach higher TEG concentration. Little impact on the optimum temperature is found at large values of m and more theoretical trays ($N = 2$ and 3). In addition, the impact of the feed gas temperature on the optimum temperature of the rich glycol feed to the regenerator is minor. The optimum temperature is slightly lower in the 310.9 K feed gas temperature case.

5.2.2. Stripping gas case

Reaching higher DPD values than the ones obtained with $N = 3$ requires the use of stripping gas. The dew point depression, the relative cost B_r and the stripping gas rate are plotted at optimum conditions as functions of m in Fig. 8(a–d) for $N = 2$, $N_B = 0-3$, $T = 100^\circ\text{F}$ (310.9 K), $P = 600$ psia (4137 kPa), gas rate = 80 MMscfd (26.2 Sm^3/s), and TEG concentrations of 99.5 and 99.9 wt%. Figs. 9c and 9d also show the effect of glycol concentration for 99.5 and 99.9 wt% TEG.

• Effect on water dew point depression and cost

Enhanced water dew point depressions, requiring higher cost, are obtained at higher TEG circulation rate m , decreasing stripping gas rate in the case $N_B = 0$ (no stripping column and stripping gas sent directly to the reboiler) as seen from Fig. 8a and constant stripping gas rate in the case N_B different from 0 (stripping gas sent above the reboiler) (Fig. 8b–d).

• Effect of lean TEG concentration

Fig. 8c and d show that, for the same dew point depression, lower cost is obtained with lower TEG concentration, 99.5 wt% TEG. At high TEG concentration (99.9 wt%), the required SGR is considerably higher, which significantly enhances the dehydration cost. As a result, the same DPD requirement can be achieved at a lower cost using a lower lean TEG concentration (Fig. 8c and d). For instance, a DPD of 45 K can be achieved at a relative cost of 1.21 for 99.5 wt% TEG and $N_B = 2$ versus $B_r = 1.38$ for 99.9 wt% TEG and the same number of theoretical trays in the stripping column (Fig. 8c).

• Effect of the theoretical number of trays in the stripper N_B

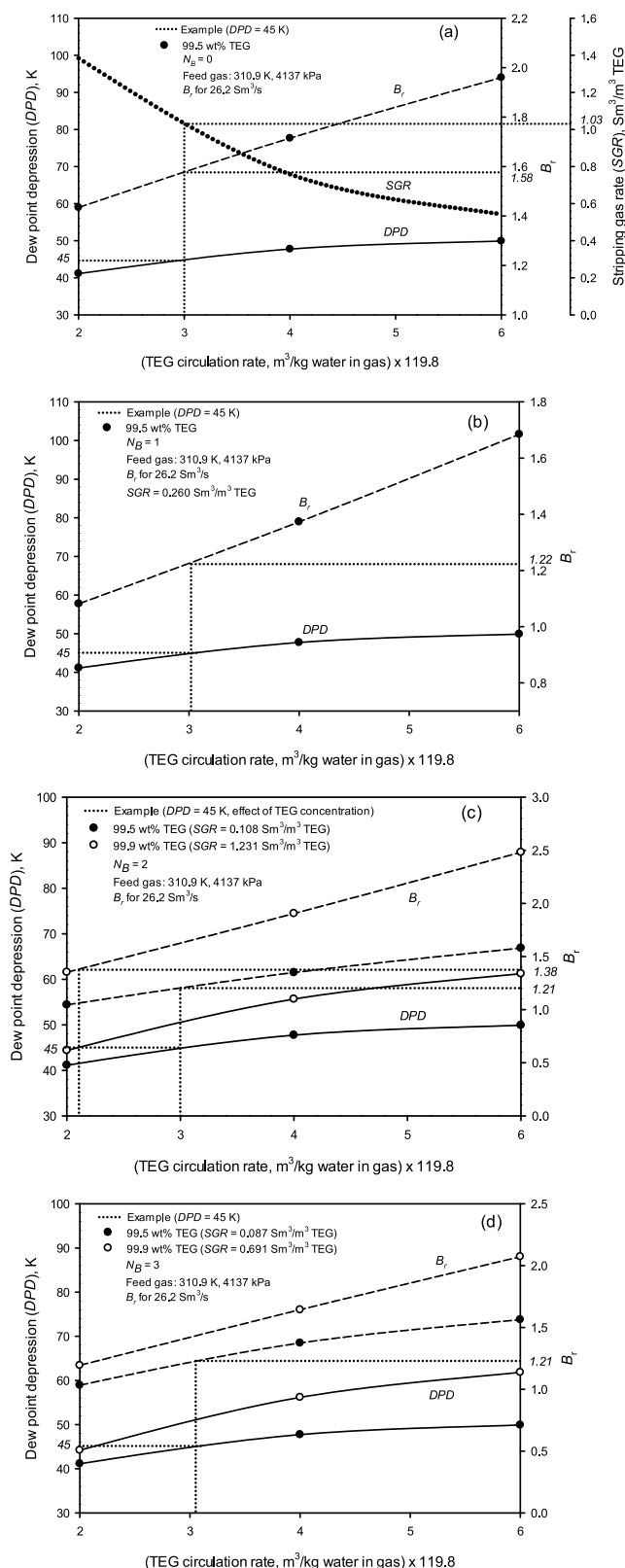


Fig. 8. Dew point depression (solid line), cost ratio B_r (low gas price, dashed line), and stripping gas rate (dotted line) at 310.9 K and 4137 kPa for $N = 2$ and (a) $N_B = 0$, (b) $N_B = 1$, (c), $N_B = 2$, and (d) $N_B = 3$.

For fixed dew point depression, Fig. 8a–d (99.5 wt% TEG) shows that increasing N_B reduces the dehydration cost. This cost

reduction is due to lower stripping gas rate requirement. For a DPD of 45 K as an example, the relative processing costs are found to be 1.58, 1.224, 1.21 and 1.21 for $N_B = 0, 1, 2$ and 3, respectively. The corresponding values of SGR are 1.03, 0.26, 0.108 and 0.087 Sm^3/m^3 TEG, respectively (Fig. 8a–d). Beyond $N_B = 2$, the reduction in cost is marginal.

6. Conclusions

TEG dehydration process was simulated using Aspen HYSYS simulator and optimizer tool along with an accurate thermodynamic model for TEG dehydration modeling, TST (Twu–Sim-Tassone equation of state). The dew point and dew point depression results are in good agreement with published values from charts and another simulation work. The present investigation provides recommendations for the minimization of the processing cost, resulting from the parametric optimization analysis to meet a specific dew point requirement. The design variables were varied for each set of parameters selected. The additional increase of the fixed capital cost due to gas stripping was found to be small. A similar conclusion is revealed in a recently published work.

The design parameters and variables were identified and optimization was performed to minimize the dehydration processing cost including capital cost and utilities needed for the reboiler, TEG pumping and stripping gas, while considering three cases: (i) no-stripping gas used, (ii) stripping gas injected into the reboiler ($N_B = 0$) and (iii) use of a stripping gas column (N_B larger than 0). Aspen HYSYS optimization was performed for different TEG circulation rates and numbers of theoretical trays (in the absorber and the stripping gas column), feed gas pressures of 300 (2068), 600 (4137) and 1400 (9653) psia (kPa) and temperatures of 80 °F (299.8 K) and 100 °F (310.9 K), in addition to two feed gas flow rates. Two levels of the gas price (low and high) were considered.

The present investigation provides the minimum processing cost, resulting from the parametric optimization analysis and including both utilities and capital cost. In addition, the analysis yields the following main conclusions based on the parametric optimization analysis: (i) Lower dehydration cost at higher pressures rendering it more cost effective only if compression followed by cooling is either unnecessary or possible at a lower cost in comparison with the reduction in the dehydration cost (ii) Lower dehydration cost at lower temperature absorber operation rendering it more cost effective only if cooling is either unnecessary, or possible at a lower cost in comparison with the reduction in the dehydration cost (iii) Reduced dehydration cost at higher TEG concentration in the no-gas stripping case with the following selection criteria for the number of theoretical trays in the absorber (contactor): $N = 1$ at low dew point depressions, $N = 2$ at intermediate DPD and $N = 3$ at high DPD (iv) Reduced dehydration cost at larger number of theoretical trays in the stripper N_B in the gas stripping case with significantly lower impact on the dehydration cost by increasing N_B from 2 to 3 (v) No substantial economy of scale for the overall cost including utilities and capital cost at higher gas flow rate.

Acknowledgment

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Appendix

The cost correlations are given in [Turton et al. \(2009\)](#). For the absorber vessel, flash separator, regenerator and gas stripping vessels, the two pumps (one standby) and heat exchangers (rich glycol/lean glycol, gas/glycol and reboiler) the bare module costs are determined from

$$C_{Bm} = C_p^0 (B_1 + B_2 F_M F_P) \quad (A.1)$$

with

$$\log_{10} C_p^0 = K_1 + K_2 \log_{10} A + K_3 (\log_{10} A)^2 \quad (A.2)$$

where A is volume in m^3 for pressure vessels, area of heat transfer in m^2 for heat exchangers, and shaft power in kW for the pumps.

The material of construction factor F_M is one for carbon steel except for pumps (1.4); the pressure factor is a function of gauge pressure and column diameter for process vessels, and a function of gauge pressure for other equipment. The pressure factor F_P is calculated for carbon steel pressure vessels (absorber vessel, flash separator and regenerator vessel) according to

$$F_P = \min \left[1, \frac{1}{0.0063} \left(0.00315 + \frac{(P_g + 1) D}{2 [850 - 0.6 (P_g + 1)]} \right) \right] \quad (A.3)$$

where P_g is gauge pressure in barg and D is the vessel diameter in m. For the heat exchangers (rich glycol/lean glycol, gas/glycol and reboiler), the pressure factors are given by

$$\log_{10} F_P = C_1 + C_2 \log_{10} P + C_3 (\log_{10} P)^2 \quad (A.4)$$

The constants B_1 , B_2 , K_1 – K_3 , and constants C_1 – C_3 needed to get F_P (non-process vessels case) are given in [Turton et al. \(2009\)](#).

The cost of the bubble cup trays were estimated as three times the cost of sieve trays ([Sinnott, 2005](#)). For both demisters and sieve trays, the following expressions are used for the bare module costs

$$C_{Bm} = C_p^0 N_a F_q \quad (A.5)$$

where N_a is the number of demisters or actual number of trays (equal to $4 \times N$), and F_q for less than 20 trays is given by

$$\log_{10} F_q = 0.4771 + 0.08516 \log_{10} N_a + 0.3473 (\log_{10} N_a)^2 \quad (A.6)$$

The cost of packing is estimated as

$$C_{Bm} = C_p^0 F_{BM} \quad (A.7)$$

where F_{BM} is 1.4 for ceramic packing.

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