Technical and economic evaluation of triethylene glycol regeneration process using flash gas as stripping gas in a domestic natural gas dehydration unit

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Abstract
Natural gas from an underground reservoir is typically saturated with water vapor. Removal of this water vapor is required in the gas processing to avoid serious problems. This study investigated the technical and economic aspects of a triethylene glycol (TEG) regeneration unit in a domestic natural gas processing plant. This study aims to improve the performance of this dehydration unit to meet sales gas specifications while minimizing the Total Annual Cost (TAC). The important variables in this work are TEG circulation mass flow and rerouting the gas from TEG Flash Drum to TEG Regenerator as stripping gas. The flash gas, instead of discharged directly to the Flare system, is routed to the TEG Regenerator and serves as stripping gas agent to improve the TEG purity. The results revealed that utilizing flash gas as stripping gas has allowed lower TEG circulation mass flow rate and reduced the reboiler duty from 1.464 to 0.934 GJ/h (a 36.2% reduction). The TAC was reduced from $296,058/year to $236,890/year (20.0% reduction). Through this work, a more economical design was obtained compared to the base case design.

KEYWORDS
dehydration unit, total annual cost, triethylene glycol

1 | INTRODUCTION

Natural gas from a reservoir is usually accompanied by a large amount of methane, heavier hydrocarbons, and water vapor, along with a considerable quantity of nonhydrocarbons, such as nitrogen, hydrogen sulfide, and carbon dioxide. Heavier hydrocarbons, acid gas components (H₂S and CO₂), and water must be removed from the natural gas to meet its sales gas qualities.¹

Water vapor within natural gas can cause problems such as hydrate formation under certain operating conditions and water condensation, which leads to internal corrosion as well as reducing combustion efficiency. Hydrate formation is regarded as the most important aspect of the natural gas treatment and transmission industry. Several studies regarding gas hydrate inhibitors have been conducted in recent years. There are several methods for dehydrating natural gas,
including solvent absorption, solid adsorption, gas condensation using refrigeration. Absorption using glycol has been widely used in the natural gas plant.

There are some benefits of glycol usage such as their low vapor pressure, high hygroscopic property, low solubility in natural gas, and high boiling point. There are four types of glycols used as a dehydrating agent in natural gas treatment, that is, tetraethylene glycol, triethylene glycol (TEG), diethylene glycol (DEG), and ethylene glycol (EG). The absorption of water using TEG is the standard method for natural gas dehydration. The use of TEG has advantages compared to the other glycols that it has lower vapor pressure and lower viscosity over its operating conditions.

A comparison of three methods for natural gas dehydration, that is, absorption by glycol, solid desiccant/adsorption, and condensation has been detailed by Netusil and Ditl. They revealed that the absorption method consumes less energy compared to the other two methods. However, for the high-pressure operation, the condensation method is determined to be the most appropriate method. The adsorption method is preferred in cases where very low water dew point in the dry gas stream is required.

The glycol dehydration process has been a subject of recent research studies. Kong et al described and listed several available methods and recent developments in the natural gas dehydration. It includes the conventional dehydration using TEG, the use of inert gas or portion of sales gas as stripping agent, the use of water exhauer (Coldfinger), the use of volatile hydrocarbon as stripping agent (such as Drizo process), as well as other emerging technologies such as membrane technologies and supersonic separator technology. In addition to the technical comparison in terms of BTEX (ie, benzene, toluene, ethylbenzene, and xylene) emissions and TEG purity, they also provided some sort of economic assessment to each dehydration strategy.

Studies on the development of the equation of state to be used in the modeling of the glycol-water system were carried out by Twu et al. They developed an advanced equation of state method for modeling the phase behavior of the TEG-water-natural gas system. The TST (Twu-Sim-Tassone) equation of state has shown good accuracy in representing the activity coefficient of the TEG-water solutions as well as the water dew point calculation of the natural gas over a wide range of pressure, temperature, and concentration typically used in TEG dehydration unit. Watanasiri et al have developed and validated the Cubic-Plus-Association (CPA) equation of state that can be used to model the gas dehydration with TEG, EG, and DEG, especially in the ASPEN HYSYS software. Bahadori et al investigated the solubility of light alkanes and acid gases in the TEG solvent. The influence of several parameters on hydrocarbon and acid gas solubility, including temperature, pressure, and solvent content, were also examined.

Rahimpour et al simulated the dehydration unit of a domestic gas processing plant using the Coldfinger system. They proposed a new mathematical model to simulate the Coldfinger system. They also studied the influence of flow rate and temperature of the stripping gas and the pressure of Coldfinger unit to the regeneration rate. Another model of Coldfinger unit was proposed by Romero et al. They used two equilibrium stages that operated at different temperatures in the presence of internal vapor recirculation to represent the Coldfinger unit.

The use of volatile hydrocarbon as a stripping agent has been studied as well. Saidi et al reviewed the application of Drizo process in a domestic gas processing plant in Iran. They studied the impact of using single Drizo compared to complex Drizo units in terms of BTEX emissions reduction. They also revealed that though n-heptane has the best performance among the other solvents, the use of recycled BTEX compounds as solvent found to reduce further the BTEX emissions.

Jokar et al simulated the natural gas dehydration plant of the Farashband gas processing plant. Their results showed that revamping of trays with structured packing could reduce outlet natural gas dew point and improve the positive effect of other parameters on the performance of the dehydration unit. Rouzbahani et al simulated a natural gas dehydration unit located in Iran and included its regeneration package of DEG in the steady-state simulation.

Arya et al used a thermodynamic property package (ThermoSystem), based on CPA, to design a natural gas dehydration process by interfacing it to any simulator that supported the CAPE-OPEN standard; their results showed that the system could be used successfully.

In addition, Anyadiegwu et al simulated natural gas dehydration using TEG and plotted the effect of the TEG flow rate on the water content.

In recent years, some studies have focused on the simulation and optimization of natural gas dehydration units. There are several studies focusing on the techno-economic reviews of the TEG dehydration units. Gad et al investigated the economic comparison between dry natural gas and nitrogen gas for stripping water from glycol in the dehydration process. The study indicated that the use of natural gas is more economical compared to nitrogen as stripping gas. The amount of natural gas used is lower than nitrogen, which leads to lower utility costs. The study, however, did not provide elaboration on the capital cost and utility cost calculations.
Chebbi et al investigated the impact of several operating parameters such as pressure, temperature, TEG circulation rate, and stripping gas flow rate to capital and operating cost.\textsuperscript{17} They presented their main conclusions based on parametric optimization analysis that lower dehydration cost can be achieved at higher pressure, at lower temperature, at higher TEG concentration without stripping gas, and at higher number of theoretical stages in the stripper.

Neagu and Cursaru investigated the technical and economic evaluations of the TEG regeneration processes by evaluating various stripping gas flowrates to improve TEG purity.\textsuperscript{18} Their work result indicated that the stripping gas configuration is a more effective way of improving the TEG purity and regeneration performance. The study used the annualized capital cost and annual cost production to compare the stripping gas drying vs the conventional gas drying.

Kamin et al developed the simulation and optimization of TEG in a natural gas dehydration process.\textsuperscript{19} In this study, they investigated the impact of lean glycol circulation rate, reboiler temperature, and the number of trays in the contactor column. The response parameters included the number of glycol losses, reboiler duty, the water moisture content in the dry gas, and the hydrate formation temperature. They used the response surface methodology to optimize the process.

Affandy et al investigated the optimization of the TEG unit by replacing the TEG Contactor internals from tray to structured packing.\textsuperscript{20} The results show that by replacing the tray column with a packed bed column can reduce the size of the column and also reduced the total annual cost (TAC) of the natural gas dehydration unit. The study used TAC that is the sum of total capital cost divided by the payback period and the total annual operating cost. The economic evaluation approach in this work was based on the simple economic analysis as described by Luyben. The approach was to consider the installed cost of main equipment and used a small 3-year payback period. It acknowledged that the total capital cost of a plant includes other expenditures beyond the main equipment installed cost (ie, site preparation and off-site units). The total investment in a typical plant is about 3-4 times the cost of major equipment. The other approach would be to increase the capital cost of the equipment by a factor of 3, then use a 9-year payback period.\textsuperscript{21}

Kong et al developed a framework to carry out a techno-economic comparison between the conventional and stripping gas dehydration process.\textsuperscript{22} The study used the use of part of sales gas as the stripping gas agent and compared it to the use of nitrogen as a stripping agent in terms of the dehydration performance to achieve the sales gas water dew point target. The study used an annual profit margin of the two stripping gas process in their economic comparisons. It was indicated that the use of part of sales gas provided better net profit compared to the use of nitrogen gas.

Kong et al also developed a parametric study of different configurations of natural gas dehydration with TEG to improve the TEG loss rate and the BTEX emissions.\textsuperscript{23} They focused on recycling the flash vapor and/or the regenerator overhead vapor into the absorber. The economic comparisons were performed using the total cost of production (TCOP) which includes the sum of annualized capital cost and the cost of production. The study indicated that the application of the recycling flash vapor and regenerator overhead vapor to the absorber did not contribute to a significant reduction in TEG loss and total BTEX emissions while requiring additional capital and operating costs.

The stripping gas mechanism in improving TEG purity is explained by Kong et al. The stripping gas works as a physical separation process according to Raoult’s law. The vapor in the TEG regenerator consists of water vapor, TEG, methane, and other vapor (such as nitrogen if we use inert gas as a stripping agent). The addition of a vapor stream (eg, dry product gas or an inert gas such as N\textsubscript{2}) is reducing the partial pressure of the water vapor in the TEG regenerator column. This reduces the mole fraction of water in TEG solution, which leads to higher purity of regenerated TEG stream.\textsuperscript{5}

This study investigated the simulation and evaluation of a dehydration unit in a domestic natural gas processing plant by using TEG, including the TEG regeneration cycle based on the work of Affandy et al.\textsuperscript{20} Property package of CPA was used in this study. A simulation resembling the plant was used as the base case. In addition, a TAC calculation was performed for the base case. Furthermore, process evaluation to utilize the flash gas as a stripping gas medium was conducted. The TAC of the evaluated process was also studied.

## 2 \ PROCESS DESCRIPTION

### 2.1 \ TEG contactor

A TEG contactor is intended to reduce water vapor in gas by contacting it with lean TEG. Wet gas is fed from the bottom of the contactor column while the lean TEG enters the contactor from the top of the column. The wet gas was brought
into contact with TEG in the TEG contactor to remove its water vapor. The dry gas comes out of the contactor from the top of the column and the rich TEG leaves from the bottom of the column. The specification of the water vapor content in the dry gas is 7 lb/MMSCF, units of measurement used in the natural gas industry. The TEG contactor is operated at 5101 kPa and 44°C (317.00 K). A chimney tray was used in the TEG contactor to improve gas distribution in the column. Figure 1 shows the process flow diagram of a domestic natural gas dehydration unit using TEG.

2.2 TEG regeneration package

This unit is used to regenerate rich TEG after being used in the contactor column. The rich TEG is regenerated, that is, reducing its water content before being reused to dehydrate the gas in the TEG contactor. Several steps in the TEG Regeneration Package are described below.

2.2.1 TEG flash drum

The TEG Flash Drum is a two-phase separator that flashes the light hydrocarbon out from the TEG solution to prevent excess hydrocarbons from entering the regeneration process in the TEG Regenerator. The TEG Flash Drum is designed to separate flash gas and liquid (rich TEG containing a small amount of water and dissolved condensate). The flashed gas is typically routed to a vent or flare system.

2.2.2 Glycol-glycol heat exchanger

After leaving flash separation, the rich TEG enters a series of the glycol-glycol heat exchanger, which is upstream of the regeneration column. These exchangers improve heat conservation by contacting hot and cold glycol streams within the regeneration process. The glycol/glycol HE-2 provides preheating steps for rich glycol stream before entering the Flash Drum and glycol/glycol HE-3 before entering the Regenerator column.

2.2.3 Regenerator column

After preheating in the glycol-glycol heat exchanger, the rich TEG is fed to the regenerator column, which is a stripper configuration. The regenerator column consists of two parts: a column and a reboiler. The regenerator column is typically operated near atmospheric pressure (103 kPa). The top products of the regenerator column are water vapor, a small amount of glycol, and volatile organic compounds, whereas the bottom product is lean TEG which contains a small
amount of water. The lean TEG from the regenerator column is routed to the Glycol Accumulator through series of the glycol-glycol heat exchanger to conserve the heat energy and then pumped back to the TEG contactor for subsequent dehydration process.

3 | THERMODYNAMIC MODEL FOR SIMULATION

A thermodynamic model is an important tool for acquiring simulation results that are in good agreement with the actual condition. Aspen HYSYS offers recommendations about thermodynamic models for gas processing, especially for gas dehydration with glycols using the CPA property package.\(^7\)

4 | RESULTS AND DISCUSSION

This section presents and discusses the simulation and evaluation results of the domestic natural gas dehydration unit. The results include base case simulation, the TAC calculation of the base case, and the evaluated result by increasing area of glycol/glycol heat exchanger 3 and rerouting the flash gas as stripping gas in the Regenerator.

4.1 | Base case simulation

The simulation resembling the plant was used as the base case. Figure 2 shows the process flow of the base case simulation. In the base case simulation, the TEG Contactor column is used to remove water by contacting wet gas (stream 1) with a lean TEG (stream 2). The water content in the dry gas (stream 3) should be less than 7lb/MMSCF. After absorbing the water in the wet gas, the lean TEG becomes rich TEG (stream 4) and therefore should be regenerated in the regeneration package that includes a TEG Flash Drum, glycol-glycol heat exchangers (HE-1, HE-2, and HE-3), and a regenerator column.

The base case simulation contains certain specifications. The TEG Contactor column has eight stages with a column diameter of 1.6 m. The Regenerator column has three stages (including the reboiler) with a column diameter of 0.4 m. The diameter of both columns was determined by the internal column sizing calculation in Aspen HYSYS. The Regenerator column was designed to regenerate the rich TEG. The areas \((A)\) of the heat exchangers HE-1, HE-2, and HE-3 are 0.6, 22.6, and 22.6 \(m^2\), respectively. The relevant specifications of TEG Flash Drum are the operating pressure and heat \((Q)\).
The operating pressure of TEG Flash Drum is 657 kPa, and it was operated adiabatically. Table A.1 in Appendix presents the stream specifications of the base case simulation. The purity of the lean TEG (stream 15) is 98.7%-wt. This purity can be achieved by setting the reboiler temperature of the Regenerator column at 204°C. Figure 3 depicts that operating the reboiler at a temperature lower than 204°C leads to the water moisture in the dry gas (stream 3) exceeding the 7 lb/MMSCF limit. It should be noted that TEG has a degradation temperature of 204°C (400°F), in which operating at that condition should be avoided.

### 4.2 TAC calculation of base case

TAC is the sum of the total capital cost (TCC) divided by a payback period plus the annual total operating cost. For this system, the operating cost consists of the cost of heating medium, cooling water, and TEG make-up. The cost of heating medium and cooling water is based on the duty of the reboiler, heater, and cooler in the base case simulation. The costs of the columns (TEG Contactor and Regenerator column), including the reboiler, Flash Drum, heat exchangers (HE-1, HE-2, and HE-3), and the cooler, are defined as the capital cost. The TAC formulas were taken from Luyben. Table 1 summarizes the formulas for all TAC calculations.

Table 2 shows the TAC of the base case simulation. The total capital cost of the base case simulation is $397436 whereas the total operating cost of the base case simulation is $163 579/year. Thus, the TAC of the base case simulation is $296 058/year. The utility cost is expensive ($143 377/year), thus it has room for further improvement. Therefore, an evaluated work is investigated, with details illustrated in the next section.

### 4.3 Process evaluation

In the conventional dehydration explained in the previous subsection, the specification of water moisture in the dry gas stream (max. 7 lb/MMSCF) can be achieved by implementing higher reboiler temperature of 204°C. To achieve lower water moisture in the dry gas stream, one of the methods explained by Kong et al is to employ the stripping gas into the regenerator system. In this work, part of the dry gas was initially added as the stripping gas agent. The stripping gas mass flow rates used in the evaluation were from 0 to 50 kg/h (with 10 kg/h increment). The reboiler temperature was lowered to 200°C. The simulation revealed that the addition of part of dry gas as stripping gas could lower the water moisture in the dry gas stream. Figure 4 depicts the water moisture content as a function of stripping gas mass flow rate with the reboiler temperature operated at 200°C. A minimum stripping gas flow rate of 15 kg/h in this application is required to achieve the 7 lb/MMSCF water moisture gas specification.

### 4.4 Flash gas rerouting to regenerator

In this work, since the flash gas (stream 9, Appendix Table A.1) has a majority of noncondensable and inert gas with small amount of water vapor (less than 7% mole), it is proposed as a stripping gas to the Regenerator to increase the TEG
### Table 1: Equipment sizing and economic basis

| Equipment                        | Formula                                                                 |
|----------------------------------|-------------------------------------------------------------------------|
| Diameter of the column \((d)\)   | Tray sizing of Aspen Hysys                                              |
| Length of the column \((l)\)     | Number of tray \((N_T)\) with 0.6096 m spacing + 20% extra length      |
| Capital cost                     | \(17640 \times (d)^{1.066} \times (l)^{0.802}\)                       |
| Flash drums                      |                                                                         |
| \(l/d\) Aspect ratio             | 2                                                                       |
| Volume                           | \(10 \text{ min} \times \text{ liquid flow rate}\)                    |
| Capital cost                     | \(17640 \times (d)^{1.066} \times (l)^{0.802}\)                       |
| Reboiler (area \([A]\) in m\(^2\)) |                                                                         |
| The coefficient of heat transfer | 568 W/(m\(^2\) K)                                                      |
| Temperature difference           | The temperature of steam-reboiler temperature \((\Delta T > 20 K)\)     |
| Capital cost                     | \(7296 \times (A)^{0.65}\)                                            |
| Heat exchangers, (area \([A]\) in m\(^2\)) |                                                                     |
| The coefficient of heat transfer | 852 W/(m\(^2\) K)                                                      |
| Temperature difference           | LMTD of (inlet temperature and outlet temperature differences)         |
| Capital cost                     | \(7296 \times (A)^{0.65}\)                                            |
| Cooler (area \([A]\) in m\(^2\)) |                                                                         |
| The coefficient of heat transfer | 852 W/(m\(^2\) K)                                                      |
| Temperature difference           | LMTD of (temperature of inlet or outlet—315 K)                         |
| capital cost                     | \(7296 \times (A)^{0.65}\)                                            |
| Heater (area \([A]\) in m\(^2\)) |                                                                         |
| The coefficient of heat transfer | 568 W/(m\(^2\) K)                                                      |
| Temperature difference           | LMTD of (temperature of steam – (temperature of inlet or outlet)       |
| Capital cost                     | \(7296 \times (A)^{0.65}\)                                            |
| Operating cost                   |                                                                         |
| Heating medium                   | $9.8/GJ                                                                 |
| Electricity                      | $16.8/GJ                                                                |
| Cooling water                    | $2.5/GJ                                                                 |
| Make-up cost                     | $2.71/kg                                                                |
| TAC                              | (Total capital cost/payback period) + annual total operating cost       |
| Payback period                   | 3 years                                                                 |

Purity, before sending the flash gas to the vent or flare system. Figure 5 shows the process flow diagram of the proposed configuration. Table A.2 in Appendix presents the stream specifications of the proposed configuration.

The amount of liberated flash gas is consistent with the work of Bahadori et al.8 Figure 6 depicts the amount of flash gas under various TEG circulation flow rate. The flash gas flow rate is about 41 kg/h at the TEG circulation flow rate of 5 m\(^3\)/h, which should be sufficient for supplying the stripping gas requirement to achieve water moisture specification. The presence of a small amount of water content in the flash gas itself will reduce the effectiveness of the stripping gas mechanism.

The TEG purity improvement using the flash gas is depicted in Figure 7, in which the TEG purity studied by Neagu and Cursaru18 is also presented as a comparison to this work. The chart indicates that the use of Flash Gas as a stripping agent to the TEG regeneration system can provide higher TEG concentration. The improvement is limited due to two main reasons, that is, (1) the amount of available Flash Gas is a function of TEG circulation rate, the TEG Contactor pressure and temperature as explained by Bahadori et al., and (2) the Flash Gas itself containing certain amount of water, therefore
TABLE 2  Itemized TAC terms of the base case simulation

| Configurations                              | Total  |
|---------------------------------------------|--------|
| Installed capital cost for TEG Contactor ($) | 120 046|
| Installed capital cost for TEG Regenerator ($) | 6642   |
| Installed capital cost for Reboiler ($)     | 55 368 |
| Installed capital cost for Flash Drum ($)   | 44 403 |
| Installed capital cost for HE-1 ($)         | 5235   |
| Installed capital cost for HE-2 ($)         | 55 368 |
| Installed capital cost for HE-3 ($)         | 55 368 |
| Installed capital cost for Cooler ($)       | 55 368 |
| Utility cost ($)                            | 143 377|
| Make-up cost ($/year)                       | 20 202 |
| TCC ($)                                     | 397 436|
| TOC ($/year)                                | 163 579|
| TAC ($/year)                                | 296 058|

FIGURE 4  Water moisture in the evaluated case simulation with reboiler temperature of 200°C and various stripping gas mass flow rate

FIGURE 5  Proposed configuration and simulation flow diagram
The mass flow rate of Flash Gas in the evaluated case simulation with reboiler temperature of 200°C and various TEG circulation flow rate.

The variation TEG concentration in the evaluated case simulation with reboiler temperature of 200°C and various stripping gas flow rate (data compared with the results from Neagu and Cursaru\textsuperscript{18}).

Water moisture in the proposed configuration simulation with reboiler temperature of 200°C, various TEG circulation flow rate and flash gas used as stripping gas reducing its effectiveness as stripping agent. This proposed configuration, that is using the Flash Gas as stripping agent, can be implemented where there are sufficient amount of Flash Gas and the water content specification in the gas can be achieved by lean TEG up to 99.0%-weight.

By rerouting the flash gas from Flash Drum (stream 9) to TEG Regenerator, the TEG purity at reboiler temperature of 200°C can be increased from 98.6%-wt and the 7 lb/MMSCF water moisture limit can be achieved at TEG flow rate higher than 2.3 m$^3$/h. The water moisture in the dry gas for the proposed configuration is depicted in Figure 8. In this work, the TEG circulation mass flow rate was reduced from 5 m$^3$/h (of 98.6%-wt of TEG) to 2.5 m$^3$/h (of 98.72%-wt of TEG). Having less TEG circulation flow rate leads to lower energy consumption, hence less Utility Cost.

In comparison with the base case, the reboiler duty is reduced from 1.464 to 0.934 GJ/h (a 36.2% reduction). The concentration of water moisture in dry gas (stream 3) is slightly reduced from 6.95 to 6.89 lb/MMSCF, which both cases are still below 7 lb/MMSCF gas moisture requirement. The TAC, as calculated using the formulas in Table 1, is reduced from $296,058/year to $236,890/year (20.0% reduction). The TAC comparison between the base case and the proposed
configuration is listed in Table 3. The proposed configuration revealed that a more economical design of a natural gas dehydration unit using TEG than the base case design was obtained.

## 5 | CONCLUSIONS

This article simulated a steady-state simulation of the TEG dehydration unit in a domestic natural gas processing plant. Process evaluation was performed to minimize the TAC of the base case by adding stripping gas to the TEG Regenerator by rerouting vapor from TEG Flash Drum, instead of venting/flaring it.

The final results show that by adding the stripping gas from the vapor of TEG Flash Drum, the TEG circulation mass flow rate can be reduced while satisfying the water moisture in the dry gas specification. Therefore, the reboiler duty is reduced from 1.464 to 0.934 GJ/h (36.2% reduction), the concentration of water moisture in dry gas is slightly reduced from 6.95 to 6.89 lb/MMSCF and the TAC is reduced from $296 058/year to $236 890/year (20.0% reduction). Therefore, a more economical design of a natural gas dehydration unit using TEG than the base case design was obtained.

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### CONFLICT OF INTEREST

The authors declare no conflict of interest.

### AUTHOR CONTRIBUTIONS

Sony A. Affandy contributed to the investigation. Adhi Kurniawan contributed to the investigation. Renanto Handogo contributed to the conceptualization, Formal analysis, investigation, methodology, and supervision. Juwari P. Sutikno contributed to the resources. I-Lung Chien contributed to the conceptualization and resources.

### NOMENCLATURE

| Symbol | Definition |
|--------|------------|
| $A$    | area (m²)  |
| $d$    | diameter of the column (m) |
| $l$    | length of the column (m) |
| LMTD   | log-mean temperature difference (K) |
| $N_T$  | number of tray (stages) |

### TABLE 3 TAC comparison between the base case and the proposed configuration

| Configurations                                  | Total          |       |       |
|-------------------------------------------------|----------------|-------|-------|
|                                                 | Base case      | Proposed configuration |
| Installed capital cost for TEG Contactor ($)    | 120 046        | 120 046 |
| Installed capital cost for TEG Regenerator ($)  | 6642           | 6642  |
| Installed capital cost for Reboiler ($)         | 55 368         | 55 368 |
| Installed capital cost for Flash Drum ($)       | 44 403         | 44 403 |
| Installed capital cost for HE-1 ($)             | 5235           | 5235  |
| Installed capital cost for HE-2 ($)             | 55 368         | 55 368 |
| Installed capital cost for HE-3 ($)             | 55 368         | 55 368 |
| Installed capital cost for Cooler ($)           | 55 368         | 55 368 |
| Utility cost ($)                                | 143 377        | 84 924 |
| Make-up cost ($/year)                          | 20 202         | 19 488 |
| TCC ($)                                         | 397 436        | 397 436 |
| TOC ($/year)                                    | 163 579        | 104 412 |
| TAC ($/year)                                    | 296 058        | 236 890 |
$P$ pressure (kPa)  
$Q$ heat (GJ/h)  
$T$ temperature (K)  
TAC total annual cost ($)  
TCC total capital cost ($)  
TOC total operating cost ($)  

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**APPENDIX A. HEAT AND MATERIAL BALANCES**

**TABLE A.1** Stream specifications of the base case simulation

| Stream | Molar flow kmol/h | 1     | 2     | 3     | 4     | 5     | 6     | 7     |
|--------|------------------|-------|-------|-------|-------|-------|-------|-------|
| Mass flow kg/h |       | 150.400 | 5629 | 150.100 | 5944 | 5944 | 5944 | 5944 |
| Temperature K |       | 317.1 | 321.4 | 319.2 | 318.8 | 320.1 | 322.6 | 377.9 |
| Pressure kPa |       | 5101 | 5237 | 5051 | 5101 | 5101 | 5101 | 5101 |
| Vapor fraction |       | 0     | 1     | 0     | 0.0211 | 0.0219 | 0.0257 |
| CO₂ mol fraction |       | 0.0267 | 0.0267 | 0.0050 | 0.0050 | 0.0050 | 0.0050 |
| N₂ mol fraction |       | 0.0183 | 0.0183 | 0.0001 | 0.0001 | 0.0001 | 0.0001 |
| Methane mol fraction |       | 0.8319 | 0.8337 | 0.0157 | 0.0157 | 0.0157 | 0.0157 |
| Ethane mol fraction |       | 0.053 | 0.0531 | 0.0028 | 0.0028 | 0.0028 | 0.0028 |
| Propane mol fraction |       | 0.0366 | 0.0367 | 0.0032 | 0.0032 | 0.0032 | 0.0032 |
| i-Butane mol fraction |       | 0.01 | 0.0100 | 0.0012 | 0.0012 | 0.0012 | 0.0012 |
| n-Butane mol fraction |       | 0.0116 | 0.0116 | 0.0018 | 0.0018 | 0.0018 | 0.0018 |
| i-Pentane mol fraction |       | 0.0042 | 0.0042 | 0.0007 | 0.0007 | 0.0007 | 0.0007 |
| n-Pentane mol fraction |       | 0.0028 | 0.0028 | 0.0006 | 0.0006 | 0.0006 | 0.0006 |
| n-Hexane mol fraction |       | 0.0017 | 0.0017 | 0.0006 | 0.0006 | 0.0006 | 0.0006 |
| n-Heptane mol fraction |       | 0.0007 | 0.0007 | 0.0004 | 0.0004 | 0.0004 | 0.0004 |
| n-Octane mol fraction |       | 0.0002 | 0.0002 | 0.0002 | 0.0002 | 0.0002 | 0.0002 |
| n-Nonane mol fraction |       | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0001 |
| n-Decane mol fraction |       | 0 | 0 | 0 | 0 | 0 | 0 |
| C₁₁+ mol fraction |       | 0 | 0 | 0 | 0 | 0 | 0 |
| H₂O mol fraction |       | 0.0022 | 0.0982 | 0.0001 | 0.3181 | 0.3181 | 0.3181 | 0.3181 |
| TEG mol fraction |       | 0 | 0.9018 | 0 | 0.6495 | 0.6495 | 0.6495 | 0.6495 |
| Total |       | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |

(Continues)
### Table A1 (Continued)

| Stream | Molar flow kmol/h | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 |
|--------|------------------|---|---|----|----|----|----|----|----|----|----|----|----|----|
| Mass flow kg/h | 41.28 | 41.28 | 5903 | 5903 | 5903 | 5903 | 288.1 |
| Temperature K | 377.9 | 377.9 | 377.9 | 377.9 | 430.1 | 419.7 | 418.9 |
| Pressure kPa | 657 | 120 | 657 | 343 | 313 | 114 | 103 |
| Vapor fraction | 1.00 | 1.00 | 0 | 0.0030 | 0.0129 | 0.0847 | 1.00 |
| CO₂ mol fraction | 0.1203 | 0.1203 | 0.0020 | 0.0020 | 0.0020 | 0.0020 | 0.0074 |
| N₂ mol fraction | 0.0037 | 0.0037 | 0 | 0 | 0 | 0 | 0 |
| Methane mol fraction | 0.5492 | 0.5492 | 0.0017 | 0.0017 | 0.0017 | 0.0017 | 0.0063 |
| Ethane mol fraction | 0.0840 | 0.0840 | 0.0006 | 0.0006 | 0.0006 | 0.0006 | 0.0024 |
| Propane mol fraction | 0.0831 | 0.0831 | 0.0010 | 0.0010 | 0.0010 | 0.0010 | 0.0039 |
| i-Butane mol fraction | 0.0270 | 0.0270 | 0.0005 | 0.0005 | 0.0005 | 0.0005 | 0.0018 |
| n-Butane mol fraction | 0.0385 | 0.0385 | 0.0009 | 0.0009 | 0.0009 | 0.0009 | 0.0033 |
| i-pentane mol fraction | 0.0129 | 0.0129 | 0.0003 | 0.0003 | 0.0003 | 0.0003 | 0.0012 |
| n-Pentane mol fraction | 0.0107 | 0.0107 | 0.0004 | 0.0004 | 0.0004 | 0.0004 | 0.0013 |
| n-Hexane mol fraction | 0.0076 | 0.0076 | 0.0004 | 0.0004 | 0.0004 | 0.0004 | 0.0016 |
| n-Heptane mol fraction | 0.0035 | 0.0035 | 0.0003 | 0.0003 | 0.0003 | 0.0003 | 0.0011 |
| n-Octane mol fraction | 0.0011 | 0.0011 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0006 |
| n-Nonane mol fraction | 0.0006 | 0.0006 | 0.0001 | 0.0001 | 0.0001 | 0.0001 | 0.0005 |
| n-Decane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C₁₁⁺ mol fraction | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H₂O mol fraction | 0.0576 | 0.0576 | 0.3250 | 0.3250 | 0.3250 | 0.3250 | 0.9634 |
| TEG mol fraction | 0.0001 | 0.0001 | 0.6667 | 0.6667 | 0.6667 | 0.6667 | 0.0050 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |

| Stream | Molar flow kmol/h | 15 | 16 | 17 | 18 | 19 | 20 |
|--------|------------------|----|----|----|----|----|----|
| Mass flow kg/h | 41.03 | 41.03 | 41.03 | 41.03 | 41.03 | 41.03 |
| Temperature K | 477.2 | 423.0 | 365.9 | 307.15 | 365.9 | 366.9 |
| Pressure kPa | 105 | 75 | 45 | 45 | 45 | 5629 |
| Vapor fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| CO₂ mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| N₂ mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| Methane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| Ethane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| Propane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| i-Butane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Butane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| i-pentane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Pentane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Hexane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Heptane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Octane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Nonane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Decane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| C₁₁⁺ mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| H₂O mol fraction | 0.0983 | 0.0983 | 0.0983 | 0.0083 | 0.0983 | 0.0983 |
| TEG mol fraction | 0.9017 | 0.9017 | 0.9017 | 0.9917 | 0.9017 | 0.9017 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |
### Table A.2 Stream specifications of the proposed configuration

| Stream | Molar flow kmol/h | Mass flow kg/h | Temperature K | Pressure kPa | Vapor fraction | CO$_2$ mol fraction | N$_2$ mol fraction | Methane mol fraction | Ethane mol fraction | Propane mol fraction | i-Butane mol fraction | n-Butane mol fraction | i-Pentane mol fraction | n-Pentane mol fraction | n-Hexane mol fraction | n-Heptane mol fraction | n-Octane mol fraction | n-Nonane mol fraction | n-Decane mol fraction | H$_2$O mol fraction | TEG mol fraction | Total |
|--------|-------------------|----------------|---------------|-------------|----------------|---------------------|---------------------|--------------------|---------------------|---------------------|-----------------------|----------------------|----------------------|----------------------|----------------------|----------------------|----------------------|----------------------|----------------------|----------------------|-------------------|-----------------|
| 1      | 7412              | 150400         | 317.4         | 5101        | 1               | 0.0267              | 0.0183               | 0.8319             | 0.053               | 0.0366              | 0.0116                | 0.0042               | 0.0028               | 0.0017               | 0.0007               | 0.0002               | 0.0001               | 0.0011               | 0                   | 0.0022             | 0.0001         | 1.0000 |
| 2      | 20.51             | 2816           | 321.4         | 5237        | 0               | 0.0267              | 0.0183               | 0.8319             | 0.053               | 0.0366              | 0.0116                | 0.0042               | 0.0028               | 0.0017               | 0.0007               | 0.0002               | 0.0001               | 0.0011               | 0                   | 0.0022             | 0.0001         | 1.0000 |
| 3      | 7396              | 150100         | 319.1         | 5051        | 1               | 0.0267              | 0.0183               | 0.8319             | 0.053               | 0.0366              | 0.0116                | 0.0042               | 0.0028               | 0.0017               | 0.0007               | 0.0002               | 0.0001               | 0.0011               | 0                   | 0.0022             | 0.0001         | 1.0000 |
| 4      | 35.48             | 3097           | 318.43        | 5101        | 0               | 0.0267              | 0.0183               | 0.8319             | 0.053               | 0.0366              | 0.0116                | 0.0042               | 0.0028               | 0.0017               | 0.0007               | 0.0002               | 0.0001               | 0.0011               | 0                   | 0.0022             | 0.0001         | 1.0000 |
| 5      | 35.48             | 3097           | 319.75        | 757         | 0               | 0.0267              | 0.0183               | 0.8319             | 0.053               | 0.0366              | 0.0116                | 0.0042               | 0.0028               | 0.0017               | 0.0007               | 0.0002               | 0.0001               | 0.0011               | 0                   | 0.0022             | 0.0001         | 1.0000 |
| 6      | 35.48             | 3097           | 322.65        | 687         | 0               | 0.0267              | 0.0183               | 0.8319             | 0.053               | 0.0366              | 0.0116                | 0.0042               | 0.0028               | 0.0017               | 0.0007               | 0.0002               | 0.0001               | 0.0011               | 0                   | 0.0022             | 0.0001         | 1.0000 |
| 7      | 35.48             | 3097           | 370.97        | 657         | 0               | 0.0267              | 0.0183               | 0.8319             | 0.053               | 0.0366              | 0.0116                | 0.0042               | 0.0028               | 0.0017               | 0.0007               | 0.0002               | 0.0001               | 0.0011               | 0                   | 0.0022             | 0.0001         | 1.0000 |
| 8      | 0.716             | 19.1           | 370.97        | 657         | 0               | 0.1185              | 0.0040               | 0.5698             | 0.0785              | 0.0317              | 0.0726                | 0.0219               | 0.0082               | 0.0096               | 0.0082               | 0.0056               | 0.0025               | 0.0007               | 0.0022             | 0.0759             | 0                 | 0.8038          | 15.06 |
| 9      | 0.716             | 19.1           | 370.97        | 105         | 0               | 0.1185              | 0.0040               | 0.5698             | 0.0785              | 0.0317              | 0.0726                | 0.0219               | 0.0082               | 0.0096               | 0.0082               | 0.0056               | 0.0025               | 0.0007               | 0.0022             | 0.0759             | 0                 | 0.8038          | 15.06 |
| 10     | 34.76             | 3078           | 370.97        | 343         | 0               | 0.0017              | 0.0017               | 0.0017             | 0.0007              | 0.0003              | 0.0002                | 0.0002               | 0.0002               | 0.0002               | 0.0002               | 0.0002               | 0.0002               | 0.0002               | 0.0002             | 0.0002             | 0.0002           | 0.0002         | 1.0000 |
| 11     | 34.76             | 3078           | 429.65        | 313         | 0.5216          | 0.5216              | 0.5216              | 0.5216             | 0.5216              | 0.5216              | 0.5216                | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216             | 0.5216             | 0.5216           | 0.5216         | 1.0000 |
| 12     | 34.76             | 3078           | 408.35        | 114         | 0.5216          | 0.5216              | 0.5216              | 0.5216             | 0.5216              | 0.5216              | 0.5216                | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216             | 0.5216             | 0.5216           | 0.5216         | 1.0000 |
| 13     | 34.76             | 3078           | 405.65        | 103         | 0.5216          | 0.5216              | 0.5216              | 0.5216             | 0.5216              | 0.5216              | 0.5216                | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216             | 0.5216             | 0.5216           | 0.5216         | 1.0000 |
| 14     | 34.76             | 3078           | 287.5         | 1.00        | 0.5216          | 0.5216              | 0.5216              | 0.5216             | 0.5216              | 0.5216              | 0.5216                | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216               | 0.5216             | 0.5216             | 0.5216           | 0.5216         | 1.0000 |

(Continues)
| Stream | Molar flow kmol/h | 15 | 16 | 17 | 18 | 19 | 20 |
|--------|------------------|----|----|----|----|----|----|
| Mass flow kg/h | 2815 | 2815 | 2815 | 0.939 | 2816 | 2816 |
| Temperature K | 473.15 | 396.65 | 339.53 | 307.15 | 339.53 | 340.5 |
| Pressure kPa | 105 | 75 | 45 | 45 | 45 | 5271 |
| Vapor fraction | 0 | 0 | 0 | 0.00 | 0.00 | 0.00 |
| CO₂ mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| N₂ mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| Methane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| Ethane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| Propane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| i-Butane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Butane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| i-Pentane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Pentane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Hexane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Heptane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Octane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Nonane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| n-Decane mol fraction | 0 | 0 | 0 | 0 | 0 | 0 |
| H₂O mol fraction | 0.0974 | 0.0974 | 0.0974 | 0.0974 | 0.0974 | 0.0974 |
| TEG mol fraction | 0.9026 | 0.9026 | 0.9026 | 0.9026 | 0.9026 | 0.9026 |
| Total | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 | 1.0000 |