Process Design and Steady State Simulation of Natural Gas Dehydration Using Triethylene Glycol (TEG) to Obtain The Optimum Total Annual Costs (TAC)

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Abstract. Natural gas is one of the most desired raw materials which can be used in petrochemical industries or as energy resource. Natural gas usually obtained from underground reservoirs and it must go through purification process so it can be utilized. All acid gas compounds such as H₂S, CO₂ and all liquids including H₂O must be removed. Most industries use triethylene glycol (TEG) dehydration unit to absorb water from natural gas streams. Therefore, further research about design and optimization of dehydration unit using TEG should be conducted since the optimization of dehydration unit using TEG affects to safety, operability, and stability of the process. In this study, the optimization of TEG Dehydration Unit process is conducted to minimize the Total Annual Costs (TAC) and improve the efficiency of the process. Aspen Plus software is used to perform the simulation of TEG dehydration process. Optimization is conducted by changing several of the base case operating conditions which have been created using existing condition to obtain the optimum conditions with minimum TAC on the existing circumstances. The variables that are used in this research is absorber column pressure (35-45 barg with 2.5 barg interval) and lean TEG temperature (39-49°C with 2°C interval). The constrain of the absorber column pressure is less than 45 barg, since the natural gas feed pressure from well is 45.16 barg. The results show that by changing some operating conditions can reduce the size of the column, reduce the energy costs including steam, cooling water, and electricity costs. Therefore, it can reduce the TAC of the natural gas dehydration unit. The validation simulation results of the steady state of the TEG Dehydration Unit using Aspen Plus and the real plant produces relatively small % error, so it can be used to create a base case using the existing data. The mole fraction of H₂O in dehydrated gas after optimization using Aspen Plus is 0.000178 while in real plant the result is 0.000200. The simulation results can reduce the TAC from 3,416,739 USD to 2,973,219 USD.

1. Introduction

Natural gas is a vital component of the world's energy supply. Natural gas is also considered as important resources for various production processes, both fuel and raw materials in various industries such as fertilizers and various kinds of chemicals [1]. Natural gas usually obtained from a natural underground reservoir and it generally contains a large quantity of methane along with heavier hydrocarbon such as ethane, propane, isobutene, normal butane, etc. Also, in the raw state, it often contains a considerable amount of non-hydrocarbons, such as nitrogen, hydrogen sulfide, and carbon dioxide. There are some traces of such compounds as helium, carbonyl sulfide, and various mercaptans. It is also generally saturated with water [2]. However, several treatments are needed to convert the raw natural gas become commercial gas which has significant different specification. All acid gas compounds, H₂S and CO₂ must be removed. Also, all the free liquids, both hydrocarbon and water, have to be removed [2]. The dehydration process usually used to remove the water content from natural gas because the content of water vapor and hydrocarbons can form gas hydrates which can inhibit the flow in pipes and other equipment, especially in the control system. Water vapor also increases the corrosion of natural gas, especially when there is acidic gas [3]. In general, many natural
Gas industries use triethylene glycol (TEG) dehydration process to remove water from natural gas flows to meet pipeline quality standards. The advantages of TEG application compared to the other dehydration process are: lower vapor pressure, higher resistance to degradation, and lower viscosity [3].

Many researchers studied about the simulation and optimization of TEG dehydration process in natural gas purification. Darwish et al. [4] studied natural gas dehydration plant which employs triethylene glycol (TEG) as the dehydrating agent has been simulated using Aspen Plus, it resulted that the dew point (water contents) of the dry gas issuing from the contactor is mostly responsive to disturbances in the sweet gas flow rate and TEG losses are mostly sensitive to the contactor pressure. Kazemi et al. [5] studied the sensitivity analysis of natural gas dehydration using TEG that shows the effectiveness parameters such as TEG circulation rate, equilibrium stages of absorption column, and operating conditions with economical considerations. Bahadori et al. [6] studied the development of simple sizing method of TEG absorber at wide operating conditions by considering TEG circulation rate and TEG purity which shows the correlation of capital cost and TEG absorber operation condition.

Based on the literature reviews that have been done, there are no studies about the simulation and optimization of TEG dehydration process by total annual cost (TAC) calculation with interest rate consideration. Therefore, in this study, the simulation and optimization of natural gas dehydration process using TEG were conducted by considering TAC coupling with interest rate. The simulation was conducted using Aspen Plus software by considering the operation of real plant. The optimization was conducted by considering the TAC in various variables of absorber column pressure and lean TEG temperature which can affect the water content of dry gas, heat duty of heat exchangers, and columns diameter. Then, all those results are used to find the optimum TAC where the annualized capital cost was found using interest rate calculation with assumption 10% interest/year for 10 years.

2. Methods
The simulation conducted in this study aims to examine theoretically the factors that influence the H2O content of the product on the absorber and heat duty cooler in the dehydration process of natural gas from triethylene glycol (TEG) solution.[8] Then, in this study, the calculation of the total model of dehydrated natural gas units is obtained by assuming a steady-state system.

2.1 Feed composition and product specification
The information of feed gas composition and gas product specification are required to develop the appropriate process design of TEG dehydration process. In this study, real plant data was used to specify the feed gas composition. Table 1 shows the feed gas composition dan lean TEG specification used in absorber column.

| Composition | Feed % mole | mass flow (kg/hr) | Lean TEG % mole | mass flow (kg/hr) |
|-------------|-------------|------------------|-----------------|------------------|
| H2S         | 0.00        | 0.00             | 0.00            | 0.00             |
| CO2         | 2.67        | 8709.56          | 0.00            | 0.00             |
| Nitrogen    | 1.83        | 3799.74          | 0.00            | 0.00             |
| Methane     | 83.19       | 98920.34         | 0.00            | 0.00             |
| Ethane      | 5.30        | 11812.44         | 0.00            | 0.00             |
| Propane     | 3.66        | 11962.47         | 0.00            | 0.00             |
| i-Butane    | 1.00        | 4308.11          | 0.00            | 0.00             |
| n-Butane    | 1.16        | 4997.40          | 0.00            | 0.00             |
| i-Pentane   | 0.42        | 2246.07          | 0.00            | 0.00             |
| n-Pentane   | 0.26        | 1497.38          | 0.00            | 0.00             |
| n-Hexane    | 0.17        | 1085.87          | 0.00            | 0.00             |
Due to the requirement for further process, the gas product specification from TEG dehydration unit are determined to has a maximum limit of dry gas water content with value of 10 lb/MMscf [9].

2.2 Simulation basis

This research used the Aspen Plus to simulate natural gas dehydration using triethylene glycol (TEG). Figure 1 shows the model simulation of natural gas dehydration using Triethylene Glycol (TEG).

| n-Heptane | 0.07 | 519.90 | 0.00 | 0.00 |
| n-Octane  | 0.02 | 169.34 | 0.00 | 0.00 |
| n-Nonane  | 0.01 | 95.06  | 0.00 | 0.00 |
| H₂O       | 0.22 | 293.76 | 8.44 | 594.93 |
| TEG       | 0.00 | 0.00   | 91.56| 53730.95|
| **Total** | 100  | 150417.43 | 100 | 54325.88 |

The equipment specification in this research is shown in Table 2.

| Equipment      | Parameter | Value |
|----------------|-----------|-------|
| Absorber Column| Internal type | Tray  |
2.3 Optimization basis

After the operating conditions on base case are obtained, the optimization is done by varying absorber column pressure (35-45 barg with 2.5 barg interval) and lean TEG temperature (39-49°C with 2°C interval) to get the optimum total annual cost (TAC). Parameters considered to obtain the optimum total annual cost are equipment cost, depreciation, and operating cost. The formulas to calculate the total annual cost were taken from literature. Table 3 summarizes the formula for all the TAC calculations.

| Equipments | Cost Parameter |
|------------|----------------|
| Columns    | NT trays with 2 ft spacing plus 20% extra length |
|            | Column length (l) |
|            | Column and other vessel (d and l are in meters) |
|            | Capital cost 17640(D)^1.066(L)^0.802 |
| Flash drums| Aspect ratio of L/D 2 |
|            | Volume 10 min x liquid flowrate |
|            | Capital cost 17640(D)^1.066(L)^0.802 |
| Reboilers  | Heat – transfer coefficient 0.568 kW/K.m² |
|            | Differential temperature Steam temperature – base temperature (AT > 20 K) |
|            | Capital cost 7296(A)^0.65 |
| Heat exchangers | Heat exchangers, liquid-to-liquid (area in m²) |
|            | Heat – transfer coefficient 0.852 kW/K.m² |
|            | Differential temperature LMTD of (inlet and outlet temperature differences) |
|            | Capital cost 7296(A)^0.64 |
| Cooler     | Heat – transfer coefficient 0.852 kW/K.m² |
|            | Differential temperature LMTD of (inlet or outlet temperature – 315 K) |
Capital cost $7296(A)^{0.65}$

| Heaters          |                      |
|------------------|----------------------|
| Heat – transfer coefficient | 0.568 kW/K.m²       |
| Differential temperature | LMTD of (steam temperature – (inlet or outlet temperature) |
| Capital cost     | $7296(A)^{0.65}$    |

| Utilities       |                      |
|-----------------|----------------------|
| High pressure steam | $9.88/GJ (41 barg, 254 °C) |
| Medium pressure steam | $8.22/GJ (10 barg, 184 °C) |
| Low pressure steam | $7.78/GJ (5 barg, 160 °C) |
| Cooling water   | $0.354/GJ            |
| Electricity     | $16.9/GJ             |

After obtaining total capital costs, then calculating annualized capital cost which is the price per year that must be spent on the equipment used. The time assumption used is 10 years, then the calculation of interest for annual capital costs is adjusted to that time period. To calculate the annual capital cost the following equation is used [7]:

\[
\text{annualized capital cost} = \frac{\text{capital cost}}{n} \times \frac{1}{\left(1 + \frac{i}{n}\right)^n - 1}
\]

Total annual cost can be formulated below:

\[
\text{Total Annual Cost} = \text{annualized fixed cost} + \text{variable cost}
\]

Where annualized fixed cost:

\[
\text{annualized fixed cost} = \frac{\text{annualized equipment capital cost}}{\text{utility cost}}
\]

The variable cost can be calculated as:

\[
\text{variable cost} = \text{utility cost}
\]

3. Result and discussion

3.1 Base case simulation and validation

The base case simulation was generated based on the process design and operating condition which described in Subsection 2.1. The base case validation is done to measure the suitability of the simulation by comparing simulation data with the real plant data. A steady state simulation that has been done shows results that have a maximum error of dehydrated gas water content 10.5%. This difference is generated because the real plant data were taken from old plant which has decreased efficiency. Occasionally, the increasing the age of the plant caused the higher water content on dehydration gas compared to the simulation results. The base case validation in this process are shown in Table 4.

| Component | Design | Simulation | Error (%) |
|-----------|--------|------------|-----------|
| H₂O       | 0.0002 | 0.000179   | 10.50     |
| TEG       | 0      | 0          | 0.00      |
| Suhu (°C) | 45.97  | 47.926     | -4.25     |

3.2 Design parameter of the columns

The design parameter is an important aspect that must be determined in this simulation. This is because the results of the design will affect the performance of the equipment which will affect the capital cost which is one of the considerations to find the optimal TAC. This process operates at pressure 35-45 barg with 2.5 barg interval and temperature 39-49°C with 2°C internal. The absorber and regenerator columns can be normally operated when weeping and flooding didn’t occur [14]. It can be checked by varying the column diameter to find the lowest ones to reduce the columns capital
cost. When the column diameter is too low, anomaly condition was occur such as flooding. The lowest diameter of absorber column used in this study is 1.81 m for condition of 45 barg and 49°C. So, if the column diameter is lower than 1.81 m, flooding will occur and decrease its performance. Meanwhile, when the column diameter is too high, the column begins to weep. So if the column diameter is too high, weeping will occur and can not be operated well. From the simulation, the column can be operated when the column diameter is 1.81 m. Figure 2 show the absorber column performance with diameter of 1.81 m for condition of 45 barg and 49°C.

![Figure 2. Trays condition at diameter 1.81 m](image)

As seen in Figure 2, the operating point doesn’t touch the weeping or flooding line which means the column can be operated at the diameter 1.81 m. The next step is to simulate again using the Aspen Plus software to get the optimal absorber and stripper column design. With the most optimal design obtained, it is expected to be able to reduce the value of capital costs for both columns. This price reduction is done by finding the smallest column diameter that meets the hydraulic plot in the simulation results using Aspen Plus software. The design parameters used are in accordance with the provisions of Aspen Plus:

| Design parameter                      | Value          |
|---------------------------------------|----------------|
| % jet flood and downcomer flood       | 80             |
| Minimal downcomer area fraction       | 0.1            |
| Maximal pressure drop                 | 0.0247 atm     |
| Maximal % jet flood                   | 100            |
| Maximal % downcomer backup            | 100            |
| Maximal % liquid entrainment          | 10             |
| Minimal weir loading                  | 4.471 cum/hr-meter |
| Maximal weir loading                  | 117.372 cum/hr-meter |
| Status warning (% limit)              | 10             |
| Foaming system factor                 | 1.0            |
| Aeration multiplier factor            | 1.0            |
| Over design factor                    | 1.0            |
3.3 The effect of absorber column pressure and lean TEG temperature on the column performances

The variables used in this study can affect the performance of the absorber and regenerator columns. These performances have impacts on changes in dry gas water content, cooler heat duty, and column diameter suppression. The optimization of the process was conducted to obtain the low water content in dehydrated natural gas and the optimum condition to get total annual cost. The variables are the absorber operating pressure and lean TEG temperature. The absorber operating pressures are varied from 35 to 45 barg with 2.5 barg interval and the lean TEG temperature are varied from 39 °C to 49°C with 2 °C interval. Figure 3 shows the effect of absorber operating column pressure and lean TEG temperature to dehydrated natural gas water content.

As seen in Figure 3, if the absorber column pressure increases, the water content of dehydrated natural gas will decrease. Meanwhile, if the lean TEG temperature increases, the water content of dry gas will increase. This is because the absorption process is more optimal at low temperatures. The lean TEG temperature in natural gas dehydration units is better if it is not cooler than 12°C to avoid condensation of hydrocarbons in the feed of natural gas [15]. The high water content of dehydrated natural gas is limited to a maximum of 10 lb/MMscf according to the specification of the sales gas [9]. The water in the pipe is a problem because it can make the flow clogged, increase corrosion, and cause gas hydrate [3].

The lean TEG temperature is varied to study the effect on the operating cost. When the output temperature of lean TEG decreases, more duty is needed and the cost will increase. Figure 4 shows the effect of lean TEG temperature to lean TEG cooler heat duty.
One of the parameters that seen in this study is the cooler heat duty. In this case, the cooler was chosen as a reviewed object because it is directly related to the independent variables used in this study. Where these two variables, absorber column pressure and lean TEG temperatures, influence the price of a cooler that functions to cool lean TEG which goes into the absorber. The absorption process is better done at low temperatures, but to reach low temperatures a high heat duty cooler is needed [15]. The inverse relationship of the lean TEG temperature and cooler heat duty affect the operating cost of the cooler which also affect the result of the TAC. Another variable that affects the absorption process is pressure, which is better done at high pressure. From Figure 4, when the pressure was higher, the heat duty cooler decreases and it will affect the finding of the optimum TAC value.

The diameter of absorber and regenerator column was also determined as parameter. The relation between absorber and regenerator column diameters with absorber operating pressure is inversely proportional. High pressure can reduce the column diameter. In order to contance the gas with the liquid properly, the gas and liquid must enter the column with a certain rate of filling material. Naturally, the liquid flowing down will choose to pass at the point where the resistance is low. If the speed or pressure of the gas is too large (because the diameter is too small), then the rate of liquid downward will be disrupted by the upward flow of gas and it will begin to collect liquid in the column (an increase in liquid hold up). In this loading condition, a slight increase in the gas rate will cause a very large increase in pressure drop. If the rate of gas continues to increase, the liquid does not flow down but collects at the top of the column. Then, the backflow occurs and the liquid spills out of the column so that flooding occurs. A column is usually operated under conditions just before loading. In this condition, the flow of liquid will be well distributed so the mass transfer is well operated. In column design, gas speeds are usually taken at 50% – 80% flooding speed [14]. By paying attention to the hydraulic plot on Aspen Plus, finding the smallest diameter column that meets the feasibility of the column. The effect of absorber operating pressure to absorber and regenerator column diameters shown in Figure 5 and Figure 6.

**Figure 4.** Effect of lean TEG temperature to lean TEG cooler heat duty in various pressure condition.
Figure 5 and Figure 6 shows the higher value of column pressure, the smaller the diameter can be found. It can reduce the capital cost of both columns the optimal TAC value until optimal results can be obtained.

3.6 Total annual cost calculation

In this study the calculation of capital cost is used in the calculation from Luyben [13], where an approach has been found to calculate the price of the tool according to its size. In addition to size, the pressure variable also affects capital costs. After obtaining total capital costs, the annualized capital costs can be obtained which is the price per year that must be spent on the equipment used. The time assumption used is 10 years, then the calculation of interest for annual capital costs is adjusted to that time period. Annualized capital costs can be known from all the prices of existing tools, for example the price of a heat exchanger can be known from the area of heat transfer value.

The annual capital cost is calculated with the following equation [7]:

$$\text{Annualized capital cost} = \text{total capital cost} \times \frac{i(1+i)^n}{(1+i)^n-1}$$

where:

- $i$ = fractional interest rate per year
- $n$ = number of years

Figure 5. Effect of absorber operating pressure to absorber column diameter

Figure 6. Effect of absorber operating pressure to regenerator column diameter
In this study the calculation of the operating cost is obtained by calculating steam requirements, cooling water, electricity, and TEG make-up flow. The utility requirement of each equipment can be obtained from the Aspen Plus simulation results. Furthermore, from each number of utilities needed the price is calculated according to those listed in Table 3.

Total annual cost (TAC) is obtained by summing annual capital cost with annual operating cost. Because the cost parameter obtained from Luyben are determined in 2011, the TAC result should be corrected based on cost index by using least square method [7]. Figure 7 shows the effect of absorber column pressure and lean TEG temperature to TAC.

![Figure 7. Effect of absorber column pressure and lean TEG temperature to TAC](image)

As seen in Figure 7, the value of TAC was decreasing while the increasing of the column pressure and lean TEG temperature. The lowest TAC result was found in the highest absorber pressure and the highest lean TEG temperature, because when the lean TEG temperature was high, it didn’t need a big amount of cooling water, so the operating cost could be reduced. The reduction in operating cost can have a bigger effect on the TAC result more than the reduction in the capital cost.

4. Conclusion

The simulation of natural gas dehydration unit using triethylene glycol (TEG) is created using Aspen Plus based on the real plant condition. The base case has been created and give the product of dehydrated natural gas with water content value of 8.514 lb/MMScf and TAC value of 3,694,601.2 USD/year. The design optimization is done by varying absorber column pressure and lean TEG temperature and give the optimum condition at absorber column pressure 45 barg and lean TEG temperature 39˚C. The optimum total annual cost of the natural gas dehydration unit is 3,240,234.9 USD/year with 8.656 lb/MMScf water content in the dehydrated natural gas.

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