INTERNATIONAL JOURNAL OF LATEST TECHNOLOGY IN ENGINEERING, MANAGEMENT & APPLIED SCIENCE (IJLTEMAS)

ISSN 2278-2540 | DOI: 10.51583/IJLTEMAS | Volume XIV, Issue I, January 2025

Simulation of Glycol Regeneration in a Natural Gas Dehydration Unit to Minimize Losses at Optimal Operation Conditions

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DOI : https://doi.org/10.51583/IJLTEMAS.2025.1401028

Received: 01 February 2025; Accepted: 07 February 2025; Published: 18 February 2025

Abstract: The focal point of this study is to analyse the workings of a glycol dehydration unit in a natural gas plant and minimize the losses in the triethylene glycol solvent at the optimal operating conditions, by using Aspen HYSYS to simulate the plant, and to suggest different optimization paths that maximize the profit or functionality of the plant.

The simulation was conducted in the steady-state mode by relying on the data provided from real data by the natural gas plant to establish a model that reflected the results seen in the actual plant accurately. On validation, the case studies were established by choosing the prominent operating conditions and gauging their effects on the characteristic process outputs of a dehydration unit.

Furthermore, to find the optimum conditions to maximize the profit or functionality, 350,640 HYSYS cases were conducted to record the data of every possible sensible combination of operating conditions, then the weighted normalized method was applied to find the optimum conditions. Six alternatives were generated depending on the optimization problem. Among the suggested alternatives, two cases were recommended based on the objective function, where the first objective function was to provide the maximum profit, with 923 $/h over the base case, and the second objective function was to provide the minimum moisture content of 0.00014% wt., which eventually provides a less profit, thus only being recommended for processes that require extreme drying.

Keywords: TEG, triethylene glycol, Aspen HYSYS, natural gas, dehydration, optimization.

I. Background and Literature

Processed natural gas is a valuable resource used in many fields and sectors. This paper is concerned with the dehydration aspect of the processing of natural gas. Knowing the composition of natural gas better showcases the importance of the dehydrating process, as the acid gas content (CO_2 and H_2S) present within it would cause side reactions and safety-related matters when contacted by liquid water, which will cause corrosion in vessels and more, mainly the pipelines, as the water in process gas can condense in the long pipelines and cause corrosion throughout the long stretch of pipes from the wells to the processing plant. Dehydrating natural gas provides the system with two benefits: it lowers the water content in the process gas as well as lowers its dew point, thus preventing condensation from happening. The plant in hand uses triethylene glycol (TEG) as its dehydrating solvent, which is known for its hygroscopic attributes, and has an affinity to absorb the water vapor from the natural gas stream, thus facilitating the dehydration process. TEG is ideal for use in conjunction with natural gas, as it has a low solubility in hydrocarbons, a low affinity for absorbing hydrocarbons and acid gases, high hygroscopicity, and a low tendency to foam or emulsify, as well as being non-corrosive [1]. Other studies considered optimizing the performance of the TEG dehydration process through simulations to predict CO_2 emissions, water content, and energy consumption [2].

There are many efforts in optimizing the design of the TEG dehydration units [3], [4], [5], [6], [7], [8]. Experiments were conducted to obtain the vapor-liquid equilibria of selected aromatic hydrocarbons in TEG [9]. Quick estimations of the absorption of the aromatic compound were also simulated for the TEG dehydration process [10], where the model is used in optimizing the dehydration units. A parametric optimization was performed using the ASPEN HYSYS simulator to minimize the processing cost, relying on several parameters including 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 [3]. The process variables, including the utilities, were also considered within the study to obtain the optimum condition. In another paper, the environmental condition was considered in the glycol dehydration process, where environmental impact was calculated [2], [11]. The study concentrated on eliminating hydrocarbons' emissions from glycol dehydrators. There are several efforts in modeling the TEG–water–natural gas system to provide accurate predictions [4], [12], [13]. A steady-state simulator was employed using TEG as the dehydrating agent [6]. The study concludes that the stripping gas is very effective in improving the regenerated TEG concentration, hence improving the performance of the dehydration plant. A sensitivity analysis using thermodynamic models was performed on the costs, where effects were noted on the operational cost [7]. A recent publication studied the optimization of TEG dehydration of natural gas, where the HYSYS simulator and optimizer tool was used to minimize the processing cost considering a group 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 [3]. The design parameters were modified to minimize the dehydration cost using a defined objective function. The parametric optimization analysis identified the key parameters to minimize the cost. There had been recent efforts for estimating TEG purity in natural gas dehydration units

ISSN 2278-2540 | DOI: 10.51583/IJLTEMAS | Volume XIV, Issue I, January 2025

using fuzzy neural networks. The efforts concentrated on key parameters such as reducing the pressure of the reboiler and gas stripping. The accurate predictions reflected the importance of these two parameters in the optimization study.

The main objective of this study is to observe the trends that affect the TEG loss in the dehydration process through Aspen HYSYS, thus facilitating an experimental background to deduce how to minimize the TEG loss in the process. Lastly, this study also aims to suggest optimization paths that minimize the losses while maximizing the profit.

Problem Setup

This section will cover the simulation specifics used to conduct this experiment in the steady state and discuss the optimization method and its application. The workings of a glycol dehydration unit (GDU) are well known and are mostly the same universally, as the unit contains an absorption section to facilitate the moisture absorption process and a regeneration section, where the water-rich TEG is returned to lean TEG. The process at hand includes a packed absorption column, a flash tank to release entrapped hydrocarbons, charcoal and sock filters to clean any debris, and degenerated TEG. In addition, it also works to absorb some of the hydrocarbons and moisture from the solvent through a packed regeneration column with a reboiler and overhead condenser, a glycol/glycol exchanger to make use of the lean TEG heat to prepare the rich TEG for regeneration, and finally recycle preparations such as cooling and pumping.

The glycol dehydration unit (GDU) absorber is a structured packed column designed to enhance gas-liquid interaction for effective water removal. The column contains equilibrium stages, with lean TEG entering at the top defined at a specific flow rate and wet natural gas introduced from the bottom at a defined flow rate. The gas exiting the absorber achieves a targeted water content of 0.01% wt., which aligns with industry standards for pipeline transmission.

The glycol regenerator includes a packed stripping column, a reboiler, and a condenser, operating under controlled conditions to maintain glycol purity. The reboiler is set to a maximum temperature of 204°C to prevent thermal degradation of triethylene glycol (TEG). A flash tank at a defined pressure is used to separate entrained hydrocarbons before the glycol enters the stripping column.

A steady-state simulation was conducted in Aspen HYSYS using real process data obtained from an industrial natural gas processing facility. The process variables, including TEG temperature, circulation rate, gas composition, and absorber pressure, were calibrated to align with actual plant operation conditions. These details ensure the reproducibility of the simulation model and align the results with real-world industrial applications.

Fluid Package

Several fluid packages were tested, such as Peng-Robinson-Stryjek-Vera Equation of State (PRSV), Non-Random Two-Liquid Model (NRTL), Cubic-Plus-Association Equation of State (CPA), and Peng-Robinson, where the latter was found to be the most accurate by comparing the main simulation results, which were the glycol purity with and without the use of a stripping gas and the amount of dry gas product, with the natural gas plant’s real data. In the selection of Peng-Robinson over alternative models, the following were performed:

Thermodynamic Model Evaluation Methodology. This included testing of multiple fluid packages, including PRSV, NRTL, CPA, and Peng-Robinson EOS, to determine the most accurate model for simulating glycol dehydration. The selection was based on comparisons of TEG purity and dry gas moisture content with real plant data, as shown in Tables 1 and 2. The Peng-Robinson EOS model produced results with the lowest deviation from actual plant data, making it the most suitable choice for the system under study.

Considering the Gamma-Phi Approach (NRTL + Peng-Robinson), additional simulations using NRTL for the liquid phase and Peng-Robinson for the vapor phase were performed. The results showed minor variations compared to the single Peng-Robinson model, but the improvements were not significant enough to justify a change in approach.

Enhancements were undergone to the Peng-Robinson EOS Model. The Peng-Robinson model in Aspen HYSYS incorporates volume translation and modified mixing rules that enhance its applicability to non-ideal systems like TEG-water-hydrocarbon mixtures. These enhancements allow for better predictions of vapor-liquid equilibrium and phase behavior over a wide range of temperatures and pressures. References detailing these improvements have been included in the revised manuscript [8], [14], [15].

Steady State Simulation

The steady-state simulation was made in Aspen HYSYS, using the provided data and specifications to construct an accurate model of the GDU, where the final simulation conditions and outputs matched those observed in the plant. Figure 1 shows the final steady-state simulation environment. The main simulation outputs are to match the TEG purity, which is set to be 99.75% wt. at minimum, and the glycol reboiler temperature, which must not exceed 204°C to avoid TEG thermal degradation.

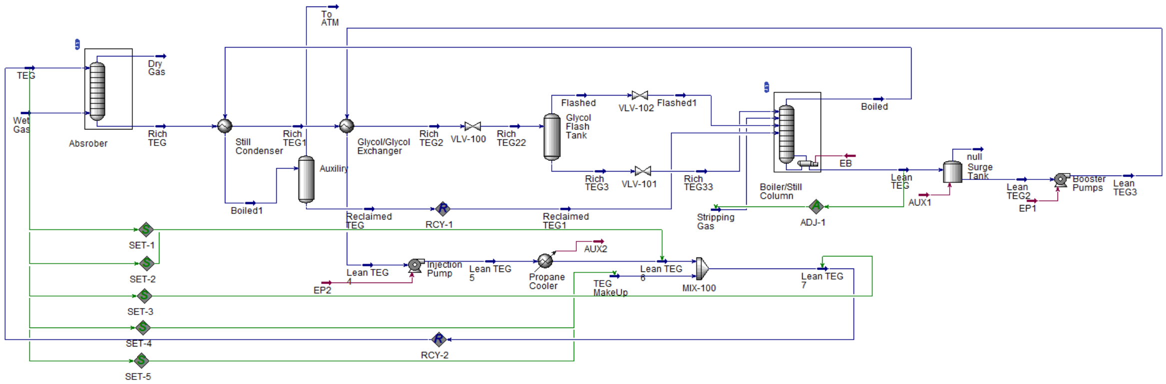

Figure 1: Steady State simulation of dehydration unit

Dynamic simulation is an extension of the steady-state process simulation whereby time-dependence is built into the models via derivative terms, i.e., accumulation of mass and energy. The time-dependent description, prediction, and control of real processes in real time have become possible with dynamic simulation. This includes startup and shutdown procedures, holdups, changes of conditions during a reaction, thermal changes, and more. The dynamics simulation is more complex than the steady-state simulation, as it requires increased specifications and calculations done by the software. This can be seen as multiple repeated steady-state simulations (based on a fixed time step) with constantly changing parameters [16]. Figure 2 shows the final dynamics simulation. The following modifications have been made to improve transparency:

The process specifications were obtained from an industrial natural gas processing facility, ensuring the accuracy of the simulation setup. Where direct plant data were unavailable, values were supplemented using Aspen HYSYS inbuilt thermodynamic databases.

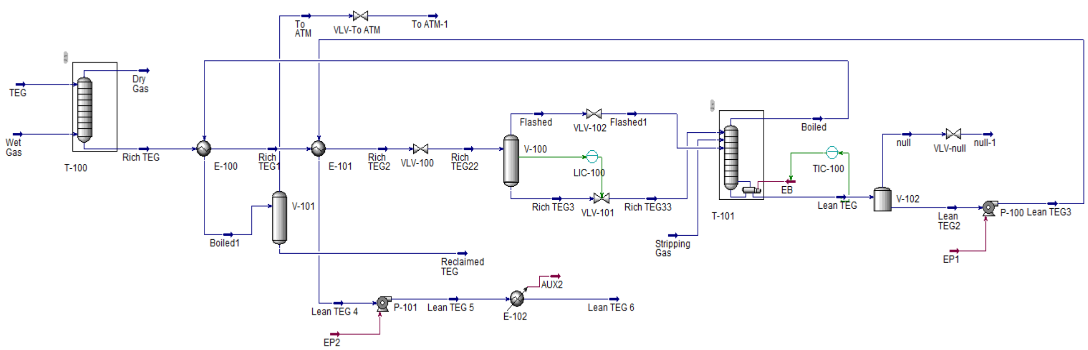

Figure 2: Dynamic simulation of dehydration unit

II. Weighted Normalized Method

Sometimes choosing the best simulation or a run can be tedious, as there are many important factors that can be overshadowed by just choosing the best-looking simulation. Hence, in this paper, the weighted normalized method (WNM) was utilized, as this method gives the benefit of implementing all the factors whilst giving every factor a desired weight. Equation 1 represents the form of the weighted normalized method formula that was used in this study, (Petropoulou, 2019).

$$Deviation = \left(\sum_{i=1}^{n} W_i \times ABS\left(\frac{Actual\ Value_i - Optimum\ Value_i}{Optimum\ Value_i}\right)\right) \quad (1)$$

Where the deviation is the difference between the tested and optimum values, n is the total number of desired variables, W_i represents the weight of the variable, $Actual\ Value_i$ is the value of the variable in the current sample, and $Optimum\ Value_i$ is the optimum value that can be obtained for that variable regardless of the remaining variables in that sample, where the optimum is either the maximum or minimum or base case value, depending on the desired outcomes. The optimization is done by applying Eq. 1 on all the samples and choosing the sample that gives the lowest deviation.

The constraints that limit this study are listed below; they vary between natural gas plant set constraints, environmental constraints, and chemical constraints:

- Lean TEG should be introduced 10-12°C above the inlet gas temperature; this allows for proper absorption of water whilst minimizing TEG loss in the absorber. Furthermore, the process should be operated at justifiably low temperatures, but not low enough that higher hydrocarbons (propane, butane, etc.) start to condense and leave in the bottom with the

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TEG, which can cause problems in the boiler afterwards. Hence, this constraint regulates the amount of TEG lost in the absorber and serves as a safety guideline for the TEG dehydration process.

- The boiler temperature must not exceed 204°C, as temperatures past that will result in the thermal degradation of the solvent TEG.
- The H_2S and CO_2 concentrations in the effluent must not exceed 10 and 5000 ppm, respectively, as per OSHA regulation.
- The TEG circulation must be at a purity of 99.75% wt. at minimum to guarantee adequate moisture removal.
- The wet gas feed temperature must not drop to 55 °C, as at those temperatures there will be liquid formation, which will affect the functionality of the absorber.

III. Results and Discussion:

Verification

The simulation results were validated with real data. A comparison was made between the selected defining process variables that were obtained from the model and the natural gas plant specifications for these variables, as displayed in Table 1.

Table 1: Steady state verification data

Condition	HYSYS Model	natural gas plant Specification	% Error
TEG Purity Without Stripping Gas	99.05% wt.	99.1% wt.	0.05%
TEG Purity with Stripping Gas	99.75% wt.	99.75% wt.	0.00%
Dry gas flow	2987 kmol/h	2988 kmol/h	0.03%

As the low percentage errors suggest, the HYSYS simulation is validated and can be used to undergo parametric optimization. This is done through conducting case studies.

To confirm that the base plant is valid and realizable and to confirm the validity of the internals of the columns that are ignored in the steady-state mode, the simulation was run in the dynamic's mode with the essential controllers adequately tuned. Table 2 shows the comparison between the dynamic mode data after they reach a steady state and the natural gas plant specifications for the same data.

Table 2: Dynamics mode verification data

Condition	HYSYS Model	Natural Gas Plant Specifications	% Error
TEG Purity	99.75% wt.	99.60% wt.	0.15%
Temperature after the still exchanger	66.26°C	67 °C	1.1%
Temperature after the glycol/glycol exchanger	146°C	160°C	8.75%
Dry gas flow	2987 kmol/h	2988 kmol/h	0.3346%

Evident by the relatively low percentage errors seen in Table 2, the dynamic simulation is accurate enough to conduct stability testing and used to conduct further case studies and optimization solutions.

Case Studies

The first part of the results represents the case studies that were conducted to demonstrate the effects of selected process variables that were deemed focal for the operation of the GDU on other process outputs that indicate the performance of the GDU.

The 7 process variables were the TEG circulation and temperature at the inlet to the absorber, the wet gas flow rate and temperature at the inlet to the absorber, the operation pressure inside the glycol flash tank, and the flow rate and temperature of the stripping gas to the boiler, while the selected process variables that were studied were the water content in the dry gas, the boiler duty, the amount of stripping gas required to achieve the minimum 99.75% wt. purity of recirculated TEG, and the glycol loss in the process, which was represented by the glycol makeup stream. The selection for the tested process variables is supported by literature [3], [4], [6], [7], [17].

It should be noted that each section will only concern one process variable; this means that the respected study was conducted by changing this one variable while keeping every other process variable constant on the base steady-state value. Another matter that should be discussed before displaying the results is how the TEG loss is represented. The results show the data of the TEG makeup stream when it does not accurately represent the amount of TEG lost in the process, as it was found to be slightly more

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than what was lost in the process. However, the TEG makeup stream accurately represents the amount of TEG that will be used to maintain the process steady on the specified circulation rate and consequently serve to calculate how much the company spends to compensate for the lost TEG, therefore validating its use. Furthermore, this mirrors how the natural gas plant compensates for TEG loss in the real plant. The following figures show the simulation that was used to conduct the case studies, and the following will contain the main results of the case studies along with the associated conclusion.

TEG Flow Rate

This part concerns the effects of the TEG circulation rate on the selected process outputs, as is demonstrated in Figures 3, 4, 5, and 6.

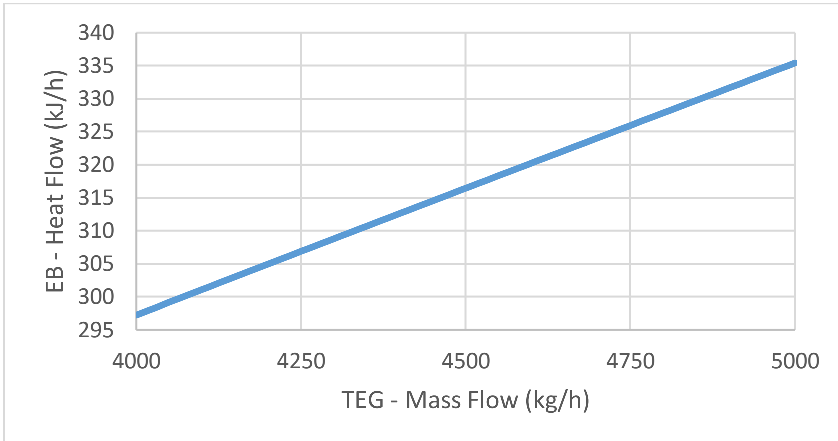

Figure 3: Dry gas water content vs. TEG circulation rate

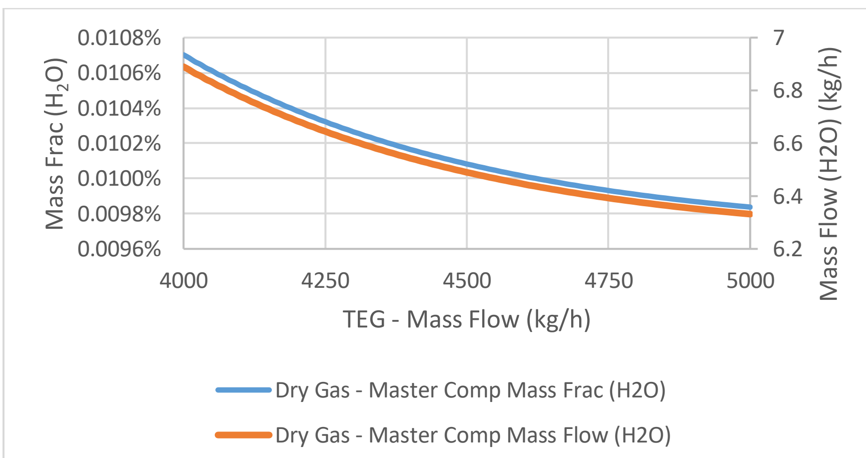

Figure 4: TEG boiler duty vs. the TEG circulation rate

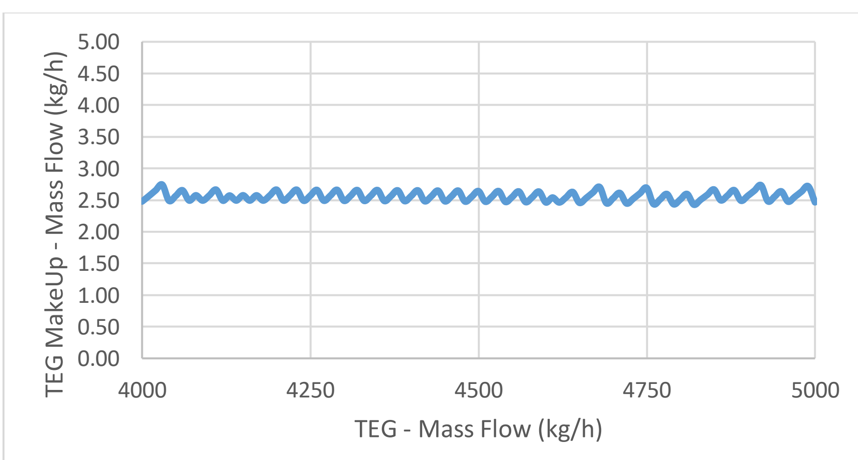

Figure 5: TEG loss vs. TEG circulation rate

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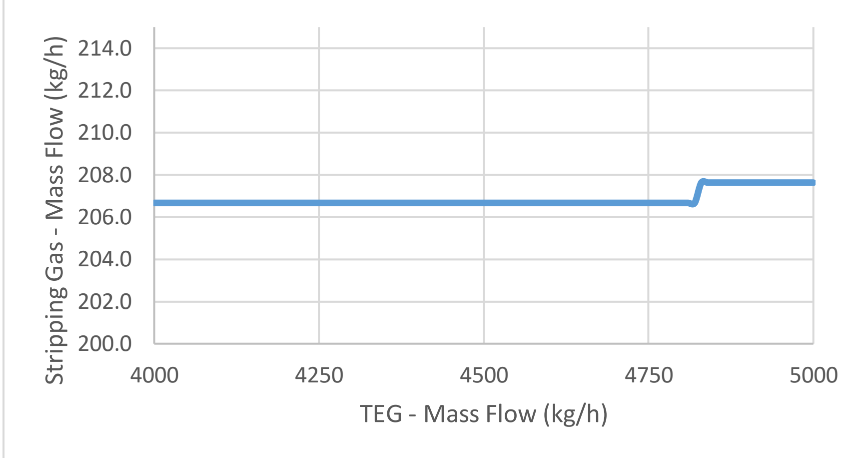

Figure 6: Stripping gas amount required for 99.75% purity of TEG vs. TEG circulation rate

As shown in Fig. 3, the water content in the dry gas decreases as the TEG circulation is increased, which is to be expected, as more TEG guarantees that there is thorough contact between the liquid and gas phases, enabling it to absorb more water before being saturated. Fig. 4 shows the direct relationship between the boiler duty and TEG's circulation; the result is expected as more heat will be needed to get the TEG to 204°C inside the boiler if the circulation is increased. As Fig. 5 suggests, the amount of TEG lost is independent from the TEG circulation rate. The areas where TEG is lost will act the same regardless of the increase in the circulation. Lastly, examining Fig. 6, it shows that it exhibits the same trend and justification seen in Fig. 5, as the number of absorbed hydrocarbons will remain constant; thus, the same amount of gas will be required to strip it.

TEG Feed Temperature

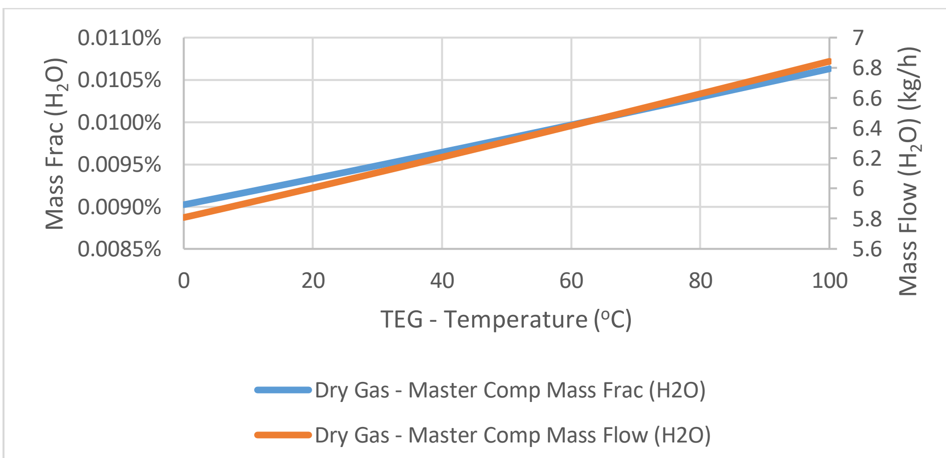

Figure 7: Dry gas water content vs. TEG feed temperature

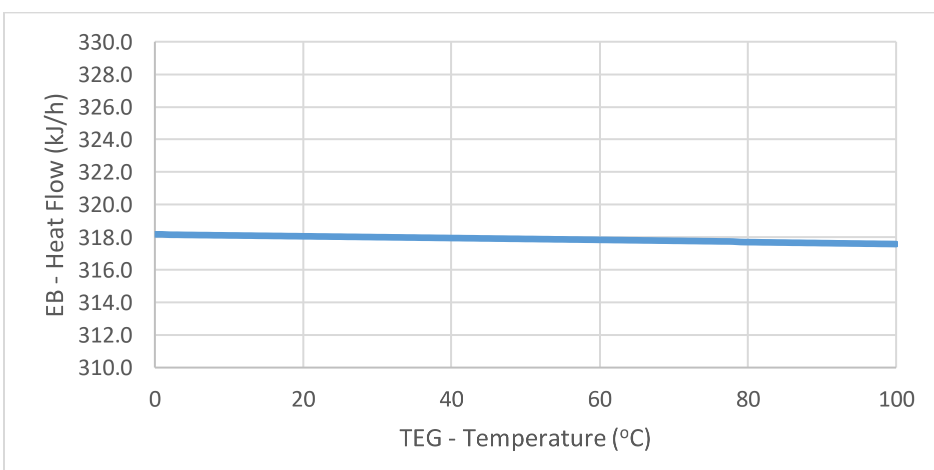

Figure 8: TEG boiler duty vs. TEG feed temperature

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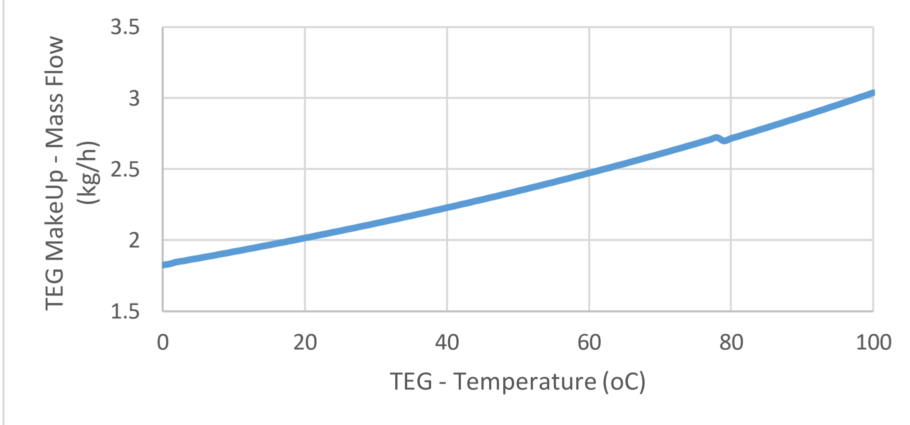

Figure 9: TEG loss vs. TEG feed temperature

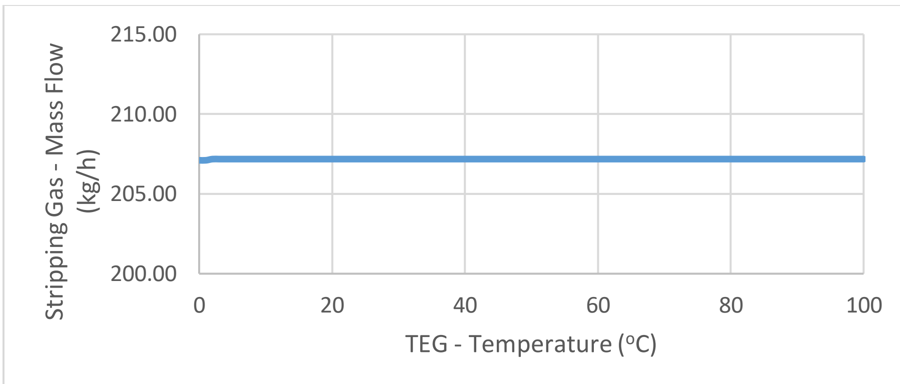

Figure 10: Stripping gas amount required for 99.75% purity of TEG vs. TEG feed temperature

This part concerns the effects of the TEG feed temperature on the absorber on the selected process outputs, as is demonstrated in Figures 7, 8, 9, and 10. It should be noted that this is meant only to demonstrate the effect of this variable, regardless of the company-specified constraints that were listed at the beginning of this paper.

As Fig. 7 suggests, the water content in the dry gas increases with the increase of temperature; this is expected as absorption favours low-temperature operations, thus making it more difficult to absorb the water from the natural gas at higher temperatures. (Ahmed Hasen Mohammed, 2016). Fig. 8 shows that the TEG feed temperature does not have a significant effect on the boiler duty, with only a slight decrease being noticed towards the end of the tested range, which is due to the boiler requiring less energy to heat the mixture up to 204°C due to it being at a higher temperature when it enters the boiler. Fig. 9 shows that the TEG lost in the process increases at higher temperatures; this is due to the higher temperature making the TEG more prone to escape to and remain in the vapor phase when flashed, which happens twice in the process, as well as making the carryover of TEG in the absorber and still columns easier. Lastly, as Fig. 10 suggests, the TEG feed temperature has no effect on the stripping gas amount to reach the minimum purity; this, along with what Fig. 6 shows, leads to the conclusion that the TEG feed conditions do not affect the stripping gas amount required to reach the minimum purity.

Wet Gas Flow Rate

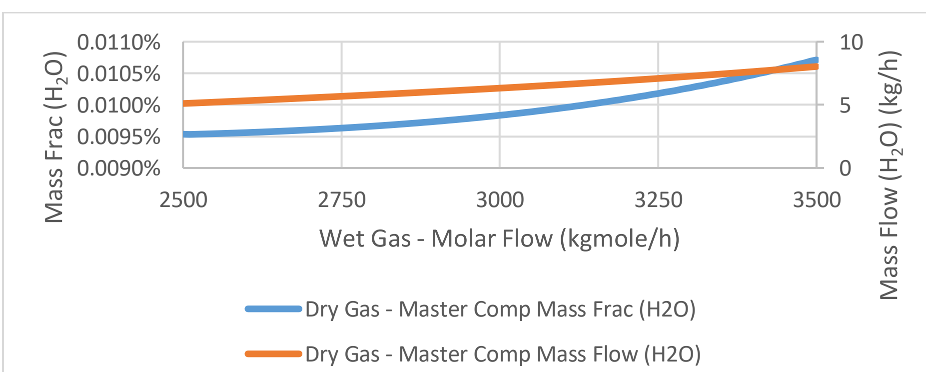

Figure 11: Dry gas water content vs. wet gas flow rate

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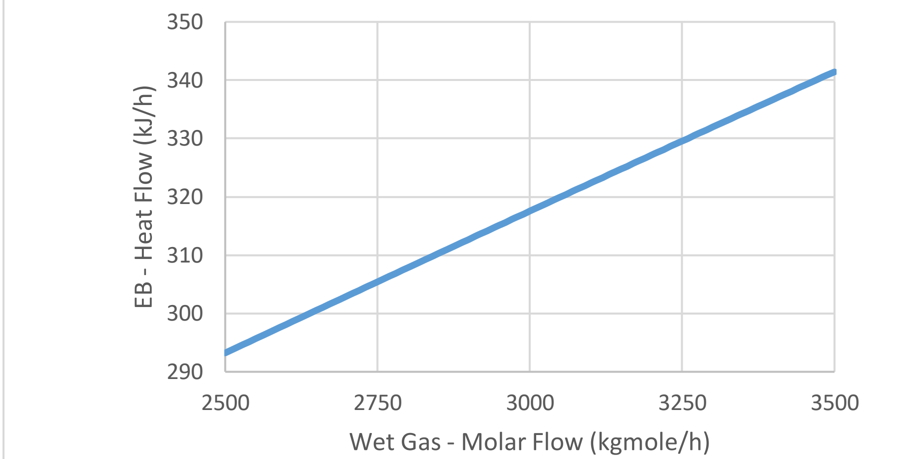

Figure 12: TEG boiler duty vs. wet gas flow rate

This part concerns the effects of the wet gas molar flow when it is introduced to the absorber on the selected process outputs, as is demonstrated in figures 11, 12, 13, and 14.

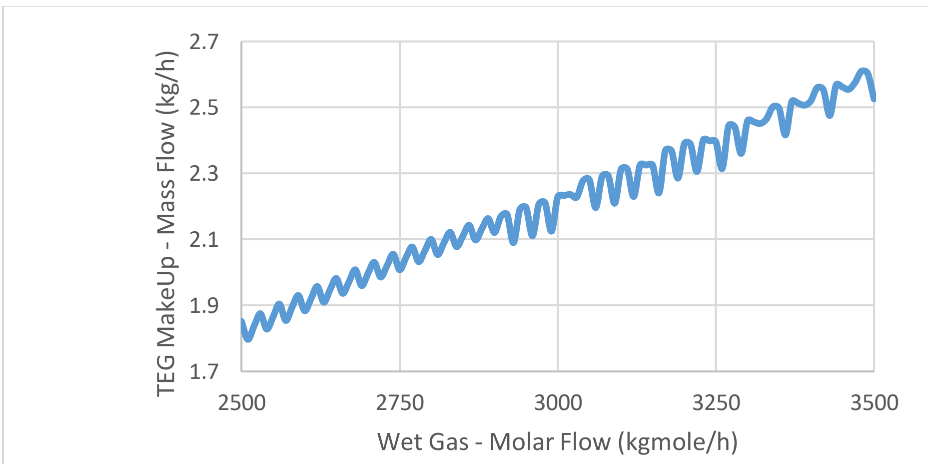

Figure 13: TEG loss vs Wet Gas flowrate

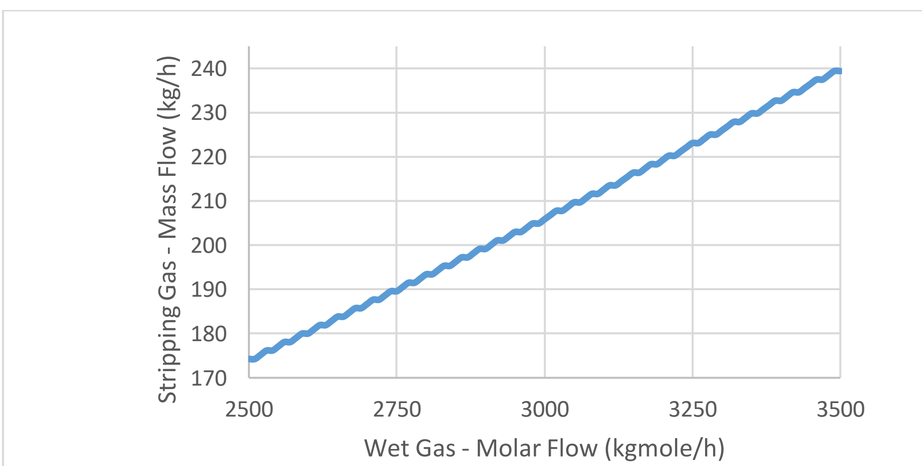

Figure 14: Stripping gas amount required for 99.75% purity of TEG vs. wet gas flow rate

Figure 11 shows the expected increasing trend in the water content of the dry gas because a higher flow rate of wet gas will naturally have a higher amount of water in it, which will subsequently need more TEG to be absorbed, which in this study was kept as a constant. Furthermore, this can be a sign of improper contact between the two fluids, caused by the increased velocity of the wet gas feed rate, which will lead to less efficient absorption. Figure 12 shows the drastic increase in the boiler duty due to the increased flow rate of the wet gas; this relates to the increased amount of water that increases the mass inside the boiler at any one time, which will render the boiler needing more heat to reach the specified 204°C. The results in Fig. 13 resemble the conclusion

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taken from Fig. 9 with regards to the TEG carry-over due to the increased velocity of the wet gas, which carries the TEG as it falls through the packing in the absorber column. Lastly, the trend in Fig. 14 has the same explanation that was discussed for Fig. 12, as more water and impurities in the mixture will require more stripping gas to be stripped out of the mixture.

Wet Gas Feed Temperature

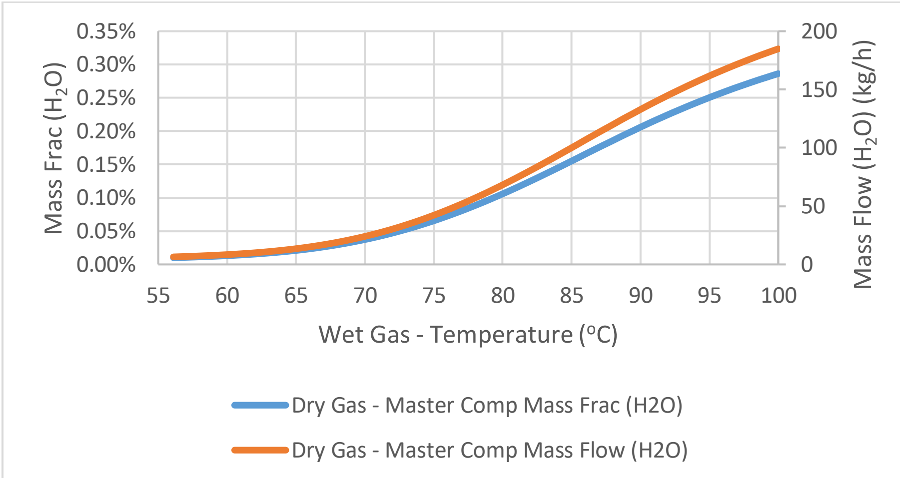

Figure 15: Dry gas water content vs. wet gas feed temperature

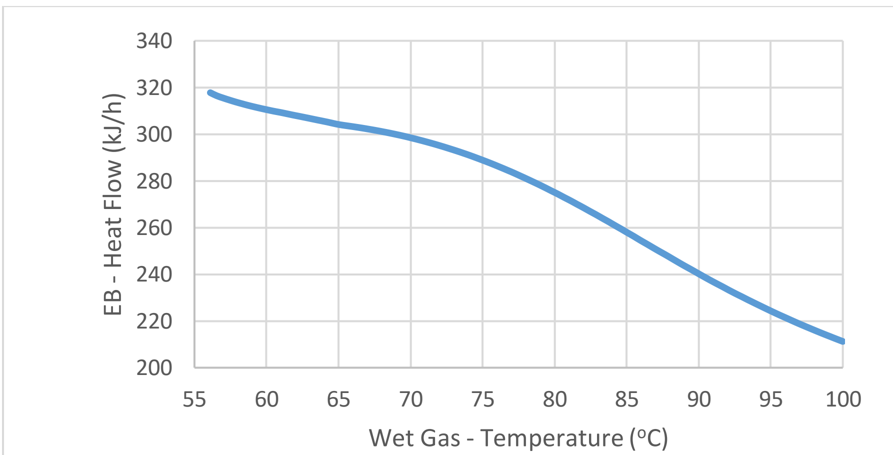

Figure 16: TEG boiler duty vs. wet gas feed temperature

This part concerns the effects of the wet gas temperature at the inlet of the absorber on the selected process outputs, as is demonstrated in figures 15, 16, 17, and 18. As it was discussed for the TEG temperature study, the acquired data only serves to demonstrate the effect of this variable and not to be within the set constraints.

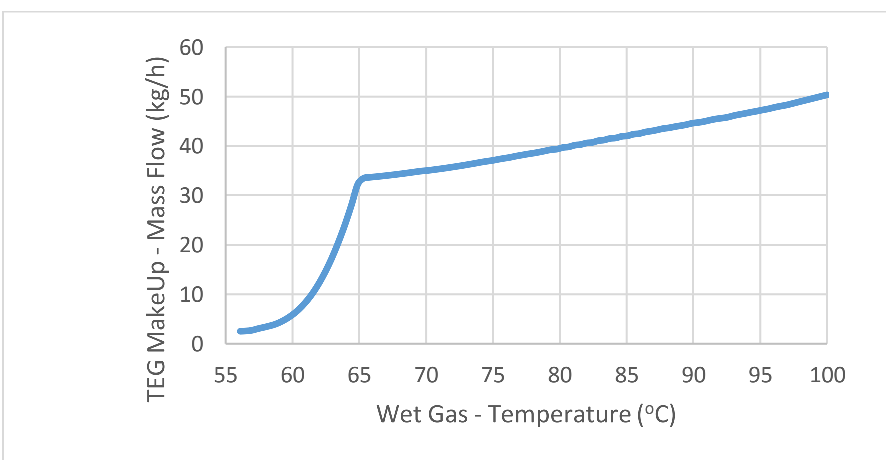

Figure 17: TEG loss vs. wet gas feed temperature

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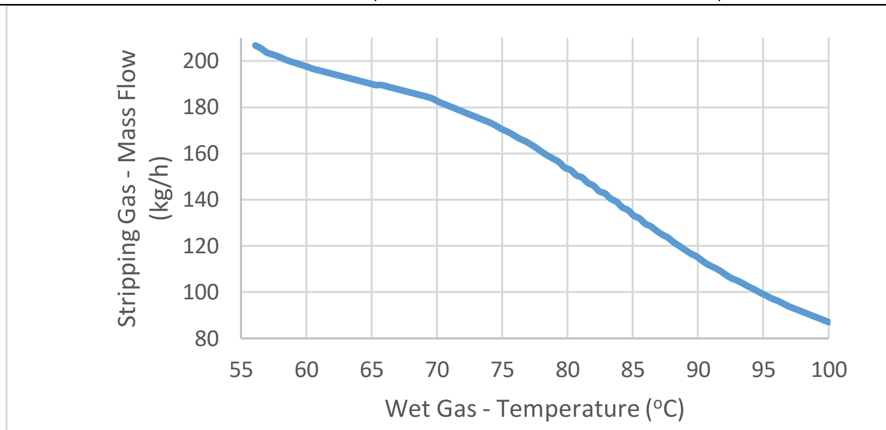

Figure 18: Stripping gas amount required for 99.75% purity of TEG vs. wet gas feed temperature

As Fig. 15 implies, the effect of increasing the wet gas temperature on the water content is the same effect seen in the increase of the TEG temperature witnessed in Fig. 7, and the reason for both is the same, as higher temperatures render water easily vaporizable, thus making it harder for TEG to absorb it. Similarly, Fig. 16 demonstrates the same trend shown in Fig. 8 but to a larger extent, as the boiler duty drastically decreases with the increase in the wet gas feed temperature; this is due to the same reason discussed earlier for Fig. 8. As Fig. 17 shows, the increase in the wet gas temperature causes a massive loss in TEG; this will mainly affect the losses due to carry-over in the absorber and losses in the flash tank, as the boiler will always remain at the same temperature, and thus the losses there shall not be affected significantly. Lastly, Fig. 18 shows how massively the wet gas temperature affects the stripping gas amount required to achieve the minimum purity for TEG, as it sharply decreases with the increase in the temperature. This was due to the TEG not being able to absorb water and impurities due to the higher temperatures, thus making it easier to reach this purity with a lower amount of stripping gas.

Flash Tank Operation Pressure

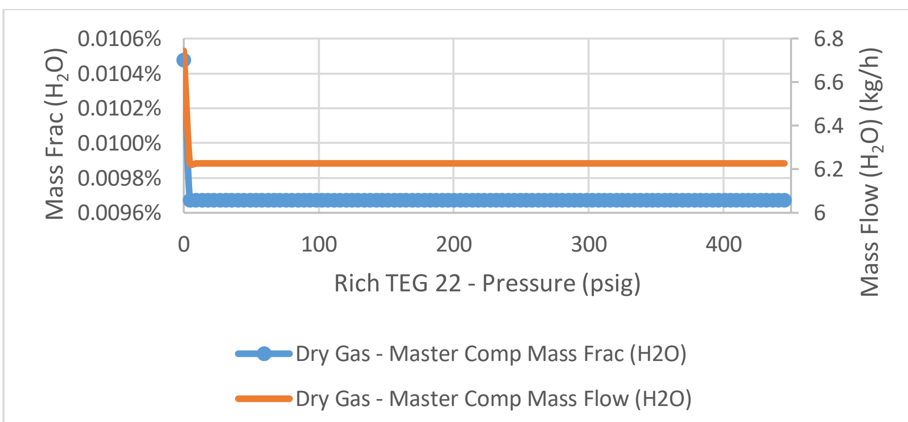

Figure 19: Dry gas water content vs. the flash tank operation pressure

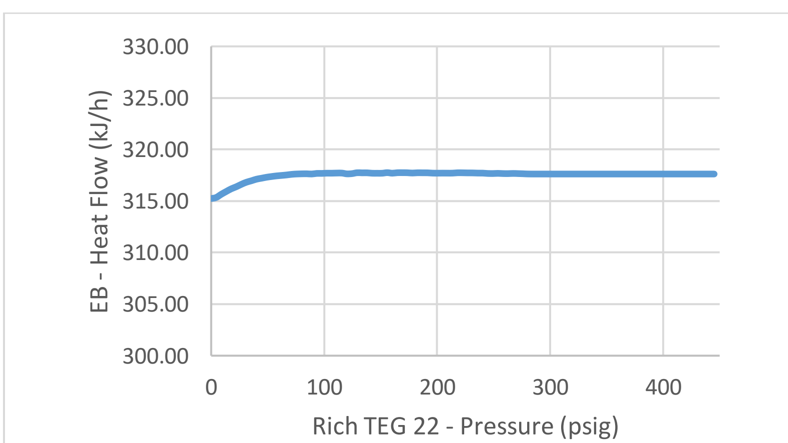

Figure 20: TEG boiler duty vs. the flash tank operation pressure

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This part concerns the effects of the operation pressure of the TEG flash tank on the selected process outputs, as is demonstrated in figures 19, 20, 21, and 22. This study does not consider the mechanical design of the equipment and only shows the effects of this variable.

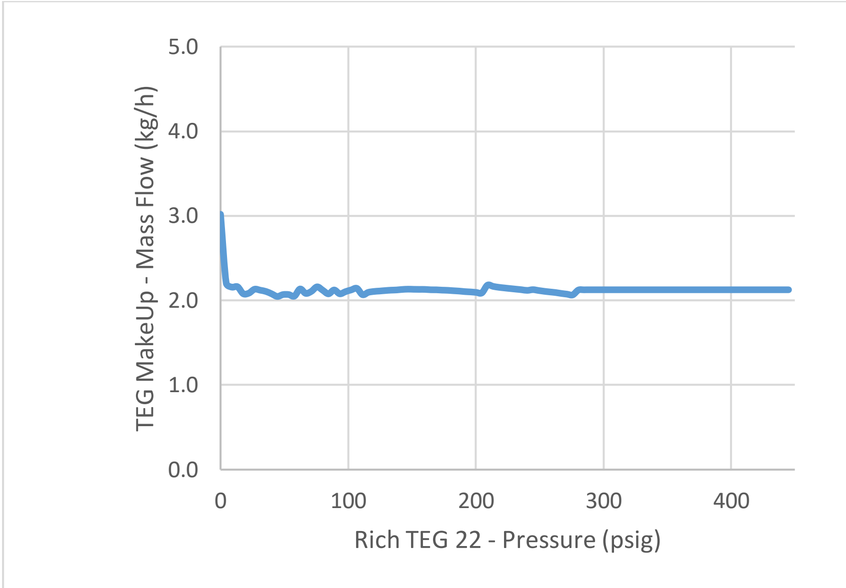

Figure 21: TEG loss vs the flash tank operation pressure

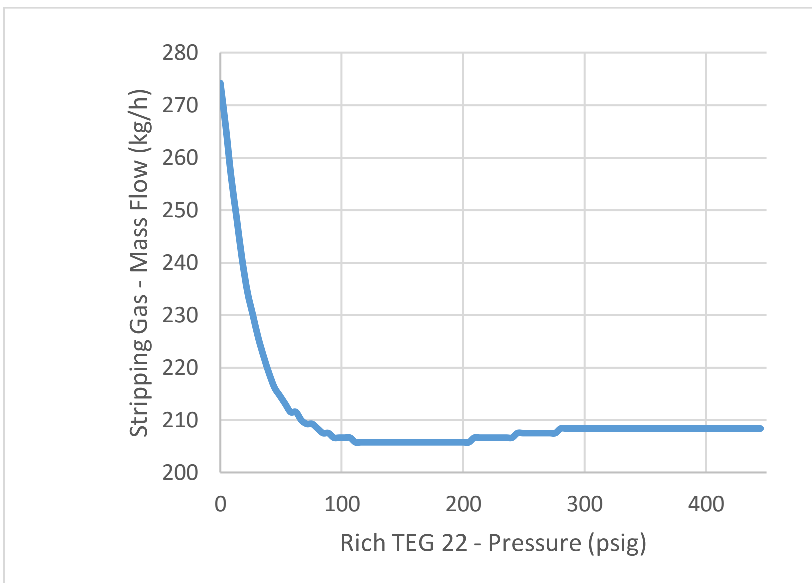

Figure 22: Stripping gas amount required for 99.75% purity of TEG vs. the flash tank operation pressure

Before beginning to discuss the results of this study, a note must be made about the observed straight line after the 280 psig mark; this is due to the fact that from this pressure onwards there will be no flashing as the pressure is not low enough. This also serves to justify the sharp pressure drop seen in the natural gas plant's flash tank, which was around 350 psig. Firstly, Fig. 19 shows that the flash tank operation pressure has no effect on the water content of the dry gas; this is expected as the flash tank comes after the absorber column and will therefore have no effect on it. Regarding the first spike in the graph, it was deemed to be a computational error, which is prone to happen when dealing with computer software, especially since HYSYS uses a solver that relies on convergence, which is a sensitive parameter. Secondly, the boiler duty is not affected significantly by the flash tank as suggested by Fig. 20, and the decrease seen in the beginning can be caused by the mixture having a higher concentration of TEG since more water and impurities are lost in the flash tank and therefore will require less power to reach 204°C. Thirdly, regarding the TEG loss due to the pressure change, Fig. 21 shows that there is no major effect from the pressure change. Lastly, Fig. 22 shows that more stripping gas is required to achieve the minimum purity of TEG when operating the flash tank at a low pressure. This is due to the same reason discussed in the boiler duty part, as the mixture will be more TEG-rich, which is viscous, and will require more stripping gas to scrub the remaining impurities out.

ISSN 2278-2540 | DOI: 10.51583/IJLTEMAS | Volume XIV, Issue I, January 2025

Stripping Gas Feed Temperature

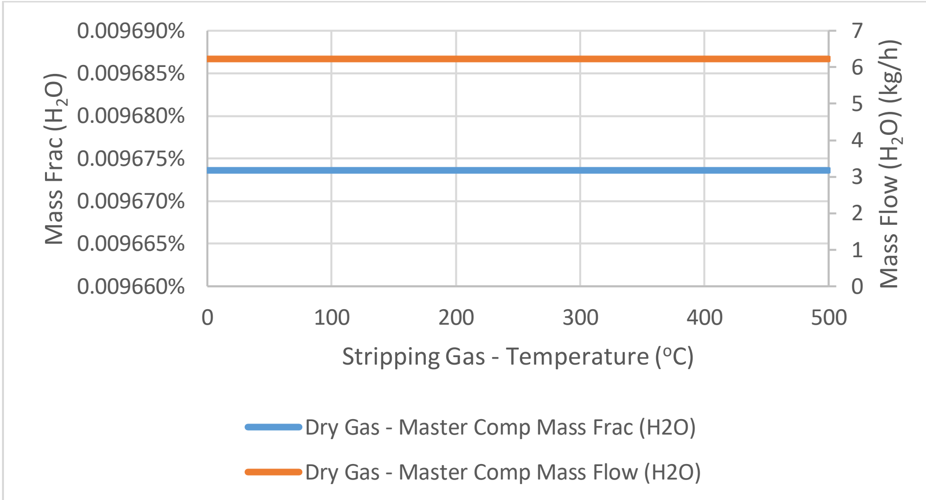

Figure 23: Dry gas water content vs stripping gas temperature

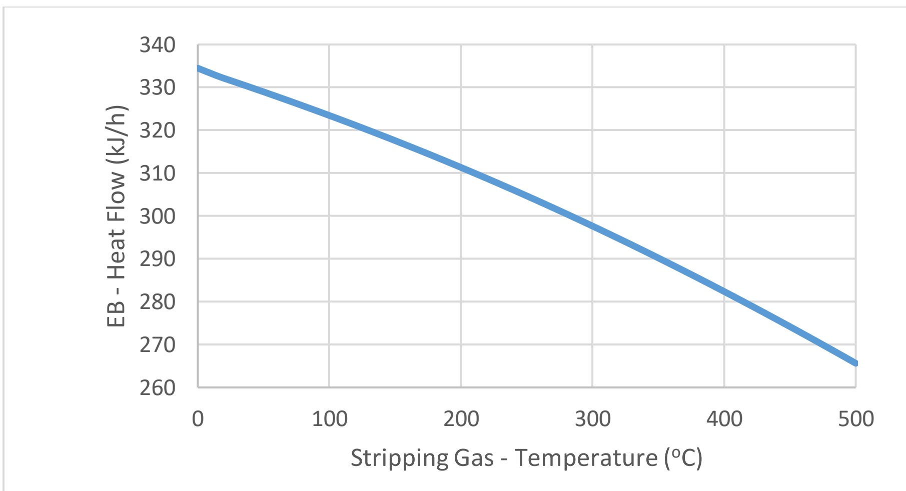

Figure 24: TEG boiler duty vs. the stripping gas temperature

This part concerns the effects of the stripping gas temperature when it enters the reboiler on the selected process outputs, as is demonstrated in figures 23, 24, 25, and 26. This study does not consider the physical constraints of the vessel, nor the utility required to heat the stripping gas to these temperatures.

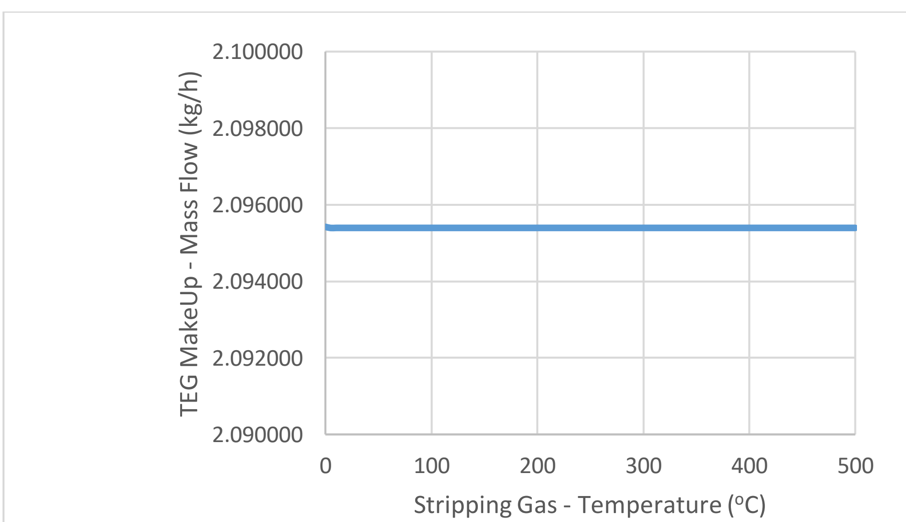

Figure 25: TEG loss vs stripping gas temperature

ISSN 2278-2540 | DOI: 10.51583/IJLTEMAS | Volume XIV, Issue I, January 2025

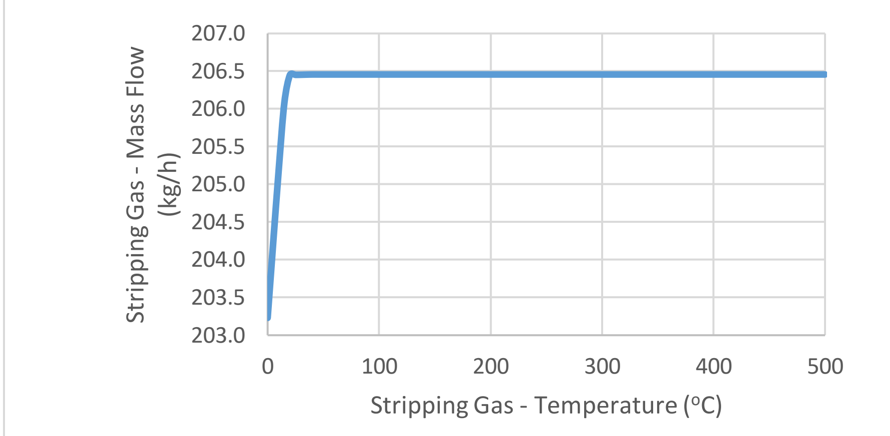

Figure 26: Stripping gas amount required for 99.75% purity of TEG vs. stripping gas temperature

As Fig. 23 suggests, the stripping gas temperature does not influence the water content in the dry gas; this is due to the stripping gas being introduced at a later stage to the absorber, therefore not affecting it. Furthermore, Fig. 24 shows that the boiler duty is heavily influenced by the stripping gas temperature; this is because when the stripping gas is introduced at a higher temperature, it will help the boiler in raising the mixture's temperature to 204°C while using less power. Fig. 25 shows that the TEG loss is not affected by the stripping gas temperature; this is expected as the TEG will not be heated to a temperature past the set 204°C, thus the loss in the boiler will stay the same throughout. Lastly, Fig. 26 shows that there is no effect on the stripping gas amount from the stripping gas temperature, and the dip seen at the beginning can be dismissed as a computational error. This is expected as the temperature gradient does not affect the actual stripping process.

Stripping Gas Mass Flow

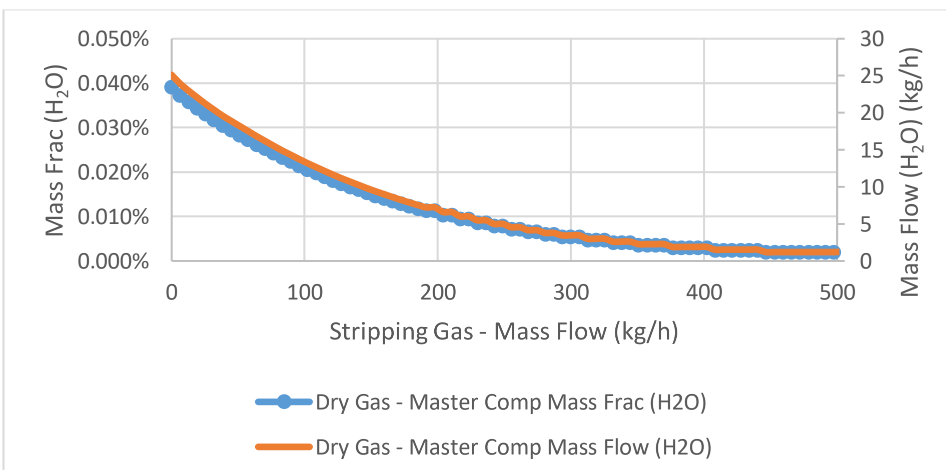

Figure 27: Dry Gas water content vs. stripping gas mass flow

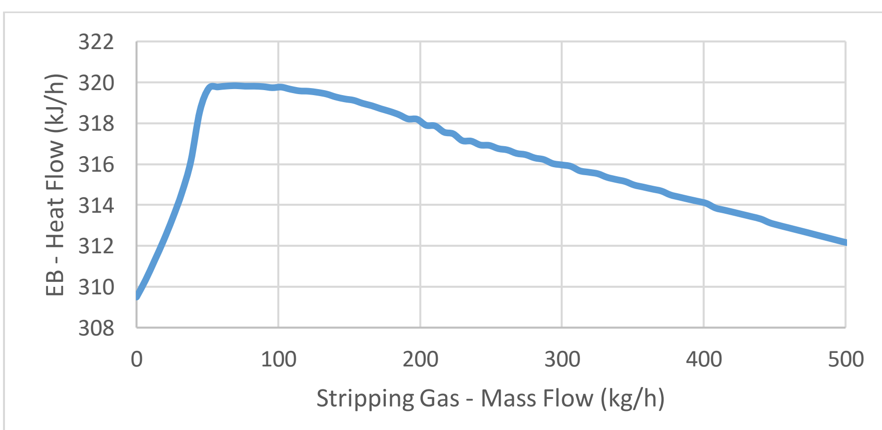

Figure 28: TEG boiler duty vs the stripping gas mass flow

ISSN 2278-2540 | DOI: 10.51583/IJLTEMAS | Volume XIV, Issue I, January 2025

This part concerns the effects of the stripping gas flow rate when it enters the reboiler on the selected process outputs, as is demonstrated in figures 27, 28, 29, and 30. This study differs from the studies shown previously, as the TEG purity is a point of interest due to it not being fixed at the minimum required value; thus, this study, though being designated to the stripping gas mass flow, will simultaneously show the effects of the TEG purity in the circulation.

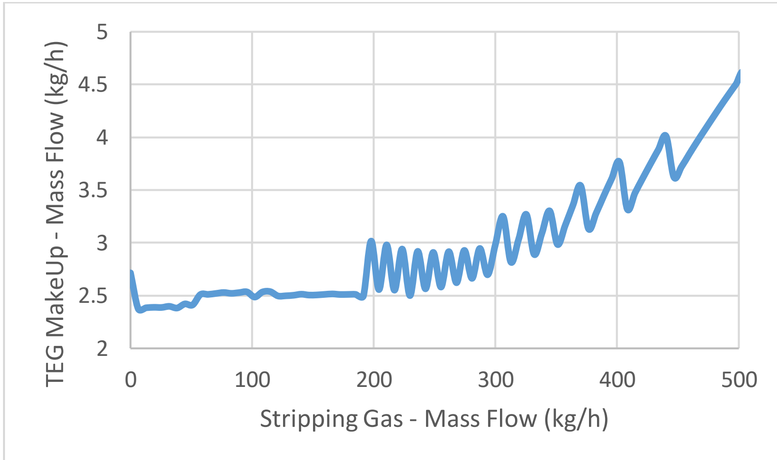

Figure 29: TEG loss vs. stripping gas mass flow

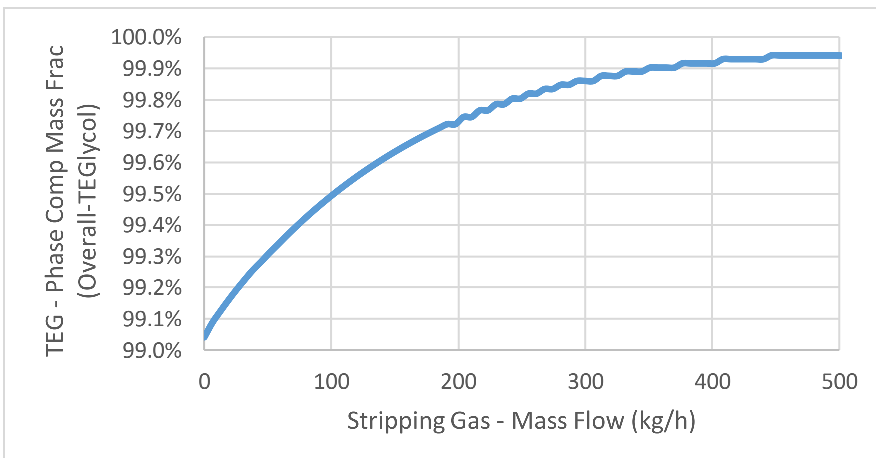

Figure 30: TEG purity vs. stripping gas mass flow

As Fig. 27 demonstrates, the water content in the dry gas is extremely reliant on the stripping gas mass flow; this is because the increase in the stripping gas flow increases the TEG purity, thus allowing for more absorption of water in the absorber column. The trend demonstrated in Fig. 28 shows a sharp increase till a peak followed by a smooth decrease. This is caused by the boiler needing more heat to heat the mixture and the stripping gas to 204°C for the first half and then not needing as much heat due to the stripping gas removing the water and impurities that are in the mixture and this leaving less material for the boiler to heat. Figure 29 shows the result of the carry-over that will occur inside the boiler and still column due to the increase in the stripping gas flow rate. Although the overall loss is not major, the trend still serves as an indication of the trade-off between the TEG's purity and loss. Lastly, Fig. 30 demonstrates the outcome that was discussed in this part thoroughly, which is that the TEG purity increases with the increase of the stripping gas amount. This is expected naturally, as the stripping gas's main duty is to increase the purity of the TEG circulation.

IV. Case Studies Summary

To help put the results of the case studies into perspective and to aid in comparing the different results, Table 3 was constructed to summarize the case study results and indicate which trends were preferred and which were not, thus establishing a general recommendation for operation conditions.

Table 3: Case study recommendations

Direct	Inverse	Direct	Direct	None	None	Inverse
None	Direct	Inverse	Direct	None	Inverse	Direct then inverse
Direct	None	Direct	Direct	None	None	Direct

ISSN 2278-2540 | DOI: 10.51583/IJLTEMAS | Volume XIV, Issue I, January 2025

None	None	Inverse	Direct	Inverse	None	-
-	-	-	-	-	-	Direct

As Table 3 suggests, it is not recommended to operate at high TEG temperatures, due to absorption preferring lower temperatures, while it is suggested to operate at higher TEG flows, as it will give better absorption at the cost of a slightly increased boiler duty. Regarding the wet gas conditions, it is recommended to operate at lower temperatures and flow rates because the increase in the water content and the TEG loss does not justify the decrease in the boiler duty and stripping gas flow, which are cheaper by comparison. Regarding the flash tank pressure, it is recommended to operate at a medium pressure, resulting in adequate flashing and stripping gas flow, especially since the other factors will not be affected significantly. Lastly, regarding the stripping gas, it is recommended to operate at higher temperatures and flow rates, thus decreasing the boiler duty and increasing the TEG purity and, by proxy, the quality of the product.

Optimization

Another goal of this study was to suggest optimization paths to minimize losses and maximize profits. The optimization was performed by using Aspen HYSYS to record the data from 359,640 optimization cases, with around 17.26 million data points. The optimization cases were constructed by finding every possible sensible combination between the operation variables while remaining within the company-set constraints and increasing the stripping gas flow incrementally for the 36,090 cases, which will lead to the entire range of the possible sensible variations to be covered. To find the optimum case, the WNM was utilized to maximize the TEG purity and the profit, which encompasses the value of the product minus the value of the TEG makeup, stripping gas, and boiler duty, and minimize the water content in the dry gas and the stripping gas spent. Furthermore, the HYSYS spreadsheet in the optimization tab was used to carry out the same optimization algorithm, as the HYSYS optimizer does not natively perform multi-objective optimization; thus, using this method within HYSYS will serve as a verification of the optimization paths presented. The results acquired from the HYSYS spreadsheet aligned with those acquired through MS Excel, thus validating the optimization path suggested and the system that was used to calculate it. Table 4 shows the suggested optimization paths along with their results compared to the base case.

Table 4: Optimization alternatives summary

Variable	Unit	Base case	HYSYS Optimizer	Alternative 1	Alternative 2	Alternative 3	Alternative 4
TEG T	°C	65.56	66	66	66	66	66
TEG F	kg/h	4536	5000	5000	5000	5000	5000
WG T	°C	56.11	56	56	56	56	56
WG F	kmol/h	3004.02	3500	3500	3500	2700	3100
Flash P	psig	100.1	142.5	114	170	285	285
SG T	°C	148.9	250	500	150	150	150
SG F	kg/h	206.5	638	638	255.2	574.2	574.2
Water Content	%wt	0.01	0.0013	0.00092	0.0092	0.00014	0.00036
TEG Purity	%wt	99.75	99.96	99.97	99.773	99.9716	99.9716
H_2S emission	ppmv	0.9543	0.8357	0.8359	0.923	0.8891	0.867
CO_2 emission	ppmv	58.659	70.912	70.932	58.66	74.091	71.776
Total Cost	$/h	22.878	63.133	61.6	27.8	58.3	57.4
Total Profit	$/h	5600.21	6487.85	6489.36	6523.74	4994.64	5744.53
1 time payment	$	-	464	464	464	464	464

As is clear from Table 4, alternative 2 is the most desirable among the suggested alternatives, evident by it having the highest profit per hour, which surpasses that of the base case by $923, with no drawbacks to speak of in the TEG purity, the dry gas water content, or the environmental impact, as well as being closer to the base case in most of the operating conditions, which means that not much effort will be needed to switch to this alternative. This makes it superior to all the alternatives presented in the table, thus making it the path that this study recommends. Regarding the HYSYS optimizer result, it clearly mimics the results of alternative 1 in the inputs as well as the outputs, as there are little to no differences between the two. This serves to confirm the validity of the optimization methods used, as the two approaches converged at the same values, which is to be expected as the same weights were used in the WNM for both approaches. Alternative 3 gives the least water content, but due to its profit being lower than that of the base case, it is not recommended, especially since the decrease in the water content does not justify the decreased profit; thus, this alternative can only be recommended for processes that require extreme drying. The results show that no environmental constraint is broken, thus making any of these alternatives valid for use by the company, especially as all of them are profitable apart from alternative 3. Furthermore, the optimization results overall show a clear outline of where the optimum state will be, as most of the operation conditions are the same for all the alternatives, which further validates the results. Lastly, all the alternatives were tested in the dynamics mode to test for stability in realistic applications where there is accumulation, and all the alternatives showed excellent stability as well as satisfactory set point tracking and disturbance rejection.

V. Conclusions

To conclude, the aims of this study, which were to optimize a GDU to minimize TEG loss and perform case studies, were successfully achieved. The study analyzed the prominent variables that affected the process outputs and recommended that the process should be run at lower inlet temperatures for the TEG and natural gas feed to the absorber, higher flow rates of TEG compared to lower flow rates of natural gas, and higher stripping gas temperature with enough flow rate to achieve the minimum required purity of 99.75% wt. or higher if wanted. The optimization findings show that these guidelines converged on where the optimums were, as the optimization results did not veer away from the case study conclusions.

Regarding the optimization, it was done by collecting the data of 359,640 HYSYS cases at different conditions to cover the entire spectrum of possible reasonable operating conditions, then applying the WNM in MS Excel to locate the best case that maximizes the profit while minimizing the losses. To confirm the optimization path acquired through applying the WNM in MS Excel, the algorithm was applied directly in the HYSYS spreadsheet to check the optimization path suggested by the HYSYS solver and optimizer, and it displayed generally the same operating condition acquired previously, thus verifying the system. The recommended optimization path gave an increase in profit on $923, without any major changes to the base case, thus making it extremely favourable. The study also recommended an alternative with less profit than the base case but exhibits extremely high levels of drying with a water content of only 0.00014% wt., which might interest other processes.

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