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Performance Evaluation of Natural Gas Dehydration Plant for Improved Water Removing Efficiency Using Aspen HYSYS

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Abstract
This study presents the simulation and performance evaluation of a natural gas dehydration system using Tri ethylene glycol (TEG) in Aspen HYSYS. The system was modelled with the dehydrator as an absorber and the regenerator as a distillation column. The objective was to enhance water removal efficiency, TEG purity, and energy performance while minimizing solvent losses and operating costs. Initial simulation inputs include a natural gas flow rate of 10,000 Nm³/hr, an inlet gas temperature of 55 °C, pressure of 60 bar, and TEG circulation rate of 10,000 kg/hr with a lean TEG concentration of 94 wt%. The dehydration performance before optimum yield of 85% water removal efficiency, TEG purity of 94%, and a reboiler duty of 180 kW. The optimum operating parameters were obtained for water removal efficiency improved from 85% to 98%, while TEG purity increased from 94% to 98.5%. The TEG circulation rate was increased from 10,000kg/hr to 12,500 kg/hr, and the inlet gas temperature was reduced to 40 °C from 55°C. Additionally, absorber stages were increased from 3 to 5, and operating pressure was reduced to 50 bar from 60 bar, improving gas-liquid interaction. For the regenerator, reboiler temperature was raised to 204 °C from 195°C, the reboiler duty was reduced to 155 kW from 180 kW. The reflux ratio attains the optimum value of 3.5 from 2.5, and the column stages increased to 10 from 8. These changes also reduced TEG losses from 25 to 10 kg/hr. The results demonstrate that there was optimum significantly improve system efficiency, solvent recovery, and energy savings. This study provides a validated process framework for improving industrial gas dehydration units, contributing to cost-effective and environmentally sustainable operations.
Keywords: *Natural Gas Dehydration, Triethylene Glycol (TEG), Aspen HYSYS Simulation, Process Optimization, Water Removal Efficiency, TEG Regeneration, Energy Efficiency, Gas Processing.*

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I. INTRODUCTION

Natural gas has emerged as one of the most essential and efficient fossil fuels in the contemporary energy landscape, offering significant advantages over coal and oil (Mokhatab et al., 2019). As global energy demands continue to rise, natural gas plays a critical role in meeting these needs while providing a cleaner alternative to other fossil fuels. Natural gas is primarily composed of methane and other light hydrocarbons, but when extracted from underground reservoirs, it typically contains various impurities, including water vapour, hydrogen sulfide, carbon dioxide, and nitrogen (Abdulrahman & Sebastine, 2013).

Among these impurities, water vapour presents particularly significant challenges in natural gas processing and transportation. The presence of water in natural gas streams can lead to numerous technical problems, including hydrate formation, pipeline corrosion, reduced heating value, and inefficient flow conditions (Abdulrahman & Sebastine, 2013). Hydrate ice-like crystalline compounds formed when water molecules combine with hydrocarbon molecules under specific pressure and temperature condition scan block pipelines and equipment, potentially causing catastrophic failures and operational hazards (Al Shehhi et al., 2019).

Dehydration, therefore, is a crucial process in natural gas treatment facilities. It involves removing water vapor from natural gas to meet specific water content specifications, typically less than 0.1 ppm, to prevent the aforementioned problems (Abdulrahman & Sebastine, 2013). Several methods are employed for natural gas dehydration, including absorption using glycol solvents, adsorption using solid desiccants, condensation by manipulating dew point temperatures, and newer techniques such as supersonic and membrane separation (Al Shehhi et al., 2019).

Among these methods, glycol absorption has become the most widely implemented technique in the gas industry due to its operational effectiveness and economic efficiency (Abdulrahman & Sebastine, 2013). The process utilizes liquid glycol desiccants, commonly triethylene glycol (TEG), diethylene glycol (DEG), or monoethylene glycol (MEG), which have a high affinity for water vapor. The glycol selectively absorbs water from the natural gas stream in a contactor tower, after which the water-rich glycol is regenerated in a reboiler system by heating to separate the water, allowing the lean glycol to be recycled back into the process (Al Shehhi et al., 2019).

While the basic principles of glycol dehydration are well-established, optimizing these systems remains an ongoing challenge for gas processing facilities. Process simulation tools like Aspen HYSYS have become invaluable for analyzing, designing, and optimizing such complex systems (Al Shehhi et al., 2019). These tools allow engineers to evaluate the effects of various operating parameters, such as glycol type, circulation rate, contactor pressure, reboiler temperature, and stripping gas rate, on system performance and efficiency without expensive and time-consuming physical experiments.

Previous studies have shown that careful selection of glycol type and optimization of process parameters can significantly improve dehydration efficiency while minimizing glycol losses and energy consumption. For instance, Abdulrahman and Sebastine (2013) demonstrated that while TEG achieved excellent water removal (down to 0.1 ppm) at circulation rates of 4000 kg/hr, it also led to higher hydrocarbon losses from the glycol regenerator compared to DEG. Similarly, Al Shehhi et al. (2019) found that DEG exhibited substantially lower losses (6.1 liters per day at 120°C regenerator temperature) compared to MEG (34 liters per day under the same conditions).

Furthermore, the configuration of the gas conditioning units can significantly impact not only water removal efficiency but also natural gas liquids (NGL) recovery. Al Shehhi et al. (2019) compared refrigeration-based chiller systems with turbo-expander systems, finding that the latter could produce significantly more NGL (1160 bbl/day vs. 750 bbl/day) under optimized conditions.

Despite these advancements, there remains considerable room for improvement in understanding the complex interactions between process variables and their combined effects on dehydration efficiency, energy consumption, and operational economics. This study, therefore, aims to evaluate and optimize natural gas dehydration-regeneration plant performance using Aspen HYSYS simulation, with particular focus on improving water removal efficiency while minimizing operational costs and environmental impact.

II.EXTENT OF PAST WORKS

Amarfio et al. (2024) conducted an in-depth study on the simulation of natural gas dehydration using TEG in ASPEN HYSYS. Their work centered around the importance of moisture removal from natural gas streams to meet commercial specifications and prevent operational issues such as pipeline corrosion, flow restrictions, and combustion inefficiencies. They highlighted the superiority of TEG over other glycols, namely ethylene glycol, diethylene glycol, and tetra ethylene glycol, owing to its higher regeneration efficiency.

The researchers developed a steady-state simulation model of a TEG dehydration unit designed to process 10 MMSCFD of natural gas at a pressure of 6200 kPa, utilizing the Peng-Robinson equation of state due to its strong VLE prediction capabilities across a wide range of conditions. Their model demonstrated substantial moisture reduction in the gas stream, from 37.98 lb/MMSCF to 2.493 lb/MMSCF, well below the industry limit of 7 lb/MMSCF.

The process flow modeled in ASPEN HYSYS included a gas-liquid separator, a contactor tower for counter-current flow between gas and lean TEG, multiple heat exchangers, and a distillation column-based regeneration section with a reboiler and condenser. The study further included detailed case studies assessing the impact of inlet gas temperature, pressure, and flow rate on dehydration performance. Findings indicated that increased gas temperature enhances moisture capacity, complicating dehydration, whereas increased pressure aids dehydration by promoting condensation. Additionally, the flow rate was shown to influence residence time, with higher flow reducing contact efficiency.

Amarfio et al., 2024 expanded their research to include an economic assessment using the Aspen Process Economic Analyzer. They incorporated real-time pricing data and derived a capital investment of approximately \$7.5 million, an annual operating cost of around \$1.85 million, and a payback period of 11.1 years within a 20-year project lifespan. These results suggest that the dehydration unit could achieve profitability in the medium term, supporting the feasibility of TEG systems from both technical and financial perspectives.

Al-Jammali (2022) explored the simulation of TEG-based natural gas dehydration using ASPEN HYSYS v8.8. His study utilized actual data from the Basra gas dehydration plant to build a validated process model. Key process units in his simulation included an inlet scrubber for impurity removal, a contactor for gas and glycol interaction, a series of heat exchangers, a flash tank for hydrocarbon and water vapor separation, and a six-tray distillation column for glycol regeneration.

Validation of simulation results with plant data yielded an average absolute error of less than 1%, underscoring the reliability of his model. A notable strength of Al-Jammali's work was his comprehensive sensitivity analysis. He investigated the influence of six operating variables, feed gas temperature, pressure, flow rate, TEG temperature, TEG flow rate, and rich glycol temperature, on four key outcomes: water content in dry gas, reboiler heat duty, glycol losses, and dry gas output.

Results from this analysis revealed complex interdependencies. For instance, higher TEG flow rates enhanced dehydration efficiency but at the cost of increased solvent losses and energy consumption. Likewise, raising the wet gas temperature led to elevated moisture content in the gas but decreased heat duty and solvent losses. Rich glycol temperature emerged as a critical parameter capable of significantly lowering reboiler duty. Interestingly, feed pressure had a minimal impact on gas dehydration but moderately increased glycol losses.

The study also incorporated Gandhidasan's empirical equation to determine dew point temperature based on outlet water content and pressure. This added a theoretical foundation that strengthened the credibility of the process analysis. Al-Jammali additionally referenced several foundational works, including those by Ranjbar et al. (2015), Rahimpour et al. (2013), Kamin et al. (2017), and Jacob (2014), all of which explored optimization strategies for TEG systems using various methodologies like relative sensitivity functions and response surface modeling. These earlier studies collectively highlighted operational limits for key parameters such as reboiler temperature and glycol flow rate, warning against issues like thermal degradation and excessive solvent loss.

Operational challenges were also a focal point in Al-Jammali's research. He cautioned against over-circulation and under-circulation of glycol, both of which can adversely affect system efficiency and environmental impact. Additionally, he noted that operating pressures beyond 100 bar introduce complexities requiring costly auxiliary equipment, thus raising both capital and operational expenditures.

In the study by Wosu et al (2024), the researchers focused on the design and optimization of a glycol-based natural gas dehydration plant through a rigorous simulation approach. Using Aspen HYSYS as the primary simulation tool, they developed a detailed model of a TEG dehydration unit operating at a constant contactor pressure. Their work systematically varied the feed gas conditions,including temperature, pressure, and flow rate, and used Microsoft Excel (Solver) coupled with differential calculus to develop and optimize mathematical relationships between the independent variables and the water content of the treated gas. The simulation results highlighted the interplay between these parameters: they found that higher feed temperatures increased the gas's ability to carry moisture, while higher pressures promoted water condensation, thus reducing the water content in the sales gas. The study identified optimum operating conditions that resulted in water contents well within pipeline specifications, confirming that a careful balance of process variables can yield significant improvements in dehydration efficiency. In addition to the process simulation, a detailed economic analysis was performed. This analysis incorporated current market prices, revealing that the optimized configuration not only met technical requirements but also promised an acceptable payback period over the plant's operational lifespan. Overall, the work by Wosu and Ezeh provided a comprehensive evaluation of how simulation and optimization techniques can be integrated to design an economically viable and efficient natural gas dehydration unit.

The study by Al Shehhi et al (2019) also utilized Aspen HYSYS to simulate and optimize a natural gas dehydration process; however, their investigation expanded the focus by considering the replacement of conventional solvents and cooling techniques. Here, the authors explored not only the use of glycols but also evaluated the potential benefits of substituting monoethylene glycol (MEG) with diethylene glycol (DEG) to reduce solvent losses and improve system performance. Their work further compared two distinct refrigeration methods, one using a conventional chiller and the other employing a turbo expander, to control the hydrocarbon dew point and maximize natural gas liquid (NGL) production. Through a series of simulations, they demonstrated that the expander technique could yield a higher production of NGL compared to the chiller, while the comparative analysis of glycol behavior indicated that DEG produced significantly lower losses than MEG at equivalent regenerator temperatures. In addition to quantifying these effects, the study offered an economic evaluation by linking the process simulation with market variables, such as product pricing, to underscore the profitability of the proposed modifications. The findings underscore that by optimizing operating conditions, ranging from inlet pressure adjustments to the selection of refrigerant systems and glycol type, the dehydration process can be tailored to enhance energy efficiency, reduce operational costs, and ultimately drive greater economic returns.

In the study by Kharisma et al. (2020), the authors present a comprehensive approach to designing and simulating a natural gas dehydration unit based on triethylene glycol (TEG). Their work is focused not only on achieving effective dehydration of natural gas by removing water and other contaminants to meet pipeline specifications, but also on minimizing the Total Annual Costs (TAC) associated with the process. Utilizing Aspen Plus as the simulation tool, the authors construct a steady-state model that replicates the performance of an industrial TEG dehydration plant. They base the simulation on real plant data, ensuring that the feed gas composition, operating conditions, and product specifications reflect practical scenarios.

The simulation begins with wet gas entering an absorber where lean TEG is used to remove water vapor from the natural gas stream. The process is further detailed by incorporating subsequent steps such as preheating of rich TEG, regeneration of the glycol through a reboiler and column system, and recycling of lean TEG to the absorber. The design parameters, including column design, tray spacing, and vessel sizing, are meticulously determined by integrating both hydraulic considerations (to avoid weeping and flooding conditions) and economic criteria. Kharisma et al. then use optimization techniques by varying key operating conditions such as the absorber column pressure and lean TEG temperature. The study shows that increasing the absorber pressure tends to reduce the water content in the dehydrated gas, while adjustments to the lean TEG temperature influence both process efficiency and the cooling duty required.

One of the distinguishing aspects of this research is its emphasis on economic evaluation through TAC analysis. By calculating equipment capital costs, operating expenses, and annualized fixed costs using methodologies derived from existing literature, the authors pinpoint an optimal operating region where the TAC is minimized. The simulation results indicate that under the optimized conditions, specifically, an absorber column pressure of 45 barg and a lean TEG temperature of 39°C, the process achieves a significant reduction in TAC, dropping from a base case value of approximately 3.7 million USD per year to an optimum value of around 3.24 million USD per year, while still maintaining acceptable water content in the dehydrated natural gas. Validation of the simulation with plant data shows a reasonably small percentage error, lending credibility to the proposed design and optimization strategy.

Overall, the work by Kharisma et al. (2020) delivers a detailed and pragmatic approach to natural gas dehydration, integrating process simulation with economic optimization. Their analysis underscores how careful adjustment of operating parameters can lead to simultaneous improvements in both technical performance and economic viability, making a strong case for the role of simulation-driven optimization in enhancing the safety, operability, and stability of TEG dehydration units. This study not only complements earlier research on TEG dehydration simulation and optimization but also extends it by explicitly incorporating total annual cost calculations into the design process, thus providing a robust framework for decision-making in industrial natural gas processing.

The study by Al Shehhi et al (2019) investigated the simulation and optimization of a natural gas dehydration process using Aspen Hysys v8.6, focusing on enhancing Natural Gas Liquid (NGL) production and operational efficiency. The authors evaluated two primary modifications: replacing monoethylene glycol (MEG) with diethylene glycol (DEG) as the dehydration solvent and substituting a chiller-based refrigeration system with a turbo expander. Additionally, they analyzed the effects of key operational parameters, such as inlet pressure, glycol regenerator temperature, and Low Temperature Separator (LTS) conditions, on process performance and economic viability.

The methodology involved process simulation using Aspen Hysys, integrated with Excel for parametric analysis. Key variables, including inlet pressure (35–98 bar), regenerator temperature (100–130°C for MEG; 111–156°C for DEG), and LTS temperature (−20°C to 0°C), were systematically adjusted to assess their impact on NGL yield, glycol losses, and hydrocarbon separation efficiency. Economic evaluations compared equipment costs (e.g., chiller vs. expander) and operational savings, assuming a crude oil price of $65 per barrel (May 2019).

Results demonstrated that the turbo expander outperformed the chiller, achieving an NGL production of 1,160 bbl/d at 82 bar inlet pressure compared to 750 bbl/d at 51 bar for the chiller. This yielded an annual profit increase of $9.7 million. Replacing MEG with DEG reduced glycol losses significantly: at 120°C regenerator temperature, DEG losses were 6.1 L/day versus MEG's 34.5 L/day, reducing operational costs. Lowering LTS temperatures increased NGL recovery (e.g., −20°C produced 1,132 bbl/d) but introduced challenges, such as higher light hydrocarbon content (1.46 m^3/h of methane/ethane) in condensate streams, necessitating reprocessing. Feed-to-export analysis showed effective removal of heavy hydrocarbons (95% reduction in C7+ components) and water (from 0.0002 to 0.000003 mole fraction), meeting pipeline specifications. Economic analysis highlighted the expander's higher initial cost but greater long-term profitability due to reduced refrigerant dependency and increased NGL output.

In conclusion, the study recommended adopting turbo expanders and DEG to optimize dehydration efficiency, maximize NGL production, and minimize operational losses. These findings provide actionable insights for improving natural gas processing economics and technical performance.

Kamin et al (2017) conducted a simulation-based optimization study on triethylene glycol (TEG) utilization in a natural gas dehydration process, aiming to minimize glycol losses, energy consumption, and water content in processed gas. The research focused on optimizing three critical parameters: lean glycol circulation rate (3,000–5,500 kg/hr), reboiler temperature (180–206°C), and the number of trays in the glycol contactor column (2–5 trays). Using Aspen HYSYS, the authors modeled the TEG dehydration unit of the Farashband gas processing plant (Iran) and validated the simulation against plant data. A central composite design (CCD) via Design Expert 7.0.0 was employed to structure 52 experimental runs, with response surface methodology (RSM) applied to analyze four key responses: glycol loss, reboiler duty, water concentration in dry gas, and hydrate formation temperature.

Results demonstrated that increasing the number of contactor trays improved water removal efficiency, with three trays identified as optimal for reducing water content to 4–7 lb/MMSCF (meeting pipeline specifications). Higher tray counts increased glycol losses marginally (34.8–35.2 kg/hr), while lower reboiler temperatures (180°C) minimized energy consumption without compromising TEG regeneration purity (>98.7 wt%). The lean glycol circulation rate of 3,944 kg/hr balanced efficiency and losses. Hydrate formation temperature decreased with optimized parameters, reducing operational risks. Regression models highlighted the quadratic effects of circulation rate and reboiler temperature on glycol losses and energy use. Compared to non-optimized conditions, the proposed configuration (3 trays, 180°C reboiler, 3,944 kg/hr circulation) enhanced dehydration performance while aligning with prior findings by Rahimpour et al. (2013) on glycol concentration impacts.

The study concluded that RSM effectively identified trade-offs between conflicting parameters (e.g., reduced reboiler duty vs. higher water content) and validated the simulation's predictive accuracy against plant data. These optimizations offer actionable strategies for improving TEG dehydration economics and operational reliability in gas processing plants.

Chidiebere et al (2023) investigated the optimization of a triethylene glycol (TEG)-based natural gas dehydration system using Aspen Hysys, focusing on minimizing water content, TEG losses, and operational costs. The study modeled an industrial gas dehydration unit (operating at 72.6 barg and 38°C) and modified it by integrating a recovery separator and reflux condenser to enhance TEG regeneration efficiency. Simulations tested TEG flow rates ranging from 3.5 to 60 m^3/h, evaluating their impact on water removal, methane recovery, and economic viability.

The methodology involved two configurations: a baseline industrial setup (Simulation 1) and a modified system (Simulation 2) with additional equipment to recover escaped TEG. Key parameters included water content in dry gas (target: ≤7 lb/MMSCF), TEG regeneration efficiency, and capital/operational costs. Results demonstrated that a TEG flow rate of 3.5 m^3/h reduced water content to 6.63 lb/MMSCF (within acceptable limits) but incurred 12% TEG loss. Higher flow rates (e.g., 25 m^3/h) further reduced water content to 0.44 lb/MMSCF and achieved 97.29% methane purity but required excessive TEG usage. The modified system (Simulation 2) achieved near-complete TEG recovery (99.98%) by incorporating a reflux condenser and separator, eliminating losses.

Economic analysis revealed that Simulation 2 had a higher initial capital cost (3.86millionvs.3.86*millionvs*.3.75 million for Simulation 1) due to added equipment. However, Simulation 2's negligible TEG losses reduced annual operating costs, yielding long-term savings. Both configurations consumed equivalent energy (114,300 kJ/h), but Simulation 2's operational efficiency offset its higher upfront investment. The study concluded that optimizing TEG flow rates (e.g., 25 m^3/h for maximal methane recovery) and integrating reflux systems significantly enhance dehydration efficiency and economic sustainability.

III.MATERIALS AND METHOD

2.1 Materials Used

The following materials and resources were utilized in the course of this study:

i. Aspen HYSYS Simulation Software
ii. Engineering Textbooks
iii. Peer-Reviewed Journals
iv. Thermodynamic Property Data Sheets.

2.2 Method Used

- The natural gas dehydration system was simulated and optimized in Aspen HYSYS using mass and energy balance principles, focusing on the absorber (dehydrator) and regenerator (distillation column).
- Appropriate thermodynamic models were applied to accurately represent phase behavior, and a detailed process flow diagram defined all streams and operating conditions.
- A representative stage j was used in the regenerator to develop generalized stage-wise balances and interpret column behavior.
- Key operating variables—TEG circulation rate, reboiler temperature, and reflux ratio—were systematically varied to assess their effects on water removal efficiency, TEG purity, and energy consumption, with results validated against literature data for optimization.

2.2.1 Model Formulation Assumptions

To develop a reliable and manageable design model for the simulation and analysis of the TEG-based natural gas dehydration system, the following assumptions were adopted:

i. Mass Transfer Occurs on Each Tray/Differential packed section
ii. Negligible Heat Loss in the Regenerator: The regenerator column is assumed to be well insulated; hence, external heat losses to the surroundings are considered negligible.
iii. Associated Gases in the Liquid Phase Are Included: Hydrocarbon gases that dissolve or associate with the rich TEG phase are taken into account in the component balance.
iv. Multicomponent Feed Stream: The feed to the regenerator consists of multiple components primarily triethylene glycol, water, and dissolved hydrocarbon gases.

v. Constant Molar Flow Rate Assumption: The variation in total molar flow across the column is assumed to be minimal; therefore, an average total molar flow rate is used for simplification.

vi. Vapor-Liquid Equilibrium (VLE): The vapor-liquid equilibrium is described using Raoults Law, and vapor pressures of individual components are calculated using Antoine Equation.

vii. Column Segmentation for Analysis: For ease of model development and interpretation, the column is conceptually divided into five envelopes typically including rectifying, stripping, feed, and reboiler sections.

viii. Representative Stage Balances: Mass and energy balances are performed on a representative stage within each of the key sections (i.e., rectifying section, stripping section, and feed stage) to reduce computational complexity while maintaining accuracy.

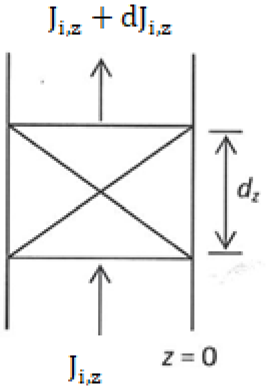

Figure 1: Elemental Packed Volume

General Mass balance equation

$$\begin{pmatrix}\text{Rate of accumulation}\\ \text{of Species/Energy}\\ \text{within the Control volume}\end{pmatrix} = \begin{pmatrix}\text{Rate of input}\\ \text{of Species/Energy}\\ \text{into Control volume}\end{pmatrix} - \begin{pmatrix}\text{Rate of output}\\ \text{of Species/Energy}\\ \text{from Control volume}\end{pmatrix} \pm \begin{pmatrix}\text{Rate of}\\ \text{deplection/generation of Species}\\ \text{by chemical reactor in Control volume}\end{pmatrix} \tag{1}$$

At steady State operation,

$$\begin{pmatrix}\text{Rate of accumulation}\\ \text{of Species}\\ \text{within the Absorber}\end{pmatrix} = 0 \tag{2}$$

Non-reacting system,

$$\begin{pmatrix}\text{Rate of}\\ \text{deplection/generation of Species}\\ \text{by chemical reactor in Absorber}\end{pmatrix} = 0 \tag{3}$$

$$\begin{pmatrix}\text{Rate of input}\\ \text{of Species}\\ \text{into Absorber}\end{pmatrix} = \vec{J}_{i,x}dydz + \vec{J}_{i,y}dxdz + \vec{J}_{i,y}dxdy \tag{4}$$

$$\begin{pmatrix}\text{Rate of output}\\ \text{of Species}\\ \text{from Absorber}\end{pmatrix} = \left(\vec{J}_{i,x} + d\vec{J}_{i,x}\right)dydz + \left(\vec{J}_{i,y} + d\vec{J}_{i,y}\right)dxdz + \left(\vec{J}_{i,z} + d\vec{J}_{i,z}\right)dxdy \tag{5}$$

Substituting the various terms into eqution (3.1), we have

$$-\left(d\vec{J}_{i,x}dydz + d\vec{J}_{i,y}dxdz + d\vec{J}_{i,z}dxdy\right) = 0 \tag{6}$$

Dividing through eqution (3.6) by unit volume of dxdydz

$$-\bar{\nabla} J_i = -\left(\frac{\partial \bar{J}_{i,x}}{\partial x} + \frac{\partial \bar{J}_{i,y}}{\partial y} + \frac{\partial \bar{J}_{i,z}}{\partial z}\right) = 0 \tag{7}$$

where,

$$\bar{J}_i = -D_{AB} \nabla C_i + C_i \bar{v} \tag{8}$$

$$\bar{v} = (v_x, v_y, v_z) \tag{9}$$

$$\bar{J}_{i,x} = -D_{AB} \frac{\partial C_i}{\partial x} + C_i v_x \tag{10}$$

$$\bar{J}_{i,y} = -D_{AB} \frac{\partial C_i}{\partial y} + C_i v_y \tag{11}$$

$$\bar{J}_{i,z} = -D_{AB} \frac{\partial C_i}{\partial z} + C_i v_z \tag{12}$$

Substituting equation (10), equation (11) and equation (12) into equation (7), gives

$$-\frac{\partial}{\partial x}\left(-D_{AB} \frac{\partial C_i}{\partial x} + C_i v_x\right) - \frac{\partial}{\partial y}\left(-D_{AB} \frac{\partial C_i}{\partial y} + C_i v_y\right) - \frac{\partial}{\partial y}\left(-D_{AB} \frac{\partial C_i}{\partial z} + C_i v_z\right) = 0 \tag{13}$$

Equation (13) is the conservative molar concentration equation simplified as

$$D_{AB} \frac{\partial^2 C_i}{\partial x^2} + D_{AB} \frac{\partial^2 C_i}{\partial y^2} + D_{AB} \frac{\partial^2 C_i}{\partial z^2} - \nabla\left(C_i \bar{v}\right) = 0 \tag{14}$$

Mass transfer in the z-direction is more pronounced than the other two directions, as such unidirection is assumed for a packed bed

$$D_{AB} \frac{d^2 C_i}{dz^2} - \frac{d}{dz}\left(C_i v_z\right) = 0 \tag{15}$$

$$D_{AB} \frac{d^2 C_i}{dz^2} - v_z \frac{dC_i}{dz} - C_i \frac{dv_z}{dz} = 0 \tag{16}$$

$$\frac{dv_z}{dz} = 0$$

but, at constant flow

$$D_{AB} \frac{d^2 C_i}{dz^2} - v_z \frac{dC_i}{dz} = 0 \tag{17}$$

let , $\varphi = {}^{v_z}\!/_{D_{AB}}$

$$\frac{d^2 C_i}{dz^2} - \varphi \frac{dC_i}{dz} = 0 \tag{18}$$

Interim of species i mole fraction

$$C_i = y_i . C \tag{19}$$

Combining equation (18) and equation (19), yield

$$\frac{d^2 y_i}{dz^2} - \varphi \frac{dy_i}{dz} = 0 \tag{20}$$

let $\alpha = \dfrac{dy_i}{dz}$ (21)

$$\frac{d\alpha}{dz}-\varphi\alpha=0 \tag{22}$$

Rearranging equation (22)

$$\frac{d\alpha}{\alpha}=\varphi dz \tag{23}$$

Integrating equation (23)

$$\alpha = C_2\exp\quad(\varphi z) \tag{24}$$

Substituting equation (21) into equation (24)

$$\frac{dy_i}{dz}=C_2\exp\quad(\varphi z) \tag{25}$$

Integrating equation (25)

$$y_i=\frac{C_2}{\varphi}\exp\quad(\varphi z)+C_3 \tag{26}$$

let, $A = \frac{C_2}{\varphi}$, $B = C_3$

$$y_i(z)=A\exp\quad(\varphi z)+B \tag{27}$$

Boundary conditions

At z=0; $y_i(z)=y_{i0}$

At z=H; $y_i(z)=y_{iH}$

Applying boundary condition to equation (27)

$$y_i(z)=\left(\frac{y_{iH}-y_{i0}}{\exp\left(\frac{v_zH}{D_{AB}}\right)-1}\right)\exp\quad\left(\frac{v_zz}{D_{AB}}\right)+\left(y_{io}-\frac{y_{iH}-y_{i0}}{\exp\left(\frac{v_zH}{D_{AB}}\right)-1}\right) \tag{28}$$

let $v_z=\vartheta$

Simplifying equation (28)

$$y_i(z)=y_{io}+\left(y_{iH}-y_{i0}\right)\left(\frac{\exp\quad\left(\frac{\vartheta z}{D_{AB}}\right)-1}{\exp\left(\frac{\vartheta H}{D_{AB}}\right)-1}\right) \tag{29}$$

Where Péclet number

$$Pe = \frac{\vartheta H}{D_{AB}} \tag{30}$$

Combining equation (29) and equation (30), we have

$$y_i(z)=y_{io}+\left(y_{iH}-y_{i0}\right)\left(\frac{\exp\quad\left(Pe\frac{z}{H}\right)-1}{\exp\ (Pe)-1}\right) \tag{31}$$

Using order-of-magnitude analysis based on the Peclet number in equation (3.31)

$Pe \ll 1$: Diffusion-dominated

$Pe \gg 1$: Convection-dominated

The mass transfer process in absorption is predominately controlled by interphase mass transfer (Diffusion)

Applying Taylor series expansion for small Pe,

$$\exp\quad(Pe) \approx 1 + Pe + \frac{Pe^2}{2!}+\cdots$$

Apply to numerator and denominator of Equation (3.30)

$$\exp\quad\left(Pe\frac{z}{H}\right)-1\approx Pe\frac{z}{H}$$

$\exp(Pe)-1 \approx Pe$

Therefore

$$y_i(z) = y_{io} + (y_{iH} - y_{io})\left(\frac{Pe\frac{z}{H}}{Pe}\right) \quad (32)$$

2.2.3 Material Balance over Tray j

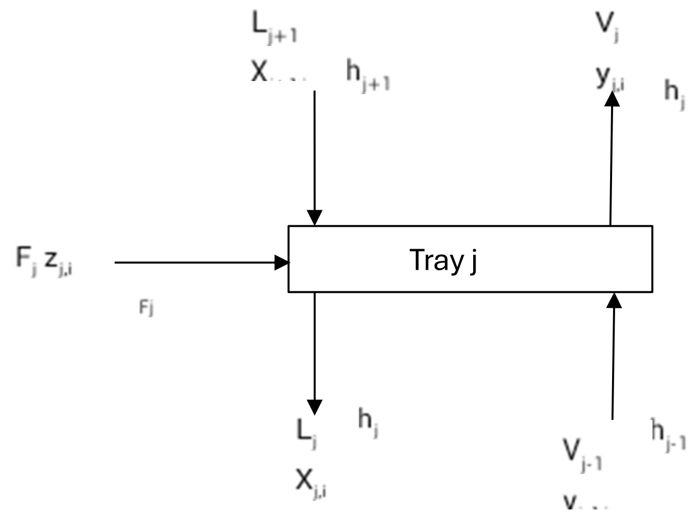

Figure 2: Representative Tray j

Material balance over tray n, from equation (1)

Rate of accumulation of material = $\frac{dM_j}{dt} = 0$

Rate of input of material = $F_j + L_{j+1} + V_{j-1} = F + L + V$

Rate of output of material = $L_j + V_j = L + V$

2.2.3.1 Overall material balance

$$F_j + L_{j+1} + V_{j-1} - L_j - V_j = 0 \quad (33)$$

Taking component balance for any species i, from equation (3.32), we have

$$FZ_{j,i} + Lx_{j+1,i} + Vy_{j-1,i} - Lx_{j,i} - Vy_{j,i} = 0 \quad (34)$$

Phase Equilibrium Relationships

$$y_{j,i} = K_{j,i}x_{j,i} \qquad (1 \le j \le Nt,\ 1 \le i \le n) \quad (35)$$

Summations of Mole Fractions

$$\sum_{i=1}^{n} x_{j,i} = 1,\ \sum_{i=1}^{n} y_{j,i} = 1 \qquad (1 \le j \le Nt,\ 1 \le i \le n) \quad (36)$$

Combining equation (34) and equation (35), we have

$$FZ_{j,i} + Lx_{j+1,i} + VK_{j-1,i}x_{j-1,i} - Lx_{j,i} - VK_{j,i}x_{j,i} = 0 \quad (37)$$

hence $K_n = K_{n+1} = K_{n-1}$ (38)

$y_n = Kx_n$ (39)

Applying equation (38) to equation (37), yield

$$FZ_{j,i} + Lx_{j+1,i} + VKx_{j-1,i} - Lx_{j,i} - VKx_{j,i} = \quad (40)$$

Applying equation (40) to all the five envelop, we have

2.2.3.2 Rectification Section

At the rectifying section of column, there is no feed flow rate, $F = 0$

$$VKx_{j-1,i}-Lx_{j,i}-Dx_D=0 \quad (41)$$

2.2.3.3 Stripping Section

At the stripping section of column, there is no feed flow rate $F = 0$

$$Lx_{j+1,i}-VKx_{j,i}-Bx_B=0 \quad (42)$$

2.2.3.4 Feed Tray

$$F_jZ_{j,i}+L_{j+1}x_{j+1,i}+V_{j-1}Kx_{j-1,i}-L_jx_{j,i}-V_jKx_{j,i}=0 \quad (43)$$

2.2.3.5 Reflux Drum

$$VKx_{Nt,i}-D(1+R)x_{D,i}=0 \quad (44)$$

2.2.3.6 Column Base

$$Lx_{1,i}-Bx_{B,i}-V_SKx_{B,i}=0 \quad (45)$$

2.2.4 Energy Balance over Tray j

Energy balance over tray n, from equation (1)

Rate of accumulation of heat = $\frac{dM_jh_j}{dt}=0$

Rate of inflow of heat = $Fh_{Fj}+Lh_{j+1}+Vh_{j-1}$

Rate of outflow of heat = $L_jh_j+V_jh_j$

$$Fh_{Fj}+Lh_{j+1}+Vh_{j-1}-Lh_j-Vh_j=0 \quad (46)$$

for any stage say k

$$h_k=C_{P_k}(T_k-T_{ref})=C_P(T_k-T_{ref}) \quad (47)$$

Substituting equation (22) into equation (21), we have

$$FC_{P_f}(T_{fj}-T_{ref})+LC_{P_{j+1}}(T_{j+1}-T_{ref})+VC_{P_{j-1}}(T_{j-1}-T_{ref})-LC_{P_j}(T_j-T_{ref})-VC_{P_j}(T_j-T_{ref})=0$$

(48)

Where:

L= Liquid rate
F= Feed rate
H= Enthalpy
Cp =Specific heat capacity
T= Temperature
D=Distillate rate
Qc= Condenser load
V=Vapour rate
R=Reflux ratio
B=Bottom rate
Q_{reb}=Reboiler load
F = Feed rate
C_{PF} = Feed heat capacity
T_f = Feed temperature
T_{ref} = Feed reference temperature
L = Liquid rate
C_{PJ+1} = Heat capacity of the liquid on the tray j+1
T_{j+1} = Temperature of the liquid on the tray j+1
T_{ref} = Reference temperature of the liquid
V = Vapour rate
C_{pv} = Heat capacity of the vapour.
T_{J-1} = Temperature of the vapour before the feed tray
T_j = Temperature of the vapour on the feed tray

For negligible change in temperature on each stage, the reference temperature

$T_{ref}=0$

Hence equation (3.47) becomes

$$FC_{P_{fj}}T_{fj}+LC_{P_{j+1}}T_{j+1}+VC_{P_{j-1}}T_{j-1}-LC_{P_j}T_j-VC_{P_j}T_j=0 \tag{49}$$

Applying equation (3.48) to all the six envelop, we have

2.2.4.1 Rectification Section

At the rectifying section of column, there is no feed flow rate $F = 0$

$$VC_{P_{j-1}}T_{j-1}-LC_{P_j}T_j-DC_{P_D}T_D=0 \tag{50}$$

2.2.4.2 Stripping Section

At the stripping section of column, there is no feed flow rate, $F = 0$

$$LC_{P_{j+1}}T_{j+1}-VC_{P_j}T_j-BC_{P_B}T_B=0 \tag{51}$$

2.2.4.3 Feed Tray

$$FC_{P_{fj}}T_{fj}+LC_{P_{j+1}}T_{j+1}+VC_{P_{j-1}}T_{j-1}-LC_{P_j}T_j-VC_{P_j}T_j=0 \tag{52}$$

2.2.4.4 Reflux Drum

$$VT_{Nt}-D(1+R)C_{P_D}T_D-Q_{COND}=0 \tag{52}$$

2.2.4.5 Column Base

$$LC_{P_1}T_1-BC_{P_B}T_B-V_SC_{P_B}T_B+Q_{REB}=0 \tag{54}$$

2.2.5 Aspen HYSYS Simulation

The process simulation for the natural gas dehydration system was carried out using Aspen HYSYS, a dynamic process modelling and optimization software widely used in the chemical and oil & gas industries. The simulation encompassed two primary units: the Dehydrator (Absorber) and the Regenerator (Distillation Column).

Simulation Setup:

Fluid Package: The Peng-Robinson Equation of State was selected due to its suitability for hydrocarbon systems involving non-ideal behaviour.

Component Selection: Natural gas was modelled with methane, ethane, propane, and water vapor, while the TEG solvent was specified as Tri ethylene glycol.

Operating Conditions:

Gas Inlet Temperature: 55°C

Gas Inlet Pressure: 60 bar

TEG Circulation Rate: Initially 10,000 kg/hr,

Lean TEG Concentration: 94 wt%

Column Configuration:

Absorber: Trayed column with 3 stages

Regenerator: Distillation column with 8 stages, reboiler and condenser specified

Optimization Targets:

Maximize water removal efficiency from the gas stream

Enhance TEG regeneration by increasing purity

Reduce reboiler duty and TEG losses

Maintain pipeline-spec water content in dry gas (≤7 lb/MMscf)

Simulation Output:

Key performance indicators such as water content in dry gas, TEG purity, reboiler duty, and energy consumption were recorded and compared before and after optimization. The model provided clear insights into how process conditions affect separation efficiency and energy usage

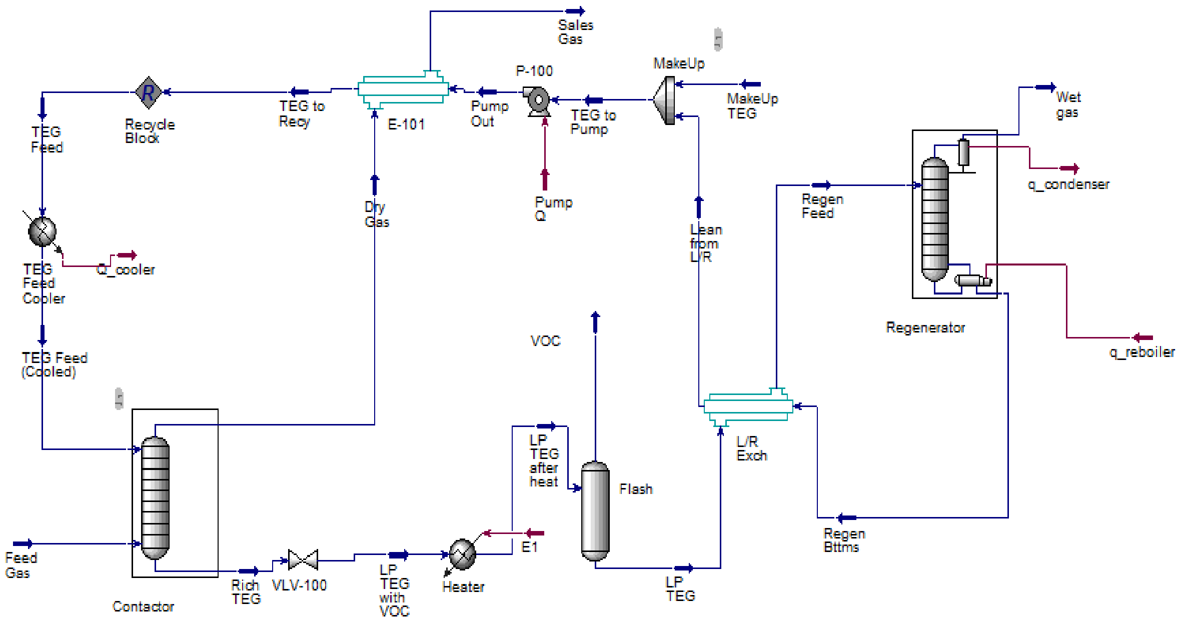

Figure 3: Natural Gas Dehydration and Regeneration Unit

IV.RESULTS AND DISCUSSION

3.1 DATA COLLECTION

Data w ere obtained from the glycol live plant as shown in table 4.1. The TEG circulation rate was 10000kg/hr with a concentration of 94%. The contactor (dehydrator) operates at a temperature of 55°C and a pressure of 60 bar with efficiency of 85%. Similarly, the distillation column recorded its parameters as reboiler duty of 180kw and column temperature of 195^{0}C. The reflux ratio was 2.5 operating with eight (8) theoretical equilibrium stages and condenser duty of 60kw with TEG purity of 94%.
In the contactor, the natural gas water content mass fraction and methane content were respectively 0.2wt% and 0.5wt%. These values were the bases for Aspen hysys Simulation.

3.2 ASPEN HYSYS SIMULATION

The process variables were used to simulate the Tri-ethylene glycol unit of the natural gas process plant. During the simulation, a flash vessel was introduced for the dissolved gases to flash off from the rich TEG solution flowing to the distillation column. The distillation (regenerator) operates at a low pressure of about 10 psia for effective regeneration. The improvement of its efficiency requires the regenerator to operate at a partial vacuum condition of about 550mmHg to 500mmHg as proven by maurice *et al*. As 100^{0}F and the regenerator top temperature kept at 110^{0}F for an outlet dew point of 10^{0}F, a TEG purity of 99.0% was obtained. Practically, a recycling mixer is not needed in the plant for a fresh glycol make up, however during the simulation, a mixer was required to raise the makeup glycol temperature to that of the rich glycol from the contactor for the contactor column to converge. The required temperature was 55^{0}C for TEG rich solution
The process of optimization involved the inputting of the simulated values in the Aspen hysys optimizer. These values are specified as 'Before optimization' in tables 4.1 and 4.2. The corresponding output values of the optimization process are also stated as 'After optimization' in the same tables.

3.3 Analysis and Evaluation of the Results of Aspen Hysys Optimization.

In Table 1, the optimization of the absorber significantly improved dehydration performance. Increasing the TEG circulation rate and the number of theoretical stages enhanced water absorption efficiency. Additionally, boosting TEG concentration and lowering inlet gas temperature improved the solvents capacity to remove water. Operating at a slightly reduced pressure also contributed to better gas-liquid contact. As a result, water removal efficiency increased from 85% to 98%, and TEG loss was reduced by more than half. These changes not only ensured pipeline-spec gas quality but also reduced operational losses, aligning with performance goals and economic sustainability.

Table 1: Optimization of the Dehydrator Unit

Parameter	Before Optimization	After Optimization
TEG Circulation Rate (kg/hr)	10,000	12,500
Number of Theoretical Stages	3	5
TEG Concentration (wt %)	94	98.5
Inlet Gas Temperature (°C)	55	40
Operating Pressure (bar)	60	50
Water Removal Efficiency (%)	85	98
TEG Loss Rate (kg/hr)	25	10

Table 2: Optimization of the Regenerator Unit (Distillation Column)

Parameter	Before Optimization	After Optimization
Reboiler Temperature (°C)	195	204
Reboiler Duty (kW)	180	155
Reflux Ratio	2.5	3.5
Number of Theoretical Stages	8	10
TEG Purity (%)	94	98.5
Condenser Duty (kW)	60	80
TEG Loss in Overhead (kg/hr)	15	5

In Table 1, the regenerator optimization focused on energy efficiency and solvent purity. Raising the reboiler temperature to 204°C enhanced water stripping, while increasing the reflux ratio improved internal separation. These changes, combined with additional stages and higher condenser duty, elevated TEG purity from 94% to 98.5%. Simultaneously, reboiler duty was reduced, reflecting improved thermal integration. TEG losses in overhead vapors dropped significantly, lowering chemical makeup costs. Collectively, these adjustments boosted regeneration effectiveness and solvent recovery, reducing waste and energy consumption, while maintaining high-quality TEG for continuous use in the absorber.

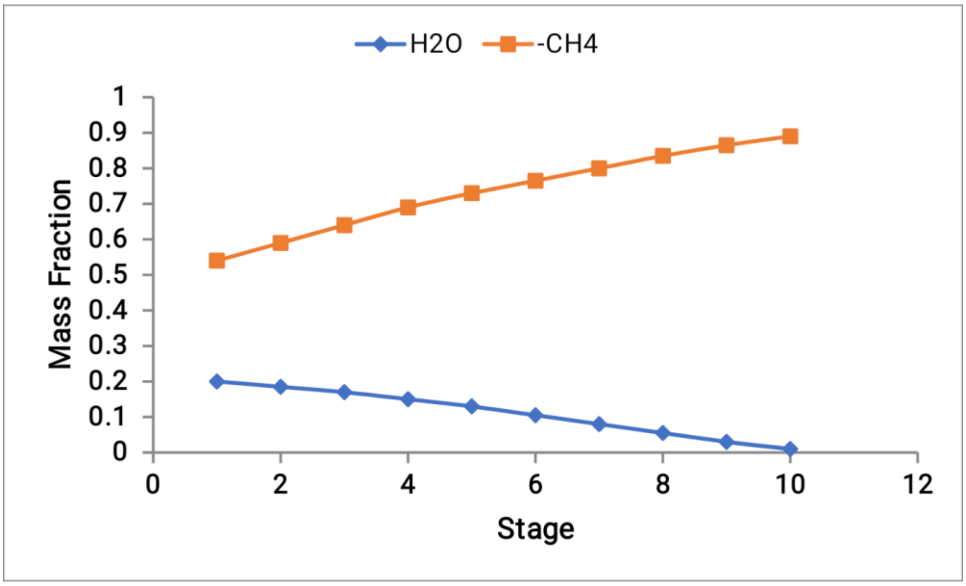

Figure 3: Mass Fractions of Methane and Water vs Stage

In Figure 1**,** the stage-wise variation of mass fractions for methane and water in the absorber provides critical insight into the mass transfer dynamics occurring in the natural gas dehydration process. In this configuration, wet natural gas composed of methane and water vapor enters from the bottom of the column, while lean Tri ethylene

glycol (TEG) is introduced at the top. As gas flows upward and TEG descends, water vapor is absorbed by the glycol, decreasing water content progressively up the column while methane concentration increases.
Initially, at the lower stages, water vapor concentration is relatively high due to fresh gas input, while methane composition is slightly lower. As the gas ascends and contacts lean TEG, water is absorbed from the gas stream, decreasing its mass fraction. Consequently, methane's mass fraction increases up the stages due to reduced dilution by water vapor. This counter-current contact allows the TEG to capture most of the water by the top stage, with the dry gas exiting at maximum methane purity. This behavior aligns with findings by Carroll (2016), who reported that stage-wise analysis is crucial for optimizing absorber performance, especially in tailoring TEG flow rates for enhanced dehydration. The observed trends also validate the model by Mokhatab et al. (2014), which emphasized the importance of mass transfer efficiency across contactor in achieving low water content in treated gas streams.

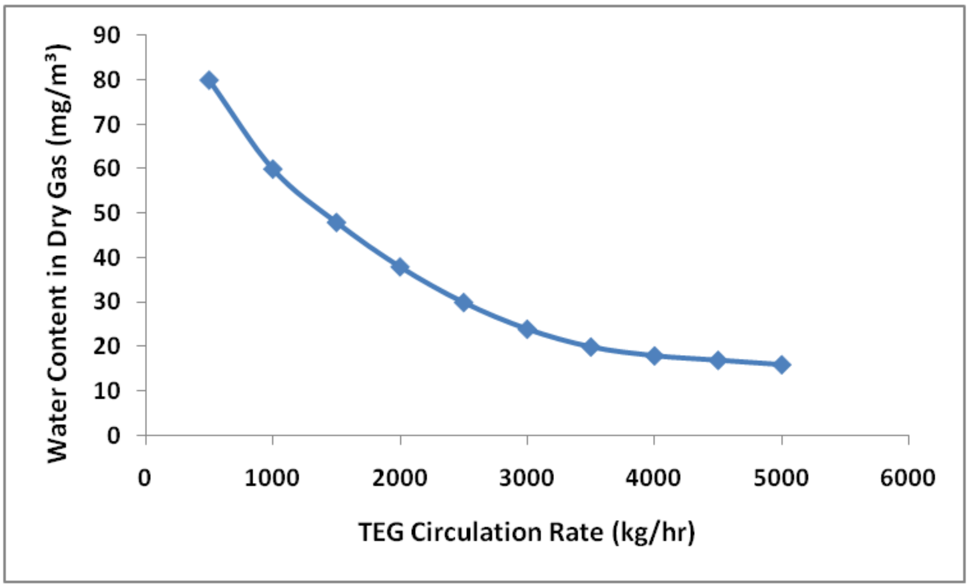

Figure 4: Water Content in Dry Gas vs TEG Circulation Rate (kg/hr)

IV. CONCLUSION

This study presents a comprehensive simulation, performance evaluation, and optimization of a natural gas dehydration system using triethylene glycol (TEG), focusing on two major units: the Dehydrator (Absorber) and the Regenerator (Distillation Column). The simulation was conducted under standard industrial conditions, and several key operating parameters were analyzed to determine their effect on system performance, energy efficiency, and product quality. In the dehydrator unit, increasing the TEG circulation rate, optimizing absorber pressure, and improving the lean TEG concentration significantly enhanced the water removal efficiency—from 85% to 98%. Reducing the inlet gas temperature and increasing the number of theoretical stages also improved mass transfer between TEG and wet gas, resulting in drier gas that meets pipeline specifications.
The study provides the following contributions to knowledge:
This research significantly advances the body of knowledge in the field of natural gas processing by providing a detailed simulation-based approach to optimizing Triethylene glycol (TEG) dehydration systems. By integrating process modeling of both the absorber and regenerator units, this study identifies key operational parameters—such as TEG circulation rate, reboiler duty, and reflux ratio—that directly influence dehydration efficiency, TEG purity, and energy consumption. The work bridges the gap between theoretical design and real-world process optimization by establishing quantifiable improvements in water removal efficiency and solvent recovery.

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