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Novel Plan for Stripping Gas Recovery in Gas Dehydration Unit: Techno-Economical Evaluation for Khark Petrochemical Company as A Case Study

Arman M. Abdehvand[1], Mahdi Haghi[1*], Aminreza Ansari[2], Ebrahim Hoseinirad[1]

[1]Research and Development department, Khark Petrochemical Company, Khark Island, Bushehr, Iran.

[2]Department of Chemical Engineering, Iran University of Science and Technology, Tehran, Iran.

[*]Corresponding Author E-mail: haghi_1321@yahoo.com

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Abstract

The presence of water vapor in natural gas presents several challenges, including corrosion, pipeline blockages, and reduced pipeline capacity. Gas dehydration is therefore a critical process for reducing water content and mitigating these issues. In this study, the dehydration unit of Khark Petrochemical Company (KPC), which utilizes triethylene glycol (TEG) as the desiccant agent, was simulated using Aspen HYSYS v11.0. The innovation of this paper lies in proposing an optimized method for recovering vent gas used in the glycol solution regeneration process, which has been thoroughly evaluated from both technical and economic perspectives. Simulation results were validated in comparison to plant Process Flow Diagram (PFD) data. A sensitivity analysis was then performed to identify and investigate the impact of various parameters on the dehydration unit's performance and the produced dry gas. Increasing the solvent circulation rate, stripping gas flow rate, and reboiler temperature improves the water removal rate, as indicated by the results. Several options for recovering around 0.7 MMSCFD of stripping gas from the regeneration tower by raising the regenerator's operating pressure were evaluated, with estimated capital and operating expenditures for each. The optimal option, which routes the recovered stripping gas to a low-pressure feed gas compressor, has an estimated capital cost of $80,000 based on Aspen ICARUS software and additional costs were calculated as a percentage of the equipment price. This option, by preventing gas waste and converting it into methanol with an operational cost of approximately $67000, generate annual revenue of $1.7 million. The amount of emission reduction achieved through gas recovery is equivalent to 48 tons of CO_2 per day.

Keywords: Dehydration, Stripping gas, Triethylene glycol, Regenerator, Aspen Hysys.

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خطة مبتكرة لاسترجاع الغاز في وحدة نزع الماء في وحدة تجفيف الغاز: تقييم تقني–اقتصادي لشركة خارك للبتروكيمياويات كدراسة حالة

الخلاصة:

يُعد وجود بخار الماء في الغاز الطبيعي من التحديات الكبيرة، لما يسببه من تآكل وانسدادات في الأنابيب وتقليل سعة النقل. لذلك تُعد عملية نزع الماء من الغاز خطوة أساسية لتقليل محتوى الماء والحد من هذه المشاكل. في هذه الدراسة، تم محاكاة وحدة نزع الماء في شركة خارك للبتروكيمياويات (KPC) التي تستخدم مادة ثلاثي إيثيلين غلايكول (TEG) كمجفف، وذلك باستخدام برنامج Aspen HYSYS v11.0. وتتمثل أهمية هذا البحث في اقتراح طريقة محسّنة لاسترجاع الغاز المتبخر المستخدم في عملية تجديد محلول الغلايكول، حيث جرى تقييمها بشكل شامل من الناحية الفنية والاقتصادية. وقد جرى التحقق من صحة نتائج المحاكاة من خلال مقارنتها مع بيانات مخطط تدفق العمليات (PFD) في المصنع. كما أجري تحليل حساسية لتحديد أثر عدد من المتغيرات على أداء وحدة نزع الماء وكفاءة الغاز الجاف المنتج. وأظهرت النتائج أن زيادة معدل دوران المذيب، ومعدل تدفق غاز التجريد، ودرجة حرارة الغلاية تؤدي إلى تحسين معدل إزالة الماء. كما تم تقييم عدة خيارات لاسترجاع ما يقارب (0.7) مليون متر مكعب قياسي من غاز التجريد من برج التجديد عن طريق رفع ضغط التشغيل، مع تقدير النفقات الرأسمالية والتشغيلية لكل خيار. وقد تبين أن الخيار الأمثل هو إعادة توجيه الغاز المسترجع إلى ضاغط الغاز المغذي منخفض الضغط، حيث بلغت الكلفة الرأسمالية المقدرة له نحو (80,000 دولار) وفق برنامج Aspen ICARUS، إضافةً إلى حساب التكاليف الإضافية كنسبة من كلفة المعدات. ويسهم هذا الخيار في منع هدر الغاز وتحويله إلى ميثانول بتكلفة تشغيلية تقارب (67,000 دولار)، مما يحقق إيرادات سنوية تصل إلى (1.7 مليون دولار). كما أن كمية الانبعاثات التي تم تقليلها عبر استرجاع الغاز تعادل نحو (48 طن من ثاني أكسيد الكربون يومياً).

1. Introduction

Natural sour gas and associated gas extracted from gas and crude oil reservoirs are typically directed to amine sweetening plants, where the final sweetened gas still contains significant amounts of water vapor, saturated at the prevailing operating temperatures and pressures [1].The presence of water vapor in natural gas poses several major challenges, including hydrate formation, internal corrosion, and reduced heating value. These issues lead to decreased performance in downstream processes due to ice formation, pressure drops, and increased maintenance costs [2][3]. Wet natural gas typically contains water in the range of milligrams per standard cubic meter (mg/Nm^3). However, pipeline specifications for permissible water content differ across regions due to varying environmental conditions, with common limits falling between 70 and 120 mg/Nm^3 [4]. Consequently, effective water removal is therefore essential to maintain operational safety and reliability throughout gas transmission and processing

Several gas dehydration methods are employed industrially to achieve this, including absorption, refrigeration, and adsorption[5]. The choice of dehydration method depends on factors such as operating conditions, required water content, and energy consumption. Among these, the absorption method is generally the most energy-efficient, making it economically advantageous

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for large-scale applications[6]. In contrast, the refrigeration method is often preferred for high-pressure operations due to its higher efficiency and lower operating costs. The adsorption method, on the other hand, is suitable for applications requiring very low water dew points, as it can achieve lower water content than absorption [7].

Liquid desiccant absorption technologies utilize various glycols, including Mono-Ethylene glycol (MEG)[7], Di-Ethylene glycol (DEG) [8], and Tri-Ethylene glycol (TEG)[9], with DEG and TEG being most prevalent. While DEG is less expensive than TEG, TEG's superior performance characteristics generally make it the more economical choice for gas dehydration. DEG exhibits higher carryover loss, provides less dew point depression, and is more challenging to regenerate to high concentration [10]. Consequently, TEG is the best option, employed in approximately 95% of glycol dehydration units[11]. TEG's favorable properties include high water affinity, excellent chemical stability, high hygroscopicity, low vapor pressure, and low solubility in natural gas, a low evaporation loss rate, and low thermal degradation during regeneration [12].

In a typical TEG dehydration process, wet gas enters the contactor column from the bottom, while lean TEG enters from the top. Water vapor is absorbed by the TEG, producing dry gas exiting from the top and a rich TEG solution exiting from the bottom. This rich TEG solution then undergoes regeneration to remove absorbed water and any entrained hydrocarbons. The regenerated TEG is then cooled in a heat exchanger and recirculates to the contactor to repeat the process [13].

Glycol regeneration can be hampered by solvent overheating, leading to glycol decomposition. Several processes mitigate this by reducing water's partial pressure in the vapor phase [14]. These include Drizo, Coldfinger, vacuum regeneration, and gas stripping. The Drizo process utilizes a volatile hydrocarbon mixture (approximately 60% aromatics, 30% naphthenes, and 10% paraffins) to increase water volatility in the glycol solution [15]. The Coldfinger system employs a cooling coil in the surge tank's vapor space, condensing a significant portion of the water vapor. This method removes trace water from both hydrophilic and hydrophobic liquids [16]. Vacuum regeneration enhances water stripping by lowering the regenerator's pressure [17]. Gas stripping is the most basic and widely used technique for improving glycol regeneration among these methods, achieving triethylene glycol (TEG) purities between 99.1 and 99.6 wt.%. In this process, a small portion of dry natural gas—typically sourced from the off-gas stream—is introduced into the regenerator. Because hot gas readily attracts water, it is passed through the heated glycol, helping to remove the remaining water content. [18].

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Numerous studies have focused on optimizing gas dehydration units. Jacob optimized a TEG dehydration plant in the Niger Delta, aiming to minimize dry gas water content. His findings indicated that increasing the number of stages reduces the required TEG circulation rate and that increasing the stripping gas rate lowers dry gas water content up to a point, beyond which no further improvement is observed [19]. Ranjbar et al. [20] simulated a TEG dehydration unit to evaluate the impact of parameters such as glycol circulation rate and absorber temperature on the dehydrated gas's water content, using a steady-state simulator. Implementation of optimized parameters minimized the water content of the dehydrated gas, glycol circulation rate, and reboiler duty. Neagu and Cursaru [21] evaluated the performance of regeneration with stripping gas, exploring various flow rates to increase TEG purity. Their study concluded that stripping gas could enhance TEG concentration to 99.22–99.85 wt.%. Chebbi et al. [22] optimized a TEG dehydration process using Aspen Hysys, focusing on minimizing processing costs, including utilities and capital expenditures. Key design parameters considered in their optimization included TEG circulation rate, feed gas pressure and temperature, gas flow rate, stripping gas rate, and the number of theoretical trays. Petropoulou et al. [23] optimized natural gas dehydration with a focus on energy saving, investigating the effect of operational parameters on the process. They determined that the stripping gas rate and reboiler/cooler duties could be lowered through optimization, leading to a significant reduction in operating costs. Kong et al.[12] compared the use of dried natural gas and nitrogen as stripping agents to achieve a target water dew point of -25°C. They found that using dried sales gas resulted in a higher net profit margin (gross profit minus total production costs). Kong et al. [24]also developed a framework to compare Drizo-based regeneration to other dehydration processes, concluding that stripping gas dehydration using dried natural gas could achieve the desired water dew point specification while maximizing gross profit margin.

The Khark Petrochemical Company (KPC) utilizes a glycol regeneration process that involves stripping gas type, with a flow rate of approximately 0.7MMSCFD. However, the stripping gas is vented to a safe location from the top of the regenerator, leading to significant energy loss and environmental pollution. To address this issue, the present study proposes a novel method for stripping gas recovery and provides a comprehensive techno-economic analysis that could significantly reduce energy loss and environmental impact in the natural gas refinery.

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2. Process Description

The KPC is one of the oldest crude oil facilities in the Iran, which was commissioned in 1969 under the license of the American company J. E. Pritchard. Initially, this facility was designed to sweeten 145 MMSCFD of sour gas associated with crude oil on Khark Island. The original process design included an amine sweetening unit, glycol dehydration, lean oil absorption, butane/propane/gasoline separation, and sulfur recovery unit from acid gas using the Claus process.

The glycol dehydration system at the KPC comprises a gas dehydration section (HP and LP gas dehydration) and a glycol circulation section (TEG regeneration loop). Figure (1) illustrates the process schematic of the dehydration unit. In the HP gas dehydration, the sweet gas cooled from temperature approximately 135°F to 85°F as it passes through the high-pressure absorber off-gas feed exchanger (E-103). Subsequently, the gas flows through a high-pressure water knockout drum (D-102) and an inline stream heater (E-104). The heater, which utilizes steam as the heating medium, is designed to prevent hydrocarbon (HC) condensation in the feed gas. The gas then enters the glycol contactor (T-101) below the bottom (ninth) tray. As the gas ascends through the contactor, it contacts counter current with TEG flowing downward. The liquid level in T-101 is maintained automatically by a level controller.

Upon exiting the top tray, the gas passes through a mist extractor to remove entrained glycol before leaving the contactor (D-103). The dry gas is then directed to the liquid recovery absorber system, where liquefied petroleum gas (LPG) and C5+ components are recovered.

The process of dehydrating the gas in the LP section is identical to the HP section which was described earlier. Sweet gas flows through the low-pressure absorber off-gas feed exchanger (E-100), a low-pressure water knockout drum (D-100), and an inline steam heater (E-101). The gas then enters the low-pressure glycol contactor (T-100). The gas flows upward through the contactor, countercurrent to the glycol, and exits the top of the tower at approximately 110°F. The dry gas is subsequently routed to the low-pressure liquid recovery absorber system for further processing.

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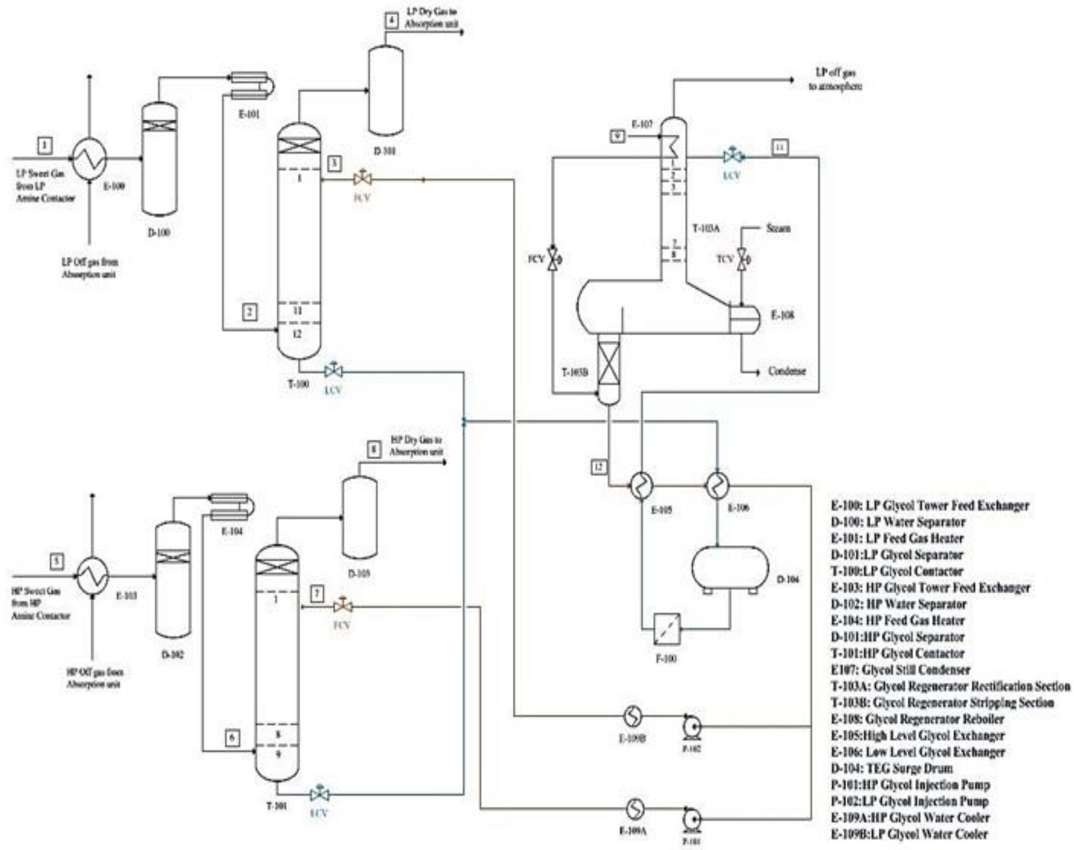

Fig. (1:) Schematic of TEG dehydration unit of KPC

The main equipment's of TEG regeneration section consists of a glycol reboiler (E-108), a glycol surge drum (D-104), two exchangers (E-105 and E-106) and a glycol regenerator (T-103). The regeneration unit is designed to process 28 gallons per minute (GPM) of TEG, achieving a water concentration of no more than 0.2% by weight. The rich glycol from both contactors (T-100 and T-101) is heated to l50 °F by passing through regenerator bottoms exchanger (E-106) and is then routed to glycol surge drum (D-104), where light HC and water vapor is flashed. The wet glycol from the bottom of the surge drum is passed through a filtration system (F-100) to remove any dirt or foreign particles that could potentially cause plugging in the level control valve or other downstream equipment. After filtration, the rich glycol is further heated by E-105 to 246°F and then enters the regenerator (T-103) for further processing.

In the T-103, the wet glycol is heated in the regenerator's bottom section using a reboiler (E-108). A slipstream of low-pressure (LP) gas is extracted and directed to the glycol still condenser (E-107) before flowing through the bottom of the regenerator, where it serves as stripping gas in the regenerator's stripping section. The flow rate of this gas is regulated by a flow controller and adjusted as needed to ensure effective glycol regeneration. The glycol descends through the tray

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tower and enters the glycol reboiler, where the water content is vaporized. The vaporized water rises through the packed tower, while any vaporized glycol is condensed and returned to the reboiler. The water-laden gas is subsequently vented to the atmosphere. A level controller maintains the glycol level in the regenerator's bottom section by operating a control valve on the glycol inlet line. The temperature of the glycol in the reboiler is maintained at approximately 375°F, a level at which glycol degradation is negligible, and the glycol concentration exceeds 99%. This temperature is controlled by a temperature controller that regulates a control valve on the heating oil line to the reboiler. Lean glycol exits the bottom of the regenerator's stripping section and flows through the shell side of the glycol feed exchangers (E-105 and E-106) for further cooling and reuse in the dehydration process. The lean glycol from the bottom of the regenerator is pumped by charge pumps (P-101 and P-102) to T-101 and T-100. This regeneration system ensures the efficient removal of water and contaminants from the glycol, maintaining its purity and effectiveness for reuse in the dehydration process. Table 1presents the design specification of main streams of gas dehydration unit including wet LP and HP gas (Stream No. 1, 5), stripping gas (Stream No. 9) and the absorption towers inlet streams (stream No. 2, 6).

Table (1): Stripping gas, HP and LP wet gas specification (Design Data)

Stream Name	**LP wet gas (1)**	**LP wet gas to T-100 (2)**	**HP wet gas (5)**	**HP wet gas to T-101 (6)**	**Stripping Gas to T-103 (9)**
Component	mol.%	mol.%	mol.%	mol.%	mol.%
H_2O	1.61	0.56	0.48	0.11	0.11
Nitrogen	0.01	0.01	0.00	0.00	0.01
CO_2	0.11	0.11	0.02	0.02	0.1
Methane	36.84	37.23	80.95	81.26	67.69
Ethane	23.33	23.57	11.23	11.26	30
Propane	24.76	25.04	4.72	4.74	1.87
i-Butane	3.58	3.62	0.64	0.65	0.02
n-Butane	6.37	6.43	1.25	1.25	0.09
i-Pentane	1.28	1.29	0.29	0.29	0.07
n-Pentane	1.45	1.47	0.34	0.34	0.02
n-Hexane	0.66	0.67	0.08	0.08	0.01
n-Heptane	0.00	0.00	0.00	0.00	0.01
Water dew point (F)	131.80	94.78	131.50	79.93	-48.28
Water content (lb H_2O/MMSCF)	765	269	231.4	51.64	4.05
Temperature (F)	135	100	135	90	166.8
Pressure (psig)	135.3	132.5	525.3	522.3	5
Molar Flow (MMSCFD)	29.22	28.91	77.41	77.11	0.688

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According to the design data, about 0.7 MMSCFD stripping gas should be used in the regenerators to strip the water vapor from TEG in regeneration column. Table 2 shows the key design parameter of KPC dehydration unit.

Table (2): Design parameters of KPC dehydration unit

Parameter	Unit	Value
Glycol flow to LP absorber (T-100)	USGPM	9
Glycol flow to HP absorber (T-101)	USGPM	18.5
LP wet gas flow to T-100	MMSCFD	28.9
HP wet gas flow to T-101	MMSCFD	77.1
Reboiler Temperature	°F	375
Stripping gas flow	MMSCFD	0.68
LP absorber pressure (T-100)	psig	130
HP absorber pressure (T-101)	psig	520
Regenerator pressure	psig	0
Glycol temperature to T-100/T-101	°F	150

3. Methodology

Following an energy assessment and routine survey of the KPC dehydration plant, potential opportunities for improvement were evaluated in consultation with the factory's engineering and technical team. In line with recent international regulations aimed at reducing global warming and greenhouse gas emissions through energy optimization, we focused on monitoring stripping gas and identifying potential scenarios for mitigating or controlling HC emissions. To investigate solutions for optimizing the dehydration plant and recovering stripping gas, the existing plant configuration of the KPC glycol dehydration unit was simulated using Aspen Hysys v.11 software. This software has the ability to model multiphase flow, chemical reactors and unit operations, and, it is widely used in oil and gas processing simulation.

The simulation employed both the Peng-Robinson equation of state and the Glycol Package, a specific thermodynamic model within Aspen Hysys designed for systems containing glycol. The Glycol Package incorporates the Twu-Sim-Tassone (TST) equation of state [14] and the Non-Random Two Liquids (NRTL) activity coefficient model [25]. This combination is recognized for accurately representing the activity coefficient of TEG-water solutions and calculating water dew points for natural gas across a wide range of pressures, temperatures, and concentrations typical in TEG dehydration units. While the Peng-Robinson equation of state was also utilized, the Glycol

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Package was primarily leveraged for its superior accuracy in modeling phase equilibrium and the properties of natural gas/water mixtures[25].

The primary objective of this study was to investigate the recovery of stripping gas from the TEG regeneration section to minimize gas loss and environmental pollution. To accomplish this goal, a series of simulations were conducted using Aspen Hysys to assess the influence of key operating parameters on the water dew point of the dry gas:

- Regenerator column pressure
- Glycol regenerator temperature
- Stripping gas flow rate
- TEG circulation rate

By systematically varying these parameters, the simulation aimed to identify optimal operating conditions that maximize stripping gas recovery while maintaining acceptable dry gas water content specifications.

4. Results and Discussion

4.1. Design simulation validation

Prior to conducting performance analysis and evaluating options for stripping gas recovery, the simulation model was rigorously validated against available plant data. This validation step is crucial to ensure accurate prediction of process behavior across a wide range of operating conditions. A simplified flow diagram of the KPC's TEG dehydration plant was constructed within the Aspen Hysys environment in Figure (2). The model incorporated the key components of the unit, including the high-pressure (HP) and low-pressure (LP) absorbers, and the TEG regenerator. The actual numbers of trays in the HP and LP absorbers are 12 and 9, respectively. Considering an overall tray efficiency of 30%, the equivalent numbers of theoretical trays were determined for the simulation. Similarly, the regenerator, which has 8 actual trays, was modeled with 4 theoretical trays based on an assumed overall efficiency of 50%. The simulation utilized the feed specifications detailed in Table (1). Based on the heat and material balance documentation for the dehydration unit, the inlet water content of the LP and HP gas streams were set at 765 and 231 lb H_2O/MMSCF, respectively. The cooling of the feed streams before entering the TEG contactors was also modeled, resulting in an approximate reduction of water content by 65% and 77% for the LP and HP wet gas streams, respectively. The temperature of the stripping gas increases from 55°F to 167°F as it exchanges heat with the regenerator overhead stream in exchanger E-107.

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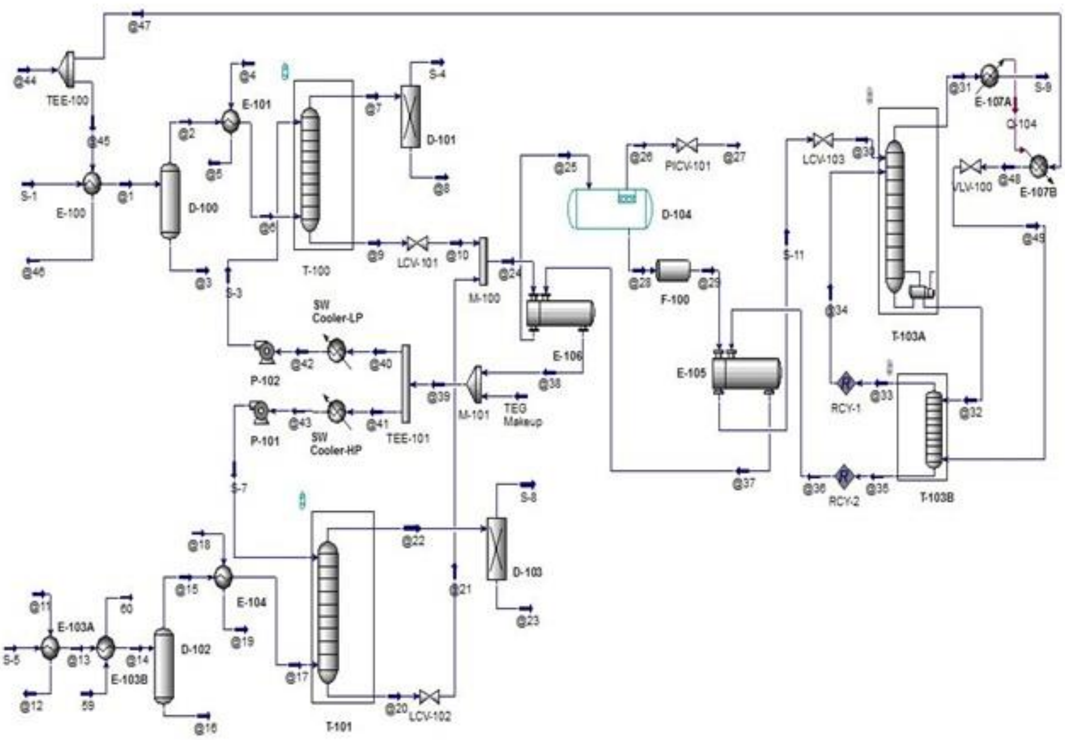

Fig. (2) Process flow diagram of KPC dehydration unit in Aspen Hysys

The simulation results were compared to the PFD data and its results presented in the Table 3. It can be seen that the deviation between the simulation results and the design data was found to be generally less than 5%. This level of agreement indicates that the simulation model provides an acceptable representation of the actual plant and is suitable for use in subsequent studies.

Table (3): Validation of simulation result in compare with design data of KPC dehydration unit

Parameter	LP Sweet Gas		LP Dry Gas		HP Sweet Gas		HP Dry Gas	
	Design	Simulation	Design	Simulation	Design	Simulation	Design	Simulation
Flow rate (MMSCFD)	29.22	29.22	28.91	28.72	77.41	77.41	77.03	77.02
Water Content (lb/MMSCF)	765	765	4.5	4.6	231.4	231.4	0.448	0.491
Water Dew point (F)	N.R.	132.6	-3.7	-3.8	N.R.	131.5	-34.3	-32.5
Temperature (F)	135	135	110	111.8	135	135	90	93.24
Pressure (psig)	132.3	132.3	130.3	130.3	525.3	525.3	520.3	520.3

N.R. Not Reported

4.2. Effect of regenerator pressure

According to the requirement of the stripping gas recovery to increasing the regenerator overhead pressure, it is important to examine the impact of regenerator column operating pressure on regenerated TEG concentration. Figure (3) depicts the effect of regenerator pressure on regenerated TEG concentration and the water removal efficiency of LP and HP dry gas streams.

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Based on design data, the dew points of LP and HP dry gas should be -3.7°F and -34.3°F, respectively; therefore, an increase in the water dew point relative to these design values serves as an indicator of changes in water removal efficiency.

Figure (3) illustrates how increasing the operating pressure of the TEG regenerator has a negative effect on the concentration of lean TEG—that is, the glycol after water has been removed when other parameters, such as reboiler duty and stripping gas flow rate are held constant. This means the TEG becomes less effective at absorbing water from natural gas in subsequent cycles [5] [26]. The results indicate that increasing the regenerator's operating pressure negatively affects the lean TEG concentration, as the system becomes less efficient at removing water from the glycol. Consequently, the regenerated TEG retains more water than desired, which compromises the dehydration performance in subsequent cycles. At a regenerator pressure of 10 psig, the water dew point of the HP dry gas increases by approximately 12°F, and that of the LP dry gas rises by about 17°F, compared to their respective design values. This increase in dew point reflects inadequate dehydration and elevated water content in the treated gas. Thermodynamically, higher operating pressure increases the boiling point of water and reduces the driving force for its evaporation, making water removal more difficult. As a result, the reboiler and stripping gas are less effective in regenerating TEG under elevated pressure, even when their operating conditions remain unchanged.

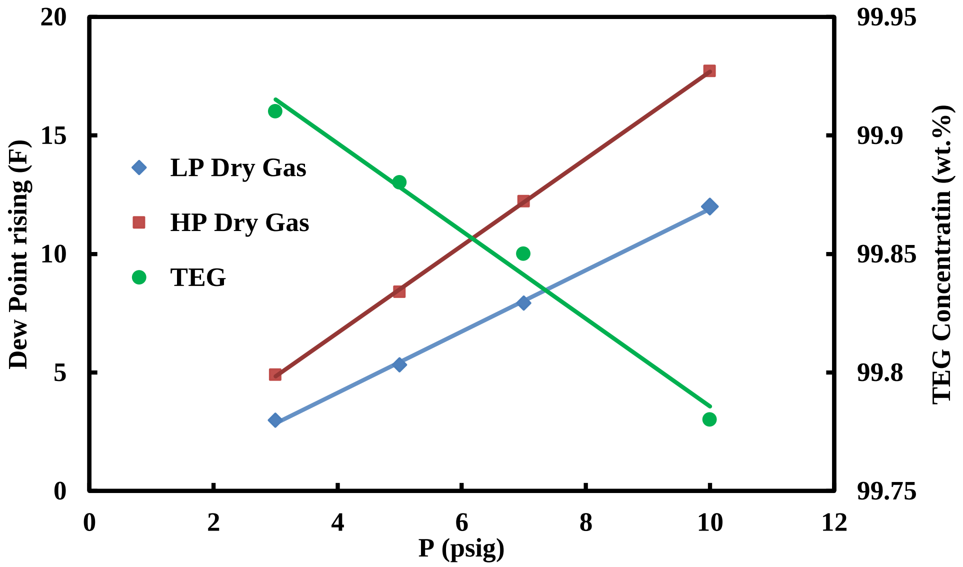

Fig. (3): Effect of regenerator operating pressure by Hysys simulation on lean TEG concentration and rising of gas stream dew point

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4.3. Effect of glycol regenerator temperature

The regenerator reboiler temperature is a significant factor affecting lean glycol purity and, consequently, the performance of the dehydration unit. This temperature governs the concentration of water in the lean glycol. It is important to note that the maximum allowable regeneration temperature for TEG should be significantly lower than its decomposition temperature of 464°F (240°C) [27]. Therefore, to prevent glycol degradation, the regenerator temperature is typically regulated between 375°F and 400°F.

Fig. 4 illustrates the effect of reboiler temperature on TEG regeneration performance at regenerator operating pressures of 5 and 10 psig, and at the TEG circulation rate and stripping gas flow rate specified in Table 2. The results show that higher regenerator temperatures enhance glycol regeneration and improve water removal from TEG. Therefore, increasing the heat input to the reboiler to elevate its temperature is an effective strategy for boosting TEG regeneration efficiency, which may also reduce the amount of stripping gas needed for optimal performance. [23].

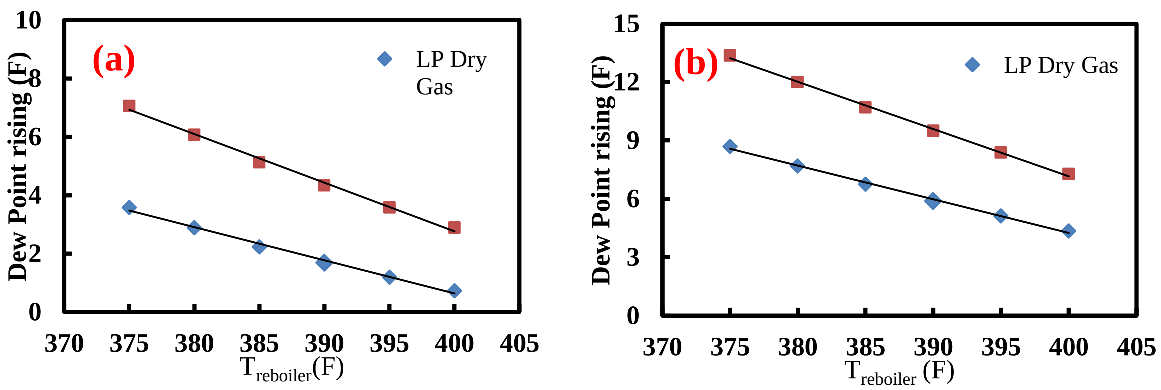

Fig. (4): The effect of regenerator reboiler temperature on dew point rising of dry gas streams by simulation at (a): P=5 psig; (b): P=10 psig

4.4. Effect of stripping gas flow rate

Figure (5) illustrates the effect of stripping gas flow rate on the water dew point of dry HP and LP off-gas streams, while maintaining constant reboiler temperature and TEG circulation rate as specified in Table (2). As depicted in Figure (5), increasing the stripping gas flow rate enhances TEG regeneration, thereby improving water removal from the wet gas in the contactor. Elevated stripping gas flow rates in the regenerator column reduce the partial pressure of water, increasing the propensity for water to transfer from the TEG to the upward-flowing gas stream within the regenerator. Consequently, with a fixed reboiler temperature, the efficiency of water removal is primarily dependent on the stripping gas flow rate. It is crucial to acknowledge that while

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increasing the stripping gas flow rate positively impacts water removal efficiency, it can also lead to increased glycol loss through the regenerator's overhead stream, potentially resulting in greater energy consumption and environmental pollution if released into the atmosphere[28].

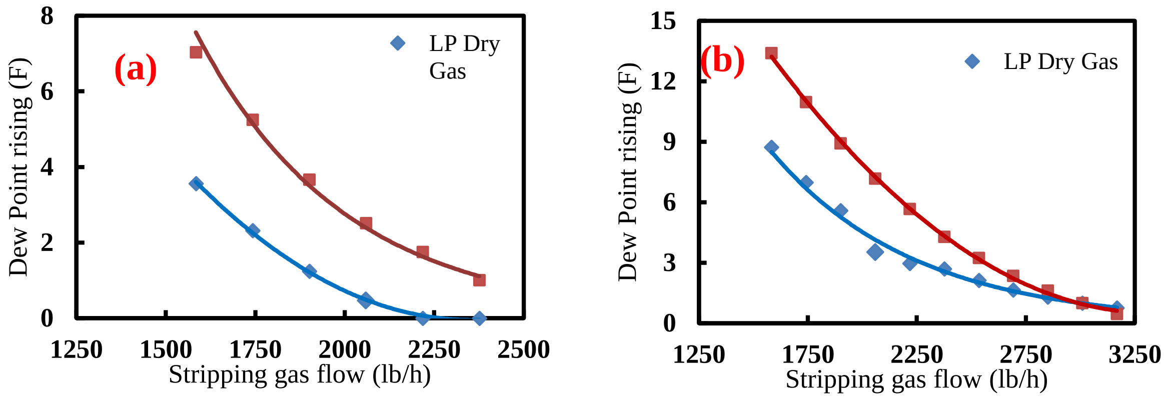

Fig. (5): The effect of stripping gas on dew point rising of dry gas streams by simulation at (a): P=5 psig; (b): P=10 psig

4.5. Effect of glycol circulation rate

Based on the design data, the total circulating TEG amount is 20.95 GPM, of which 9 GPM equal to 33% is sent the HP gas contactor (T-101) while 18.5 GPM (67%) is contacted with wet gas in the LP contactor (T-100). To investigate the effect of glycol flow rate on dehydration performance, simulations were conducted with glycol flow rates up to 25% above the normal operating values based on design data, considering the power and flow rate limitations of the TEG circulation pumps. Throughout these simulations, the HP to LP glycol flow ratio was maintained at approximately 1:2, consistent with the design ratio.

Figure (6) illustrates the impact of TEG flow rate variations on the water dew point of the dry HP and LP off-gas streams, while maintaining stripping gas flow rate and reboiler temperature at the design values listed in Table (2). As shown in Figure (6), increasing the TEG flow rate initially improves the water absorption process in the glycol contactors. However, TEG regeneration becomes less effective when other influencing parameters are held constant. While the dry gas dew point initially decreases with increasing TEG flow rate, further increases eventually lead to a rise in the water dew point. This is because the TEG concentration gradually decreases due to reduced regeneration efficiency[19]. Although increasing the reboiler temperature or stripping gas flow rate could potentially maintain the TEG concentration at the desired range, this approach is not ideal as it results in higher energy consumption. Therefore, minimizing the TEG flow rate is a more energy-efficient strategy.

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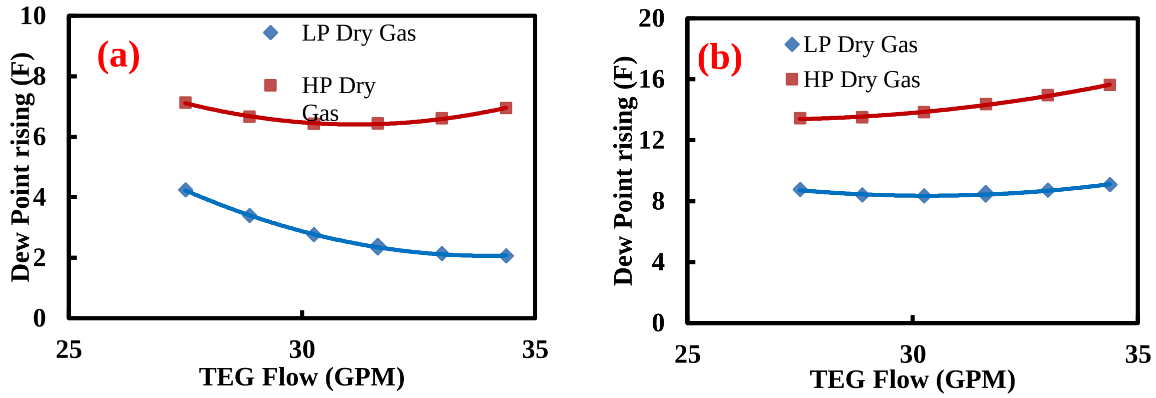

Fig. (6) The effect of TEG circulation rate dew point rising of dry gas streams at (a): P=5 psig; (b): P=10 psig

4.6. Proposed stripping gas recovery options for KPC dehydration unit

There are three options proposed for the stripping gas recovery from the dehydration unit of KPC as following:

- **Option 1:** Stripping gas compression and use as fuel.
- **Option 2:** Compressing and recirculating the stripping gas in the closed loop.
- **Option 3:** Increased regenerator pressure and integration with LP feed gas.

All stripping gas recovery scenarios require an increase in regenerator pressure. The implementation of each recovery option is contingent upon establishing sufficient positive pressure. Simulations for Options 1 and 2 were conducted at 5 psig, while Option 3, based on pressure drop calculations between the glycol regenerator and LP booster compressor (LP gas feed streams before amine sweetening), required approximately 10 psig. It's important to note that increasing pressure can negatively impact the dew point of LP and HP gases. Therefore, technical evaluations of each scenario included adjustments to TEG regeneration parameters to maintain the dry gas dew point within design specifications. Based on sensitivity analyses, stripping gas flow, reboiler temperature, and TEG circulation rate were identified as key adjustable parameters. Stripping gas flow was the primary adjustment, limited to a 20% increase over the design flow rate to minimize glycol loss. Reboiler temperature was then adjusted, with a maximum limit of 400°F to prevent TEG degradation. Finally, TEG flow rate was adjusted within a 25% increase over the design value to achieve adequate water removal from the HP and LP dry gas streams.

Option 1: Stripping Gas Compression and Use as Fuel

This option involves maintaining a positive pressure of 5 psig at the top of the glycol regeneration tower using a pressure control valve. Operational parameters, such as stripping gas flow rate and

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reboiler temperature, are adjusted to maintain a consistent regenerated glycol concentration compared to design data. The overhead vapor from the TEG regenerator is compressed to 60 psig via a two-stage compressor and then used as fuel in incinerators or boilers. The simulation flowsheet for this option is shown in Figure (7). The stripping gas from the regenerator overhead enters a knockout drum for liquid separation. The gas is compressed to 21 psig, increasing the temperature to 251°F. After compression, the gas is cooled to 113°F in an inter-stage cooling water cooler. The gas stream is then directed to the second-stage suction drum, where any remaining water is separated before final compression to 61 psig, the required pressure for the fuel gas header. The gas is then routed to the fuel stop tank.

Option 2: Compressing and recirculating the stripping gas in the closed loop

In this option, the regenerator column pressure is increased to 5 psig. The overhead vapor is routed to a knockout drum before entering the compressor suction drum, where it's compressed to 18 psig. The gas stream then passes through a cooling water cooler and a propane refrigerant cooler, reducing the temperature to 41°F. The gas is then directed to a discharge drum for bulk water separation. The partially dry gas is sent to a small TEG contactor column for final moisture removal before being recirculated back to the glycol regenerator column in a closed loop. The simulation schematic for this option is shown in Figure (7). Unlike Option 1, where the regenerator overhead gas is used as fuel, this option dehumidifies and reuses the gas in a closed loop, which is the key differentiator.

Option 3: Increased regenerator pressure and integration with LP feed gas

As shown in Figure (7), this option avoids compression of the regenerator overhead gas. Instead, the regenerator pressure is increased to 10 psig, and the gas is passed through a propane refrigerator cooler. After cooling, the gas is routed to the LP booster compressor suction drum. To maintain the lean TEG concentration at the bottom of the regenerator, both the stripping gas flow rate and the boiler temperature are increased. This prevents negative impacts on the LP and HP dry gas water removal performance.

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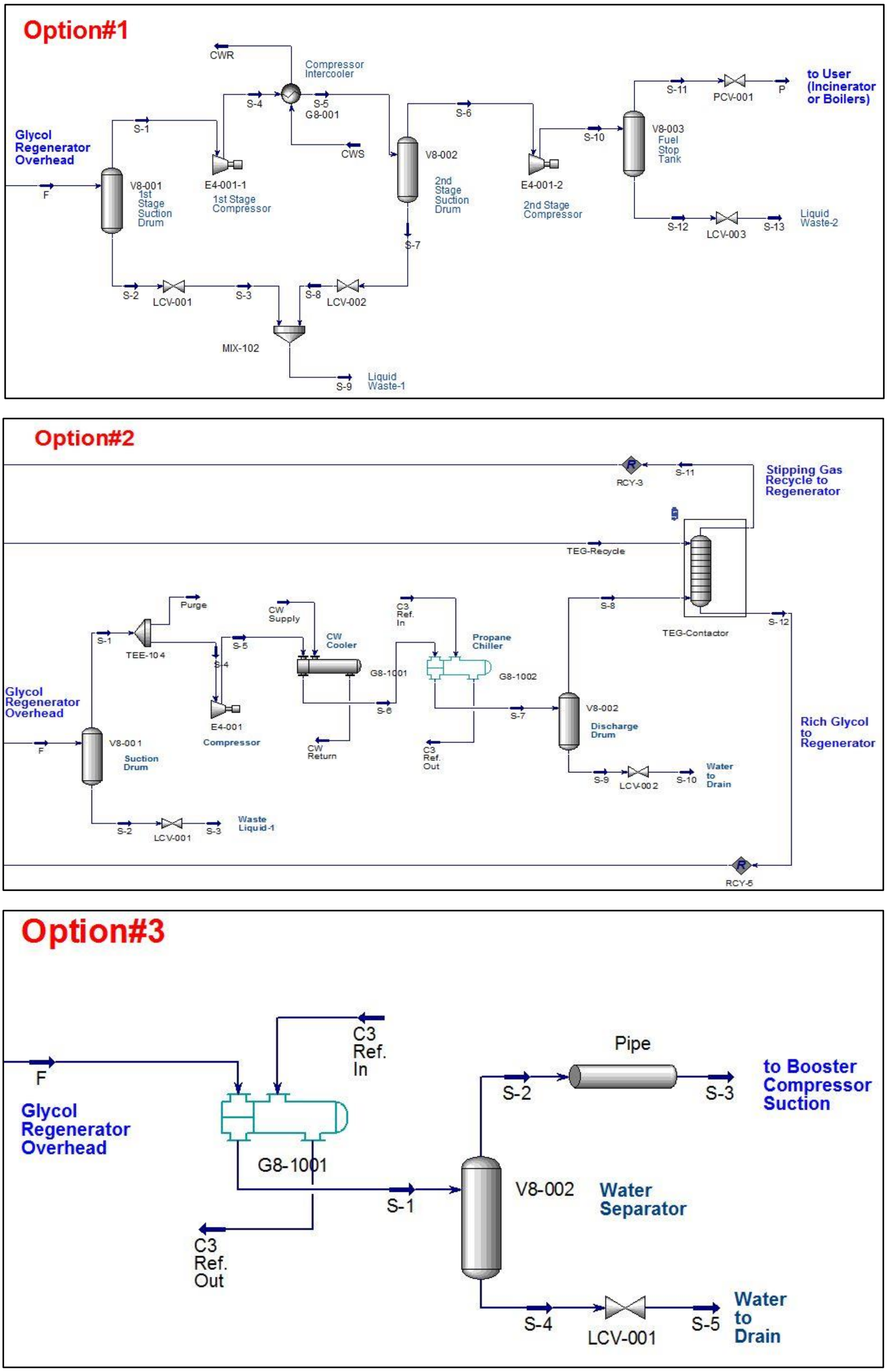

Fig. (7): Sketch of options proposed for the stripping gas recovery

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4.7. Comparison of Proposed Options

4.7.1 Technical Analysis

Table (4) summarizes the process simulation results and economic evaluation for each option. The simulations indicate that increasing the regenerator pressure to 5 psig (Options 1 and 2) necessitates increasing the stripping gas flow rate, reboiler duty, and TEG circulation rate by 20%, 10%, and 3%, respectively, to maintain water removal efficiency. For Option 3, where the TEG regenerator pressure is 10 psig, the stripping gas flow rate, reboiler duty, and TEG circulation rate must be increased by 25%, 30%, and 15%, respectively, to prevent an increase in the water dew point of LP and HP dry gas.

Table (4): key parameters of dehydration unit after implementing the different options for stripping gas recovery

Stream	Case Study	Design	Options 1,2	Option 3
	Parameter	P=0 psig	P=5 psig	P=10psig
LP Dry Gas	Water Dew point (F)	-3.8	-3.95	-4.32
	Water Content (lb /MMSCF)	4.59	4.547	4.464
HP Dry Gas	Water Dew point (F)	-32.46	-34.25	-33.93
	Water Content (lb/MMSCF)	0.488	0.449	0.456
TEG	Glycol Concentration (wt. %)	99.96	99.95	99.92
	Glycol loss (USGPM)	0.028	0.032	0.026
	LP Glycol Flow(USGPM)	18.5	19.05	21.28
	HP Glycol Flow(USGPM)	9	9.82	11.72
Regenerator	Stripping Gas Flow (lb/h)	1584	1900	2000
	Reboiler Duty (Btu/h)	1.64E+06	1.79E+07	2.13E+06
	Reboiler Temp. (F)	375	385	395
	Water loss (USGPM)	0.965	0.812/0.937	0.943

4.7. 2. Economic analysis

The investment cost evaluation for each stripping gas recovery scenario encompasses a range of expenditures, including equipment, piping, electrical systems, instrumentation, construction, execution, and installation. The necessary equipment—such as separators, compressors, and heat exchangers—has been appropriately sized based on standard engineering practices. Equipment costs were then estimated using Aspen Icarus software to ensure accuracy and consistency across scenarios.

The specifications of required equipment for each scenario of stripping gas recovery are shown in Table (5).

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Table (5): Specification of equipment required for each options

Equipment	Specification	Unit	Value
	Option 1		
Compressor	Power	kW	74.58
Intercooler	Duty	kW	175.3
Cooling Water	Flowrate	m3/h	15.2
First Suction Drum	Diameter / Height	m	0.5 /2.3
Second Suction Drum	Diameter / Height	m	0.35 /2.2
Fuel Stop Tank	Diameter / Height	m	0.5 /1.5
	Option 2		
Compressor	Power	kW	22.4
After Cooler	Duty	kW	146.2
Cooing Water	Volume Flowrate	m3/h	12.6
Propane Chiller	Duty	kW	41.1
Propane Refrigerant	Mass Flowrate	ton/h	0.38
Compressor Suction Drum	Diameter / Height	m	0.5 / 2.3
Compressor Discharge Drum	Diameter / Height	m	0.4 / 2.2
New Glycol Regenerator	Diameter / Height	m	0.3 / 5.0
New Reboiler	Duty	kW	568
	Option 3		
New Reboiler	Duty	kw	625
Propane Chiller	Duty	kW	196.7
Propane Refrigerant	Mass Flowrate	ton/h	1.83
Water Separator	Diameter / Height	m	0.35 / 2.2

The equipment cost calculation takes into account various factors such as equipment depreciation, material and labor cost, and other overheads. The final investment cost for each scenario is calculated as the sum of the equipment cost plus the other associated costs such as piping, electrical, instrumentation, construction, execution, and installation, which are estimated by multiplying the total equipment cost with the corresponding coefficient factor (F_i) values are given in the Table (6)[29].

Table (6): Cost estimation coefficient for CAPEX

Row	Description	Cost Estimation Coefficient (F)
1	Piping	30%
2	Electrical	10%
3	Instrument	10%
4	Insulation	7%
5	Construction and Installation	30%
6	Contingencies	10%

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This total cost would provide an estimate of the capital expenditure (CAPEX) required for implementing the selected option

$$C_{CAPEX} = \sum_{i=1}^{N} (Equipemtn\ Cost).F_I \qquad (1)$$

Based on the above explanations, the total capital investment cost for different stripping gas recovery options was estimated, and the results are presented in Table (7). As indicated, the investment costs for options 1, 2, and 3 are estimated at $1,330,000, $152,000, and $81,000, respectively. Therefore, option 3, which requires less equipment and lower capital investment, is more suitable option.

To estimate the operating costs, it is necessary to determine the consumption of steam consumption for the reboiler, electricity for the compressor, and cooling systems for each option. Table (8) presents the required utility values for each option based on the simulation results. According to the unit prices of the utilities provided in Table (7), the total cost of the utilities has been calculated.

Table (7): Utility prices used for the calculation of the operating cost

Utility	Price	Unit
Electricity	0.04	$/ kWh
Steam	8.33	$/ ton
Cooling Water	0.034	$/ ton
Propane for chilling	2.94	$ /ton

Table (8): Utility consumption of for different options

Utility Consumption	Unit	Option 1	Option 2	Option 3
Power	kW	74.58	22.4	0
Steam	ton/h	0.1	0.1	0.3
Cooling Water	ton/h	15.2	12.6	0
Propane Chilling	ton/h	0	0.4	1.83

The total operating expenditure (OPEX) costs for each option, is calculated as the sum of the utility consumption costs, maintenance and contingency expenses. The results are presented in Table (9). As observed, the OPEX costs for options 1, 2, and 3 are estimated at $61,000, $27,000, and $67,000 per year, respectively.

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Table (9): Operating cost estimation for different options

Operating Cost	Option 1	Option 2	Option 3
Power	26250	7884	0
Steam	5428	5428	17732
Cooling Water	4103	0	0
Propane Chilling	0	8849	42497
Total Utility Price	35780	22161	60229
Maintenance	20278	2311	1239
Contingencies	5605.9	2447.2	6146.7
Total Operating Cost	61664	26919	67614

The KPC has a methanol production unit adjacent to the KPC gas refinery plant which converts the LP and HP off gas from gas refinery to the valuable methanol product. The recovered stripping gas from TEG dehydration unit will be used as methanol plant feed. The revenue obtained from this project depends on the amount of methanol produced by adding the stripping gas, which is equivalent to 22.4 tons per day of methanol. The Figure (7) illustrates the sensitivity of the annual revenue of this project to a ±20% variation in methanol price relative to its current average value (230 $/Ton). As shown, the sales revenue of this proposal can vary between $1.4 million and $2.1 million, depending on the methanol market price.

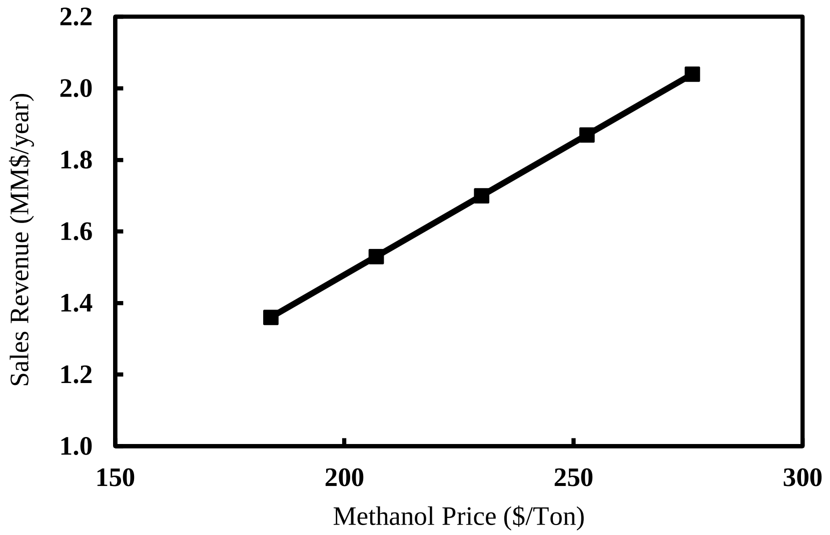

Fig. (7): Sales revenue sensitivity to methanol price variations

Economic analysis summary for different options proposed for stripping gas recovery is shown in Table (10). It can be observed, that the option 1 has a higher capital cost compared to other options and also, the longest payback period (calculated the revenue divided to CAPEX). This is due to

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the fact that in option 1, the recovered stripping gas is utilized as fuel rather than being used as feedstock for methanol production. As a result, the revenue generated from increased methanol production is not realized in option 1. On the other hand, option 2 and 3 provide significant revenue increases due to the recovered stripping gas being used as feedstock for methanol production. While option 3 has the lowest investment cost among the three options, and the highest the OPEX, but it provides the shortest payback period. Therefore, based on the economic analysis results, option 3 would be the recommended option for the recovery of stripping gas from TEG dehydration unit.

Table (10): Economic analysis of proposed options

	Option 1	Option 2	Option 3
CAPEX (1000 USD)	1331.6	151.8	81.4
OPEX (1000 USD/Year)	61.7	26.9	67.6
Revenue (1000 USD/Year)	1700	1700	1700
Payback period (Month)	19.5	2.2	1.2

4.7. 3. Environmental analysis

The CO_2 emissions occur via three primary pathways. The first is direct emissions, which come from process streams, while the second and third are indirect emissions, resulting from electricity and heat consumption. To estimate the CO_2 emissions from 0.7 MMSCFD LP gas with the given composition (stream 9 in the Table 1), we need to calculate the combustion-based CO_2 emissions for each component and sum them up[30]. Additionally, the indirect CO_2 emissions from electricity are associated with the compressors, cooling water pump and propane chilling, while those from heat is due to the higher steam consumption in the regeneration section's reboiler should be quantified.

According to, generating 1 MWh of electricity releases 0.596 tons of CO_2, whereas producing 1 MMBtu of heat emits 205.3 pounds of CO_2, the net CO_2 emissions for all options have been calculated[31]. Table (11) compares the emission reduction results from the three proposed options. As can be seen, the amount of emission reduction achieved through gas recovery is nearly identical across all options, equivalent to 48 tons of CO_2 per day.

Table (11): CO_2 emissions of for different options

CO_2 emission source	Unit	Option 1	Option 2	Option 3
Power	Ton/h	0.038	0.019	0.034
Steam	Ton/h	0.018	0.018	0.054
LP stripping gas	Ton/h	-2.06	-2.06	-2.06
Net	Ton/d	-48.09	-48.55	-47.33

5. Conclusions

The study focused on optimizing the KPC glycol dehydration unit to recover stripping gas, reduce hydrocarbon emissions, and improve energy efficiency in alignment with global efforts to mitigate greenhouse gas emissions. Using Aspen Hysys v.11, the existing plant configuration was simulated. The simulation evaluated the impact of key operating parameters—regenerator pressure, reboiler temperature, stripping gas flow rate, and TEG circulation rate—on the water dew point of dry gas. The simulation was validated against plant data, showing less than 5% deviation, confirming its reliability for further analysis. The study identified that increasing regenerator pressure negatively affects TEG concentration and water removal efficiency, while higher reboiler temperatures and stripping gas flow rates improve TEG regeneration. However, excessive TEG circulation rates reduce regeneration efficiency, highlighting the need for balanced optimization.

Three stripping gas recovery options for recovering approximately 0.7 MMSCFD of LP gas, were proposed as following:

Option 1: Compression and use as fuel.

Option 2: Compression and recirculation in a closed loop.

Option 3: Increased regenerator pressure and integration with LP feed gas.

Economic analysis revealed that option 3 is the most viable, with the lowest CAPEX and the highest revenue potential due to the integration of recovered stripping gas into methanol production. This option also aligns with environmental goals by minimizing hydrocarbon emissions and energy consumption. Also, the environmental analysis revealed that the CO_2 emission reduction achieved through stripping gas recovery amounts to approximately 48 tons per day.

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Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper

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Abbreviations

CAPEX	Capital expenditure
DEG	Diethylene glycol
HC	Hydrocarbon
HP	High Pressure
KPC	Kharg Petrochemical Company
LP	Low Pressure
LPG	Liquefied petroleum gases
MEG	Monoethylene glycol
MMSCFD	Million standard cubic feet per day
NRTL	Non-Random Two Liquids
OPEX	Operating expenditure
PFD	Process Flow Diagram
TEG	Triethylene glycol
TST	Twu-Sim-Tassone
USGPM	United states gallon per minute
USD	United stated dollar

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