RESEARCH PAPER

Optimal solvent definition for gas treatment units with the aim of acid gas enrichment

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ABSTRACT

Increasing the concentration of H2S in the acid gas entering the sulfur recovery unit is one of the key parameters for enhancing the sulfur recovery efficiency. This study aimed to determine the optimal solvent for a gas treatment unit with a focus on enriching hydrogen sulfide gas in the feed of the sulfur recovery unit. Data from the design of an industrial-scale gas-treating unit was utilized to achieve the research objectives. For simulation, analysis of results, and plotting graphs, ASPEN-HYSYS V11 and DESIGN EXPERT V10 softwares were used, respectively. The solvents available in the GPSA handbook were utilized for the solvent and additives selection. In the formulation of a composite solvent, DEA and MDEA solvents were considered as the base solvents. Regarding the weight percentage range used in the gas processing industry, and other solvents were used as additives to the composite solvent. Simulation results were analyzed by comparing them with the standards of the Iranian National Gas Company. The results showed that the combination of MDEA with 42.5wt% as the base solvent and Sulfolane with 7.5wt% as an additive is the optimum solvent. The reason for this selection is that there will be the maximum H2S concentration in the acid gas while adhering to the mentioned standards. The optimum solvent can increase the concentration of H2S from 33.6 mol% to 41.84 mol% in the acidic gas to the Sulphur recovery unit.

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1. INTRODUCTION

The acid gas enrichment process has received rising attention because the operating units of the gas processing and refining industry are facing numerous problems, including low-quality sulfur production, high fuel and energy consumption, and the release of polluting gases such as sulfur dioxide into the atmosphere. To better understand the importance of the enrichment process, in the first step, it is necessary to investigate some of the problems caused by the low concentration of hydrogen sulfide in the acid gas flow.

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1.1. Reducing the temperature of the combustion chamber

Based on studies, the optimal temperature in the combustion chamber is in the range of 1100-1200 °C [1]yword>Operating Cost<. Increasing the CO2 concentration in the sour gas will decrease the furnace temperature. This temperature decrease will increase fuel consumption and the amount of additional air required in the combustion chamber, while also reducing the burning of hydrogen sulfide in the process. Therefore, the quality, amount and recovery rate of the produced sulfur will decrease [1].

1.2. BTEX are stable at low temperatures

The analysis of the acid gases shows the presence of impurities such as heavy and aromatic hydrocarbon compounds, BTEX[1], in the sour gas. These components do not burn in the low-temperature combustion chamber, and they will cause poisoning of the catalysts in the catalytic section of the sulfur recovery unit [2]. Catalyst poisoning will cause a decrease in performance and subsequently an increase in operating costs due to the premature replacement of the catalyst, an increase in the amount of pollutants entering the atmosphere, and ultimately a decrease in the sulfur recovery factor in the unit. Based on the diagram presented in Figure 1, the relationship between the concentration of hydrogen sulfide in the acid gas flow and the decomposition of aromatic and heavy hydrocarbon impurities with the temperature of the combustion chamber is shown. The flame temperature will be increased along with the increase in the H2S concentration in the acid gas flow.

Natural gas is a crucial fossil fuel resource, containing hydrocarbon compounds such as methane, ethane, etc., and non-hydrocarbon compounds such as carbon dioxide, nitrogen, oxygen, and hydrogen sulfide [3]. The presence of non-hydrocarbon compounds such as H2S and CO2 not only poses environmental challenges but can also lead to severe corrosion in refining facilities and the formation of gas hydrates in gas distribution and transmission networks [4]. Various processes have been proposed for separating acid gas and other impurities from sour gas. The most common method for natural gas sweetening in Iran is the use of amine solvents. In this process, the acid gas is separated from the sour gas using amines. Several parameters, such as the amine type, the sour gas temperature, pressure, and composition, have an impact on the gas sweetening process. Kulivand

1.Benzene, toluene, ethylbenzene, xylenes

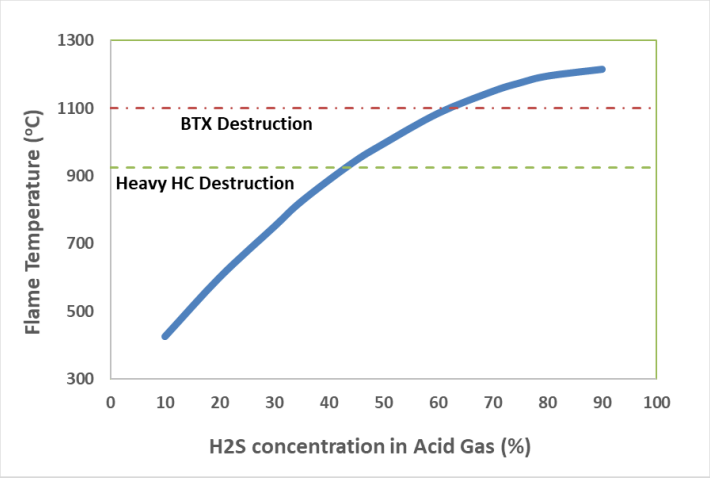

Figure 1. The temperature of the reaction furnace with the molar percentage of hydrogen sulfide in the acid gas stream [1]

et al., [5] investigated and determined the optimal temperature of the amine solvent entering the regeneration column for better separation of acid gas from rich amine. Shahsavand et al., [6] suggested reducing the temperature of both the sour gas and the amine solvent simultaneously by installing a heat exchanger to improve the efficiency of the acid gas absorption process. Changing the type of sweetening solvent also significantly affects the separation and, consequently, the concentration of acid gas. Soltani Panah et al., [7] developed a novel approach for modeling acid gas solubility in alkanolamine aqueous solutions. They tested different amines such as MEA[1], DEA, TEA[2], MDEA, and DGA[3]. If the separation of sulfur-containing compounds such as carbonyl sulfide and carbon disulfide from sour gas is considered in the sweetening process, a mixture of amines can be used [8]. For the simulation of acid gas treatment, Aspen software, such as Aspen Hysys and Aspen Plus, are among the best softwares and the results from the simulation are consistent with the real plant data. This software has a special thermodynamic fluid package for acid gas treatment. Much research has been done in this field with these two softwares. Kamnetochi et al., [9] used Aspen Hysys for natural gas treatment units with different types of Amine. Based on the results from the simulation compared to the real plant data, they concluded that Aspen HYSYS can be used to study the process of acid gas treatment successfully. Al-Lagtah et al., [10] simulated and optimized Lekhwair natural gas sweetening plant using Aspen HYSYS. They made some modifications to this plant to reduce the energy consumption. They also investigated the equipment capacity and limitations for operational optimization. They simulated and discussed two modifications (conventional split-loop and modified split-loop). They concluded by the conventional split-loop that they can save up to 50% of the current operating expenses around £175,000 extra investment, and a penalty of 1.0 ppm of H2S concentration in the sweet gas, which is still well below the pipeline gas specification. Chew et al., [11] optimized an industrial GTU[4] focused on the energy consumption. They used Aspen Hysys software for the process simulation and LINGO software for the optimization. Some operational parameters, such as amine flow rates, absorber pressure, and amine temperature, have been taken into account in the optimization. They also did a multi-subject optimization by using Pareto analysis to maximize the profit and minimize the CO2 content. They concluded that it is possible to increase the profit by 9.15% in the same CO2 content in sweet gas (0.77%) and 23.3% by a higher acceptable CO2 content (1 mol%). They reported the optimized parameters for their study.

Milind M Vaidya et al., [12] proposed a new process to enrich the H2S in the gas from a usual GTU. They proposed a process including an absorber, a regenerator, a compressor, and a membrane unit at the end of their proposed process. This new plant will be installed between the GTU and SRU[5]. The acid gas from GTU with less than 35 H2S mol% enters the new process. A highly concentrated CO2 from the top of the absorber and most of the H2S are absorbed by an amine stream. The rich amine is sent to the regenerator and a gas stream with H2S concentration less than 60 mol% is taken from the top of the regenerator. This gas can be sent to a membrane module to increase the H2S concentration. They claimed that it is possible to increase the H2S concentration up to 90 mol%. Even though they stated that this process can enrich to a very high level, it is predicted that their offered process will not be economical. The reason is that this process is like a new GTU, in addition to a compressor and a membrane unit. Moreover, the price of the compressor for such an acid gas will be very expensive.

Ghavami et al., [13] attempted to enrich the H2S concentration of acid gas from GTU to SRU. They added an acid gas enrichment tower to a normal GTU. In addition, they investigated the ef-

1.Monoethanolamine
2.Triethylamine
3.Diglycolamine

4. Gas Treatment Unit
5.Sulfur Recovery Unit

fect of different parameters, such as MDEA flow rate, MDEA temperature, acid gas flow rate, and AGE[1] tower pressure on the H2S and CO2 concentration of the acid gas to SRU. Furthermore, they studied the energy consumption of GTU. They could increase H2S concentration in the enriched acid gas to 62.45 mole%, at optimum conditions. This means a 24.8% increase in H2S concentration compared with existing plant conditions. The energy consumption was decreased by 25.05 GJ/hr, and CO2 concentration in off-gas was 94.94 mole. Garmroodi Asil and Shahsavand [14] focused on increasing the H2S content of acid gas from GTU to SRU. They tested different solvent mixtures based on having the maximum H2S concentration in the acid gas to SRU. They used Aspen Hysys software for the process simulation and MLP artificial neural network technique for the optimization. They selected the solvent with 45 wt% MDEA and 37% Solfinol as the optimum solvent. They claimed they could increase the H2S concentration in acid gas from 33.5 mol% to 57 mol%. The CO2 content in the sweet gas is 4.5 mol%, which is not acceptable in many standards. Xue et al., [15] did an experimental study focused on selective H2S absorption. They investigated the performance of 35 wt% MDEA and 5 wt% TBEE[2] as composite solvents with MDEA performance. Based on their experiment, they concluded that a blend of MDEA and TBEE is an efficient solvent for selective H2S removal. The absorption was done at atmospheric pressure and 40 °C. The selected blended solvent obtained a 99.79% H2S removal rate and a 22.68% CO2 co-absorption rate, while the MDEA alone solvent achieved a 98.33% H2S removal rate and a 23.52% CO2 co-absorption rate. Pirouzfar et al., [16] designed a CO2 capture and liquefaction to enrich H2S in SRU feed gas from GTU. They could increase the H2S/CO2 ratio to two and investigated the economy of the process by Aspen Economic Analyser. Cui et al., [17] investigated the effect of different parameters on the selective absorption of H2S and CO2 from a syngas to enrich the acid gas to SRU. They designed a new process for selective H2S capture and increased H2S concentration to SRU. They also claimed that they could decrease energy consumption and operational cost. Darani et al., [18] used Aspen Hysys software to optimize an industrial GTU. They used the L9 orthogonal array Taguchi method to design the experiment. The optimization parameters were amine concentrations (25, 28, and 30 wt%), temperatures (40, 50, and 60 °C), and circulation rates of lean amine (220, 240, and 260 m3/h). The Kent−Eisenberg thermodynamic model was used to predict CO2 absorption by an aqueous DEA solution. The optimum condition happened in a DEA concentration of 30 wt%, a temperature of 40 °C, and a circulation rate of 260 m3/h.

A review of the literature shows that many studies have been conducted in different fields of GTU and SRU. The activity related to H2S enrichment is usually focused on new processes such as adding an enrichment tower. To the best of our knowledge, few research has been done on the composite solvents. In the previous studies in this field, some limitations, such as maximum amin concentration and solvent flow rate have not been considered. In the current study, the optimum composite solvent has been defined considering these limitations.

2. Research Method

2.1. Process description

The sour gas enters the unit through a separator to separate heavy hydrocarbons (mostly butane and heavier components), water, and other impurities from the bottom. The gas enters the absorber column at the bottom. In the absorber column, the lean amine solvent, entering from the top, reacts with the sour gas in a counter-current motion, absorbing impurities, mainly carbon dioxide and hydrogen sulfide, from the sour gas. Thus, the main product of the gas refinery, which is the sweet gas, exits from the top of the column and moves towards the dehydration units. Figure 2 shows a general overview of a gas treatment unit with an amine solven.

1.Acid Gas Enrichment
2.2-tertiary butylamine-2-ethoxyethanol

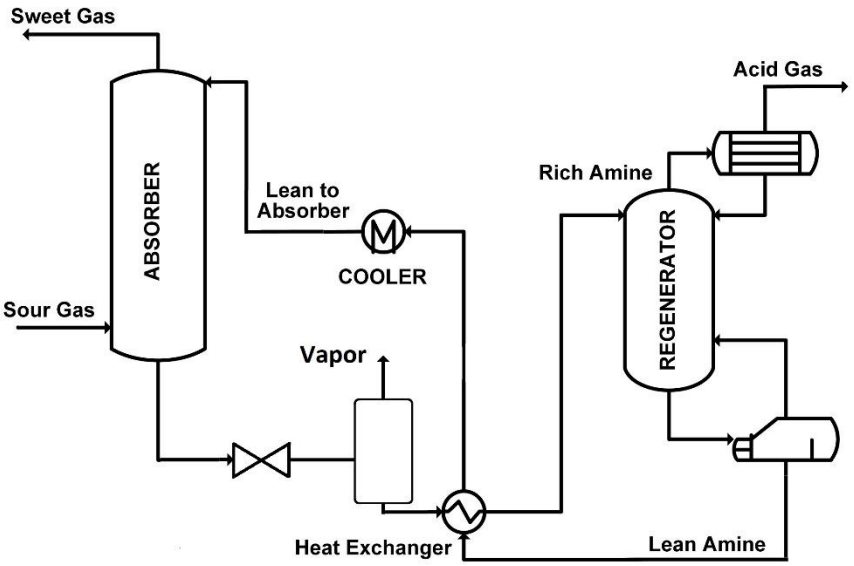

Figure 2. Gas treatment unit

The sour gas enters the unit through a separator to separate heavy hydrocarbons (mostly butane and heavier components), water, and other impurities from the bottom. The gas enters the absorber column at the bottom. In the absorber column, the lean amine solvent, entering from the top, reacts with the sour gas in a counter-current motion, absorbing impurities, mainly carbon dioxide and hydrogen sulfide, from the sour gas. Thus, the main product of the gas refinery, which is the sweet gas, exits from the top of the column and moves towards the dehydration units. Some important design conditions are shown in Table 1 below.

The reaction between CO2 and H2S with a water-amine solution is exothermic. Regardless of the type of amine, H2S will chemically react with amines according to equation 1 by direct proton transfer and the formation of amine hydrosulfide.

$$R_1R_2R_3N + H_2S \leftrightarrow R_1R_2R_3NHHS \quad \text{Eq. 1}$$

Table 1. Some important operational conditions

Parameter	Value	Parameter	Value
Sour Gas Temp (°C)	21	Amine Recycle m3/hr	903.8
Treated Gas Temp. (°C)	53.2	Lean Amine Temp. To Absorber (°C)	53
Absorber pressure (kPa)	7295-7329	Acid Gas H2S to SRU	33.6
Treated Gas H2S (ppm)	4.1	Acid Gas Temp to SRU (°C)	52
Treated Gas CO2 %	0.02	Lean Amine Temp. From Sripper (°C)	121

Unlike H2S, the reaction of amines with CO2 is more complicated and depends on the type of amine, and the reason for this complexity lies in the dual mechanism of the reaction of CO2 with amine. In the first mechanism, bicarbonate will be formed by CO2 hydrolyzing in water, during a slow reaction. Bicarbonate will carry out the following general reactions.

$$CO_2 + H_2O \leftrightarrow H_2CO_3(carbonic\ Acid) \quad \text{Eq. 2}$$

$$H_2CO_3 \leftrightarrow H_2 + CO_3(Bi\ carbonate) \quad \text{Eq. 3}$$

$$H + R_1R_2R_3N \leftrightarrow R_1R_2R_3NH \quad \text{Eq. 4}$$

$$CO_2 + H_2O + R_1R_2R_3N \leftrightarrow R_1R_2R_3NHHCO_3 \quad \text{Eq. 5}$$

The rich amine, now containing acid gas and other impurities, exits from the bottom of the absorber column. The rich amine is then depressurized by a valve, forming a two-phase mixture and enters a flash drum. The gas stream from the upper part, after passing through a packed column and contact with lean amine to reduce the amount of impurities, is then sent to an incinerator for the final disposal. The liquid phase moves towards a shell-and-tube heat exchanger. In the shell-and-tube exchanger, the rich amine, with low temperature, is preheated with lean amine that returns from the regeneration column, which has a high temperature, to reduce the energy consumption of the unit. Then, it enters the regeneration column from the top. In the regeneration column, due to the contact between the rich amine and the hot vapors, the bond between the amine and acid gas is broken, and thus, two main products, i.e., clean amine and acid gas are produced from the bottom and the top of the regeneration column, respectively. It should be noted that some water and amine are lost in the process through output streams or amine degradation. To balance the mass and maintain a constant concentration of the lean amine in the absorption column, the required amount of water and amine is added to the solvent storage tank at various time intervals.

2.2. Simulation

ASPEN-HYSYS V11 software is used for the simulation of an industrial GTU. As mentioned in the previous section, many researchers have used this software for the simulation in this field, and they mentioned that the simulation results are in good agreement with the plant data. The acid gas-chemical solvents thermodynamic package was used to predict the equilibrium between vapor and liquid. This package provides very high accuracy in the gas treatment process with the help of solvents and significantly reduces errors related to the thermodynamic calculations.

2.3. Design and Operational Information

The operational data is needed to be able to simulate the industrial GTU plant. Different data, such as sour gas operating pressure and temperature, the absorber feed composition, and the recycled amine flow rate, pressure, and temperature in the unit are shown in Tables 2, 3, and 4.

Table 2. Sour gas composition to absorber column [19].

Component	Mole (%)	Component	Mole (%)
N2	0.368	C4H10	0.028
CH4	81.901	iC5H12	0.010
CO2	6.459	C5H12	0.014
C2H6	0.509	C6H14+	0.035
H2S	3.588	Benzene	0.015
C3H8	0.058	E-Benzene	0.002
iC4H10	0.008		

Table 3. Sour gas design and operational condition [19].

Parameter	Design	Operational
Temperature (°C)	56	62
Pressure (kPa)	7329	7258
Flow rate (kg/hr)	166500	113228

Table 4. Lean amine design and operational condition [19].

Parameter	Design	Operational
Temperature (°C)	51.5	59.47
Pressure (kPa)	7171	7584
Flow rate (kg/hr)	450300	360070

3. Results and Analysis

3.1. Validation

To verify the accuracy of the simulation results, operational data from an industrial unit were used for the MDEA solvent. The simulation results closely matched the operational data, confirming the accuracy of the results. For example, the temperature profiles of the absorption and regeneration columns are shown in Tables 5 and 6.

3.2. Definition of Solvent Mixture

DEA and MDEA are widely used in the gas processing industry (Gas Processors Suppliers & Gas Processors, 2012). Therefore, in defining mixtures, these two solvents were selected as the base solvent, considering the permissible weight percentage range in the industry, as shown in Table 7.

Table 5. Temperature profile of the absorber column

Absorber column	Simulation temperature (°C)	Operational temperature (°C)	Error %
Tray No. 5	62.7	62.12	+0.009
Tray No. 10	67.63	65.22	-0.031
Tray No. 15	74.93	73.78	+1.56
Tray No. 20	82.16	80.50	+2.06

Table 6. Temperature profile of the regeneration column

Absorber column	Simulation temperature (°C)	Operational temperature (°C)	Error %
Tray No. 6	116.8	108.28	+7.87
Tray No. 12	117.7	117.75	-0.04
Tray No. 18	118.4	119.00	-0.51

Table 7. Permissible Range for Base Solvent in GPSA Reference

Solvent	Allowed range (wt%)
DEA	30-40
MDEA	40-50

Then, other solvents such as TEA[1], DGA, DIPA[2], Piprazine, and Sulfolane were added as additives to the base solvent to achieve a total amine weight percentage of 50%. The remaining amount in the water-amine solution will be attributed to water. It should be noted that in the process of adding additives to the base solvent, minimum, intermediate, and maximum amounts of the base solvents were assigned as constant numerical weight percentages. The remaining amount up to the ceiling of 50 wt% was then allocated to the additives. This is a practical issue that many previous researchers have not paid attention to. The solvent with high concentration has some problems, such as corrosion, foaming, and solvent degradation.

Designing experiments according to features

1.Triethanolamine
2.Diisopropylamine

such as various methods of experiment design, statistical analysis of data, mathematical models, and optimization, the proposed models, DESIGN EXPERT software for experiment design, was used and Aspen-Hysys software version 11 was also used simultaneously. Although the Design Expert software supports a variety of methods for optimization comprising RSM[1], Taguchi Design, K-Optimal Design, D-Optimal Design, Desirability Function Approach, Statistical Analysis, etc., in the present research, Factorial Design optimization technique was used. Factorial Design is a systematic technique employed to examine the effects of multiple factors on a response variable. This method entails conducting experiments across all possible combinations of the factor levels. The design can be either full factorial, where every possible combination is tested, or fractional factorial, where a subset is selected to reduce the number of experiments while still capturing the essential information. This approach enables the identification of main effects and interactions between factors, making it widely used for its efficiency in exploring the relationships between factors and responses in a structured and statistically rigorous manner.

The aim of the study was the simultaneous evaluation of the performance of the combined solvents in the process of sweetening and acid gas enrichment. The experiments were designed based on two factors: the concentration of the basic amine (MDEA) in the combined solvent and the circulating amine flow rate. In addition, in making the composite solvent, considering that the maximum weight percentage of the basic solvent in its industrial application is 50wt%, the weight percentage of water in the water-amine solution will be equal to 50wt%. According to the variation range of the base solvent between 35 and 50wt%, the weight percentage of the additives in the composite solvent will vary between 0 and 15% by weight. To prevent foaming and flooding of the absorption column, a permissible range for the amine flow rate in the circulation was established in the designed experiments using Design-Expert software, with values ranging from 440 to 985 m3/hr, as presented in Table 8.

1.Response Surface Methodology

Table 8. Designed experiments by design-expert software for evaluating the performance of the composite solvent

Absorber column	Simulation temperature (°C)	Operational temperature (OC)	MDEA wt% in composite solvent	Amin Flow (m3/hr)	Additives and weight %
42.5	985	DGA, 7.5%	35	440	Sulfolane, 15%
42.5	712.5	DGA, 7.5%	50	712.5	-
50	440	-	42.5	712.5	DIPA, 7.5%
42.5	712.5	DEA, 7.5%	42.5	712.5	Sulfolane, 7.5%
35	712.5	Piprazine, 15%	50	985	-
42.5	712.5	Piprazine, 7.5%	35	985	TEA, 15%
42.5	440	Sulfolane, 7.5%			

3.3. Characteristics and Operational Constraints in Simulation

To assess the impact of the designed mixtures on the sweetening process, it is necessary to introduce characteristics used in the unit and then compare the results with them in all simulation stages. The primary characteristic in the mentioned unit is the standards set by the National Gas Company of Iran as the purchaser of sweet gas, which includes 4 ppm of hydrogen sulfide and a maximum of 2 mol.% carbon dioxide in the delivered sweet gas [19].

Another characteristic relates to the gas stream sent to the incinerator, where the amount of hydrogen sulfide in it should be less than 100 ppm according to environmental standards. The third characteristic under consideration is the composition of the acid gas percentage as the feed to the sulfur recovery unit, and its design data are reported in Table 9.

Table 9. H2S and CO2 Levels in the Feed to the Sulfur Recovery Unit

Component	Mol. %
CO2	56.3
H2S	33.6

The fourth characteristic is related to the maximum amine flow rate in the circulation, which should not exceed 985 cubic meters per hour. The last characteristic is the maximum allowable loading of amine in the amine path, which should be a maximum of 5.0 based on the industrial information (SGPC, 2004).

Considering the constraints and operational conditions mentioned in the previous sections, the design of mixtures based on flow rate and the permissible range of MDEA and the base solvents were carried out using the DESIGN EXPERT experimental design software. Figure 3 illustrates the changes in the mole percentage of hydrogen sulfide in the acid gas for the optimal mixture. The experimental design software suggested 100 mixtures at different flow rates. After simulating all 100 cases using the HYSYS software and comparing the results with the constraints, the MDEA and Sulfolane mixture was identified as the optimal solvent because it provides the highest H2S concentration in the acid gas to SRU unit. Below, some of the cases and results are analyzed. Based on Figure 3, some other points provide more H2S concentration in the SRU feed gas, but for all of them, some constraints, such as H2S or CO2 amount in sweet gas or H2S concentration in off-gas, were violated. Therefore, they are not acceptable operating points.

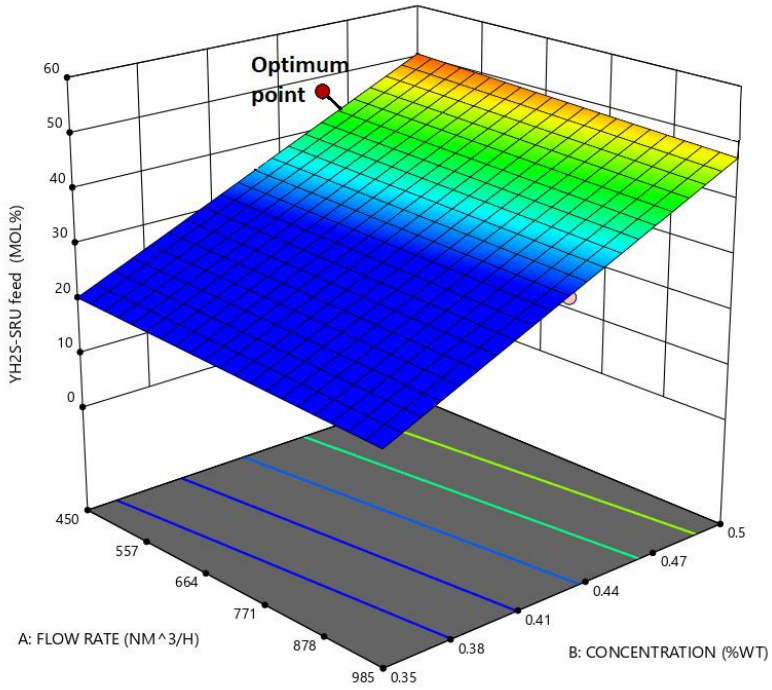

Figure 3. Changes in the Mole Percentage of Hydrogen Sulfide Gas in Acid Gas

Adding TEA with 7.5 wt% to the MDEA base solvent in the condition that the amine flow rate in circulation was 440 m3/hr and the MDEA concentration in the combined solvent was 42.5 wt%, could sweeten the sour gas with the desired characteristics and increased the amount of H2S enrichment in the acid gas stream from 33.6 to 39.93. The addition of piperazine in the aforementioned conditions caused a decrease in H2S enrichment from 33.6 to 32.29 mol% in acid gas flow.

Therefore, reducing the amount of enrichment of the combined solvent made based on this additive was not considered. The simulation results of the combined solvent obtained by adding DGA with 7.5 wt% to the basic solvent of 42.5 wt% MDEA and water at a flow rate of 440 m3/hr showed the negative performance of this solvent in the sweetening process of sour gas despite an increase in the enrichment rate. In this combined solvent, the amount of H2S enrichment has increased from 33.6 mol% to 39.95 mol%, but the sweet gas exiting from the top of the absorption tower has characteristics beyond the maximum standards, i.e., 8 ppm of hydrogen sulfide and 2.12 mol% carbon dioxide. Therefore, according to the stated results and due to the weakness of the combined solvent mentioned above in the sweetening of sour gas, this combined solvent could not be considered as an optimal solvent.

In the conditions where the concentration of MDEA solvent in the combined solvent is 42.5wt% and Sulfolan substance with 7.5wt% was used as an additive, the amount of hydrogen sulfide in the sweet gas stream was 0 ppm and the amount of carbon dioxide was 2.3% by mole. The amount of H2S enrichment in the acid gas stream also increased from 33.6 to 41.84 mol%. These results show that due to the good performance of sulfolane additive in increasing the enrichment of hydrogen sulfide, this additive can be considered as a suitable additive in the production of the mixed solvent.

Table 10. shows some comparisons between the selected composite solvent and the two base solvents.

Solvent	Lean Amine Flow (m3/hr)	Treated Gas CO2 %	Treated Gas H2S (ppm)	Acid Gas H2S %
DEA 35%	452	0.02	4.1	33.6
MDEA 50%	452	1.07	4.18	38.7
MDEA 42.5%, Sulfulane 7.5%	455	1.95	4.12	41.84

Based on standard treated gas, the CO2 concentration can be up to 2 mol% and H2S about 4 ppm. One of the parameters that compsite solvent can increase the H2S concentration of acid gas is the selectivity. As the results in Table 10 show, a composite solvent can absorb less CO2 and it is the reason that H2S concentration increases in the acid gas from stripper.

4. Conclusion

The low concentration of hydrogen sulfide gas in the acid gas results in a decrease in the efficiency of the sulfur recovery process, leading to increase the operational costs and environmental pollution. With the continuous tightening of the environmental standards, the need for acid gas treatment to improve efficiency and reduce energy consumption becomes more crucial. In this study, the impact of different amine mixtures on the sweetening of a gas treatment unit was investigated. The key parameter for the optimum solution was the H2S concentration of acid gas to the Sulphur Recovery Unit. The higher value can increase the Sulphur recovery and decrease the pollution to the atmosphere. To select the optimum case, some operational constraints such as sweet gas H2S and CO2 concentration, amine recycle flow rate, and off-gas H2S concentration were investigated.

As stated, the addition of Sulfolane by 7.5wt% to the MDEA base solvent at a 42.5wt%, improved the concentration of hydrogen sulfide in the acid gas stream by 24.5%. The composite solvent can improve the Acid gas H2S concentration because of higher selectivity to H2S compared to the base solvent. The results showed that the treated gas CO2 concentration is increased when the composite solvent is used and it is the main reason that H2S concentration increases in acid gas.

The results of this study revealed that it is pos-

sible to increase the H2S concentration to SRU with the minimum cost by using a composite solvent instead of a current solvent. The cost for this solution is much less than other solutions, such as adding an enrichment tower for increasing H2S concentration in acid gas.

Acknowledgment

In this study, the industrial data from Khangiran gas refinery was used.

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مقاله تحقیقاتی

تعیین حلال بهینه واحدهای تصفیه گاز به منظور غنی‌سازی گاز اسیدی

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چکیده

افزایش غلظت S2H در گاز اسیدی ورودی به واحد بازیابی گوگرد، یکی از پارامترهای کلیدی برای افزایش راندمان بازیابی گوگرد است. هدف این تحقیق تعیین حلال بهینه برای یک واحد تصفیه گاز با تمرکز بر غنی‌سازی گاز سولفید هیدروژن در خوراک واحد بازیابی گوگرد است. برای دستیابی به اهداف تحقیق، از داده‌های طراحی یک واحد تصفیه گاز در مقیاس صنعتی استفاده شد. برای شبیه‌سازی، تجزیه و تحلیل نتایج و رسم نمودارها، به ترتیب از نرم‌افزارهای 11V SYSYH-NEPSA و 01V TREPXE NGISED استفاده گردید. حلال‌های موجود در کتابچه راهنمای ASPG برای انتخاب حلال و افزودنی‌ها استفاده شدند. در فرمولاسیون یک حلال ترکیبی، حلال‌های AED و AEDM با توجه به محدوده درصد وزنی مورد استفاده در صنعت فرآوری گاز، به عنوان حلال‌های پایه و سایر حلال‌ها به عنوان افزودنی به حلال ترکیبی در نظر گرفته شدند. نتایج شبیه‌سازی با مقایسه آنها با استانداردهای شرکت ملی گاز ایران مورد تجزیه و تحلیل قرار گرفت. نتایج نشان داد که ترکیب AEDM با ۲۴/۵ درصد وزنی به عنوان حلال پایه و enalofluS با ۷/۵ درصد وزنی به عنوان افزودنی، حلال بهینه خواهد بود. دلیل این انتخاب، وجود حداکثر غلظت S2H در گاز اسیدی ضمن رعایت استانداردهای ذکر شده است. حلال بهینه می‌تواند غلظت S2H را از ۳۳/۶ درصد مولی به ۱۴/۴۸ درصد مولی در گاز اسیدی ورودی به واحد بازیابی گوگرد افزایش دهد.

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