ABSTRACT: The presence of carbon dioxide in natural gases can lower the quality of natural gas and can cause CO₂ freezing problems. Therefore, using reliable techniques for the reduction and elimination of carbon dioxide from natural gases is necessary. The aqueous diethanol amine (DEA) solution’s ability to simultaneously absorb H₂S and CO₂ from sour natural gases makes it possible to use this solution in the natural gas sweetening process. The goal of this work was to determine the maximum amount of the removed CO₂ by an aqueous DEA solution in one of the gas sweetening plants of the National Iranian South Oilfields Company (NISOC). For this purpose, based on the obtained designed experiment results using the L9 orthogonal array Taguchi method, the experiments were conducted and three levels of 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⁻¹) were considered as the key operational parameters on CO₂ removal. To evaluate the ability of the HYSYS simulation software and the Kent–Eisenberg thermodynamic model to predict CO₂ absorption by an aqueous DEA solution in the gas sweetening process, the field data were compared with the results of the simulation. It was observed that the maximum removal of CO₂ is achieved at a lean amine concentration of 30 wt %, a temperature of 40 °C, and a circulation rate of 260 m³ h⁻¹. Also, the experimental results indicate that the effects of the selected process variables on CO₂ absorption are not linear and the most effective parameter on carbon dioxide removal is the concentration of amine in an aqueous solution and the temperature of the lean amine has the least effect. Besides, the obtained simulation results are in the range of the unit design basis but have some deviations from field data. The findings of this study can help in better understanding of the selection of the effective variables in the natural gas sweetening process and obtaining their appropriate values to achieve the highest efficiency.
that can include about 70% of the total cost of operation of the gas purification process.\(^9\) Besides, vapor–liquid equilibrium (VLE) modeling of acid gas–aqueous amine systems is necessary for the synthesis, design, and analysis of gas sweetening units. There are two categories of VLE models for the description of gas–aqueous amine systems: the empirical models based on the Kent–Eisenberg model and activity coefficient- or excess Gibbs energy ($G^\mathrm{ex}$)-based models.

Kent and Eisenberg proposed a VLE model to predict the equilibrium partial pressures of $\text{H}_2\text{S}$ and $\text{CO}_2$ in aqueous monoethanolamine (MEA) and diethanolamine (DEA) solutions.\(^10\) Jou et al. applied such an approach for the correlation of $\text{H}_2\text{S}$ and $\text{CO}_2$ solubilities in aqueous methyl diethanolamine (MDEA) solutions.\(^5\) Moreover, Chakma and Meissen extended the Kent–Eisenberg approach for the system of $\text{CO}_2$–DEA–$\text{H}_2\text{O}$.\(^11\) According to Weiland et al., the Kent–Eisenberg correlation results show a good agreement with experimental data only in the loading range of 0.2 to 0.7 moles acid gas per mole of amine, and the model gives inaccurate results for mixed acid gases.\(^1\) Haji-Sulaiman et al. extended the Kent–Eisenberg model to estimate the $\text{CO}_2$ loading in the aqueous mixtures of DEA, MDEA, and DEA–MDEA, and it was observed that this model forecasts a relatively accurate carbon dioxide loading over a wide range of operating conditions.\(^1\) Ebenezer evaluated the HYSYS capability to estimate the $\text{CO}_2$ removal at operating conditions of minimizing hydrocarbon and chemical losses.\(^14\) Also, Aliabad and Mirzaei studied the accuracy of HYSYS and ASPEN simulators in gas sweetening forecasting by aqueous amine solvents.\(^15\) The Aspen HYSYS software was also applied to simulate and optimize the gas sweetening process using several amine types and blends, for example, DEA,\(^16\) MEA,\(^16\) MDEA,\(^16–18\) DGA,\(^19\) and MEG\(^20,21\) aqueous solutions and the effects of operating conditions such as circulation rate, concentration, and inlet temperature of amine on the regeneration reboiler temperature and duty were studied.

So far, no comprehensive study has been conducted on the effects of process parameters on increasing the efficiency of natural gas sweetening units in Iran, and it seems that the study on this issue is very important to reduce production and natural gas sweetening units in Iran, and it seems that the study conditions.\(^13\) Ebenezer evaluated the HYSYS capability to simulate and optimize the gas sweetening process using several simulators in gas sweetening forecasting by aqueous amine systems is

\[\text{The magnitude of the values in the last column of Table 3. The average value of the maximum and minimum amounts of S/N ratios and is listed in Table 3. The magnitude of the values in the last column (I}_{max} - I}_{min} \right) shows the influence of the factors and the minus}

\[\text{have prevented a more complete study by investigation on further variables with a wide range of variations.}

In this article the definitions are presented. Then, the operating conditions and control are presented. In the Experimental Procedure and Sampling section, different stages of the experiments and sampling are described, and in the Steady-State Simulation and Optimization section, the capabilities of Aspen HYSYS and the Kent–Eisenberg thermodynamic model in the $\text{CO}_2$ absorption estimation by an aqueous DEA solution in this case study are investigated. The results of the experimental and modeling study are reported as tables and figures in the Results and Discussion section and their evaluations are presented in this section. In the Conclusion section, an overview of the obtained results is provided.

### RESULTS AND DISCUSSION

#### Experimental and Analysis Results

The measured $\text{CO}_2$ concentration in the sweet natural gas for each trial and sample is listed in Table 1. It can be observed that trial 2 has the highest variation in its results and also the proposed operational conditions in trial 7 including a temperature of 40 °C, a DEA solution concentration of 30 wt %, and an amine circulation rate of 260 m$^3$ h$^{-1}$ lead to the best $\text{CO}_2$ removal result. The results of the signal-to-noise ($S/N$) ratio for each trial are presented in Table 2.

The effect of each factor was estimated by calculating the average value of the $S/N$ ratios at the total levels of the factor. The factor effect was the arithmetic difference between the maximum and minimum amounts of $S/N$ ratios and is listed in Table 3. The magnitude of the values in the last column ($I_{max} - I_{min}$) shows the influence of the factors and the minus
the CO2 concentration in sweet natural gas as the target, performed under controlled conditions.22,23 did not vary during the experiments, and the experiments were best estimate, it could be assumed that unconsidered factors percent contribution of error is low, 11.414%, and with the lable factors causing it. The ANOVA results show that the experiments and experimental error, interactions, or uncontrol-percentage of error contribution points to the accuracy of the weight % (level 1 to 2) leads to an increase in the CO2 absorption concentration of the aqueous DEA solution from 25 to 28 strongly. Also, it can be observed that the increasing amine has the highest impact on the response and an increase in lean amine concentration causes CO2 removal to increase in Table 4. Analysis of Variance

| trial number | amine concentration (wt %) | temperature (C) | circulation rate (m3h-1) | results (CO2 concentration in sweet natural gas) (ppm, mole) |
|--------------|-----------------------------|-----------------|--------------------------|---------------------------------------------------------------|
| 1            | 25                          | 40              | 220                      | 1, 6                                                          |
| 2            | 25                          | 40              | 240                      | 1, 8                                                          |
| 3            | 25                          | 50              | 260                      | 1, 8                                                          |
| 4            | 28                          | 40              | 240                      | 1, 4                                                          |
| 5            | 28                          | 50              | 260                      | 1, 4                                                          |
| 6            | 28                          | 60              | 220                      | 1, 19                                                        |
| 7            | 30                          | 40              | 260                      | trace (0.005)                                                 |
| 8            | 30                          | 50              | 220                      | 1, 05                                                        |
| 9            | 30                          | 60              | 240                      | 1, 3                                                          |

Table 2. S/N Ratios of Trials

| trials | S/N ratio |
|--------|-----------|
| 1      | -16.990   |
| 2      | -31.623   |
| 3      | -16.484   |
| 4      | -12.042   |
| 5      | -10.970   |
| 6      | -25.347   |
| 7      | 46.020    |
| 8      | -5.374    |
| 9      | -6.990    |

Table 3. Main Effects of the Individual Factors

| factor          | level 1 | level 2 | level 3 | Lmax−Lmin |
|-----------------|---------|---------|---------|-----------|
| concentration   | -21.699 | -16.119 | 11.919  | 33.618    |
| temperature     | 5.663   | -15.289 | -16.274 | 21.937    |
| Rate            | -15.204 | -16.885 | 6.189   | 23.074    |

design indicates a decrease in the S/N ratio as a variety of levels. Preliminary reviews show that the amine concentration has the greatest difference in the levels, and consequently, it can affect the CO2 concentration in sweet natural gas as the target, significantly.

The evaluation of the influence of the individual factor was obtained from the analysis of variance (ANOVA) and is shown in Table 4. It can be revealed that the amine concentration with a contribution percent of 46.14% plays the most significant influence on the process of CO2 removal. The percentage of error contribution points to the accuracy of the experiments and experimental error, interactions, or uncontrollable factors causing it. The ANOVA results show that the percent contribution of error is low, 11.414%, and with the best estimate, it could be assumed that unconsidered factors did not vary during the experiments, and the experiments were performed under controlled conditions.

To investigate the main effects of the factors on the trail conditions, the S/N ratio average for each parameter was plotted versus the various levels and is illustrated in Figure 2.

According to the slope of the lines, the concentration of lean amine has the highest impact on the response and an increase in lean amine concentration causes CO2 removal to increase strongly. Also, it can be observed that the increasing concentration of the aqueous DEA solution from 25 to 28 wt % (level 1 to 2) leads to an increase in the CO2 absorption gradually and increases the absorption steeply from 28 to 30 wt %.

Thus, it is concluded that the effect of lean amine concentration on CO2 absorption is not linear and it can be a source of error for estimating Yexp. The second section of Figure 2 describes the response graph as a function of the lean amine circulation rate. As can be seen, such as lean amine concentration, an increase in the circulation rate increases the carbon dioxide concentration but with a lower slope. The circulation rate from 220 to 240 m3 h-1 (level 1 to 2) has no weighty impact on the target, but from 240 to 260 m3 h-1 (level 2 to 3), the CO2 absorption increases sharply. The effect of the lean amine temperature changes on the CO2 absorption is pictured in the third section of Figure 2. Raising the lean amine temperature causes the S/N ratio to decrease. It is expected that temperature increase reduces gas absorption. However, reduction of the S/N ratio in the alteration temperature from 40 to 50 °C (level 1 to 2) is sharp and in another temperature alteration from 50 to 60 °C is almost constant. The effects of the last two factors are nonlinear too (as lean concentration) and cause errors.

Estimated Results at the Optimum Condition. To calculate the expected results at the optimum condition, the grand average of performance, R0, was evaluated. All factors in this work were significant and the performance at the optimum condition should be calculated using all of them. According to Table 3, a high level of lean amine concentration factor, a low level of temperature factor, and a high level of lean amine circulation rate factor were given at the highest S/N values and were considered as the optimal condition. The contribution of each factor was calculated and is presented in Table 5 along with the optimum settings and levels of the various factors.

Since the S/N ratio is used, the estimated result at the optimum condition can be converted back to the scale of original observation units. In this case, the expected result in

Table 4. Analysis of Variance

| factors       | DOF | S-S    | V     | F     | S'    | percent P (%) |
|---------------|-----|--------|-------|-------|-------|---------------|
| concentration | 2   | 1947.419 | 973.709 | 17.172 | 1834.017 | 46.14         |
| temperature   | 2   | 921.154 | 460.577 | 8.122  | 807.751 | 20.321        |
| Rate          | 2   | 992.87  | 496.435 | 8.755  | 879.468 | 22.125        |
| Error         | 2   | 113.402 | 56.701 | 17.172 | 19 (0.005) | 11.414        |
| Total         | 8   | 3978.848 |        |       |       | 100           |
terms of the S/N ratio is 41.036. This is equivalent to an average performance $Y_{\text{exp}} = 0.009$, which is calculated using eqs 1 and 2.

$$\text{MSD} = 10^{\left[-\frac{S}{N}\right]/10}$$  \hspace{1cm} (1)

$$Y_{\text{exp}} = (\text{SD} \times 0.5)$$  \hspace{1cm} (2)

**Interaction Study.** The difference between the experimental and the estimated values can be related to the interaction between the control factors. Hence, an interaction study was performed with the software Qualitek-4. The intensity of the presence of interactions was measured in terms of a numerical quantity via the angle between the two lines of the selected factors. Table 6 shows the interacting pair factors and their severity index (S-I). The last column indicates the desirable levels to achieve the optimum condition. As can be observed, temperature and the circulation rate have the highest interaction with an S-I of 21.44%. Also, it can be found that the interaction indices are negligible, which is predictable because the factors are independent and could not interact with each other.

**Simulation Results.** To evaluate the capability of Aspen HYSYS in estimating the CO₂ absorption by an aqueous DEA solution in the gas sweetening process, the field data were compared with the simulation results. Thus, after running each trial and sample analysis, the operating conditions of each trial were fed to Aspen HYSYS and the process was stimulated by a selection of amine packages and the Kent–Eisenberg thermodynamic model.

The input data for the absorption column in the simulation was as follows: the pressure at the top of the tower was 28 bar, the pressure at the lower part was 29.75 bar, and the number of trays was 20. For the stripping column simulation, the pressures at the top and the lower part of the column were 1.7 and 2.2 bar, respectively. Also, the number of trays was 24, and to converge the stripping column, the temperatures of the condenser and reboiler were specified with the values of 60 and 125 °C, respectively.

According to the unit design basis, the lean amine loading must be less than 0.02 mole (CO₂ + H₂S)/mole DEA. The amount of lean amine demonstrates the quality operation of the regeneration unit and increasing the amount of lean amine loading leads to a decrease in acid gas absorption. As mentioned earlier, after running each trial, lean amine was analyzed, and keeping the lean amine loading below the designed quantity leads to the operation of the reboilers of the regeneration unit in the maximum duty. Besides, rich amine loading was controlled by adjusting the amine circulation rate.

## Table 5. Optimum Condition of CO₂ Removal by an Aqueous DEA Solution

| control factor | level description | level | contribution |
|----------------|-------------------|-------|--------------|
| concentration  | 30                | 3     | 20.551       |
| temperature    | 40                | 1     | 14.296       |
| circulation rate| 260               | 3     | 14.822       |

## Table 6. Table of the Test of Interactions

| interacting factor pair | column interaction | S-I % | optimum level |
|-------------------------|--------------------|-------|---------------|
| temperature × rate      | 2 × 3              | 21.44 | [1,3]         |
| concentration × rate    | 1 × 3              | 17.99 | [3,3]         |
| concentration × temperature| 1 × 2          | 10.11 | [3,1]         |

## Table 7. Comparison between Field and Simulation Data for CO₂ Content and Loading

| trial No | CO₂ concentration in sweet natural gas (ppm mole) | loading (mole CO₂ + mole H₂S)/mole DEA | lean amine | lab | simulation | error % | rich amine | lab | simulation | error % |
|----------|--------------------------------------------------|----------------------------------------|------------|-----|------------|---------|------------|-----|------------|---------|
| trial 1  | 6                                                | 216                                    | 0.0134     | 0.4929 | 0.4739     | −3.85   | 0.0124     | 0.4011 | 0.4216     | 5.11    |
| trial 2  | 35                                               | 70                                     | 0.0120     | 0.4016 | 0.4142     | 3.14    | 0.0090     | 0.3602 | 0.4752     | 31.93   |
| trial 3  | 5                                                | 30                                     | 0.0115     | 0.4639 | 0.3948     | −14.90  | 0.0103     | 0.3475 | 0.4530     | 30.36   |
| trial 4  | 4                                                | 175                                    | 0.0106     | 0.3922 | 0.4139     | 5.53    | 0.0122     | 0.3453 | 0.3992     | 15.61   |
| trial 5  | 3                                                | 48                                     | 0.0111     | 0.3280 | 0.3558     | 23.54   | 0.0111     | 0.2880 | 0.3558     | 23.54   |
According to the basic design of the unit, the amount of rich amine loading must be kept at 0.5 mole (CO₂ + H₂S)/mole DEA. The high acid gas loading enhances steel corrosion and consequently increases the iron sulfide content, which amplifies foaming tendency. The experimental and estimated CO₂ contents in the sweet natural gas and amine loadings are compared in Table 7. It is found that the estimated CO₂ concentration in the sweet natural gas by Aspen HYSYS is much greater than that of field data. Hence, the Kent−Eisenberg model in HYSYS is not an appropriate model for this process. Consequently, simulated amine loadings have some deviations from field data, but both of them are in the range of design basis of the plan.

The experimental data and simulated results of H₂S content in the sweet gas are presented in Table 8. As can be seen, there is a significant difference between the experimental and simulated values of H₂S content in the sweet natural gas, while the operating conditions are controlled to achieve a value of less than 4 ppm of H₂S in the sweet natural gas. Also, it is observed that regardless of the circulation rate, in the same level of lean amine concentration, such as trials 1 to 3 or trials 4 to 6, increasing the amine temperature leads to the reduction of the estimated CO₂ content and an increase in the simulated H₂S content in sweet natural gas (Figures 3−5). It seems that at a higher temperature, the kinetic effect is stronger than solubility decrease, and the H₂S concentration in the sweet natural gas increases monotonically with lean amine temperature due to the decreasing solubility.

CONCLUSIONS
In this study, the experimental design method was applied to formulate the experimental layout and optimize the operating conditions of a natural gas sweetening unit to remove the maximum CO₂ by an aqueous DEA solution. For this purpose, an L9 orthogonal array Taguchi method as a statistical experimental design was applied, and the effects of lean amine concentration, temperature, and circulation rate in three levels as the key control factors on CO₂ absorption by an aqueous DEA solution were investigated. The experiments were carried out in the Amak GTP at amine concentrations of 25, 28, and 30 wt %, the temperatures of 40, 50, and 60 °C, and the lean amine circulation rates of 220, 240, and 260 m³ h⁻¹. Furthermore, the accuracy of the Aspen HYSYS Kent−Eisenberg thermodynamic model was evaluated and field data and simulation results were compared with each other. From this study, the following conclusions can be drawn:

1. It is revealed that the lean amine concentration is the most significant control factor on CO₂ absorption using aqueous DEA solution with a contribution percent of about 46.14%, while the lean amine circulation rate and temperature have a contribution of 22.125 and 20.321%, respectively.

2. Field data indicate that the CO₂ content for all the trials is less than 50 ppm mole.

3. The ANOVA-calculated error percent is lower than 15% (11.414%). This means that all of the key operating parameters are considered and no significant control factors are left out from the experimental condition.

4. The interaction study indicates that there is no significant interaction between the control parameters.

5. The maximum removal of CO₂ from natural gas by an aqueous DEA solution is achieved at a lean amine concentration of 30 wt %, a temperature of 40 °C, and a circulation rate of 260 m³ h⁻¹.

6. The effects of concentration, temperature, and circulation rate of lean amine on CO₂ absorption are not linear.

7. The estimated results from the Kent−Eisenberg thermodynamic model in the Amine Package of Aspen HYSYS process simulator show deviations from field data.

### Table 8. Comparison between Field and Simulation Data for CO₂ and H₂S Content in the Sweet Natural Gas

| trial no | CO₂ concentration in the sweet natural gas (ppm mole) | H₂S concentration in the sweet natural gas (ppm mole) |
|----------|-----------------------------------------------------|-----------------------------------------------------|
|          | lab 1 | lab 2 | Simulation | lab 1 | lab 2 | simulation |
| trial 1  | 6     | 8     | 216        | 0     | 0     | 1.04       |
| trial 2  | 35    | 41    | 70         | 0     | 0     | 1.63       |
| trial 3  | 5     | 8     | 30         | 0     | 0     | 2.98       |
| trial 4  | 4     | 4     | 175        | 1     | 1     | 0.75       |
| trial 5  | 3     | 4     | 48         | 1     | 1     | 1.50       |
| trial 6  | 19    | 18    | 35         | 0     | 0     | 2.56       |
| trial 7  | 0.005 | 0.005 | 147        | 0     | 0     | 0.58       |
| trial 8  | 0.5   | 2     | 68         | 0     | 0     | 1.90       |
| trial 9  | 1     | 3     | 22         | 0     | 0     | 2.29       |

![Figure 3. Lean amine temperature vs. acid gas concentration in sweet natural gas at a lean amine concentration of 25 wt %](https://doi.org/10.1021/acsomega.1c00744)
data. Thus, this simulator is not a proficient simulator for the determination of CO₂ concentration in sweet natural gas in the natural gas sweetening process.

(8) Regardless of the circulation rate, in the same level of lean amine concentration, increasing the amine temperature leads to the reduction of the estimated CO₂ content and increase of the simulated H₂S content in sweet natural gas.

(9) A decrease in solubility leads to an increase in the concentration of hydrogen sulfide in the sweet natural gas with the increasing temperature of lean amine.

(10) Simulated lean and rich amine loadings show some deviations from field data, but both of them are in the range of the unit design basis.

It seems that if there are no operational restrictions in changing the parameters affecting the process, more findings would be obtained and the results would be more generalizable.

**Definition. Process Description.** The unit was designed to reduce the H₂S content in the sweet gas and sweet liquid to below 4 and 50 ppmv, respectively. As can be observed in Figure 5, the unit is divided into three sections: slug catchers, gas sweetening–amine trains A and B, and liquid sweetening. The feed gases are different sour gases from different stations that are mixed and enter the slug catchers SC-801A/B. In this stage, the mixture is condensed and separated. The obtained liquid is fed to the liquid sweetening unit for sulfur removal utilizing a stripping stream and the gas portion is fed to a gas sweetening unit for H₂S and CO₂ removal by using an aqueous DEA solution. The treated gas, which contains less than 4 ppmv of H₂S and 0.4% mole of CO₂, is sent to the NGL 700/800 unit and some of it is used as a stripping medium in the liquid sweetening unit to reduce the H₂S content in the treated liquid up to less than 50 ppmv and enters the NGL 700/800 unit. The produced acid gas goes to the compression station for further treatment.

**Process Chemistry.** The acid gas absorption is not a physical process only. Only one fraction of H₂S ionizes in water to hydrogen ions and sulfide ions:

\[
H_2S + H_2O \leftrightarrow (H_3O^+) + (HS^-) \tag{3}
\]

DEA is a weak base and ionizes in water to form amine ions and hydroxyl ions:

\[
R_2NH + H_2O \leftrightarrow (R_2NH_3^+) + (OH^-) \tag{4}
\]

where R indicates the ethanol group CH₂CH₂OH. While H₂S dissolves into a solution containing amine ions, a weakly bonded salt of the acid and the base is produced as below, and then, the sulfide ion is absorbed by the amine solution:

\[
(R_2NH_3^+) + (HS^-) \leftrightarrow R_2NH_2SH \tag{5}
\]
Based on reaction 5, the reaction of salt formation is not complete. As the arrows indicate, an equilibrium level of $\text{H}_2\text{S}$ remains in the hydrocarbon flow. Thus, the overall reaction can be summarized as follows\(^{25}\)

\[
\text{R}_2\text{NH} + \text{H}_2\text{S} \leftrightarrow \text{R}_2\text{NH}_2\text{SH}
\]

(6)

Operating variables are set to raise forward reaction 6 in the absorption step and enhance the reverse reaction in the regeneration step. The CO$_2$ absorption is achieved according to the following reactions\(^{25}\)

\[
\text{CO}_2 + \text{H}_2\text{O} \leftrightarrow \text{H}_2\text{CO}_3
\]

(7)

\[
2\text{R}_2\text{NH} + \text{H}_2\text{CO}_3 \leftrightarrow (\text{R}_2\text{NH})_2\text{CO}_3
\]

(8)

\[
2\text{R}_2\text{NH} + 2\text{H}_2\text{CO}_3 \leftrightarrow (\text{R}_2\text{NH})\text{HCO}_3
\]

(9)

The CO$_2$ absorption is slower than the H$_2$S absorption because reaction 7 is carried out slowly and occurs first. The rate of all absorption reactions is enhanced at high pressures and low temperatures, and also, high H$_2$S and CO$_2$ contents shift the equilibrium reactions toward the right side. The amine regeneration is performed at low pressures and high temperatures and shifts the equilibrium of the mentioned reactions to the left side. On the other hand, the low H$_2$S and CO$_2$ partial pressures of the generated stripping vapor in the reboiler lead to a high driving force for the H$_2$S and CO$_2$ mass transfer.

**Operating Conditions and Control.** The absorption of H$_2$S/CO$_2$ into the aqueous amine solution is increased by five factors: low temperature, low acid gas loading, high amine concentration, high H$_2$S/CO$_2$ partial pressures in the feed stream, and intimate contact. In general, the fourth and fifth factors are not operating variables and are fixed by the unit design criteria and choosing equipment in the absorbers’ design. Furthermore, low feed rates may, however, cause poor tray efficiency and thus somewhat a poor H$_2$S/CO$_2$ removal in comparison with achievable at or near design flow rates. In general, decreasing the temperature of the lean amine solution causes an increase in H$_2$S/CO$_2$ removal. Besides, the lean amine temperature must be maintained at 10 °C higher than the temperature of the gas feed stream to avoid any possible condensation of the hydrocarbon vapors. The lean amine is cooled typically by air to about 60 °C.

It should be noted that the acceptable acid gas removal efficiency depends on good aqueous amine solution regeneration and restricting the H$_2$S/CO$_2$ loading in the rich amine to favor the forward direction of reaction 6. The H$_2$S/CO$_2$ loading of the aqueous amine solution is controlled by adjusting the amine circulation rate.

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**EXPERIMENTAL PROCEDURE AND SAMPLING**

As mentioned earlier, this study was undertaken on the GTP of Amak. To measure accurate and reliable data, all flow instruments of sour gas, sweet gas, lean amine, and rich amine were calibrated. After calibration, operating variables, including lean amine concentration, temperature, and circulation rate of the trial, were set and each trial was run. When a steady state was achieved, flow sampling was performed. The CO$_2$ content of two obtained samples from sweet gas was measured, and the analysis of two samples of sour gas as a feed was performed to simulate a process with field data. The carbon dioxide and hydrogen sulfide contents in the gas samples were analyzed by the gas chromatography method. For this purpose, a calibrated Agilent 6890 series gas chromatograph equipped with a DB-1 capillary column was
applied. The following chromatographic conditions were set to provide accurate results: manual splitless injection, an inlet temperature of 105 °C, a total gas flow rate of 30 mL min⁻¹, a column gas flow rate of 2 mL min⁻¹, and a detector outlet temperature of 200 °C. Helium was used as a carrier gas. The H₂S + CO₂ loading of lean amine must be less than 0.02 and it was checked for two samples after running each trial. It should be noted that the reboiler in the regeneration unit of GTP operated at the maximum duty to achieve the minimum amount of acid gas loading for lean amine and the acid gas loading of rich amine shall not exceed 0.5 mole H₂S + CO₂ per mole of amine and it was checked for two samples of rich amine.

**STEADY-STATE SIMULATION AND OPTIMIZATION**

The capability of the Aspen HYSYS and Kent–Eisenberg thermodynamic model in the CO₂ absorption estimation by an aqueous DEA solution in the Amak gas sweetening process was evaluated, and the field data were compared with the simulation results, as mentioned earlier.

The Taguchi method and the software Qualitek-4 were applied to optimize the operating variables of the CO₂ removal process from sour gas including lean amine concentration, temperature, and circulation rate. The levels of the selected parameters are listed in **Table 9**. In conclusion, an L₉ orthogonal array for three-level factors was selected as the experimental layout to design the trials and determine the effects of various parameters on the CO₂ removal process yield using the aqueous DEA solution in GTP. To evaluate and analyze the results of the Taguchi design, CO₂ concentration in the sweet gas was considered as the main target value with the thermodynamic model in the CO₂ absorption estimation by an aqueous DEA solution in GTP. To evaluate and analyze the results of the Taguchi design, CO₂ concentration in the sweet gas was considered as the main target value with the thermodynamic model in the CO₂ absorption estimation by an aqueous DEA solution in GTP.

**REFERENCES**

1. Adib, H.; Kazerooni, N.; Falsafi, A.; Adhami, M. A.; Dehghan, M.; Golnari, A. Prediction of Sulfur Content in Propane and Butane after Gas Purification on a Treatment Unit. Oil Gas Sci. Technol. 2018, 73, 1–9.
2. Abdulrahman, R.; Immanuel, S. Natural Gas Sweetening: Process Design and Simulation. LAP Lambert Academic Publishing, Saarbrucken, 2012.
3. Kohl, A.; Nielsen, R. Gas Purification, 5th ed.; Gulf Professional Publishing: Houston, 1997.
4. Quek, V. C.; Shah, N.; Chachuat, B. Modeling for Design and Operation of High-Pressure Membrane Contactors in Natural Gas Sweetening. Chem. Eng. Res. Des. 2018, 132, 1005–1019.
5. Abdel-Aal, H. K.; Aggour, M.; Fahim, M. A. Petroleum and Gas Field Processing; Marcel Dekker: New York, 2003.
6. Omar, N. M. Simulation and Optimization of Gas Sweetening Process at Mellitah Gas Plant Using Different Blends of Amines; University Bulletin, 2017, ISSUE, 2017, No.19, Vol. (1).
(7) Romeo, L. M.; Minguell, D.; Shirmohammadi, R.; Andrés, J. M. Comparative Analysis of the Efficiency Penalty in Power Plants of Different Amine-Based Solvents for CO2 Capture. *Ind. Eng. Chem. Res.* 2020, 59, 10082–10092.

(8) Mokhtab, S.; Poe, W.; Mak, J. *Handbook of Natural Gas Transmission and Processing*; Gulf Professional Publishing: Houston, 2006.

(9) Mandal, B. P.; Bandypadhyay, S. S. Simultaneous Absorption of Carbon Dioxide and Hydrogen Sulfide into Aqueous Blends of 2-Amino-2-Methyl-1-Propanol and Diethanolamine. *Chem. Eng. Sci.* 2005, 60, 6438–6451.

(10) Patil, P.; Malik, Z.; Jobson, M. Prediction of CO2 and H2S Solubility in Aqueous MDEA Solutions Using an Extended Kent and Eisenberg Model. *Institution of Chemical Engineers, 2006, 152, 498–510.

(11) Chakma, A.; Meisen, A. Improved Kent-Eisenberg Model for Predicting CO2 Solubility in Aqueous Diethanolamine (DEA) Solutions. *Gas Sep. Purif.* 1990, 4, 37–40.

(12) Weiland, R. H.; Chakravarty, T.; Mather, A. E. Solubility of Carbon Dioxide and Hydrogen Sulfide in Aqueous Alkanol Amines. *Ind. Eng. Chem. Res.* 1993, 32, 1419–1430.

(13) Haji-Sulaiman, M. Z.; Aroua, M. K.; Benamor, A. Analysis of Equilibrium Data of CO2 in Aqueous Solutions of Diethanolamine (DEA), Methyldiethanolamine (MDEA) and Their Mixtures Using the Modified Kent Eisenberg Model. *Trans Chem. Eng. Res. Des.* 1998, 76, 961–968.

(14) Ebenezer, S. A. Removal of CO2 from Natural Gas for LNG Production; Institute of Petroleum Technology Norwegian University of Science and Technology: Trondheim, 2005.

(15) Aliabad, Z.; Mirzaei, S. Removal of CO2 and H2S Using Aqueous Alkanol Amine Solutions World Academy of Science. *Int. J. Chem. Biol. Eng.* 2009, 49, 194–203.

(16) Abdulrahman, R. K.; Sebastine, I. M. Natural Gas Sweetening Process Simulation, and optimization: A case study of Khurmala field in Iraqi Kurdistan region. *J. Nat. Gas Sci. Eng.* 2013, 14, 116–120.

(17) Omar, N. M. Simulation and Optimization of Gas Sweetening Process at Mellitah Gas Plant Using Different Blends of Amines; University Bulletin, 47- ISSUE No.19, March 2017; Vol. 1.

(18) Abd, A. A.; Naji, S. Z. Comparison study of Activators Performance for MDEA Solution of Acid Gases Capturing from Natural gas: Simulation-Based on a Real Plant. *Environ. Technol. Innovation* 2020, 17, 100562.

(19) Al-Amri, A.; Zahid, U. Design Modification of Acid Gas Cleaning Units for an Enhanced Performance in Natural Gas Processing. *Energy Fuels* 2020, 34, 2545–2552.

(20) Alnili, F.; Barifcani, A. Simulation Study of Sweetening and Dehydration of Natural Gas Stream Using MEG Solution. *Can. J. Chem. Eng.* 2018, 96, 2000–2006.

(21) Sulaiman, M. M.; Matloub, F. K.; Shareef, Z. N. Simulation and Optimization of Natural Gas Sweetening Process: A Case Study of Ng Sweeting Unit Designed by Chen Group in the Gulf of Mexico Green Design and Manufacture: Advanced and Emerging Applications. AIP Conference Proceedings; 2030, 2018.

(22) Roy, R. K. *Design of Experiments Using the Taguchi Approach (16 Steps to Product and Process Improvement)*; John Wiley & sons: New York, 2001.

(23) Zeinali, E. *Design of Experiments with Taguchi Method Using Qualitek Software*, Petrochemical Research & Technology Company: Tehran, 2008.

(24) Fogler, H. S. *Elements of Chemical Reaction Engineering*, 3rd ed.; Prentice Hall PTR Inc: New Jersey, 1999.

(25) Speight, J. G. *Chemical Process and Design Handbook*; McGraw-Hill Companies, Inc: New York, 2002.

(26) Littel, R. J.; Filmer, B.; Versteeg, G. F.; Van Swaaij, W. P. M. Modelling of Simultaneous Absorption of H2S and CO2 in Alkanolamine Solutions: The Influence of Parallel and Consecutive Reversible Reactions and the Coupled Diffusion of Ionic Species. *Chem. Eng. Sci.* 1991, 46, 2303–2313.