INTRODUCTION

Oil, also known as industrial life-blood, is the world’s largest source of energy with high demand, low extraction rates, and uneven global distribution. Unfortunately, a sustainable production of oil often leads to a low oil recovery rate in nonconventional oil reservoirs. Thus, increasing the oil recovery efficiency is the key for ensuring the stability of oil production. Generally, the recovery rate of low-permeability oil reservoirs ranges from 20% to 25% and the proportion of such low-permeability reservoirs in proven oil reserves has been increasing every year. EOR from these oil fields has become an important research subject. Many flooding patterns are available, of which CO$_2$-EOR is more desirable than flooding with water, nitrogen, air, or flue gas, as it involves low cost and generates high-quality natural gas.

CO$_2$-EOR technology injects CO$_2$ into the oil recovery layer, which can reduce the viscosity of crude oil, increase its volume, improve the fluidity ratio of crude oil to water, and increase the crude oil recovery rate, thus achieving 43% of the original oil in place (OOIP) after properly defining the production and injection scenarios. For low-permeability...
reservoirs with large oil and gas resources, CO₂-EOR is a preferable recovery method.

A fraction of the CO₂ injected into the subsurface for flooding will be released via the production of natural gas as the associated petroleum gas. The natural gas accompanied with CO₂ causes pipeline corrosion during subsequent transportation, and is particularly detrimental.9,10 In addition, the presence of CO₂ in natural gas will reduce its calorific value, increasing the transportation cost per unit of energy.11 At present, four methods are mainly used for CO₂ capture from the associated gas: chemical absorption, variable pressure adsorption, membrane separation, and low temperature fractionation.12,13 Among them, the mature and efficient adsorption, membrane separation, and low temperature fractionation.12,13 Among them, the mature and efficient chemical absorption method, with a high CO₂ removal efficiency and scale-up feasibility, is most commonly used.14-16 In addition, flue gas CO₂ capture is undoubtedly a short-cut for the associated petroleum gas. The difference between the two methods is the influence of the characteristics of the raw material on the process parameters. The associated petroleum gas is pressurized, with the flue gas at atmospheric pressure, so that the driving force for the flue gas is lower than that for natural gas.17 Associated petroleum gas typically has low amount of methane (50%-70%), high amounts of nitrogen and carbon dioxide (2%-20%), and abundant hydrocarbons.18 Specific characteristics of the associated petroleum gas are discussed in detail in the next section.

During CO₂ removal from associated petroleum gas, selection of proper solvent is essential for a good chemical absorption and it also influences the removal efficiency and energy consumption of the overall process. An aqueous amine solvent is commonly composed of monoethanolamine (MEA), diethanolamine (DEA), triethanolamine (TEA), and methyl-diethanolamine (MDEA). In contrast to the use of a single alkylamine solvent, recent research has shown that blended amine solvents can overcome the problem of lower absorption efficiency and higher reboiler duty, obtaining a greater balance between the two. In the past 2 years, many new chemical solvents have been developed in order to improve the stability and activity of the absorption process and reduce the energy consumption in the desorption process. Zhang et al.19 deemed that a low energy consumption and less secondary pollution in the carbonic anhydrase (CA) absorption process are the advantages of replacing the traditional amine-based CO₂ capture method and also carried out research on the activity and stability of a related surfactant to industrial calcium enzyme. Zhang et al.20,21 and Fang et al.22 by comparing the performances of various absorbents in the chemical absorption method, showed that blended amines and phase-change solvents are some of the most promising new solvents. To date, the mixed solvent of MEA and MDEA has been considered ideal for the absorption of an acidic gas.23,24 Thus, we used a blended solvent containing MEA and MDEA in 1:3 ratio in the simulation.

To solve the persistent problem of energy consumption, various process flowsheet modifications have been proposed in literature. Researchers have focused on process modification and superstructure approaches, including the use of multiple solvent feeds to the absorber column, addition of auxiliary equipment such as heat pumps, and absorber intercooling.25,26 Moullec et al.27 proposed a large number of possible configuration patterns that can be used to enhance the overall performance and reduce the reboiler duty. Cho et al.28 reported the design and optimization of a novel amine-based sweetening system that reduced the cost and energy consumption by 15.9%. The incorporation of a variety of configurations to study the performance and energy consumption is difficult in practice; so, the best method for measuring the energy consumption is simulation. Zhang et al.29 and Lu et al.30 used the rate-based model embedded in Aspen Plus to simulate the process and suggested that it was superior to the equilibrium-stage model, which was more useful for the design of CO₂ capture systems. Muhammad et al.31 used Aspen HYSYS to simulate the effect of the acidic gas content on the total cost of the procedure. Roy et al.32 performed a simulation of the Bakhrabad gas processing plants using the Aspen HYSYS process simulator to validate the proposed model. By comparing these two types of simulation software (Aspen Plus and Aspen HYSYS), Gutierrez et al.33 showed that the simulations did not produce any significant difference in results, with errors of less than 15%. Based on the validation of the model embedded in Aspen Plus, it is now the most commonly used simulation method in research.

This study was performed to accurately analyze associated petroleum gas and propose a CO₂ capture system. The parameters of the process equipment suitable for the pressurized associated petroleum gas to ensure normal operational and absorption efficiency in the chemical absorption method were determined while studying the CO₂ capture process through system simulation. This will be of guiding significance for the CO₂ capture in associated petroleum gas.

2 | BACKGROUND

2.1 | Associated petroleum gas

The content of the associated petroleum gas flooded by CO₂ is highly fluctuational, with a large CO₂ content and high partial pressure, which presents technical difficulties in CO₂ separation and treatment.

The components of the associated petroleum gas in a single well in the Shengli oilfield are shown in Figure 1. Combined with other wells in the field, the gas exhibits the following characteristics: (a) High carbon content of 91%; (b) high hydrocarbon content (C₅+) of 1.12%, accounting for 8.6% of the total hydrocarbons; (c) no oxygen or sulfide gas; (d) due to the liquid crude oil, gas separation should be
performed before adding to the separation process system; (e) no water vapor components; and (f) the pressure range is 0.3-0.6 MPa(G).

On analyzing the composition of associated petroleum gas, a high CO₂ content (>15%) was observed. A set of technological processes applicable to all large-scale stations in the oilfield with a scale of 100 000 Nm³/d was developed. According to Table 1, CO₂ and CH₄ contents in the feed gas were 20 mol% and 80 mol%, respectively.

### 2.2 | Simulation process

Most CO₂ capture processes are similar in terms of absorption and desorption columns, although some modifications are possible to reduce the energy consumption. A complete flow process of the CO₂ capture system was generated by Aspen Plus (Version 8.6) and is shown in Figure 2.

The extraction gas (material 1) is first processed through the pretreatment system which includes an oil and gas separation column, a heavy hydrocarbon removal system, a filter, and a gas preheater, to remove the oil and heavy hydrocarbon constituents. The gas (material 2) that is passed through the pretreatment system is then sent to an absorption column and the CO₂ in the associated petroleum gas is absorbed by the solvents. The exhaust gas (material 3) is subsequently discharged from the top of the absorption column into the processing network.

The rich amine loaded with CO₂ (material 4) from the bottom of the column is pumped into the rich and lean amine heat exchanger, and the high-temperature rich amine (material 5) is subsequently returned to the desorption column. Here, the dissolved CO₂ and water vapor are cooled, separated, and removed. Simultaneously, 99% pure (dry basis) CO₂ gas (material 6) can be obtained, which is then transported to the sequence section. The lean amine (material 7) is discharged from the desorption column after the CO₂ is released. The lean liquid at high temperature (material 8) and the rich liquid at low temperature (material 4) pass through the pump and cooler after the heat transfer in the heat exchanger. A new lean liquid (material 10), which is a mixture of low-temperature lean liquid (material 9) and the liquid of CO₂ product gas, enters the absorption column for a new absorption process. The process of continuous absorption and desorption by the solvent is thus formed.

Based on the design information provided, the main parameters related to the absorption and desorption columns are shown in Table 2. This information was used as a basis for the Aspen Plus simulation. MELLAPAK 250Y of SULZER company was chosen as the structured packing material to achieve a minimum pressure drop. A minimum energy consumption was required with many theoretical stages of separation and the packing parameters used in the simulation are listed in Table 3.

### 2.3 | Reaction principle

MDEA (methyldiethanolamine, C₅H₁₃NO₂) is stable and does not corrode carbon steel. Since MDEA is weakly alkaline, it is easier to desorb after absorbing acidic gases and regeneration can be performed by flash evaporation at low pressures, resulting in significant energy savings. However, the CO₂ desorption rate is low for MDEA. MEA (monoethanolamine, C₂H₇NO) exhibits rapid absorption rate with a small absorption capacity and some corrosive effect. This results in larger energy consumption and easy oxidization by SO₂ and O₂.

In various applications, CO₂ capture is achieved by a single amine or amine/piperazine (PZ)-promoted adsorbents. However, the removal rate of the blended MEA/MDEA absorbent was higher than that of the single component absorbent, and had additional economic benefits. The reaction of MDEA (R₂ = C₄H₁₀O₂) and CO₂ can be written as follows:

\[
\text{CO}_2 + H_2O \leftrightarrow H^+ + HCO_3^-
\]  
(R1)
Combining Equations R1 and R2:

\[
\text{H}^+ + \text{R}_2\text{CH}_3\text{N} \leftrightarrow \text{R}_2\text{CH}_3\text{NH}^+ 
\]

(R2)

Combining Equations R3 and R5:

\[
\text{R}_3\text{NH} + \text{CO}_2 \leftrightarrow \text{R}_3\text{NCOO}^- + \text{H}^+ 
\]

(R4)

where, Equation R3 is controlled by Equation R1.

For the blended solvents (when \( \text{R}_3\text{NH} (\text{R}_3 = \text{C}_2\text{H}_6\text{O}) \) is added to MDEA), the absorption of \( \text{CO}_2 \) takes place in the following steps:

\[
\text{R}_2\text{CH}_3\text{N} + \text{CO}_2 + \text{H}_2\text{O} \leftrightarrow \text{R}_2\text{CH}_3\text{NH}^+ + \text{HCO}_3^- 
\]

(R3)

Combining Equations R4 and R5:

\[
\text{R}_3\text{NH} + \text{CO}_2 \leftrightarrow \text{R}_3\text{NCOO}^- + \text{H}^+ 
\]

(R4)

\[
\text{R}_3\text{NCOO}^-\text{H}^+ + \text{R}_2\text{CH}_3\text{N} + \text{H}_2\text{O} \leftrightarrow \text{R}_3\text{NH} + \text{R}_2\text{CH}_3\text{NH}^+ + \text{HCO}_3^- 
\]

(R5)

The reaction described by Equation R6 is controlled by Equation R4, where (R4) is the secondary reaction and the reaction rate constant \( K_c = 10^4 \text{ L/(mol·s)} \) at 298.15 K. When \( \text{R}_3\text{NH} \) is added, the free amine \( N_{[\text{R}_3\text{NH}]} \) is \( >10^{-2} \text{ mol/L} \).
The reaction in Equation R4 is much faster than that in Equation R1.

To summarize, the addition of MEA modifies the process of absorption of CO$_2$ in the MDEA solvent. The CO$_2$ absorbed by MEA is continuously transferred to MDEA, thus imparting high absorption and desorption rates.

The simulated calculations were performed with a lean amine CO$_2$ content of 0.1 mol CO$_2$/mol MDEA and 0.28 mol CO$_2$/mol MEA, and the detailed parameters are shown in Table 4. These parameters ensured that the simulation could be performed, providing the conditions for an accurate optimization.

### 3 | MODEL DESCRIPTION

#### 3.1 | Thermodynamic model

The embedded models for the MEA-MDEA-CO$_2$-H$_2$O system provided by Aspen Plus were combined to develop a thermodynamic model. Binary interaction parameters and electrolyte pairs recommended by Aspen Plus from its physical property data bank were used for the MEA-MDEA-CO$_2$-H$_2$O system. The Ehrlich equation of state and the physical property data bank were used for the MEA-MDEA-CO$_2$-H$_2$O system, the reaction kinetics of CO$_2$ with aqueous MEA and MDEA, and the Radfrac model parameters governing mass and heat transfer. The kinetic chemistry is summarized below:

\[
\text{EQUIL } 2\text{H}_2\text{O} \leftrightarrow \text{H}_3\text{O}^+ + \text{OH}^- \quad \text{(R7)} \\
\text{EQUIL } \text{MEAH}^+ + \text{H}_2\text{O} \leftrightarrow \text{MEA} + \text{H}_3\text{O}^+ \quad \text{(R8)} \\
\text{EQUIL } \text{HCO}_3^- + \text{H}_2\text{O} \leftrightarrow \text{H}_3\text{O}^+ + \text{CO}_3^{2-} \quad \text{(R9)} \\
\text{EQUIL } \text{MEAH}^+ + \text{H}_2\text{O} \leftrightarrow \text{MDEA} + \text{H}_3\text{O}^+ \quad \text{(R10)}
\]

The temperature dependence of the chemical equilibrium constants of the reactions can be expressed as follows:

\[
\ln(K_{eq}) = A + B/T + C \ln(T) + D \cdot T \\
+ E \cdot ((P - P_{ref}/P_{ref}) \quad \text{(R11)}
\]

where $K_{eq}$ is the equilibrium constant, Henry's Law constant, or salt precipitation equilibrium constant of the reaction eq, $T$ is the temperature in K; and $P_{ref}$ is the reference state pressure in Pa. The constants $A, B, C, D,$ and $E$ are retrieved from the Aspen Plus databank.

#### 3.2 | Rate-based model

A rate-based model was also developed using Aspen Plus to simulate the combined MEA + MDEA absorption under the associated petroleum gas parameters listed in the previous section. The rate-based model was developed using the RadFrac module embedded in Aspen Plus, which partitioned the absorption column into stages and calculated the mass transfer, heat transfer, chemical equilibrium, reaction kinetics, hydraulic characteristics, and interfacial behavior at each stage.

To properly calculate the complex absorption process using the aqueous MEA/MDEA absorbent, the rate-based model must account for the thermodynamics of the MEA-MDEA-CO$_2$-H$_2$O system, the reaction kinetics of CO$_2$ with aqueous MEA and MDEA, and the Radfrac model parameters governing mass and heat transfer. The kinetic chemistry is summarized below:

\[
\text{KINETIC } \text{MEA} + \text{CO}_2 + \text{H}_2\text{O} \rightarrow \text{MEACOO}^- + \text{H}_3\text{O}^+ \quad \text{(R12)} \\
\text{KINETIC } \text{MEACOO}^- + \text{H}_3\text{O}^+ \rightarrow \text{MEA} + \text{CO}_2 + \text{H}_2\text{O} \quad \text{(R13)} \\
\text{KINETIC } \text{MDEA} + \text{CO}_2 + \text{H}_2\text{O} \rightarrow \text{MDEAH}^+ + \text{HCO}_3^- \quad \text{(R14)} \\
\text{KINETIC } \text{MDEAH}^+ + \text{HCO}_3^- \rightarrow \text{MDEA} + \text{CO}_2 + \text{H}_2\text{O} \quad \text{(R15)}
\]

The reduced power law expressions were used for the kinetically controlled reactions:

\[
r_j = k_j T^n \exp \left( - \frac{E_j}{RT} \right) \prod_{i=1}^{N} (C_i)^{\theta_i}
\]

where $r_j$ is the rate of the reaction; $k_j$ is the pre-exponential factor; $T$ is the absolute temperature; $n$ is the temperature exponent; $E_j$ is the activation energy; $R$ is the gas constant; $N$ is the number of components in the reaction; $C_i$ is the concentration of component $i$; and $\theta_i$ is the stoichiometric coefficient of component $i$. 

### Table 4 | Methyldiethanolamine/MEA blended absorbent

| Material parameters | Numerical value |
|---------------------|----------------|
| Absolute pressure (MPa) | 0.30-0.60 |
| Temperature (K) | 308.15 |
| MEA to MDEA ratio | 1:3 |
| Amines to water ratio | 2:3 |
| Component (mol%) | |
| CO$_2$ | 0.018 |
| H$_2$O | 0.873 |
| MEA | 0.043 |
| MDEA | 0.066 |
In addition, in order to simulate the steady state process, the model makes the following assumptions:\(^37,38\):

- The convection process involves well-mixing of the liquid and gas phases.
- The vaporization of the blended solvents is neglected.
- Ambient heat loss is ignored.
- Both heat and mass transfer conform to the two-film theory.

### 3.3 Model validation

The rate-based model is verified by comparing the experimental data with the simulated data, with its main index as CO\(_2\) concentration. Borhani et al.\(^39\) verified the reliability of the rate-based model by capturing the acidic gases, CO\(_2\) and H\(_2\)S, in the sweet gas in two cases, which showed that there was almost no variation between the experimental and simulation results. In the CO\(_2\) capture process of natural gas, Salvinder et al.\(^38\) and Emmanuel et al.\(^37\) adopted the rate-based model for the energy consumption and process analysis.

Therefore, it can be concluded that the rate-based model can reliably predict the CO\(_2\) capture of the CO\(_2\)/CH\(_4\) system by the amine solvent. In the simulation, the CO\(_2\) capture rate is set to reach 90\%, with the main objective of analyzing the influence of key process parameters on the lean liquid flow and energy consumption.

### 4 INFLUENCE OF OPERATING PARAMETERS ON THE CIRCULATION FLOW

#### 4.1 Effects of the absorption column packing height and size

The temperature of the prefetch absorber inlet lean solvent was set to 313.15 K. To reach the gas purity requirements, the amount of solvent needed for circulation was analyzed as a function of the packing height under different operating pressures.

As shown in Figures 3 and 4, the flow rate of the lean solvent decreases with the CO\(_2\) loading of the rich amine from 1-4 m, indicating that the solvent is not in full contact with CO\(_2\) because of the short contact time between the gas and liquid. Therefore, when the packing height and reaction time increased, the circulation of the solvent and CO\(_2\) loading of the rich amine changed significantly. When the packing height was >8 m, the reaction time in the absorption column was sufficiently long. With the increasing packing height, the reaction time changed only negligibly. Therefore, the flow of

| Variables | Units | Value |
|-----------|-------|-------|
| Absorption pressure | MPa | 0.30-0.60 |
| Feed gas temperature | K | 308.15 |
| Lean amine flow rate | t/h | 34-39 |
| Feed gas flow rate | t/h | 4 |
| L/G ratio | | |
| Desorption pressure | MPa | 0.10-0.20 |
| Desorption column inlet stream temperature | K | 358.15 |
| Heat duty of reboiler | kW | 1200 |

![Figure 3](image-url) Effects of the packing height on circulation flow
the lean amine and CO₂ loading of the rich amine remained largely unchanged after that point.

According to the variation curve of the parameters under different pressures in Figures 4 and 5, as the pressure increases, the circulation of the solvent decreases and CO₂ loading of the rich amine increases, indicating that a higher pressure increases the solvent absorption of CO₂. In addition, the flow curves shifted to the left and the lower packing heights could meet the minimum requirements of the absorption reaction, demonstrating that a high pressure improves the absorption rate. Therefore, the optimum packing height was determined to be 10 m.

4.2 | Effects of the inlet temperature of the lean amine

To eliminate the limiting effects of the packing height on the absorption process, 10 m packing height was selected and the influence of the absorption temperature on the circulation of the solvent and CO₂ loading in the rich amine were analyzed. It was assumed that the absorption process was largely unaffected by the packing height.

Figures 5 and 6 show the variation curves of the solvent circulation and CO₂ loading with the absorption temperature. Because the reaction of CO₂ with the alcohol amine solvent is exothermic, low temperatures are advantageous. The amount of circulation solvent increases as the absorption temperature rises. When the absorption temperature increased from 298.15 to 348.15 K, the solvent circulation amount increased by approximately 1.55 times. As the absorption temperature increases, the equilibrium concentration of CO₂ in the amine solvent decreases, and the concentration of CO₂ in the rich amine decreases. In other words, the loading of CO₂ increases as the pressure increases and the amount of circulation flow decreases.

It can also be seen that at 308.15 K, both circulation flow and CO₂ loading exhibit inflection points where the circulation flow reaches a minimum. Thus, a temperature of 308.15 K was chosen for the lean solvent inlet.

4.3 | Effects of the pressure of the absorption column

Based on the above analysis, we chose 10 m as the packing height of the absorption column and 308.15 K as the temperature of the lean amine inlet. By changing the operating pressure of the absorber, the inlet pressure of the gas and lean amine changed. By holding these parameters constant, we can analyze the trends of the amount of CO₂ in the rich solvent and solvent cycling.

As shown in Figures 7 and 8, as the operating pressure increased from 0.3 to 1.0 MPa, the circulation of solvent decreased from 40.2 to 30.2 t/h, and the CO₂ loading in the rich amine increased from 0.55 to 0.65 mol/mol of solvent. High pressure causes the alcohol amine solvent to absorb CO₂ fully, resulting in a significant reduction in the amount of circulating solvent. The absorption pressure should be selected by considering the influence of the pressure change on the nature of the associated petroleum gas, energy consumption, and cost of the equipment. Therefore, the optimal operational parameters should be determined after analyzing the effect of absorption pressure on the energy consumption.

**FIGURE 4** Effects of the packing height on CO₂ loading of the rich amine
4.4 | Effects of CO₂ loading of the lean solvent

The CO₂ content in the lean solvent directly affects the amount of CO₂ absorbed by the solvent, which in turn affects the circulation flow of the lean solvent in the absorption column. Figure 9 shows the relationship between the circulation flow and CO₂ loading of the lean solvent. From a loading of 0.13 to 0.28 mol/mol of the lean solvent, the amount of circulating solvent increased from 32.2 to 51.51 t/h.

The loading of CO₂ in the lean solvent influences the desorption energy consumption and circulation flow. Lower CO₂ loading of the lean solvent decreases the degree of regeneration of the solvent, but increases the circulation flow. Thus, a higher loading of CO₂ in the lean solvent will lower the circulation flow, but desorption would become more difficult, requiring higher desorption temperatures, leading to a higher energy consumption. Therefore, for the optimization of the CO₂ loading in the lean solvent, both circulation flow and energy consumption should be considered.
5.1 Effects of the absorption column pressure

From Figure 10, it is clear that over the pressure range shown, with the increasing absorption pressure, the reboiler duty decreases rapidly from 1.59 to 1.21 MW, and the unit energy consumption decreases from 3.68 to 2.8 GJ/t CO₂.

It is apparent that the operating pressure of the absorber has a significant influence on the energy dissipation of the system. When the absorption pressure was increased from 0.2 to 1 MPa, the reboiler duty decreased by 23.8% and the energy consumption per unit of product decreased by a similar degree.

Therefore, selection of the operating pressure of the absorber should consider the nature of the associated petroleum gas, operational cost, and energy consumption of the system. Herein, we chose 0.6 MPa as the optimal operating pressure.
of the absorption column for a further process design and system simulation.

5.2 Effects of the desorption column pressure

The operating pressure of the desorption column affects the equilibrium solubility of CO2 in the solvent. With the increasing pressure, the reaction balance moves toward the direction of absorption. However, with more difficult desorption, more energy will be consumed.

From Figure 11, the bottom temperature and desorption pressure exhibit a mostly linear relationship. When the desorption pressure was increased from 0.1 to 0.2 MPa, the bottom temperature increased from 373.86 to 389.18 K.

Figure 12 shows the relationship between the pressure of the operating desorption column and the reboiler duty and energy consumption per unit product of CO2. When the desorption pressure was increased from 0.1 to 0.2 MPa, the increased pressure made the desorption difficult and required higher desorption temperatures; so, the temperature of the tower rose. The reboiler load increased from 1.28 to 1.71 MW, and the energy consumption per unit of product increased from 2.94 to 4GJ/t CO2. Considering the lower energy consumption and desorption pressure control precision,
a desorption pressure of 0.12 MPa was chosen with a power consumption of 3.2 GJ/t CO₂.

### 5.3 Effects of CO₂ loading of the lean solvent

The relation between the unit energy consumption and CO₂ content in the lean solvent is shown in Figure 13. When the CO₂ content was increased from 0.10 to 0.20 mol/mol of lean solvent, the desorption of CO₂ was facile. The reboiler duty and unit energy consumption decreased from 3.33 to 3.14 GJ/t CO₂, but the circulation flow increased from 29.85 to 40.5 t/h. To ensure an efficient absorption and a reasonable circulation flow, a CO₂ loading of 0.2 mol/mol of the lean solvent was selected.

### 6 Conclusion

Based on the chemical method of CO₂ capture, a complete set of EOR gas extraction process parameters were determined.
for a novel absorbent pair. A 100 000 Nm$^3$/d specification was used as a reference to obtain the detailed operational parameters. A blended solvent containing 30wt% MDEA and 10wt% MEA was used to result in a reduced energy consumption compared to the single component solvent systems. In the simulation, rate-based models were used to calculate the mass transfer, heat transfer, chemical equilibrium, reaction kinetics, hydraulic characteristics, and interfacial behaviors at each stage for both, absorption and desorption columns.

Circulation flow and energy consumption are the two main parameters that were considered during optimization. For the circulation flow, the main influencing factors were the packing height of the absorption column, absorption pressure, absorption temperature, and CO$_2$ loading of the lean solvent. For the absorption pressures between 0.30 and 0.60 MPa, a packing layer height of 10 m of the absorption column and the absorption temperature of 308.15 K were reasonable. The maximum quantity of solvent required was approximately 39.15 t/h. In terms of energy consumption, the main influencing factors included the absorption pressure, desorption pressure, and CO$_2$ loading of the lean solvent. A comprehensive analysis showed that the optimal desorption pressure was 0.12 MPa and the lean fluid load was 0.2 mol/mol, which resulted in an energy consumption of approximately 3.16 GJ/t CO$_2$. This work provides a basis for improving the gas process for the subsequent chemical process. In future studies, other devices will also be optimized.

**ACKNOWLEDGMENTS**

This work was supported by National Key R&D program of China (No. 2017YFB0603300).

**CONFLICT OF INTEREST**

The authors declare no conflict of interest.

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How to cite this article: Liu B, Zhang M, Wang T, Jia W. Process simulation of CO₂ capture from CO₂-EOR associated petroleum gas with aqueous MEA and MDEA solvents. Energy Sci Eng. 2019;7:663-675. https://doi.org/10.1002/ese3.308