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# Low-energy-consumption $CO_2$ capture via coupling process of direct organic fluid stripping with biphasic solvents

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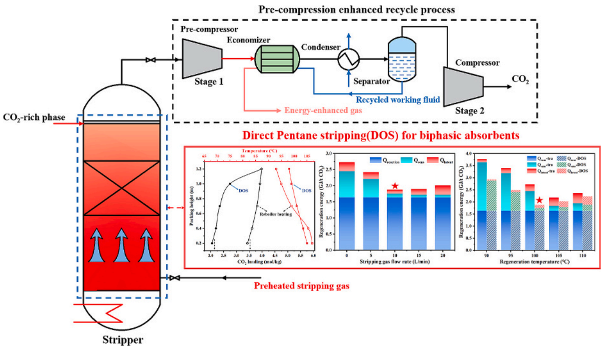
## HIGHLIGHTS

- Coupling process of direct organic fluid stripping(DOS) with biphasic solvent is proposed.
- Pentane is identified as the optimal stripping fluid for various biphasic solvents.
- DOS reduces energy consumption by up to 46.1 % compared to conventional reboiler heating.
- 180-h long-term tests show >90.1 % $CO_2$ removal ratio and < 6.5 % amine loss in coupling process.
- Pre-compression recycle process boosts energy recovery and pentane recovery (99.7 %).

## GRAPHICAL ABSTRACT



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## ABSTRACT

Amine-based $CO_2$ chemisorption technology has currently gained significant attention in the field of $CO_2$ capture. However, high energy consumption remains one of the key challenges limiting large-scale applications in flue gas $CO_2$ chemisorption. To tackle this, the coupling process of direct organic fluid stripping (DOS) with biphasic solvents was innovatively proposed in this study. Various typical biphasic absorbents were applied in a laboratory-scale continuous flow system. Pentane was identified as the ideal organic fluid for solvents due to its low vaporization heat requirement and outstanding enhancement in $CO_2$ regeneration. Compared to conventional reboiler heating(CRH) process, DOS process effectively mitigated temperature non-uniformity within the stripper and provided extra mass transfer driving force for $CO_2$ desorption of biphasic solvents. Multiple operational parameters were systematically analyzed to establish a comprehensive impact mechanism on the energy-saving efficiency of DOS process. It was revealed that DOS process promoted efficient regeneration of biphasic solvents at temperatures below 100 °C, achieving up to a 46.1 % reduction in regeneration energy compared to CRH process. Moreover, the coupling process demonstrated excellent operational stability in 180-h continuous absorption-desorption cycling tests. While strengthening $CO_2$ regeneration within the stripper, DOS process notably suppressed amine loss in solvents(<6.5 %) compared to CRH process, maintaining high $CO_2$ removal ratio (>90.1 %) and stable phase separation of biphasic solvents. To further enhance energy efficiency and

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pentane recovery in the system, the pre-compression enhanced recycle process was applied and simulated in Aspen Plus, achieving a maximum recycled energy of 0.65 GJ/t $CO_2$ and a pentane recovery ratio of 99.7 %.

## 1. Introduction

$CO_2$ capture technology holds significant potential to reduce $CO_2$ emissions, bolstering global initiatives to address environmental challenges [1,2]. Among various technologies, $CO_2$ chemisorption using amine-based solvents provides a reliable strategy for the flue gas $CO_2$ capture due to its application maturity, high capture rate, and technical reliability [3,4]. The technology mainly consists of two processes including $CO_2$ absorption in the absorber and $CO_2$ regeneration in the stripper. Under conventional reboiler heating(CRH) which only provides the thermodynamic driving force, most solvents require relatively high temperatures, typically exceeding 105 °C, to achieve effective $CO_2$ desorption [5]. This leads to substantial energy consumption, primarily from the sensible heat needed to increase the solution's temperature and the latent heat for water evaporation in the stripper [6,7]. Represented by traditional 30 % MEA [8], the elevated energy consumption of solvents has severely hindered large-scale applications in $CO_2$ chemisorption.

To address this issue, current researches on $CO_2$ chemisorption aim to reducing energy consumption through both solvent development and energy-saving process design. A range of novel absorbents, such as water-lean solvents [9], biphasic solvents [10], and ionic liquids [11], have been developed. Among them, biphasic absorbents have exhibited superior performance in optimization of regeneration energy. The solvent typically consists of a reactive amine, a phase separation agent, and water, which would undergo phase separation under specific conditions, such as increased $CO_2$ loading or temperature changes, forming the $CO_2$-rich phase and the $CO_2$-lean phase. Only the $CO_2$-rich phase is directed to the stripper for desorption, achieving a great decrease in regeneration energy [10]. By reducing the water content of the solution and the volume of the $CO_2$-rich phase, while using more easily regenerable active amines, related studies have effectively reduced the solution regeneration energy consumption to 2.0–2.4 GJ/t $CO_2$ [12–14]. However, the drawbacks of CRH process, including limited driving force for $CO_2$ desorption and the high temperatures in the stripper, remain significant obstacles to the application of biphasic solvents. These issues contribute to a bottleneck in optimizing energy consumption, as well as solvent losses, including amine degradation [15] and evaporation [16].

Therefore, the development of energy-saving processes for regeneration enhancement holds significant potential in chemical absorption [17,18]. Fang et al. [19,20] developed a direct steam stripping process (DSS). Stripping steam is introduced directly at the bottom of the stripper, which mainly provides the driving force in the gas-phase $CO_2$ mass transfer for $CO_2$ desorbing of absorbents without supplying additional heat. By optimizing the process parameters, DSS achieved approximately a 23.2 % reduction in energy consumption compared to CRH process for the 30 wt% MEA solution [8]. Based on it, Frimpong et al. [21] applied organic solvents with lower boiling points and latent heat of evaporation to replace steam and proposed the direct organic working fluid stripping process. It can further reduce the low-grade heat input required for vapor generation. Wang et al. [22] applied the direct organic fluid stripping for the $CO_2$ regeneration process of 30 wt% MEA and reduced the regeneration energy to a minimum of 2.38 GJ/t $CO_2$, which was 38.4 % and 15.2 % lower than that under CRH and DSS processes, respectively.

Based on it, the coupling optimization of direct organic fluid stripping process(DOS) with biphasic absorbents presents a promising avenue for further advancements in $CO_2$ chemisorption technology. Current research [22] has only conducted short-term operations for 30 % MEA solution within the stripper, while the $CO_2$ regeneration process of biphasic solvents under DOS process at various conditions, including gas flow rate and regeneration temperature, and the underlying mechanism remained unclear. Moreover, the impact of DOS process on the absorption-desorption and phase separation performance of biphasic solvent, as well as the operation stability of the coupling process needs to be systematically evaluated through extended cyclic testing. Furthermore, for practical applications, efficient recovery of organic fluids is essential to minimize operational costs. However, conventional water-cooling methods have been found insufficient to meet the required recovery standards [21,22]. Therefore, the development and application of DOS process still face several challenges that need to be addressed.

In this study, the coupling process of DOS with biphasic absorbents was proposed. Various types of typical biphasic solvents reported in our previous works, including amide-based solvent DETA/DEA/DMAC (2DE1AC) [13] and lipophilic solvent DEEA/AEEA (DAH) [12], were applied in a laboratory-scale continuous flow system(LSCF) to reveal the optimization characteristic of DOS process. Pentane was identified as the optimal organic stripping fluid through preliminary screening for biphasic solvents. The $CO_2$ desorption process of solvents within the stripper under DOS process and underlying mechanism were revealed. Multiple operating conditions, including stripping gas flow rate and regeneration temperature, were considered to investigate the systemic impact mechanisms on the efficiency of DOS process. Compared to CRH process, the DOS was proven to promote efficient regeneration of biphasic solvents at lower temperatures. Moreover, 2DE1AC and DAH solvents underwent a 180-h continuous absorption-desorption cycling test under CRH and DOS processes to evaluate the comprehensive impact of the DOS on biphasic solvents and assess the operational stability of the coupling process. To further enhance the recovery efficiency of system energy and organic fluid, the pre-compression enhanced recycle process integrated with the DOS was proposed and simulated in Aspen Plus V11. It was indicated to efficiently recycle the system energy and improve the condensation recovery ratio of the organic fluid, which further strengthened system stability and economic viability.

## 2. Experiments

### 2.1. Materials

Comprehensive information for each component in the 2DE1AC and DAH solutions, including purity, supplier, structure, and concentration, was detailed in Table S1. Pure $CO_2$ gas was purchased from Hangzhou Jingong GAS Co., Ltd., Hangzhou, China. The dry air was generated by a piston-type air compressor.

### 2.2. The experiment of the coupling process of DOS with biphasic absorbents

#### 2.2.1. Insolubility test

Since the solution sent to the stripper refers to the aqueous phase of the biphasic solvents after phase separation, pure $CO_2$ was absorbed to DETA/DEA/DMAC(2DE1AC) [13] and DEEA/AEEA(DAH) [12] solutions to obtain $CO_2$ saturated solutions. The saturated aqueous phase was subsequently obtained using a separatory funnel. 20 mL of different liquid organic fluid and 20 mL of the saturated aqueous phase were placed into a test tube, followed by magnetic stirring for 15 min. The mixture in the tube was maintained to stand for 30 days for observation under 1 atm and 25 °C. The photographs of the mixture states in the test were recorded in Table. S4.

#### 2.2.2. Condensation recovery test

The condensation recovery ratio(%) of the organic working fluid at different condensation temperatures was measured using the apparatus shown in Fig. S1. The liquid organic solvent in the tank was pumped into a heat exchanger, heated by an oil bath set at 100 °C, to generate organic vapor. The vapor then entered a 450 mm long coiled condenser. Cooling water at various temperatures, which was controlled by a water bath, circulated through the condenser to control the condensation temperature. The condensation ratio of the organic working fluid was calculated using Eq. 1.

$$\sigma = \frac{m_{condense}}{m_{total} - m_{remaining}} \times 100\% \tag{1}$$

Where $m_{condense}$ represents the mass of the condensed organic solvent collected in the condensation flask in the test, kg. $m_{total}$ is the initial mass of the liquid organic solvent in the storage tank, kg. $m_{remaining}$ is the mass of the remaining liquid organic solvent in the storage tank after the test, kg.

### 2.2.3. Laboratory-scale continuous flow(LSCF) system

Laboratory-scale continuous flow(LSCF) system was designed to analyze the absorption-desorption as well as phase separation performance of biphasic solvents at various conditions, as depicted in Fig. 1. The mixed gas of $CO_2$ and air ($CO_2$ vol% = 12 %) was introduced at the bottom of the absorber, where it counterflowed with the absorbent. The gas flow rate was controlled at 45 L/min using mass flow controller. The absorber had an internal diameter of 0.12 m and a height of 1.2 m. The absorption temperature was controlled at 313 K, with pressure set at 1 atm. The mixed gas after $CO_2$ absorption was discharged from the top of the absorber, dried through a drying tube, and then the $CO_2$ concentration was measured using a $CO_2$ infrared analyzer. The $CO_2$-rich solution was released at the bottom of absorber and separated in a demixer. After phase separation, the $CO_2$-rich phase was pre-heated in the lean-rich solution heat exchanger and entered the stripper at the top. The temperature difference for heat exchange was set to 10 K.

The stripper had an internal diameter of 0.1 m and stood 1.2 m tall, filled randomly with Dixon rings measuring 3 mm × 3 mm, which was designed to be insulated to minimum heat loss. The $CO_2$-rich phase underwent $CO_2$ desorption under CRH and DOS process within the stripper at various conditions(detailed in the Supplementary Material), with pressure set at 1 atm. Five solution sampling ports were distributed at different heights along the stripper(at heights of 0.2, 0.5, 0.7, 1.0, and 1.2 m) to obtain the $CO_2$ loading and amine concentration of the samples under various conditions within the stripper, through acid-base titration and Cation Ion Chromatography (IC), respectively. Five thermocouples were applied to monitor the corresponding temperatures at various packing heights. An electric heater at the bottom of the stripper was used to provide the thermodynamic driving force for $CO_2$ desorption and to control the regeneration temperature(average of temperatures at different heights within the stripper) by adjusting its heating power, with a maximum output power of 3.6 kW. After gas purification through condenser and active carbon adsorber, the flow rate and temperature of desorbed $CO_2$ were obtained using a wet-gas flow meter and thermocouple, respectively.

The $CO_2$-lean solution exiting from the bottom of the stripper was transported by a gear pump. After passing through the lean-rich solution heat exchanger, it was mixed with the organic phase from the upper place of the demixer and then sent back to the absorber for continuous operation. The solution circulation flow rates in the system were uniformly set to 0.6 L/min. In the experiments, the water mass ratio of the $CO_2$-rich phase and $CO_2$-lean solution at the inlet and outlet of the stripper, respectively, under different operating conditions was measured using the Karl Fischer method, with the $CO_2$ loading of solvents obtained using acid-base titration.

The LSCF system has been validated by comparing the energy consumption data for 30 wt% MEA measured using the experimental setup with literature values [8,19,23]. The validation method and results are detailed in Supplementary Material.

The calculation methods for various performance parameters of biphasic solvents in the LSCF system, including $CO_2$ absorption-desorption, phase separation and amine stability under CRH and DOS processes are detailed in the Supplementary Material.

### 2.3. Calculation of $CO_2$ regeneration parameters in the LSCF system

$CO_2$ regeneration performance parameters include $CO_2$ regeneration rate, cyclic capacity, and regeneration energy consumption.

The $CO_2$ regeneration rate (mol $CO_2$/min) was calculated as depicted in Eq. 2.

$$r_{CO2} = \frac{P_{CO2} \bullet V_{CO_2}}{R \bullet T_{CO2}} \tag{2}$$

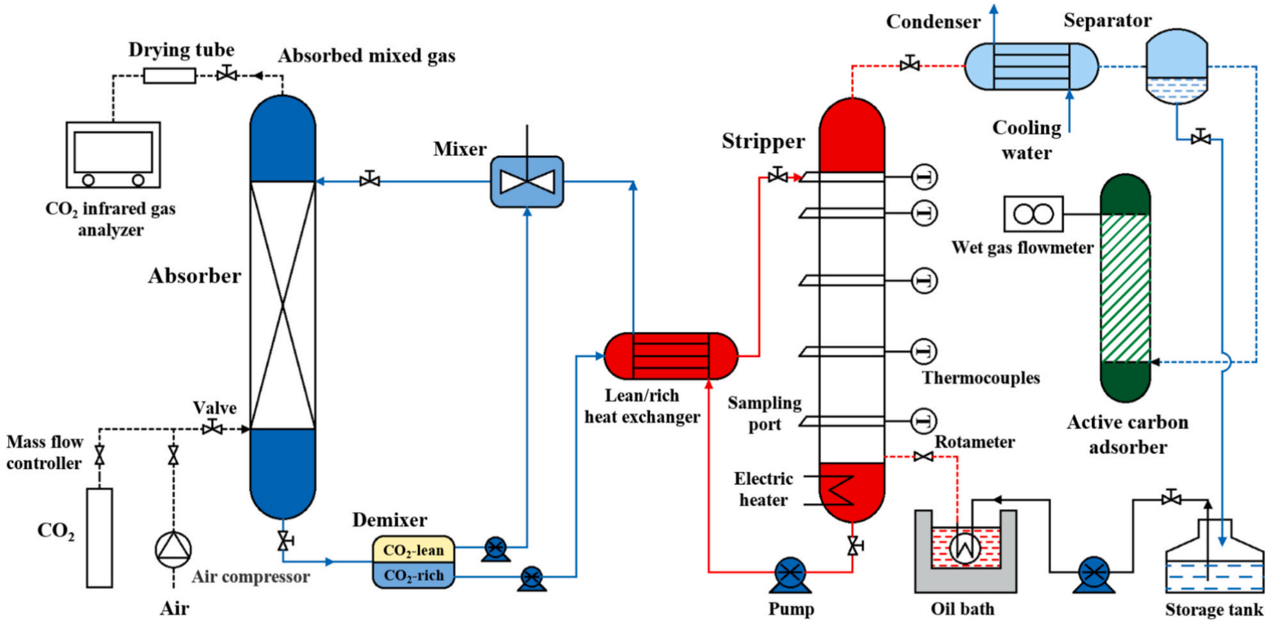

**Fig. 1.** The LSCF system schematics.

Where $P_{CO2}$ denotes the pressure of the $CO_2$ after the purification, set to 1 atm. $V_{CO2}$ is the volume flow rate of purified $CO_2$ obtained through the wet gas flow meter, $m^3$/min. R is the universal gas constant. $T_{CO2}$ is the temperature of $CO_2$, K.

Since only the aqueous phase solution underwent regeneration in the stripper, the cyclic capacity of the solvent($\alpha_{cyclic}$, mol/kg) under different conditions was determined by the difference in $CO_2$ loading between the $CO_2$-rich phase and the lean solution at the inlet and outlet of the stripper, respectively, which was obtained using acid-base titration.

Regeneration energy consumption ($Q_{reg}$, GJ/t $CO_2$) contains three parts: reaction heat ($Q_{reac}$), sensible heat ($Q_{sens}$), and latent heat ($Q_{latent}$), as shown in Eq. 3 [24,25].

$$Q_{reg} = Q_{reac} + Q_{sens} + Q_{latent} \tag{3}$$

Each part of energy consumption (GJ/t $CO_2$) of different biphasic absorbents during regeneration processes in the LSCF system under both CRH and DOS processes was calculated separately, as detailed in the Eq. S(6–15).

### 2.4. Aspen plus simulation

Aspen Plus V11 was applied to simulate the pre-compression enhanced recycle process proposed in this work on a 4 $Nm^3$/h scale(to align with the experimental study). The simulation process flow schematic, flow information, and design parameters [26–28] for each component are provided in the Supplementary Material. The composition of the mixed gas at the outlet of the stripper, along with the operational parameters of each stream, were determined according to the LSCF experimental validation.

The amount of recycled energy(GJ/ t $CO_2$) from the mixed gas at the outlet of stripper and the pentane condensation ratio(%) under different operating conditions were obtained. The calculation method was detailed in Supplementary material. As the process simulation only involves the post-treatment stage of the stripper, parameter settings for the stripper components were not included.

## 3. Results and discussion

### 3.1. Preliminary screening for organic fluid

In the preliminary screening for optimal organic fluid applied in the DOS for biphasic solvents, the main selection criteria for organic solvents were as follows: the water solubility, thermodynamic properties, chemical stability and safety, and $CO_2$ regeneration enhancement effect on the solutions.

Pentane, hexane, and cyclohexane have been identified as the optimal organic fluid in DOS for the regeneration process of 30 wt% MEA solution among various alkanes in our previous work [22]. The thermodynamic and physical properties of pentane, hexane, and cyclohexane are provided in Table S2.

As the boiling point and latent heat of vaporization of the organic solvent rise [29,30], the additional external heat required for complete vaporization increases. Based on Table S2, the ranking for extra heat consumption for vaporization among the three solvents was: cyclohexane > hexane > pentane. Furthermore, pentane, hexane, and cyclohexane all exhibited remarkably low water solubility, approximately $10^{-5}$ mol/mol, throughout the temperature range of 20 to 80 °C [31,32]. Additionally, these three organic solvents were mixed with the saturated aqueous phases of the biphasic solvents 2DE1AC and DAH. The mixtures were stored at a temperature of 25 °C for 30 days. As demonstrated in Table S4, the liquid-liquid phase boundary between the organic solvent and the biphasic solution remained stable throughout the insolubility test, with the organic fluid consistently remaining in the upper phase due to its lower density. It confirmed excellent immiscibility between the selected organic fluid and the amine aqueous solutions.

Additionally, the condensation recovery ratio of the organic vapor is one of the most critical criteria in DOS process. A high condensation recovery ratio of organic vapor ensures the stability of system operation while reducing operational costs. To evaluate the condensation recovery ratio of solvent candidates under different condensation temperatures, cooling water at 3, 7, 15, 20, and 25 °C was used in the condensation test. The condensation recovery ratio of each organic vapor is shown in Fig. S5. At each condensation temperature, the ranking of condensation recovery ratio for the three solvents was: cyclohexane > hexane > pentane. The maximum condensation recovery ratio for cyclohexane, hexane, and pentane reached 91 %, 86 %, and 83 %, respectively, at 3 °C. At a cooling temperature of 25 °C which was commonly used in industrial cooling water towers, the condensation recovery ratio for cyclohexane, hexane, and pentane decreased to 86 %, 80 %, and 77 %, respectively.

Furthermore, to assess the enhancement effects of DOS processes applying different organic fluid on the $CO_2$ regeneration of biphasic absorbents, continuous absorption-desorption tests were conducted on 2DE1AC and DAH solvents in the LSCF system under both CRH and DOS processes. The flow rates of pentane, hexane, and cyclohexane were set at 5, 10, 15, and 20 L/min in DOS process. The $CO_2$ regeneration rates of 2DE1AC and DAH under CRH(gas flow rate of 0 L/min) and DOS processes at various stripping gas flow rates were obtained, as indicated in Fig. 2. The electric heating power was controlled at 1.44 kW(opening ratio of 40 %) to ensure consistent thermal driving forces between CRH and DOS processes. The superheat of organic fluid was set at 20 °C.

With the introduction of different organic fluid, the $CO_2$ regeneration rates of 2DE1AC(Fig. 2a) and DAH(Fig. 2b) solutions gradually increased. Among them, pentane stripping exhibited the greatest enhancement effect on the $CO_2$ regeneration process of solutions. Specifically, compared to regeneration process under CRH, the DOS regeneration using pentane, hexane, and cyclohexane at 20 L/min enhanced the $CO_2$ regeneration rate of 2DE1AC by 78.4 %, 58.8 %, and 37.3 %, respectively. Additionally, the $CO_2$ regeneration rate of DAH significantly increased by 65.1 % under DOS using pentane stripping at flow rate of 20 L/min, which was 13.9 % and 26.3 % higher than that achieved under the same DOS conditions with hexane and cyclohexane. Therefore, compared to hexane and cyclohexane, applying pentane as the stripping working fluid in DOS process demonstrated a significant advantage in promoting $CO_2$ regeneration of the 2DE1AC and DAH biphasic absorbents.

Therefore, for the DOS of 2DE1AC and DAH biphasic absorbents, although cyclohexane has the highest condensation recovery ratio, its effectiveness in enhancing $CO_2$ regeneration for biphasic absorbents was revealed to be inadequate. Hexane, while having a condensation recovery ratio similar to that of pentane across various cooling temperatures, was less effective in enhancing $CO_2$ regeneration. The DOS using pentane required the least extra heat input for vaporization and showed a significant advantage in enhancing the $CO_2$ regeneration rate of solutions. Additionally, pentane has a boiling point significantly lower than the temperature within the stripper, allowing it to remain in a stable gaseous state in the stripper and be efficiently separated from the solutions throughout the $CO_2$ capture system. Therefore, pentane was identified as the optimal organic fluid applied in the DOS for the 2DE1AC and DAH biphasic absorbents in this study.

### 3.2. The enhancement mechanism of the DOS on $CO_2$ regeneration

To explore the enhancement mechanism of DOS process on the $CO_2$ desorption process of biphasic solvents, the $CO_2$ loading of each 2DE1AC and DAH solution at different packing height was determined, and the corresponding temperatures at these locations were recorded, as indicated in Fig. 3. The gas flow rate of pentane was set to the optimal DOS regeneration conditions for 2DE1AC and DAH (gas flow rate of 10 and 15 L/min, respectively) to highlight the differences in $CO_2$ desorption of solvents under DOS and CRH process. The electric heating power was set

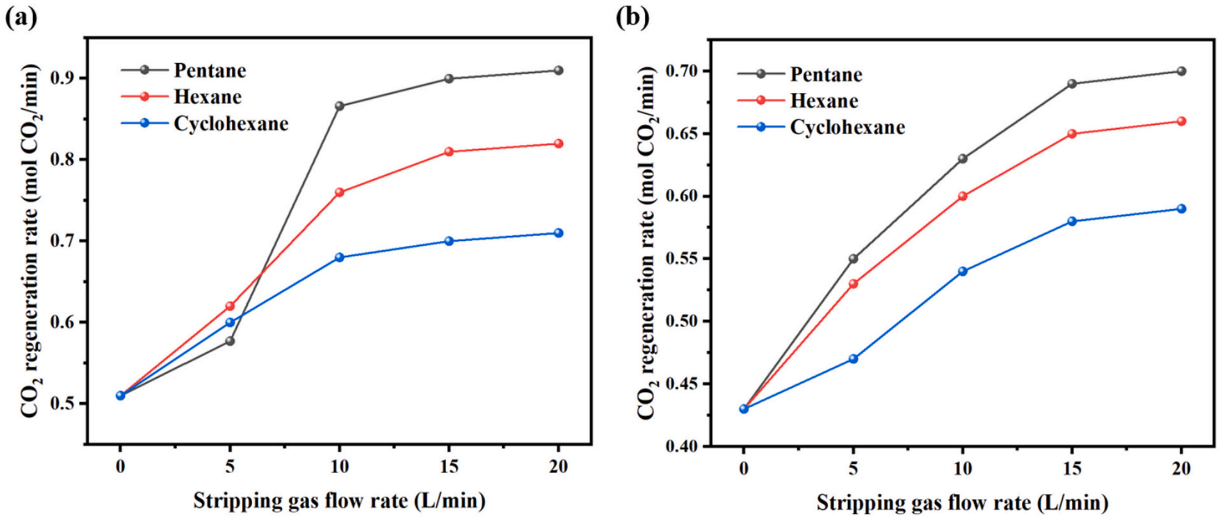

**Fig. 2.** The CO$_2$ regeneration rates of biphasic solvents under CRH and DOS processes using different organic fluids. (a) 2DE1AC solvents. (b) DAH solvents.

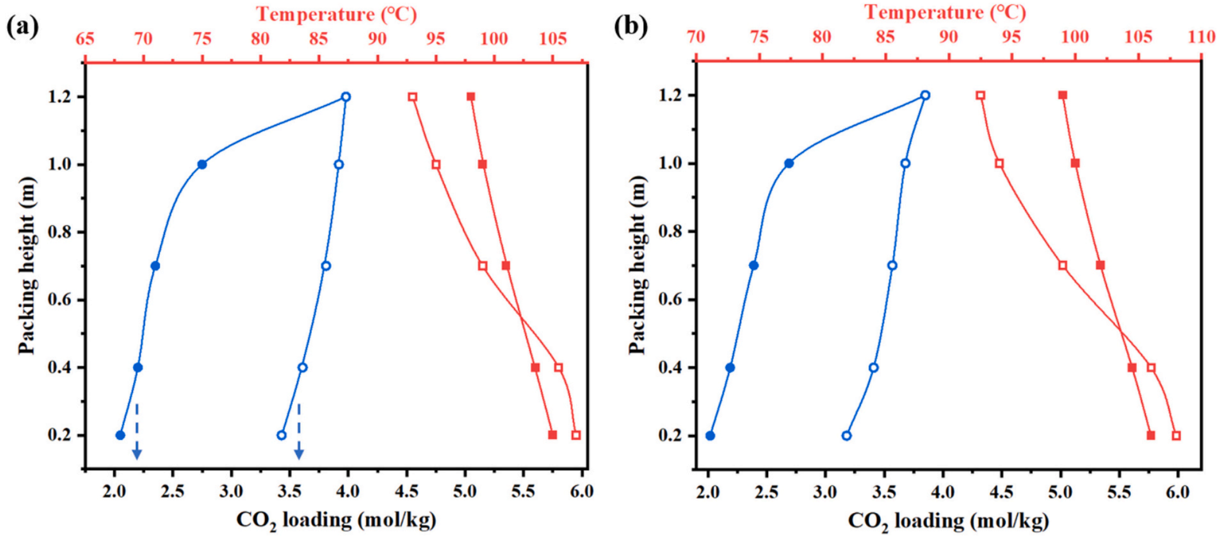

**Fig. 3.** The variation in CO$_2$ loading of the 2DE1AC and DAH solutions and temperature at different heights within the stripper, under both CRH and DOS processes (solid lines represent DOS process, and hollow lines represent CRH process). (a) 2DE1AC (pentane flow rate of 10 L/min under the DOS). (b) DAH (pentane flow rate of 15 L/min under the DOS).

at 1.44 kW, and the gas superheat was 20 °C.

As indicated in Fig. 3, both of the 2DE1AC and DAH solvents exhibited consistent CO$_2$ loading in the CO$_2$-rich phase at the stripper inlet under both CRH and DOS processes, which is close to the theoretical optimum value of each solvent [12,13]. At this point, the solvents demonstrated outstanding phase separation efficiency in the demixer (Table S5). However, at the bottom of the stripper, the CO$_2$ loading in the 2DE1AC and DAH solutions under the CRH was significantly higher than that under the DOS, indicating a lower cyclic capacity of solvents. Analyzed from Fig. 3, for different solutions, the temperature within the stripper decreased to varying extents with increasing packing height in both DOS and CRH process. In CRH process, a 13.8 °C temperature drop from 0.2 m at the bottom of the stripper to 1.2 m at the top was observed in Fig. 3a in the 2DE1AC regeneration process. The phenomenon was similarly observed during the DAH regeneration process, with a temperature fluctuation of 15.5 °C from 0.2 to 1.2 m(Fig. 3b). Due to the temperature fluctuations, the temperature in the top section of the stripper (1.0–1.2 m) was relatively low for 2DE1AC and DAH. Specifically, the regeneration temperatures at inlet of the stripper were only 93.3 and 92.5 °C for 2DE1AC and DAH, respectively, leading to limited thermodynamic driving forces for CO$_2$ desorption. Consequently, even though the solutions at the inlet of stripper had the highest CO$_2$ phase loading and vapor pressure, the CO$_2$ loading variations of 2DE1AC and DAH within the stripper under CRH process remained small.

However, the introduction of pentane stripping efficiently alleviated the uneven temperature distribution within the stripper of 2DE1AC and DAH solutions for the enhanced heat transfer process attributed to the gas flow disturbance of pentane stripping. Specifically, the temperature fluctuations for 2DE1AC and DAH in DOS process were significantly reduced to 6.8 and 6.7 °C from 0.2 to 1.2 m, respectively, which was beneficial for system stability in applications. At the bottom of stripper (height of 0.2 m), the regeneration temperatures of 2DE1AC and DAH in DOS process were slightly lower by 2 °C compared to CRH process. It was due to the lower temperature of the pentane vapor at the inlet of the stripper absorbing some of the heat. The reduced temperature at the bottom inhibited the evaporation of water vapor in the solution, decreasing the latent heat consumption and saving the demand for high-quality heat [33,34]. At the bottom and mid-section(0.2–0.7 m) of the regenerator, both 2DE1AC and DAH solutions showed similar trends in CO$_2$ loading variations under CRH and DOS processes.

Attributed to the efficient heat transfer within the stripper, the regeneration temperatures at the inlet maintained higher values of 98.3 and 99.3 °C for 2DE1AC and DAH in DOS process, respectively. Notably, $CO_2$ loading in the solutions near the top place of stripper(1.0–1.2 m) significantly decreased in DOS process. Specifically, the $CO_2$ loading of 2DE1AC dropped from 3.98 mol/kg to 2.75 mol/kg as the packing height decreased from 1.2 m to 1.0 m, while the $CO_2$ loading of DAH decreased from 3.85 mol/kg to 2.69 mol/kg. It indicated that biphasic absorbent occurred a phenomenon similar to flash vaporization at the top of the stripper in this case. Analyzed from Fig. 3, this was mainly due to that the direct pentane stripping not only increased the $CO_2$ equilibrium partial pressure in the liquid phase of the inlet solution which was associated with higher temperature, but also lowering the $CO_2$ partial pressure in the gas phase within the stripper, resulting in the maximum pressure gradient and remarkably stronger mass transfer driving force for $CO_2$ desorption from the solution at the top of the stripper compared to CRH process. It was revealed as the main contributor to the significant enhancement on $CO_2$ regeneration rates (Fig. 2a and b) and cyclic capacities(Fig. 3) of biphasic solvents under DOS process compared to CRH process [8], which was advantageous for the reduction of capital cost of the stripper [35].

### 3.3. Multiple-condition impact mechanism on the efficiency of DOS process

#### 3.3.1. The impact of stripping gas flow rate

Pentane gas flow rates of 5, 10, 15, and 20 L/min were selected in DOS process for $CO_2$ regeneration of 2DE1AC and DAH biphasic absorbents. The regeneration energy (including reaction heat, latent heat, and sensible heat) and the water evaporation ratio of solutions in the stripper under various conditions were measured and compared with CRH process, as shown in Fig. 4. The electric heating power was set at 1.44 kW, with regeneration temperature at various conditions recorded in the Fig. S6. The gas superheat was 20 °C.

The energy consumption of 2DE1AC and DAH solution under CRH process (Fig. 4a and c, stripping gas flow of 0 L/min) was found to be relatively high, reaching 2.73 and 2.93 GJ/t $CO_2$. The main reason was assumed to the limited thermodynamic driving force provided by CRH process at a regeneration temperature of 100 °C, which resulted in a relatively low $CO_2$ desorption rate(Fig. 2a and b) and a high demand for the regeneration sensible heat(Fig. 4a and c). With the introduction of pentane stripping, for the 2DE1AC solution, it provided extra driving force for $CO_2$ desorption which was strengthened as flow rate increased from 0 to 10 L/min, leading to greater amount of desorbed $CO_2$(Fig. 2a) under the same regeneration temperature in DOS process and significant reduction in the regeneration sensible heat. At a gas flow rate of 10 L/min, the heat required to raise the temperature of the $CO_2$-rich phase within the stripper($Q_{sens}$, Eq. S9) of 2DE1AC was reduced by 85.2 % compared to CRH process. Meanwhile, pentane flow rate from 0 to 10 L/min exhibited a certain inhibitory effect on water evaporation in the solution(Fig. 4b), which could attributed to lower temperatures at bottom of the stripper and uniform temperature distribution(Fig. 3a). At a gas flow rate of 10 L/min, the introduction of pentane reduced the water evaporation by 1.07 wt% in the 2DE1AC solution compared to the CRH, reducing the latent heat of water evaporation($Q_{water}$, Eq. S10) by 0.15 GJ/t $CO_2$. Simultaneously, the total heat consumption of pentane vapor

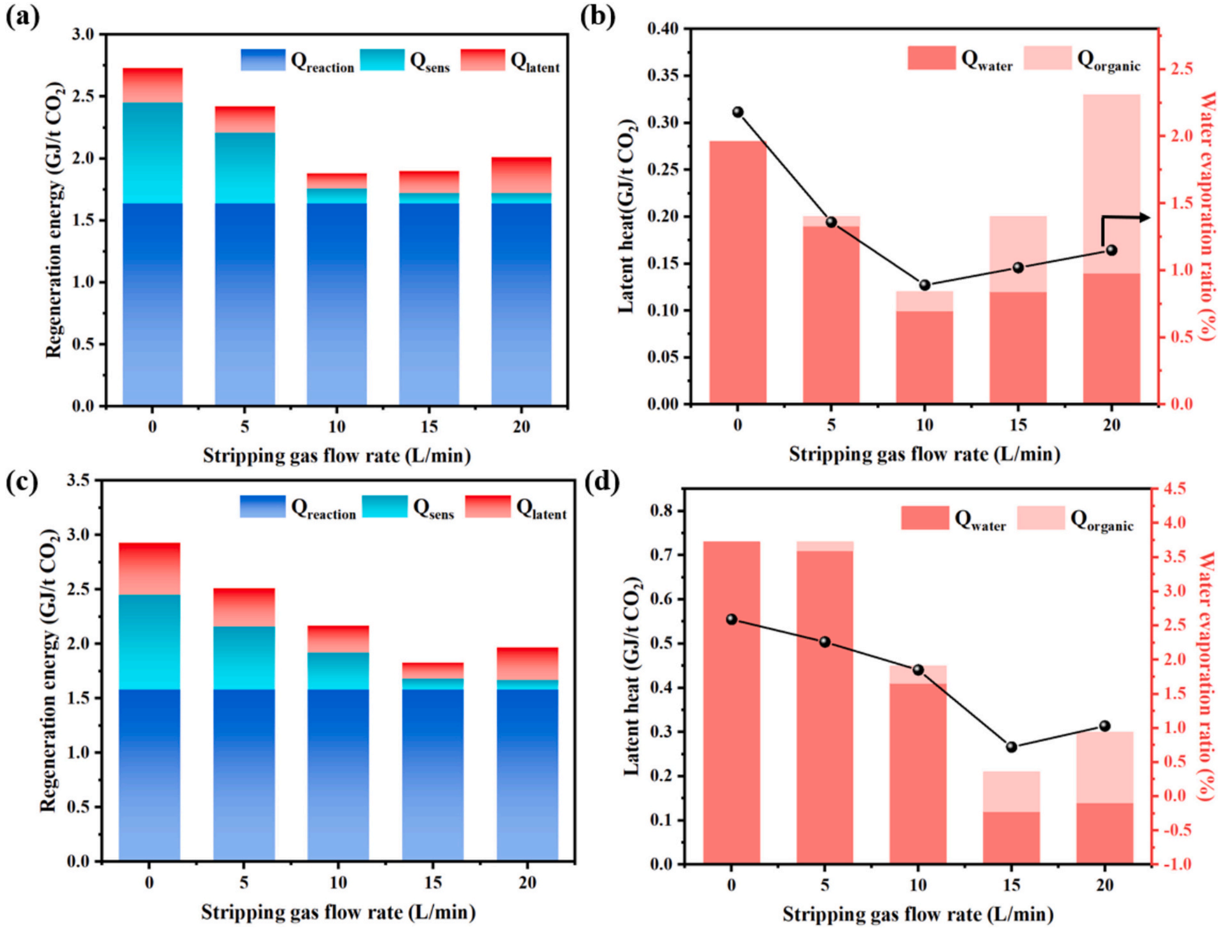

**Fig. 4.** The overall regeneration energy, latent heat, and water evaporation ratio of 2DE1AC and DAH solution under CRH and DOS processes. (a) and (b) 2DE1AC. (c) and (d) DAH.

within the stripper and condenser($Q_{organic}$, Eq. S13) remained relatively stable with increasing gas flow rate(Fig. 4b). Therefore, the total energy consumption of 2DE1AC under DOS with gas flow rate of 10 L/min reduced to 1.88 GJ/t $CO_2$, being 31.1 % lower than that under CRH process. However, as the gas flow rate rose from 10 to 20 L/min, it was found a rise of 6.9 % in the overall regeneration energy consumption of the 2DE1AC. Analyzed from Fig. 2a, it was mainly because that the $CO_2$ regeneration rate of the solution has already reached its peak at a gas flow rate of 10 L/min. Due to the limitations of the $CO_2$ desorption capacity of 2DE1AC, the promotion effect of further increase in pentane flow rate on the $CO_2$ desorption of the solution became limited, resulting in a relatively stable $Q_{sens}$. Meanwhile, elevated latent heat was observed as gas flow rate from 10 to 20 L/min, especially at a range of 15 to 20 L/min(Fig. 4b), which was mainly due to increased heat consumption of organic fluid($Q_{organic}$) during DOS process and water evaporation ratio in the solvents. Specifically, the latent heat increased by 0.17 GJ/t $CO_2$, becoming the primary factor responsible for the deterioration in the overall regeneration energy of 2DE1AC.

A similar phenomenon of the variation for regeneration energy with rising pentane gas flow rates was observed in the DAH solution. With a rising gas flow rate from 0 to 15 L/min, the pentane stripping increased regeneration temperature(Fig. S6b) and facilitated $CO_2$ regeneration rate(Fig. 2b). Meanwhile, the water evaporation in the solvent was inhibited. These led to a continuous decrease in the $Q_{sens}$ and $Q_{latent}$ of the solution(Fig. 4c). At a gas flow rate of 15 L/min, the total regeneration energy consumption of DAH significantly reduced to 1.83 GJ/t $CO_2$, being 37.5 % lower than that under CRH process. However, when the gas flow rate further increased from 15 to 20 L/min, the regeneration energy of the DAH solution under DOS process slightly increased 7.6 %, which was mainly due to the limited enhancement effect of DOS process on the $CO_2$ regeneration of solvents as well as rising water evaporation ratio and heat consumption of organic fluid. Therefore, 10 L/min and 15 L/min were identified as the optimal stripping gas flow rates for 2DE1AC and DAH solutions, respectively, under DOS process.

### *3.3.2. The impact of regeneration temperature*

Regeneration temperature of 90, 95, 100, 105, and 110 °C were selected in this section to analyze the regeneration characteristics of biphasic solvents under CRH and DOS processes at different temperatures within the stripper. To ensure the efficiency of $CO_2$ regeneration in DOS process for 2DE1AC and DAH solutions at different conditions, the pentane flow rates were set to their respective optimal values, which are 10 L/min and 15 L/min, respectively, with a gas superheat of 20 °C.

As indicated in Fig. 5, DOS process enhanced the $CO_2$ regeneration rate and cyclic capacity of biphasic absorbents at different regeneration temperatures compared to CRH process, leading to a reduction in the energy consumption of the solutions(Fig. 6). However, the optimization effect of the DOS on the solvents altered at various temperature.

As the regeneration temperature elevated from 90 to 95 °C, for CRH process, the $CO_2$ regeneration of 2DE1AC and DAH solutions relied solely on thermodynamic driving forces. Analyzed from Fig. 5, it was indicated that $CO_2$ desorption has not yet begun in the solutions in this case, resulting in extremely high sensible heat consumption for regeneration, which was assumed to be the primary reason for the elevated regeneration energy demand(Fig. 6a and Fig. 6c).

However, the pentane stripping promoted the $CO_2$ desorption process of both 2DE1AC and DAH solvents, as indicated in Fig. 5. For DAH solution, the $CO_2$ desorption rate and cyclic capacity of DAH under DOS process increase by 6.1 and 2.7 times, respectively, compared to that under CRH process at regeneration temperature of 95 °C(Fig. 5b). It indicated that pentane stripping enabled efficient $CO_2$ regeneration of DAH solvents at relatively low temperature of 95 °C, which could be beneficial for the reduction of amine thermal degradation [36] and equipment corrosion [37]. In this case, sensible heat consumption of DAH under DOS significantly reduced by 1.43 GJ/t $CO_2$ compared to the CRH(Fig. 6c). However, as shown in Fig. 5a, the improvement force of pentane stripping on $CO_2$ desorption rate and cyclic capacity of 2DE1AC solvent was limited at 95 °C. It could be primarily due to the higher $CO_2$ desorption reaction heat of DETA in the 2DE1AC compared to AEEA as a secondary alkanolamine in the DAH [38,39], which increased the difficulty of the $CO_2$ desorption in the 2DE1AC solvents. As a result, the pentane stripping led to a reduction of 0.78 GJ/t $CO_2$ in the sensible heat for the regeneration process of 2DE1AC solution at 95 °C(Fig. 6a) compared to CRH process. On the other hand, at a relatively low temperature of 90–95 °C below the boiling point of water, the latent heat for water evaporation and pentane heat losses of 2DE1AC(Fig. 6b) and DAH (Fig. 6d) solutions remained similarly low value under both CRH and DOS processes. Overall, the regeneration energy consumption of DAH and 2DE1AC solutions decreased under DOS process as the temperature rose from 90 to 95 °C. For DAH, with the pentane stripping, the regeneration energy consumption reached the lowest value of 1.78 GJ/t $CO_2$ at 95 °C, being 46.1 % significantly lower than that under the CRH at the same temperature(Fig. 6c).

As the regeneration temperature increased to 105 °C, under CRH process, $CO_2$ desorption gradually began in both 2DE1AC(Fig. 5a) and DAH(Fig. 5b) solutions with increasing regeneration rate and cyclic capacity, which peaked at 105 °C. It led to a significant reduction in the sensible heat for regeneration of solvents. Although $Q_{latent}$ gradually increased with rising regeneration temperatures for the elevated water evaporation ratio(Fig. 7), the total energy consumption for the 2DE1AC

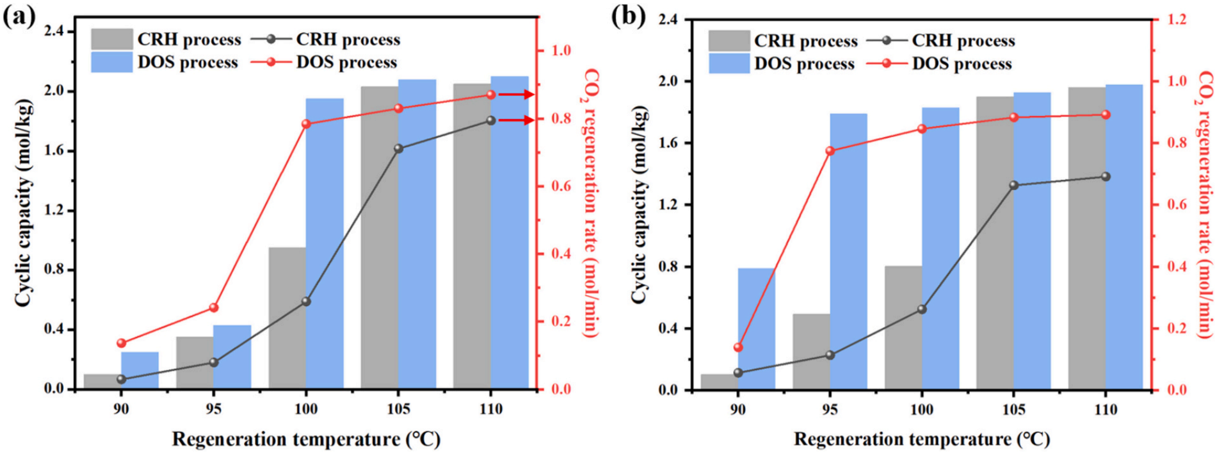

**Fig. 5.** The cyclic capacity and $CO_2$ regeneration rate of 2DE1AC and DAH solution under CRH and DOS processes at various regeneration temperature. (a) 2DE1AC. (b) DAH.

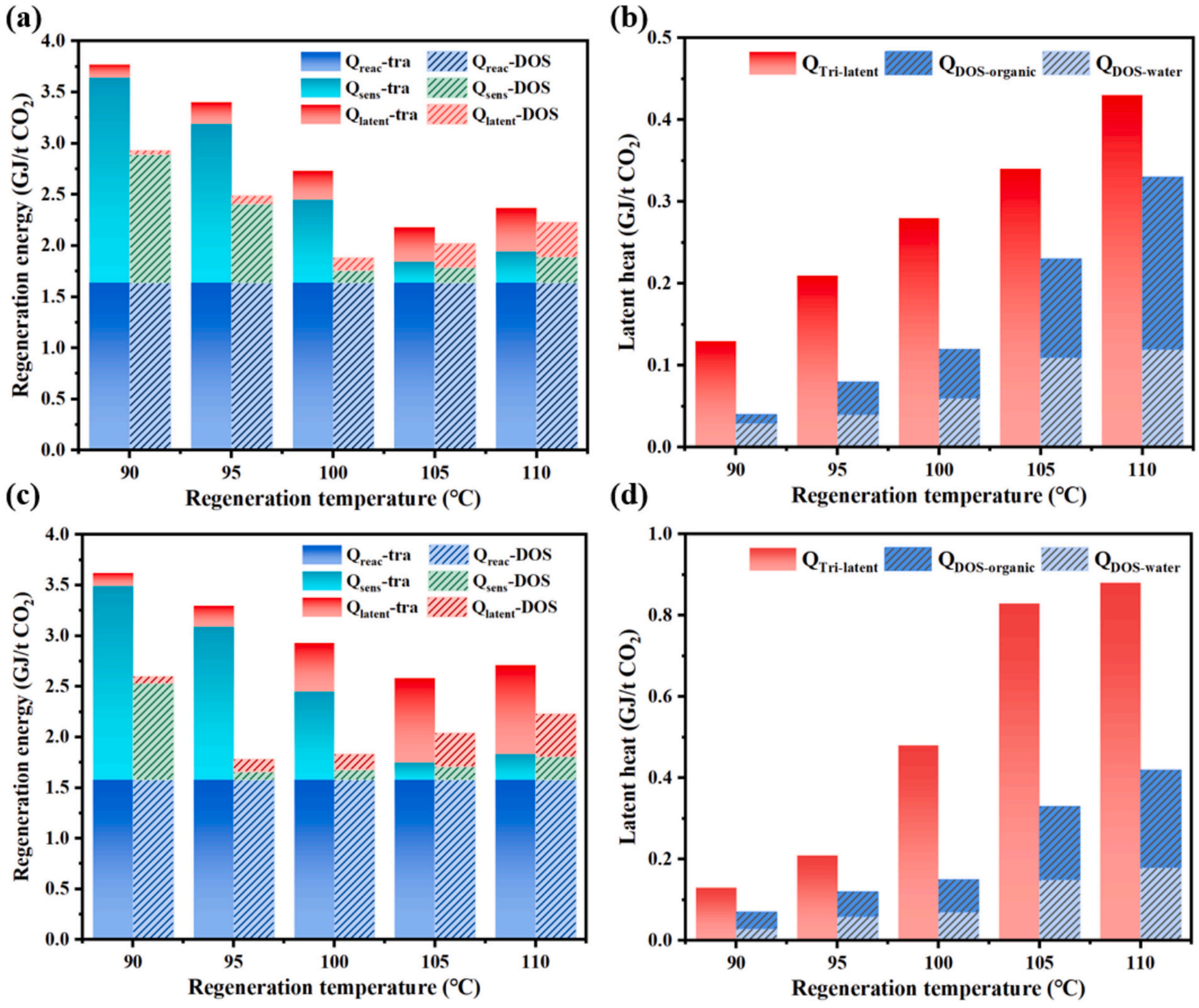

**Fig. 6.** The overall regeneration energy and latent heat of 2DE1AC and DAH solution under CRH and DOS processes at various regeneration temperatures. (a) and (b) 2DE1AC. (c) and (d) DAH.

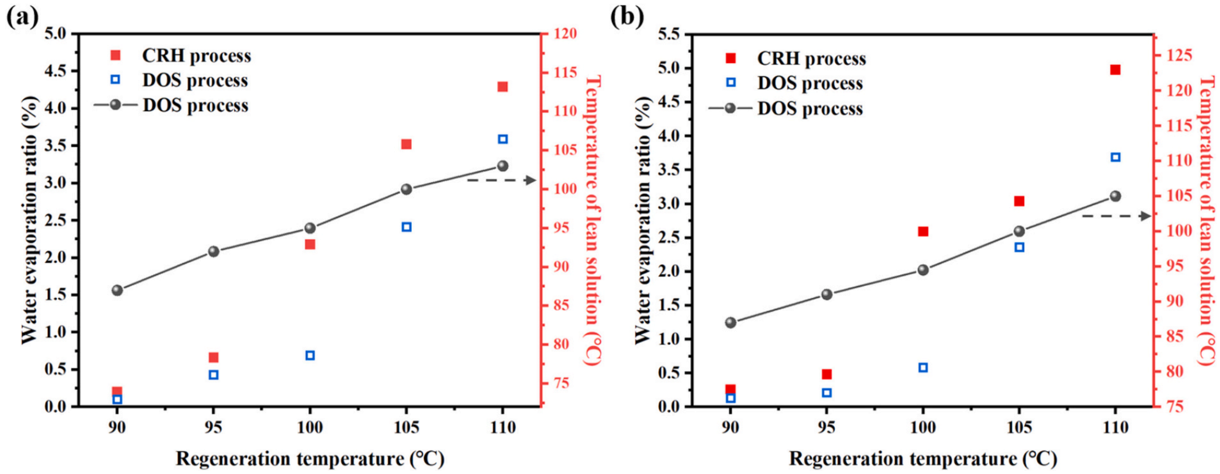

**Fig. 7.** The water evaporation ratio and temperature of lean solution of biphasic solvents under CRH and DOS processes at various regeneration temperature. (a) 2DE1AC. (b) DAH.

and DAH solutions under CRH process kept decreasing with temperature from 95 to 105 °C and reached its lowest values of 2.18 and 2.58 GJ/t $CO_2$ at 105 °C, respectively.

Under DOS process, for the DAH solution, both the sensible and latent heat consumption were observed to increase as the temperature rose from 95 to 105 °C(Fig. 6c). It was due to that the $CO_2$ desorption rate and cyclic capacity had already reached peak values under pentane stripping at 95 °C(Fig. 5b). In this case, further enhancement effect of the DOS on the $CO_2$ desorption in the DAH solution became limited as the regeneration temperature continued to rise, while the heat required for heating the lean solution to the regeneration temperature gradually rose, leading to a slight increase in $Q_{sens}$ of the DAH. Meanwhile, as the regeneration temperature increased from 100 to 105 °C, with temperature of the lean solution exceeded 94.7 °C, the rising temperature caused a significant rise in the water evaporation ratio(Fig. 7b), and thereby exacerbated $Q_{water}$ under DOS process for the solution(Fig. 6d). Overall, the regeneration energy consumption of DAH solution under DOS process increased to 2.04 GJ/t $CO_2$ as the regeneration temperature rose from 95 to 105 °C, which was only 20.9 % lower than that under the CRH. This indicated that the energy-saving efficiency of pentane stripping for the DAH weakened.

For the 2DE1AC, when the regeneration temperature rose from 95 to 100 °C, the energy consumption under the DOS maintained to decrease (Fig. 6a). It can be primarily attributed to the continuous enhancement effect of pentane stripping on the $CO_2$ desorption process of the 2DE1AC solution(Fig. 5a). Specifically, the $CO_2$ regeneration rate and cyclic capacity of the solution under DOS process was respectively 2.1 and 1.1 times higher than that under CRH process as temperature reached 100 °C, leading to a reduction of 0.65 GJ/t $CO_2$ in the sensible heat of solution under the DOS. Meanwhile, pentane stripping demonstrated an inhibitory effect on water evaporation compared to CRH process, while incurring minimal heat loss from pentane. This resulted in a reduction of latent heat consumption by 0.16 GJ/t $CO_2$ compared to CRH process. Overall, DOS process optimized the energy consumption of 2DE1AC reached its lowest value of 1.85 GJ/t $CO_2$ at lower temperature of 100 °C, being 32.2 % lower than that under CRH process at the same temperature(Fig. 6a). However, when the regeneration temperature continued to increase to 105 °C, with temperature of the lean solution exceeded 95.1 °C(Fig. 7a), similar to the DAH solution, the regeneration energy of 2DE1AC under the DOS increased by 0.22 GJ/t $CO_2$, which was mainly due to the increasing latent heat for water evaporation and pentane heat losses in DOS process.

When the regeneration temperature reached 110 °C, it was found that the sensible heat and latent heat for both 2DE1AC and DAH solutions within the stripper further increased under both CRH and DOS processes. Although the addition of pentane can still reduce the regeneration energy consumption of 2DE1AC and DAH solutions compared to CRH process, the rising sensible and latent heat of solvents within the stripper under DOS process deteriorated its effectiveness in optimizing energy consumption at excessively high temperatures. For instance, at 110 °C, the addition of pentane only resulted in a reduction of 5.9 % and 17.7 % in energy consumption for the 2DE1AC(Fig. 6a) and DAH (Fig. 6c), respectively, compared to CRH process. Therefore, to optimize the efficiency of DOS process and ensure the economic viability of the system, it is essential to maintain the stripping gas flow rate and regeneration temperature within an appropriate range in practical applications.

### 3.4. The operational stability of the coupling process of the DOS with biphasic solvents

To evaluate the operational stability of the coupling process, and the comprehensive impact of the DOS on the absorption-desorption cycling system of biphasic solvents in the long-term application, 2DE1AC and DAH solvents were conducted in 180-h continuous tests under DOS and CRH processes in the LSCF system. The overall performances, including $CO_2$ absorption-desorption and phase separation parameters of biphasic solvents were monitored over the 180-h period. Additionally, the amine loss ratio of solvents under both DOS and CRH processes was compared. In the experiments, DOS and CRH processes were operated at optimal regeneration condition parameters for each solution, as obtained in section 3.3.

During the operation of DOS and CRH processes, the systems reached stability within 1–2 h. Subsequently, the fresh solutions demonstrated efficient $CO_2$ absorption-desorption and phase separation performance under both CRH and DOS processes within 30-h operation. In this case, the $CO_2$ removal ratio(Fig. 8a and Fig. 9a) of solvents in the absorber was beyond 94.0 % and total $CO_2$ loading(Fig. S7) was at approximately 1.76 and 2.15 mol/kg for the 2DE1AC and DAH solvents, respectively. When entering the demixer, the 2DE1AC and DAH solvents efficiently separated into biphasic solvents within 1.92 and 2.02 mins(Fig. S9), respectively, with both the $CO_2$-rich phase loading and phase separation ratio(Fig. S8) approaching their theoretical optimal values. The high enrichment of $CO_2$ in the aqueous phase and the rapid phase separation process ensured the regeneration efficiency of the solution. The regeneration energy consumption(Fig. 8c and Fig. 9c) of 2DE1AC and DAH maintained lowest value of 2.17 and 2.58 GJ/t $CO_2$, respectively, under CRH process, which was further reduced to approximately 1.85 and 1.78 GJ/t $CO_2$, respectively under the DOS, exhibiting stable enhancement effect of DOS process on the regeneration performance of biphasic solvents. Moreover, the $CO_2$ loading of the 2DE1AC and DAH mixed solution(Fig. S7) entering the absorber remained well below the phase separation point of each solvent [12,13], which ensured an efficient absorption-desorption cycle [40].

However, it is noteworthy that the overall performance of the 2DE1AC and DAH solvents exhibited intermittent deterioration since 30 and 33 h, respectively, under CRH process, whereas this was delayed to 35 and 37 h under DOS process. This intermittent variation trend was also observed in the amine loss ratio (Fig. 8b and Fig. 9b) of the solvents, with acceleration of amine loss occurring before and closely coinciding with the onset of overall performance deterioration for each solution under both processes. Therefore, after ruling out the possibility of side reactions between pentane and amine solutions during the tests (detailed in the Supplemental Material), the continuous increase in amine loss, which was attributed to different degradation pathways within the biphasic solvents, was considered the primary factor contributing to the gradual deterioration in the overall performance of 2DE1AC and DAH solutions.

Under CRH process, the amine loss ratio of the 2DE1AC and DAH solutions reached 20.3 % and 19.7 %, respectively, at 180 h. The substantial active amine losses severely impacted the absorption-desorption as well as phase separation performance of the solutions. For the DAH solution, the total $CO_2$ loading(Fig. S7c) and $CO_2$ removal efficiency (Fig. 9a) decreased by 22.3 % and 12.4 %, respectively throughout the test under CRH process. Meanwhile, the $CO_2$-rich phase loading (Fig. S8c) dropped by 19.3 %, reaching only 2.89 mol/kg at 180 h. Although the phase separation time shortened with decreasing total $CO_2$ loading, the insufficient $CO_2$ enrichment in the aqueous phase led to a 33.5 % reduction in the cyclic capacity of the DAH solution(Fig. S7c) at 180 h, with regeneration energy consumption(Fig. 9c) increasing by 41.5 %. Additionally, the increasing amine loss had a more severe impact on the overall performance of 2DE1AC for its relatively low initial amine concentration. In the absorber, the total $CO_2$ loading (Fig. S7a) and $CO_2$ removal efficiency(Fig. 8a) of 2DE1AC significantly decreased by 28.9 % and 18.4 %, respectively throughout the test. In the later stage of testing (170–180 h), phase separation of the solvent disappeared in the demixer due to the inability of 2DE1AC to reach the phase separation point within the absorber(Fig. S7a). In this case, $CO_2$ in the 2DE1AC could no longer be enriched in the aqueous phase, resulting in a sharp deterioration in cyclic capacity(Fig. S7a) and regeneration energy consumption(Fig. 8c) within the stripper.

Compared to CRH process, the 2DE1AC and DAH solvents exhibited limited amine loss under DOS process throughout the test, which was benefited from the efficient mass transfer and the lower regeneration temperature within the stripper. It was indicated that the amine loss ratio of the 2DE1AC and DAH solvents during the 180-h operation decreased to 5.9 % and 6.5 % under the DOS, respectively, which represented a significant reduction of 14.4 % and 13.2 % compared to CRH process. In this case, the 2DE1AC and DAH maintained efficient absorption-desorption and phase separation performance during the

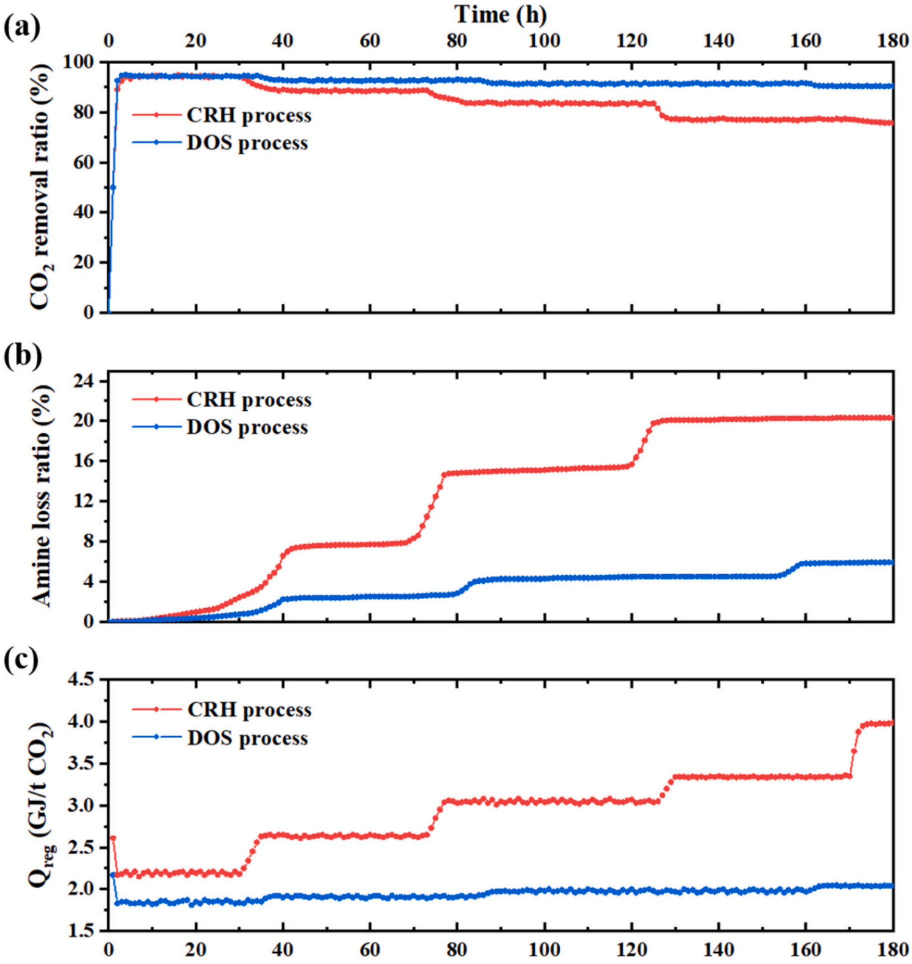

**Fig. 8.** The operational parameters of the 2DE1AC solvent under CRH and DOS processes. (with a regeneration temperature of 100 °C and a pentane flow rate of 10 L/min under DOS process, and a regeneration temperature of 105 °C under CRH process).

tests. In the absorber, the $CO_2$ removal efficiency of 2DE1AC and DAH solutions remained above 90.1 %, while the total $CO_2$ loading(Fig. S7b and Fig. S7d) stayed above 93.2 % and 93.8 %, respectively, of the initial $CO_2$ capacity of each solution. This ensured phase separation stability and enhanced the phase separation efficiency for biphasic solvents, particularly for 2DE1AC, during the long-term application. At 180 h, the $CO_2$-rich phase loading(Fig. S8b and Fig. S8d) of 2DE1AC and DAH solutions exceeded 3.61 and 3.33 mol/kg, respectively, with phase separation time(Fig. S9) within 1.46 and 1.54 min.

Within the stripper, DOS process consistently enhanced the $CO_2$ regeneration of 2DE1AC and DAH, maintaining their cyclic capacity (Fig. S7b and Fig. S7d) at approximately 1.81 and 1.63 mol/kg at 180 h. In this case, the regeneration energy consumption(Fig. 8c and Fig. 9c) of the 2DE1AC and DAH remained below 2.04 and 1.93 GJ/t $CO_2$, respectively, which was 48.7 % and 47.1 % significantly lower than that under the CRH. Meanwhile, it was found that the introduction of DOS process could mitigate the adverse impact of increasing amine loss on $CO_2$ regeneration performance of biphasic solvents. For instance, under CRH process, when the amine loss ratio of 2DE1AC reached 5.5 % at 39 h, its cyclic capacity decreased by 18.7 % compared to the fresh solution at the beginning of the test, with a rise of 21.6 % in regeneration energy. However, under DOS process, the same level of amine loss in the solvent at 170 h resulted in only 6.7 % reduction and 9.7 % increase in cyclic capacity and regeneration energy, respectively, compared to the fresh solution. A similar trend was also observed in the tests for DAH solution.

Moreover, 2DE1AC and DAH solution samples at different heights within the stripper and the outlet of absorber were continuously monitored during the 180-h tests under the DOS, with no liquid pentane was detected in all solutions. It further validated efficient separation of organic fluid and amine solution within the absorption-desorption cyclic system.

### 3.5. Pre-compression enhanced recycle for DOS

The mixed gas at the outlet of the stripper, primarily composed of $CO_2$ and pentane, along with a small amount of water, contained significant energy input from low-grade heat sources and the stripper. However, the original process of direct water-cooling process(Fig. 1) lacked of measurements for energy recycle, leading to rising energy wastes in $CO_2$ capture system. Moreover, due to the low boiling point and high vapor pressure of pentane, the condensation recovery ratio by original process remained insufficient(Fig. S5), which would result in higher operational costs in practical applications.

To address these issues, a pre-compression enhanced recycle process was proposed and simulated on a 4 $Nm^3/h$ scale using Aspen Plus V11 to align with the experimental study (simulation information detailed in the Supplementary Material). As indicated in Fig. 10, the mixed gas from the top of the stripper first underwent a stage of pre-compression for energy upgrading and then passed through an economizer, where it preheated the returning liquid pentane from the bottom of the separator. In this case, the energy of the pre-compression mixed vapor, including the energy from the low-grade heat source, the stripper, and pre-compressor were effectively recycled. After passing through the economizer, the mixed gas continued to contact with cooling water, where the

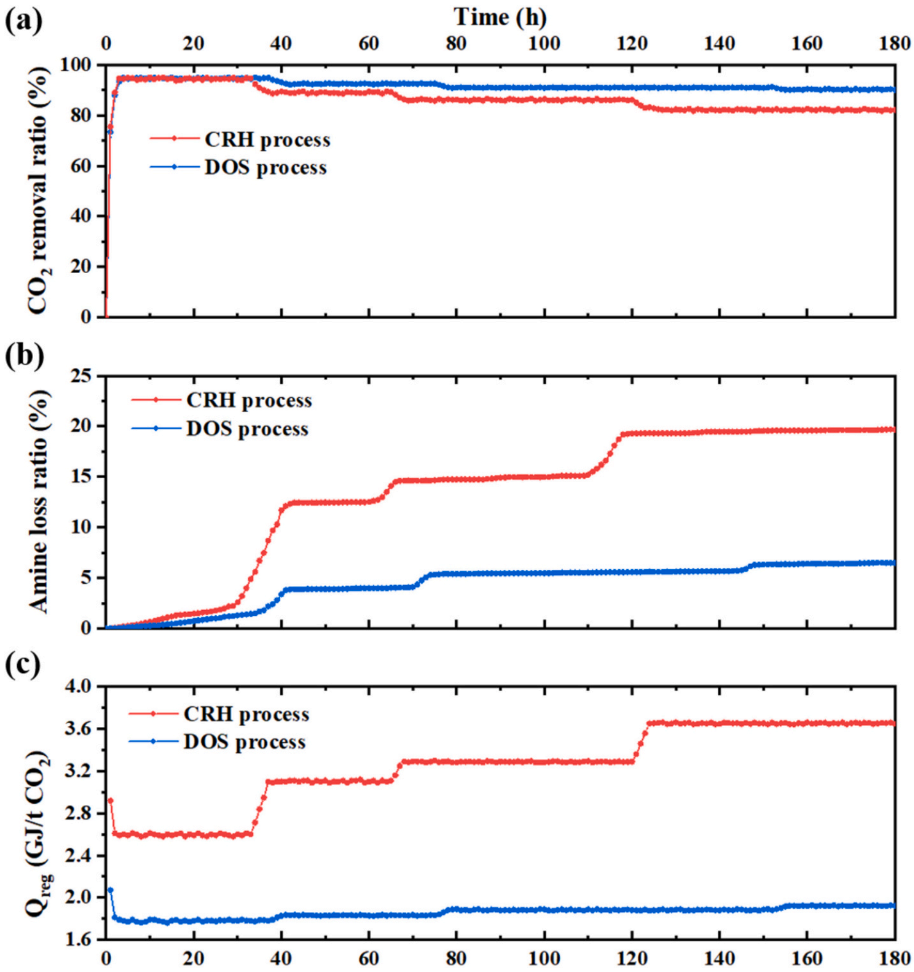

**Fig. 9.** The operational parameters of the DAH solvent under CRH and DOS processes (with a regeneration temperature of 95 °C and a pentane flow rate of 15 L/min under DOS process, and a regeneration temperature of 105 °C under CRH process).

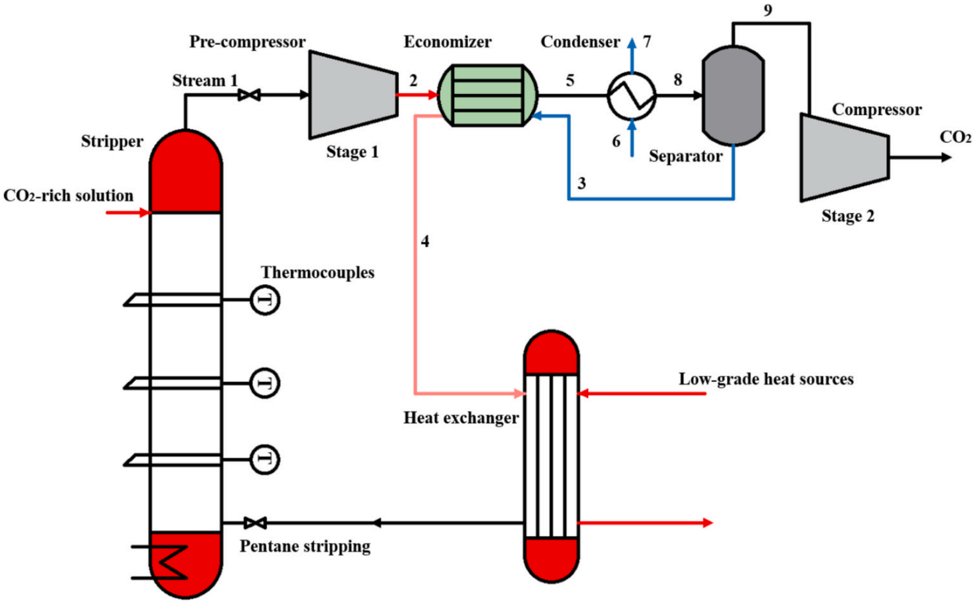

**Fig. 10.** The system diagram of the pre-compression enhanced recycle process integrated with DOS process.

increased temperature difference between the pressurized gas and the cooling water enhanced the pentane condensation recovery ratio. Finally, after the pentane separation, the remaining gas, primarily $CO_2$, underwent a second stage of compression for pressurized $CO_2$ storage. Based on the cost assessment methodology from McCollum et al. [41], the capital cost of the compressor is unlikely to significantly rise under the co-compression of pentane and $CO_2$ in stage 2. The amount of recycled energy(GJ/t $CO_2$) from the mixed gas at the outlet of stripper (Fig. 11a) and the pentane condensation recovery ratio(%) under different operating conditions(Fig. 11b) were obtained. Since pre-compression energy constitutes an additional energy input in the process, the recycled energy presented in Fig. 11 has excluded the energy consumption of pre-compression (Eq. S19) to facilitate a direct comparison with the original DOS process.

As shown in Fig. 11a and b, with the increase in pre-compression pressure(absolute pressure), both the recycled energy and the pentane condensation ratio at various condensation temperatures gradually rose in the economizer and condenser, respectively. At a pre-compression pressure of 8 atm and a condensation temperature of 263 K, the recycled energy reached the maximum value of 0.65 GJ/t $CO_2$, with a condensation recovery ratio of 99.7 %, representing a 13.7 % improvement over the original process at same condensation temperature.

However, the pentane condensation ratio was found to significantly decrease when the temperature exceeded 293 K under different pre-compression pressures, with a maximum of only 83.9 % at a condensation temperature of 303 K and a pre-compression pressure of 8 atm, which fails to meet the pentane recovery requirements. Additionally, this led to a lower mass flow rate and a higher temperature of the liquid pentane at the separator outlet (stream 3), which was assumed to be the primary reason for the lower recycled energy in the processes at condensation temperatures higher than 293 K(Fig. 11a). On the other hand, further decrease in condensation temperature from 293 to 263 K exhibited limited improvement effect on pentane recovery efficiency (Fig. 11b) as well as the recycled energy at various pre-compression pressures. For example, when the pre-compression pressure was 8 atm, reducing the condensation temperature from 323 to 293 K resulted in a significant increase in recycled energy of 0.56 GJ/t $CO_2$, while the temperature changed from 293 to 263 K only led to a 0.05 GJ/t $CO_2$ increase in recycled energy. Conversely, excessively low condensation temperatures can increase operational costs and reduce economic benefits.

Meanwhile, with a condensation temperature of 293 K, increasing the pre-compression pressure from 1 to 5 atm significantly enhanced recycled energy and pentane condensation, achieving a condensation ratio and energy recovery of 92.4 % and 0.51 GJ/t $CO_2$, respectively, at a compression pressure of 5 atm. It indicated an excellent heat exchange efficiency between the pre-compressed vapor with stream 3 and cooling water in the economizer and condenser, respectively. Additionally, a pentane recovery ratio of over 90 % enabled the removal of the activated carbon tower in practical applications, which can further enhance overall process economics. However, it was worth noting that pre-compression pressure of mixed vapor from 5 to 8 atm only slightly increased the condensation ratio and recycled energy by 4.8 % and 0.09 GJ/t $CO_2$, respectively. Since operating at excessively high pressures could led to many problems including increasing operational costs and system instability, pre-compression pressure beyond 5 atm was assumed unsuitable for the DOS.

Therefore, a pre-compression pressure at stage 1 of 5 atm along with a condensation temperature of 293 K was considered ideal for efficiently recovering system energy and stripping working fluid while ensuring system stability in applications.

## 4. Conclusions

A low-energy consumption $CO_2$ chemisorption strategy via coupling process of direct organic fluid stripping(DOS) with biphasic solvents was proposed in this study. Pentane was identified as the ideal organic fluid for biphasic solvents due to its low vaporization heat requirement and outstanding performance in $CO_2$ regeneration enhancement. Compared to the conventional reboiler heating(CRH), DOS process effectively eliminated temperature non-uniformity within the stripper and provided an extra mass transfer driving force for the $CO_2$ regeneration of biphasic solvents, enhancing the $CO_2$ desorption rates as well as cyclic capacity of solvents. Increasing the stripping gas flow rate and regeneration temperature could strengthen the enhancement effect of DOS on the $CO_2$ regeneration of solvents, while excessive gas flow rate and high regeneration temperature would weaken the efficiency of DOS process. The DOS achieved minimum regeneration energy consumptions of 1.85 GJ/t $CO_2$ for the DETA/DEA/DMAC biphasic solvent and 1.78 GJ/t $CO_2$ for the DEEA/AEEA biphasic solvent at temperatures below 100 °C, representing significant reductions of 32.2 % and 46.1 %, respectively, compared to CRH process under the same conditions. The operational stability of the coupling process was validated in 180-h absorption-desorption cycling tests. While strengthening $CO_2$ regeneration within the stripper, DOS process significantly suppressed amine loss in solvents (<6.5 %) compared to CRH process, maintaining high $CO_2$ removal ratio (>90.1 %) and phase separation stability of biphasic solvents.

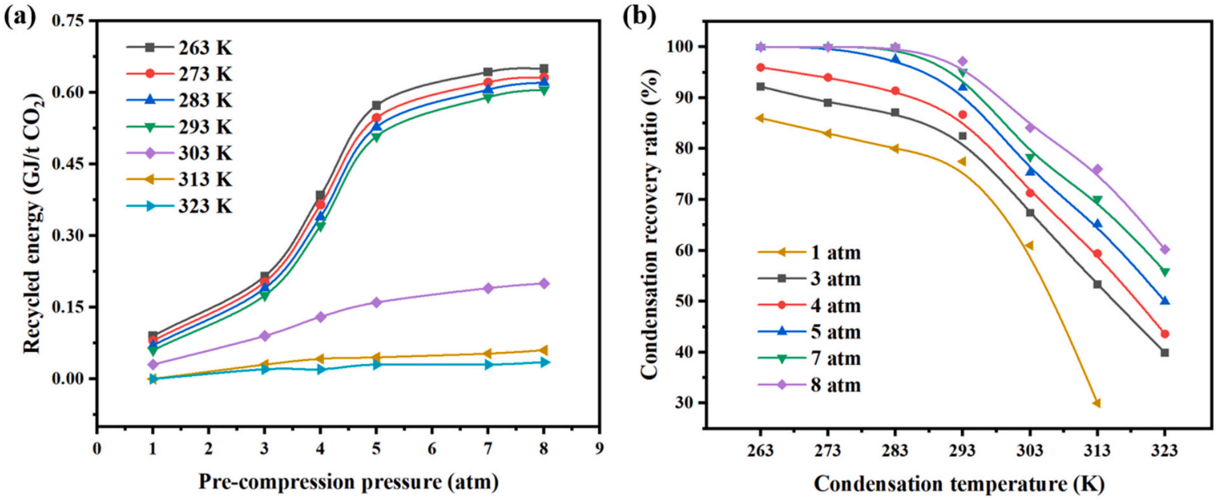

**Fig. 11.** Process simulation results of pre-compression enhanced recycle process under various conditions. (a) Recycled energy of the mixed gas at the stripper outlet. (b) Pentane recovery ratio.

Furthermore, the pre-compression enhanced recycle process integrated with the DOS enhanced the energy efficiency and pentane recovery, achieving a maximum increase of 0.65 GJ/t $CO_2$ in recycled energy compared to conventional recycling process and a pentane recovery ratio of 99.7 %.

## CRediT authorship contribution statement

**Zhipeng Chen:** Writing – original draft, Software, Methodology, Investigation, Formal analysis, Data curation, Conceptualization. **Zhengrong Wang:** Funding acquisition. **Kailiang Wang:** Funding acquisition. **Tao Wang:** Writing – review & editing, Supervision, Resources. **Luchang Sun:** Funding acquisition. **Mengxiang Fang:** Writing – review & editing, Supervision, Resources. **Long Jiang:** Resources. **Yan Huang:** Software. **Wei Chen:** Resources. **Ximing Hu:** Resources.

## Declaration of competing interest

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

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## Appendix A. Supplementary data

Supplementary data to this article can be found online at https://doi.org/10.1016/j.apenergy.2025.126012.

## Data availability

I would like to submit my data/code if necessary.

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