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Design Methodology of Absorption Process (use of MEA absorbent) based on Self-heat Recuperation Technology

Akira Kishimoto, Yasuki Kansha, Chihiro Fushimi, Atsushi Tsutsumi*
Collaborative Research Center for Energy Engineering, Institute of Industrial Science,
The University of Tokyo, 4-6-1 Komaba, Meguro-Ku, Tokyo, 153-8505, Japan
a-tsu2mi@iis.u-tokyo.ac.jp

A chemical absorption process is the most commercially used process for CO_2 separation in flue gases. However, the process needs a huge amount of energy to strip CO_2 in a stripper. To overcome this problem, a new gas separation process using chemical absorption method based on self-heat recuperation technology is proposed in this study for energy saving. In this process, the regeneration energy of absorbent is provided from the exhausted heat of absorber and stripper by using a compressor and heat exchangers. The process is divided into two modules (absorber and stripper), in which the internal heat circulation is maximized. Then, the process is reconstructed by combining these modules. We evaluated the amount of energy consumption of the process as compared with the conventional gas separation process for CO_2 by using a commercial process simulator (PRO/II). From the simulation results, energy consumption of the proposed process decreased to one-third at that of conventional heat recovery process. Thus, the proposed process based on self-heat recuperation technology is a very promising process for energy saving of gas separation.

1. Introduction

The concern over the CO_2 separation technology has risen because of Global warming. Absorption, adsorption, cryogenic separation and membrane separation are generally used to capture CO_2. These separation methods are commercially available to capture not only CO_2 but also natural gas in refinery plants and waste gases in other industries.

The critical issue of gas separation process using the chemical absorption is that the process needs a huge amount of energy to strip CO_2 in a stripper. Nowadays, it has been often discussed that amine-based chemical absorption processes to improve the effectiveness and efficiency of CO_2 capture (Goto et al., 2009). One of the most famous amine for CO_2 capture is a monoethanolamine (MEA). It is well known that MEA required a large amount of energy (4.0 GJ/t-CO_2) for CO_2 stripping. Therefore, Goto et al. (2009) developed a new absorbent to reduce the regeneration energy in the stripper. In addition, Leites (2003) analyzed some industrial CO_2 chemical absorption processes and proposed the optimization methods for process energy based on examination of driving force and quasi-static analysis, but the study still has not proposed optimizing the system configuration. They have paid less attention to process configuration. In

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contrast, Barchas and Davis (1992) proposed an energy saving process. In their study, they installed a flash column to CO_2 rich stream between an absorber and a stripper. However, the process lost the effect of reduction energy while CO_2 partial pressure and the absorption process pressures were low.

Figure 1 shows a schematic diagram of the conventional CO_2 separation process (The heat recovery process) which uses amine absorption method. The process equips a heat exchanger (HX1) for reboiler heat recovery. We analyzed the heat input and output of the processes, and found that the required energy is consumed in the following three places; i) process after cooler of the S3, ii) the lean solvent cooler between absorber and stripper, and iii) the reflux condenser at the top of stripper.

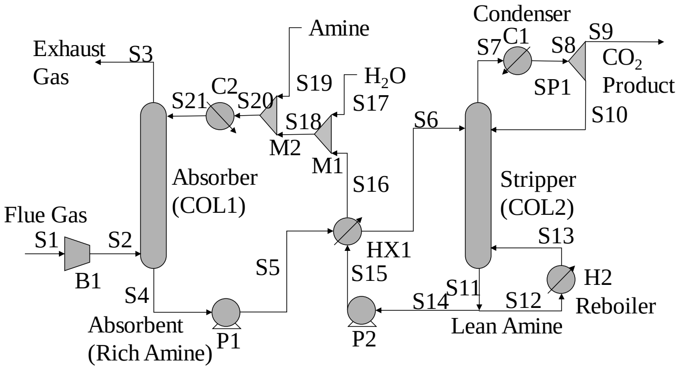


Figure 1: Base MEA stripping process (The heat recovery process)

Next, we analyzed the heat recovery system of the conventional absorption process from the energy and material balance. Recovered heat is only a part of the exothermic reaction heat in the absorber and the vaporization heat in the stripper. This means that the conventional process could afford to recover more exhausted heat in the process. Kansha et al. (2009, 2010) proposed the design technology for process energy saving, in which not only the latent heat but also the sensible heat of the process stream can be circulated without any heat addition. Based on the study, we propose CO_2 separation process of lower energy consumption as an industrial application in this study.

2. Proposed process based on self-heat recuperation

CO_2 separation process using chemical adsorption method based on self-heat recuperation has been proposed. The proposed process is designed by the following important steps.

2.1 Process Division

The CO_2 separation process consists of an absorber and a stripper. The process shown in Figure 1 is divided into two simple processes, and then the heat balance of each simple process is analyzed. Therefore, energy input and output of the CO_2 separation process can be distinguished. In addition, heat-balanced simple processes are named absorption and stripping modules. A feed and effluent in each module are set at 25 °C and 1 atm as the standard condition.

2.2 Design of self-heat recuperative module

In the aforementioned analysis, pairs of feed and effluent streams are selected to recover the heat. To recover the heat among selected pairs, the self-heat recuperation technology is adopted. In the self-heat recuperation technology, the stream condition is changed by means of the compression of the process stream and the stream can provide self heat to stream of the pair in the heat exchanges, leading to that the self heat of the process stream is recirculated based on exergy recuperation to reduce the process energy consumption. Following series of steps, self-heat recuperative module is constructed.

3. Simulation

In this paper, process simulation with PRO/II (Invensys plc.) was conducted to calculate the process energy consumption of absorption process. We used a standard amine package model of PRO/II (Invensys plc.) in the software for this simulation. In thermodynamics data, liquid enthalpy is calculated by adding a correction for a heat of reaction. The liquid enthalpy and liquid phase density is calculated by ideal method.

Table 1. Concentration of each gas *in flue gas*

Flow rate of flue gas		100	kmol/h
Concentration of CO_2 in flue gas	N_2	77	mol%
	CO_2	10	mol%
	H_2O	10	mol%
	O_2	3	mol%
Degree of CO_2 removal		>99	%
Purity of recovery CO_2		90	%

Vapor phase enthalpy, entropy and density are calculated by SRKM. We assumed that the flue gas was a natural gas, and that the amount of flue gas and CO_2 were 100 and 10 kmol/h, respectively. (Table 1).

4. Results and Discussion

4.1 The conventional heat recovery process

The conventional heat recovery process and its energy consumption are shown in Figure 2. The inlet flue gas (S0) was set to the standard condition. S0 was heated to 42.2 °C, 101.3 kPa. The heated flue gas (S1; 42.2 °C, 101.3 kPa) was compressed, and supplied to an absorber (S1→B1→S2, 45.0 °C, 101.3 kPa). In the absorber, aqueous MEA absorbs CO_2 from the flue gas (COL1), and the remaining gases are discharged from the top of the absorber (S3). The influent rich amine into the absorber is pumped, accordingly the influx is kept liquid phase (S4→S5→S6). The conditions of S4 and S6 were 55.5 °C and 111.4 kPa, and 107.3 °C and 501.3 kPa respectively. The heat of S4 includes a part of exothermic reaction heat, which is effectively used as a part of CO_2 stripping energy. The stripper column (COL2) needs a huge amount of energy. The heat of the rich amine sorbent from the bottom of stripper is exchanged with the heat of lean amine solvent (S15→S16), and the rich amine sorbent is heated by recovery (S5→S6). The absorbed CO_2 is stripped by a part of heat of exothermic reaction and the reboiling heat in the reboiler. S6 is fed into the stripper, and then CO_2 gas and steam are discharged from the top of the stripper (S7). CO_2 gas was cooled, and the steam was condensed by a condenser (C1; 45 ^{o}C, 186.3 kPa), then CO_2 and the condensed water (S8) were separated by a separator (SP1). The condensed water is returned to the stripper as the reflux flow (S10). CO_2 gas is released through the separator (S8→SP1→ S9). Aqueous MEA is boiled, and water changed to the steam in a bottom reboiler (H2) of the column, and the vapor stream is returned to the stripper (S11→S12→S13→ COL2, 121.4 ^{o}C, 186.3 kPa). Lean amine absorbent was pumped to the same pressure as the absorber (S11→S14→P2→S15→S16). In mixers (M1, M2), MEA and H_2O are mixed with the lean amine absorbent (S16→S17→M1→S18→S19→M2→S20). The aqueous MEA was cooled by a cooler (C2) (S20→C2→S21; 45 ^{o}C, 186.3 kPa). The total energy consumption of the process was 507.1 kW. The heat recovery process aimed to recover the exhausted heat of the C2. However, the C2 still wastes a large amount of heat of reboiler duty. In addition, C1 is wasted heat in the condenser. The sum of C1 and C2 duties was 389.6 kW. The wasted heat is larger than recovered heat in HX1. It can be understood that we can achieve more energy saving if these consuming energies are recirculated into the process. In this simulation, the flue and effluent gases were set at 25 ^{o}C and 1 atm as the standard condition in order to compare with the energy consumption of self-heat recuperative process under the same condition.

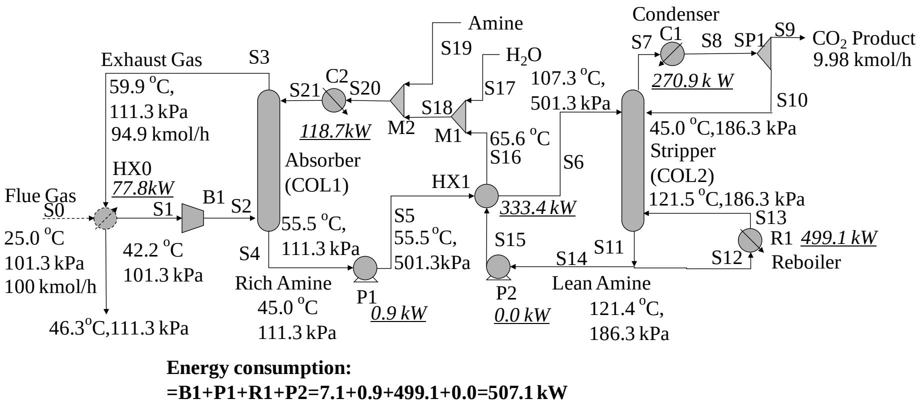


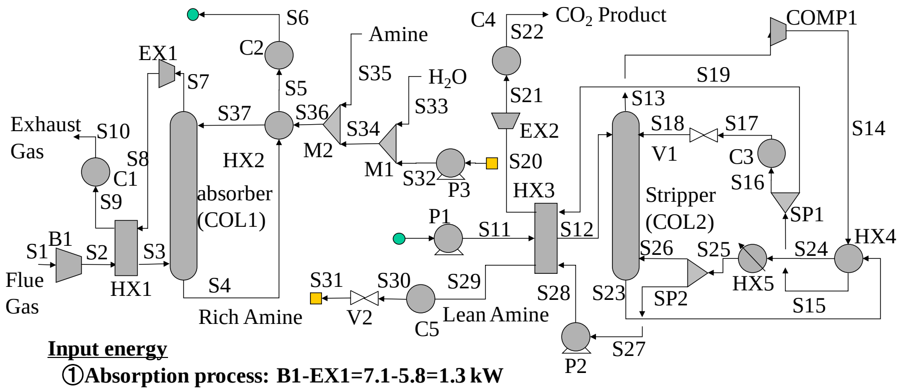
Figure 2. Energy consumption of heat recovery process

4.2 The proposed process

At first, the process shown in Figure 1 was divided into two processes for absorption and stripping. We considered the enthalpy balance of each process in which the conditions of the feed and effluent were set at 25 ^{o}C and 1 atm. In these processes, influent enthalpy is same as effluent enthalpy without reaction heat, that is, the absolute value of heat of exothermic reaction is equal to the absolute value of heat of endothermic reaction. Under the condition, the total energy consumption is expressed by the following equation;

$$Q_{\text{total}} = Q_{\text{ab}} + Q_{\text{st}} \tag{1}$$

where Q_{ab} is energy consumption in the absorption process, Q_{st} is energy consumption in the stripping process, respectively. We designed self-heat recuperative modules from these simple processes. The proposed process, which is reconstructed from the absorption and stripper self-heat recuperative modules, is shown in Figure 3. In these modules, streams recovered waste heat to maximize the recovered energy.



Input energy

①Absorption process: B1-EX1=7.1-5.8=1.3 kW

②Desorption process: P1+COMP1-EX2+P2=0.2+95.5-21.3+0=74.4 kW

③Heat pump: 82.0 kW (COP=2)

①+②=75.7 kW ①+②+③ =157.7 kW

Figure 3. Self-heat recuperative process (proposed process)

The inlet flue gas (S1 25 $^{\circ}$C, 1 atm) was compressed by a blower (B1) and heated by a heat exchanger (HX1) to supply a process heat to the absorption column (COL1) (S1→B1→S2→HX1→S3). The remaining gases are discharged from the top of COL1 (S7: 59.9 $^{\circ}$C, 111.3 kPa), expanded by an expander (EX1), and then the vapor temperature decreased by adiabatic expansion (S7→EX1→S8 56.7 $^{\circ}$C, 111.3 kPa). S2 was heated, and S8 was cooled by HX1 (S2; 28.4 $^{\circ}$C→HX1→S3; 45 $^{\circ}$C, S8; 56.7 $^{\circ}$C→HX1→S9; 46.2 $^{\circ}$C). Afterwards, the gas stream (S9) was discharged from the process and cooled (S9→C1→S10; 25 $^{\circ}$C, 111.3 kPa). The absorbent (S4; 55.5 $^{\circ}$C) was discharged from the bottom of COL1, and then S4 was cooled by the lean amine in a heat exchanger (HX2) and a cooler (C2) (S4→HX2→S5 35.8 $^{\circ}$C→C2→S6 25 $^{\circ}$C, 101.3 kPa). S6 is connected to a pump (P1), and is heated in a heat exchanger (S11→HX3→S12; 107.3 $^{\circ}$C, 501.3 kPa). CO_2 and other gases (S13) were compressed by a compressor (COMP1, 95 kW) (S13; 108.5 $^{\circ}$C, 186.3 kPa→COMP1→S14; 363.4 $^{\circ}$C, 1300.0 kPa). The absorbent (S23, lean amine) were discharged from the bottom of the stripper, and the absorbent were heated by a heat exchanger (HX4; 121.4 $^{\circ}$C). To minimize the exergy loss, only latent heat should be recovered in HX4. However, in the absorption process, gas from COL2 which consists of vapor and CO_2 (S13) provides not only latent heat but also sensible heat in HX4, results in exergy loss. The absorbent (S27) was heated by a heat exchanger (HX5) by exothermic reaction heat transported from absorber (S26→HX4→S27→HX5→S28). HX4 and HX5 have a role of a reboiler. Thus, all of the exhausted heat from cooler is expressed (C1+C2+C3+C4+C5). In contrast, in the proposed process, almost of all exhausted heat is used as the heat of endothermic reaction in the stripper. If we assume that the heat of exothermic reaction is transported to heat of endothermic reaction. The heat required on the reboiler is provided from the heat of exothermic reaction by using heat pump. A performance of the heat pump is expressed the coefficient of performance (COP). Here, the COP is expressed as received heat from exothermic reaction heat divided by supplied compressor work.

The total energy consumption of this process was 157.7 kW (Figure 3) while the COP of heat pump was 2. It can be seen that the proposed process decreased the energy consumption by using the compression of the process stream as compared with the conventional process and that waste heat reflects to decrease the energy consumption.

Table 2. The comparison of energy consumption in the total process

	Absorber	Stripper	Overall
Heat recovery process	7.1	500.0	507.1
Proposed process	1.3	74.4	75.7 (157.7)

5. Conclusion

In this process, we proposed a new process using a chemical absorption method based on self-heat recuperation technology to reduce amount of energy consumption and we evaluated the amount of energy consumption of the process as compared with the conventional gas separation process for CO_2 by using a commercial process simulator. From the simulation results, energy consumption of the proposed process decreased to one-third at that of conventional heat recovery process.

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