Process Simulation and Energy Consumption Analysis for CO_2 Capture with Different Solvents

Boyang Xue, Yanmei Yu, and Jian Chen

1 Introduction

Carbon dioxide capture and storage (CCS) technology is considered to be the most effective technology to reduce greenhouse gas emissions in the future (Jones and Wigley 1990). The cost of CCS technology applied in fossil fuel power plant is about 40–60 \$/t CO_2 (Fei et al. 2005). And the electricity price will increase by 45% when coupled with CCS (Le Moullec and Kanniche 2011b). Among all the parts of the cost, the CO_2 capture process plays a major role (Fei et al. 2005). Thus, the most crucial problem CCS faces now is that the energy consumption of capture and separation process is tremendous. The absorption with amine solutions is the most reliable and efficient method of CO_2 capture, which is widely applied in fossil fuel power plants at present. Several studies are found in the literature that discuss the two main paths to reduce energy consumption in CO_2 capture process, developing new solvents and optimization of the process configurations (Oyenekan and Rochelle 2007; Aroonwilas and Veawab 2007; Le Moullec and Kanniche 2011a; Cousins et al. 2011).

Many kinds of amine have been studied in CO_2 capture process, such as monoethanolamine (primary amine, MEA), diethanoamine (secondary amine, DEA) (Diab et al. 2013), methyldiethanoamine (tertiary amine, MDEA) (Zhang and Chen 2010) and aminomethylpropanol (sterically hindered primary amine, AMP) (Li et al. 2013), piperazine (heterocyclic amine, PZ) (Li et al. 2014), and so on.

But at present MEA is still considered to be the main solvent in aqueous alkanolamine-based capture processes because of its high absorption rate and low

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solvent cost, as well as the fact that MEA is easy to regenerate (Aaron and Tsouris 2005).

As the heat of reaction with CO_2 in MEA is quite high (around 85 kJ/mol CO_2), it leads to a high energy requirement for stripping. DEA is also suitable for low-pressure operations and has a lower heat of reaction with CO_2 (around 70 kJ/mol CO_2). Secondary amines, like DEA, are much less reactive to sulfur components and their reaction products are not particularly corrosive. All these factors make DEA an attractive option for CO_2 capture. However, a disadvantage of DEA is that it exhibits slow kinetics (Kohl and Nielsen 1997; Carson et al. 2000; Gabrielsen et al. 2005; Galindo et al. 2012; Warudkar et al. 2013).

Le Moullec et al. (2014) reviewed 20 different process modifications, which are almost exhaustive, for numerous publications so far. However, most studies evaluate process modification for MEA solvent only and the interaction between solvent and process is ignored. Therefore, it is worth investigating the energy consumption of different amine solvents in different process.

As a result, this work proposes a comparative study on CO_2 capture process flow sheet modifications between MEA and DEA to decrease their energy consumptions. Including the conventional capture process, 10 process flow sheet modifications are evaluated at first step, which are inspired by the work of Le Moullec et al. (2014). The simulations are carried out with commercial software to calculate energy consumption for different process flow sheets for a power plant and with a CO_2 compression process. For further study and discussion, a detailed analysis is presented to study the effect of some significant parameters in capture process and the total energy consumptions of each condition are evaluated and compared.

2 Simulation Hypothesis

As many commercial simulation softwares perform well in process simulation such as Aspen Plus, ProMax, PRO/II, CO2SIM, and so on, PRO/II (version 9.0) is selected to be the simulation software in this work due to the simplicity and usability that fulfill the purpose of this study. With the amine packages in PRO/II, results obtained for MEA and DEA are accurate enough for use in final design work, because the parameters have been regressed from a large number of sources for MEA and DEA systems, resulting in good prediction of phase equilibrium. The accuracy of the simulations using PRO/II can also be validated in the following part. In the simulation, the property system uses the amine packages, which were already implemented in PRO/II, and electrolyte algorithm is calculated in both vapor and liquid phases. An equilibrium stage is assumed in absorber and stripper of all the processes. Although equilibrium models are known to give qualitatively different results from rate-based models, equilibrium models are less complex to solve than rate-based models. On the other hand, the reaction rates of amines like MEA and DEA are fast enough when large theoretical stages or packed height were implemented. Kinetics has little effect and the deviations between simulation and experiment are acceptable. Thus, the first step of approximation to study the optimization strategy in capture process is equilibrium models and proposed methodology will be then extended to rate-based models (Rodriguez et al. 2011).

2.1 Chemical Equilibrium of Amine System

The chemistry of aqueous primary and secondary amines scrubbing CO_2, like MEA and DEA, behave similarly in thermodynamics. In aqueous solutions, CO_2 reacts in an acid–base buffer mechanism with alkanolamines. The acid–base equilibrium reactions in PRO/II are written as chemical dissociations following the approach taken by Kent and Eisenberg (1976):

$$H_2O \leftrightarrow H^+ + OH^- \tag{1}$$

$$CO_2 + H_2O \leftrightarrow HCO_3^- + H^+ \tag{2}$$

$$HCO_3^- \leftrightarrow CO_3^{2-} + H^+ \tag{3}$$

$$REACOO^- + H_2O \leftrightarrow REAH + HCO_3^- \tag{4}$$

$$REAH + H^+ \leftrightarrow REAH_2^+ \tag{5}$$

where R represents an alkyl group, and here REA equals to MEA, DEA. The chemical equilibrium constants for the dissociation reactions are represented by polynomials in temperature as follows:

$$\ln K_i = A + \frac{B}{T} + \frac{C}{T^2} + \frac{D}{T^3} \tag{6}$$

2.2 Conventional CO_2 Capture Process Simulation Validation

A typical CO_2 capture process, shown in Fig. 1, mainly consists of absorber, stripper, and heat exchanger. As the figure shows, the flue gas enters into the bottom of absorber and contacts with the countercurrent lean CO_2 loading solvent flow introduced from the top of the column. After CO_2 absorption, amine solvent becomes a rich CO_2 loading flow, which then exits absorber from the bottom and is pumped to stripper to desorb CO_2. Before being injected into stripper, the cold rich solvent will be preheated by hot, lean solvent exiting from the bottom of stripper. Heated rich solvent enters into stripper to release CO_2 and then becomes lean solvent again. Pure CO_2 flow can be collected from the top of stripper for further processing and the amine solvent is cycled in two columns to capture CO_2 continuously.

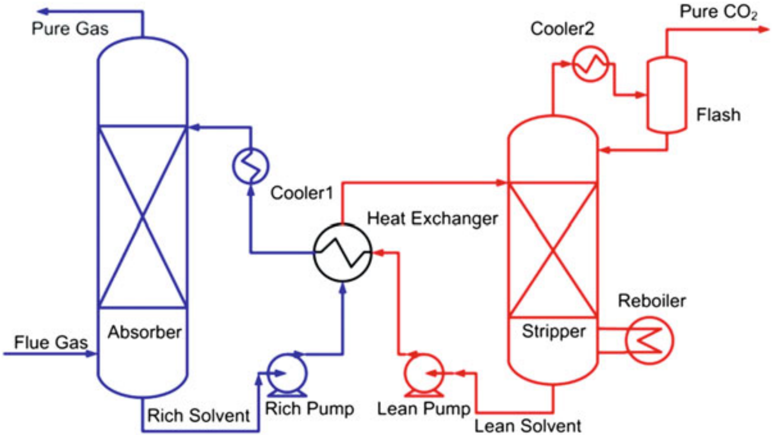

Fig. 1 Conventional CO_2 capture process

The work of Cousins et al. (2012) presented a lot of experimental data on CO_2 capture pilot plant. This work selected the pilot plant data of 1/02/2011, 24/03/2011, 31/03/2011 to validate the simulation process, and the comparison of simulation and experiment is shown in Table 1. As all the parameters were kept the same as in the literature, good agreements on rich loading and reboiler temperature are obtained. Because of the neglect of kinetics, around 10% of deviation on CO_2 capture ratio and reboiler duty is acceptable. These results can validate the accuracy of process simulation to some extent.

In process simulation of this work, the flue gas is made up of 10% CO_2, 6% H_2O, and 84% N_2 in volume, and the flue gas enters at 40 °C, 1.2 bar. As mentioned before, ideal equilibrium stages are used in simulating both absorber and stripper, and the stage number is 10, which is a feasible amount of stages proved in previous work (Warudkar et al. 2013). The operating pressure of absorber is 1 bar, stripper is 1.5 bar, and 0.1 bar pressure drops in the two column. The temperature pinch of heat exchanger is 10 °C. For reference simulation, 30 wt% MEA and 40 wt% DEA aqueous amine are used to capture 90% CO_2 of flue gas, and the CO_2 lean loading is set at 0.25 mol CO_2/mol MEA and 0.1 mol CO_2/mol DEA, which are all at their typical concentrations in various literatures and practice. All of these parameters are kept constant in the simulation of following process modifications in order to make effective comparison in energy consumption and optimization strategy. For further study and discussion in this work, parameter changes will be highlighted individually.

Table 1 Comparison of pilot plant results with the simulation results for MEA

Date	Rich loading (mol CO_2/mol MEA)	$T_{reboiler}$ (GJ/t CO_2)	CO_2 capture (%)	Q_{reb} (GJ/t CO_2)
1/02/2011				
Literature	*0.466*	*116.9*	*75.5*	*4.0*
Pro/ii	*0.477*	*116.1*	*80.1*	*3.6*
24/03/2011				
Literature	*0.486*	*117.1*	*77.7*	*4.2*
Pro/ii	*0.481*	*116.2*	*79.7*	*3.6*
31/03/2011				
Literature	*0.472*	*116.5*	*72.2*	*3.9*
Pro/ii	*0.475*	*115.8*	*75.1*	*3.4*

3 Process Modifications Description

Some studies and reviews of process modification have already been published in open literatures (Oyenekan and Rochelle 2007; Le Moullec and Kanniche 2011a; Cousins et al. 2011; Ahn et al. 2013; Le Moullec et al. 2014), which contain a variety of amine-based capture process modifications for the purpose of energy consumption reduction. Le Moullec et al. (2014) reviewed 20 different process modifications, which are almost exhaustive, for numerous publications so far. However, most studies evaluate process modification for MEA solvent only and the interaction between solvent and process needs to be considered. Therefore, it is worth investigating the energy consumption of different amine solvents in different processes. In this work, nine different process modifications are simulated using MEA and DEA solvent to study the energy consumption, and comparing them with the reference process.

3.1 Intercooled Absorber (ICA)

Intercooled absorber is a widely studied and used modification (Aroonwilas and Veawab 2007; Karimi et al. 2011). Absorption of CO_2 is an exothermic process that will lead to the temperature rise in the absorber. This has a negative effect on thermodynamic driving force for absorption and it results in lowering the solvent absorption capacity. Figure 2 illustrates that this modification is to remove a part or all of the liquid flow from the absorber at one of its stages, cooling it, and then injecting it back at the same part. Intercooled absorber is efficient in control of the temperature in the absorber column, which can increase the carrying capacity of the solvent and hence reduce the required amount of recycling solvent as well as the

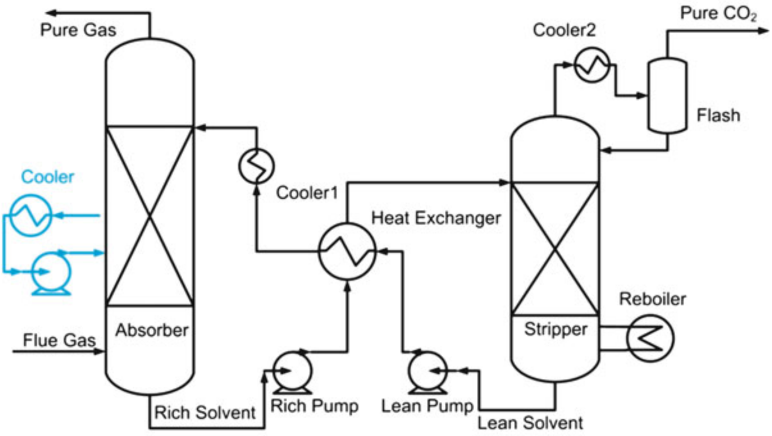

Fig. 2 Intercooled absorber (ICA)

size of equipment. In simulation work, the fifth stageis cooled to the temperature of 45 °C for MEA and DEA. As a result, the rich CO_2 loading reaches 0.492 mol CO_2/mol MEA, which is 0.465 mol CO_2/mol MEA in conventional process. For DEA, 0.468 mol CO_2/mol DEA obtained as only 0.447 mol CO_2/mol DEA in reference. It is found that the recycled lean amine solvent is reduced by 11.5% for MEA and 4.7% for DEA. Thus, 7.1% of reboiler duty is saved by MEA, and DEA gains 2.8%. ICA is more efficient for MEA than DEA because the heat of reaction with CO_2 is higher for MEA. In such favorable process in thermodynamics, MEA gains more benefits by cooling in absorber.

3.2 Flue Gas Precooling (FGP)

Flue gas precooling is a simple modification discussed in the work of Tobiesen et al. (2007) and Le Moullec and Kanniche (2011a). As Fig. 3 shows, flue gas is cooled to a lower temperature before being introduced to absorber. The principle of flue gas precooled is similar to that of the intercooled absorber to some extent, which also lowers the temperature of vapor–liquid mixture in absorber and enhances CO_2 absorption in thermodynamic aspect. Thus, higher rich loading solvent and less reboiler duty are foreseeable. Flue gas is cooled to 30 °C in our simulation, and around 5% reduction in reboiler duty is achieved with MEA, compared with a 2% saving with the DEA case.

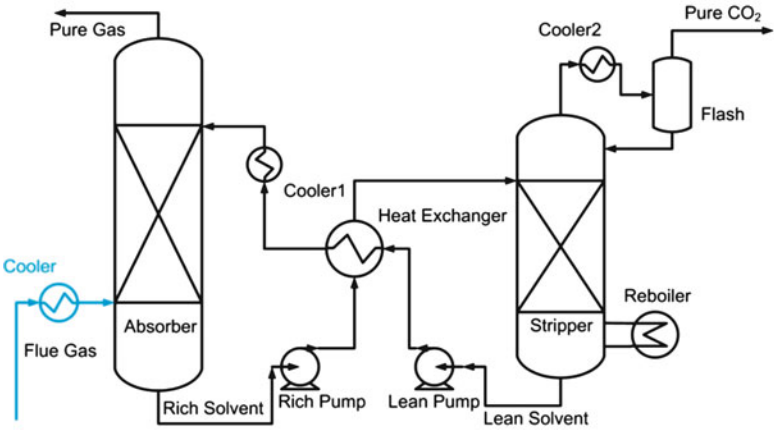

Fig. 3 Flue gas precooling (FGP)

3.3 Rich Solvent Split (RSS)

This process modification was suggested way back by Eisenberg and Johnson (1979). In Fig. 4, it splits the cold rich loaded solvent into two flows, and the split one remains unheated when it enters the top of stripper, while the other one is heated in the lean/rich heat exchanger and it is injected at lower stage. With the rich split modification, the heated rich solvent can reach a higher temperature at which CO_2 can desorb more easily. Meanwhile, the vapor released from the rich solvent meets with the cold solvent injected above, which is able to strip a little CO_2 from it. Thus, there is a reduction in reboiler duty. At first, 10% of the rich solvent unheated is split to the top of stripper in our study. There is a saving in reboiler duty of 7.7% from reference in MEA and 7% in DEA. RSS has neutral effect on rich loading and solvent required as the absorption process remains the same.

In published literatures, rich solvent split often combines with other modifications such as rich solvent preheating and split flow, which is discussed in the following. All these process modification combinations have similar principle and reduce energy consumption in the same way.

3.4 Rich Solvent Pre-heating (RSP)

As Herrin (1989) proposed, the cold rich solvent can be heated by the hot vapor exiting the stripper, as Fig. 5 shows, which can make use of the latent heat and reduce the cooling water required in stripper condenser. It seems to be efficient because the rich solvent can be heated twice. However, due to the temperature of the hot vapor is exactly similar with the rich solvent temperature after heated by hot lean solvent, even a little lower, the heat transfer cannot exist if all rich solvent is

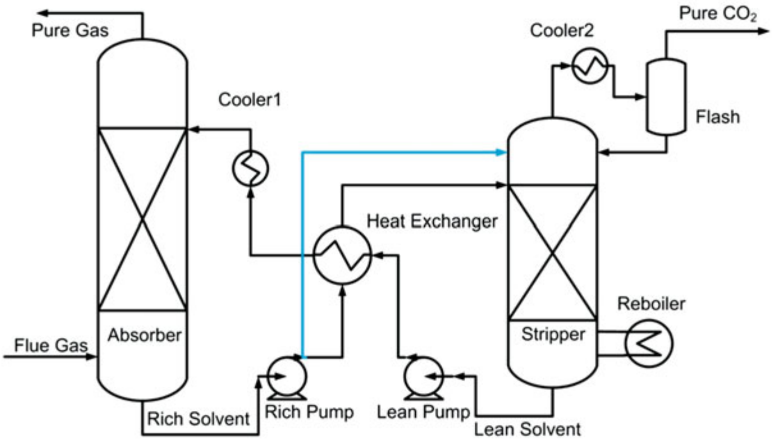

Fig. 4 Rich solvent split (RSS)

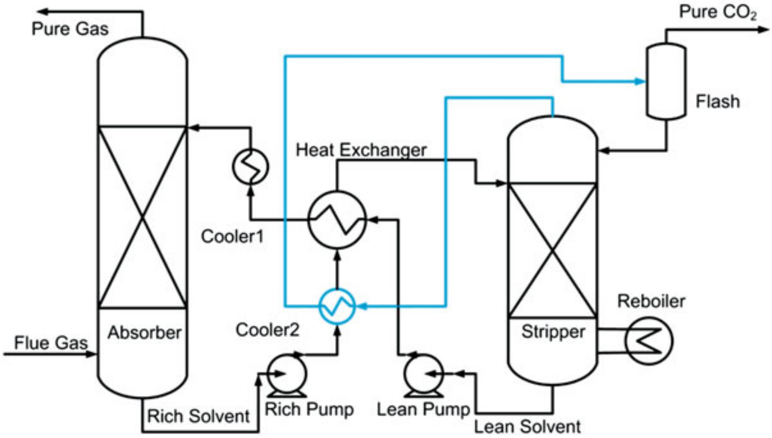

Fig. 5 Rich solvent preheating (RSP)

heated. No energy reduction is obtained in the simulation of MEA or DEA. But obvious benefits are gained if combining rich solvent preheating with rich solvent split (Ahn et al. 2013); contact with a fraction of cold rich solvent can break the heat transfer limit. Then the wasted heat can be used and other principles of energy saving are the same with rich solvent split – no more tautology or simulation here.

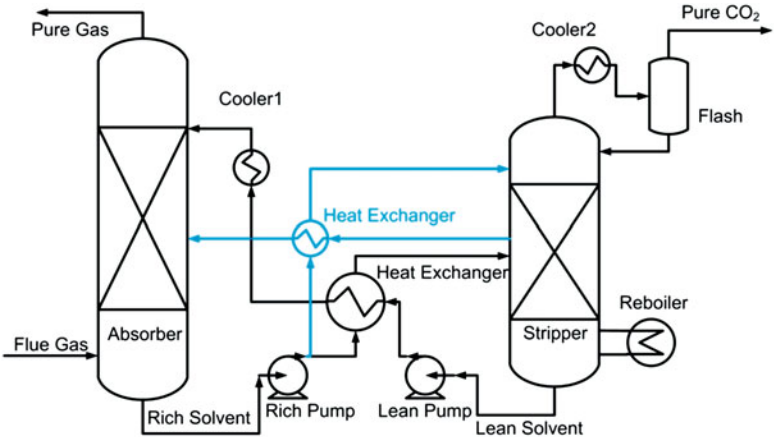

Fig. 6 Solvent split flow (SSF)

3.5 *Solvent Split Flow (SSF)*

The modification of split flow was first proposed by Shoeld (1934), consisting of a partial regeneration cycle of lean solvent. A flow of semi-lean solvent is taken from the middle of the stripper, having heat exchange with the cold rich solvent and is injected to the middle of absorber. Among all the variants of split flow modifications, the most common one is described by Leites et al. (2003) and Aroonwilas and Veawab (2007), as shown in Fig. 6. It is a combination of two modifications: simple split flow and rich solvent split. Furthermore, as the semi-lean solvent is cooled down before entering absorber, it also takes a little bit advantage of ICA. Many parameters need to be taken into account to reach a minimal energy consumption; for example, the stages to draw off semi-lean solvent from stripper and inject into absorber, the flow rate of cold rich solvent split fraction and semi-lean solvent, and the introduced stage of hot rich solvent. In principle and simulation, the semi-lean stream is drawn off from the middle of stripper to provide the cold rich solvent split with more heat. Since less rich solvent contacts with the hot lean solvent leaving stripper, hot inlet stream reaches higher temperature, and then if it is injected at lower part of stripper, energy saving is further allowed. Optimal energy savings are found in simulation when taking all these factors into account. As a result, simulation shows that SSF can lead to a 7.6% cut in reboiler duty in MEA case, correspondingly 7.8% in DEA case.

It is worth mentioning that the required amount of circulating solvent becomes larger in the solvent split flow modification than in the conventional process because the average solvent working capacity is lowered. Bigger equipments such as columns and pumps are required to match with the flow rate.

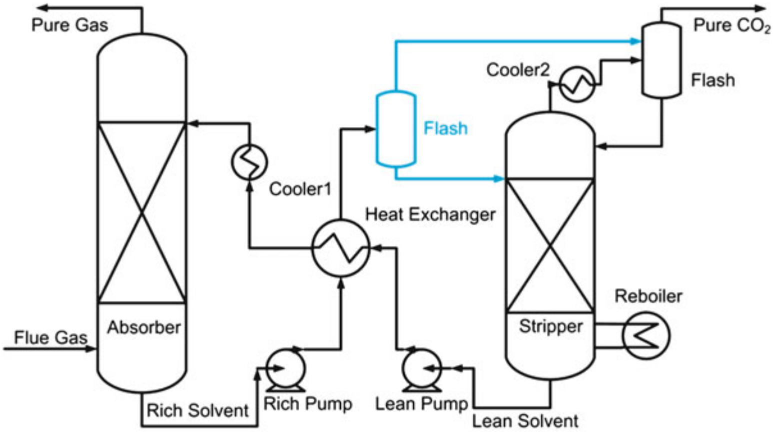

Fig. 7 Rich solvent flashing (RSF)

3.6 *Rich Solvent Flashing (RSF)*

The principle of the modification of rich solvent flashing is to flash the inlet stream of stripper before entering, as Fig. 7 illustrates. By flashing the hot rich solvent, a little more CO_2 is gained, whereas vaporization lowers the temperature of liquid phase. In fact, this flashing process is just like completing separation process once at an ideal stage, so the phenomenon in which occurs happens in the top stage in stripper. As a result, this modification does not obviously reduce energy consumption except providing one more stripping stage. Simulation result in this work is the same as what Le Moullec and Kanniche (2011a) claimed.

3.7 *Stripper Condensate Bypass (SCB)*

In the modification of stripper condensate bypass, the condensate liquid is not fed back to the top of stripper. Instead, this stream is directly injected to the absorber. This modification is used in the work of Oexmann and Kaher (2009) as Fig. 8. The simulation of this work provide a 0.6% reboiler duty saving with MEA and 0.4% with DEA, that is, stripper condensate bypass almost makes no difference in limiting energy consumption. Because of the small flow rate of condensate, the duty saving for heating it in stripper is restricted.

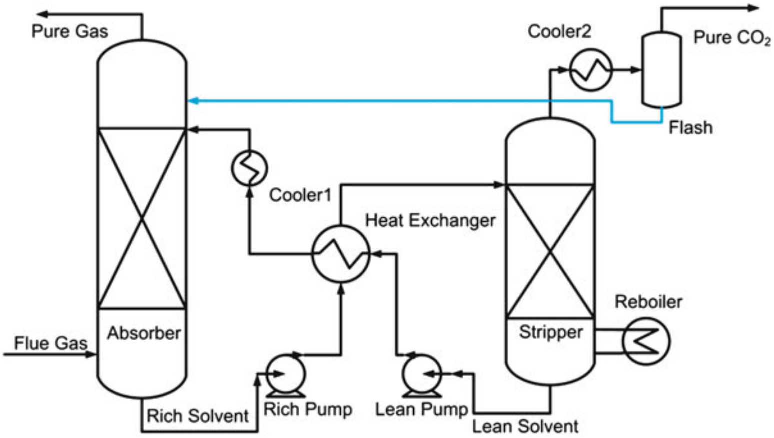

Fig. 8 Stripper condensate bypass (SCB)

3.8 *Stripper Condensate Heating (SCH)*

The modification of stripper condensate heating is proposed and studied in Aroonwilas and Veawab (2007) and Ahn et al. (2013) as Fig. 9. As vapor temperature in the top of the stripper is high, stripper condensate heating is to make use of this to heat the stripper condensate, and then feeding the hot condensate back to the bottom of stripper to provide a little heat recovery. Nevertheless, it has been proved by theoretical analysis and simulation in this work that there is insignificant gain in energy consumption. Only 1% of reboiler duty is reduced both for MEA and DEA.

3.9 *Lean Vapor Compression (LVC)*

Lean vapor compression is one of the most widely suggested modifications in a variety of literatures and patents, such as Batteux and Godard (1983), Reddy et al. (2007), and Woodhouse and Rushfeldt (2008). As Fig. 10 shows, the principle is to flash the hot lean solvent at a lower pressure, then compress the hot vapor generated and reinject it into the bottom of stripper. As the vapor benefits from the sensible heat of hot lean solvent as well as recompression, it can reach a very high pressure and temperature, which can provide additional steam and heat in the column for stripping. In the simulation, the hot lean solvent is flashed to the atmospheric pressure and this modification shows significant savings in reboiler duty. With MEA, a 12.8% of reduction is obtained, and as for DEA, LVC allows a gain of 11.9% of reduction in reboiler duty. However, it should be noted that as a compressor is introduced here, it leads to the additional electricity consumption that cannot be neglected. The adiabatic efficiency of the compressor is 80% in

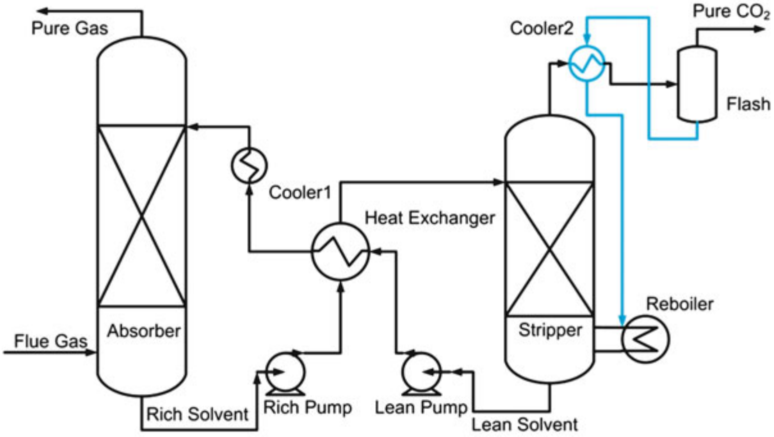

Fig. 9 Stripper condensate heating (SCH)

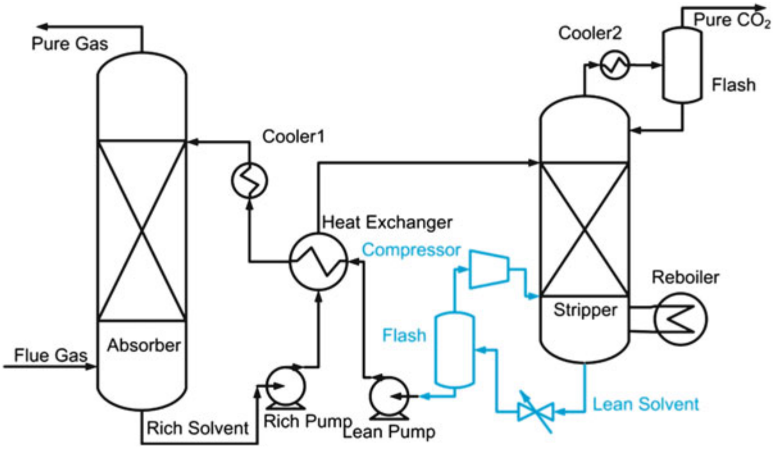

Fig. 10 Lean vapor compression (LVC)

simulation, and the performance of total energy saving compared with the conventional process will be discussed in detain in the following.

4 Results and Discussions

Preliminary simulation results have been presented in the previous part; process modifications description, and detailed simulation results and further discussion will be demonstrated in the following paragraphs, including process operating parameter adjustments, and total energy consumption is calculated for comparison.

4.1 Total Work Calculation

As mentioned before, the process modification of LVC introduces a compressor to generate vapor with high pressure and temperature, and the electricity consumption should not be neglected. Therefore, it is essential to investigate the total energy consumption to make a global comparison with the conventional process.

The equivalent work (W_{eq}) is commonly used to evaluate the process configuration performance to unify the thermal and electrical energy consumptions. As there are a variety of expressions in calculating the total equivalent work, such as Le Moullec and Kanniche (2011a), Ahn et al. (2013), and Van Wagener et al. (2013), we finally calculate the total equivalent work for this work by the following equation from Van Wagener and Rochelle (2011) and Liang et al. (2015):

$$W_{eq} = 0.75 \times Q_{reb}\left(\frac{T_i + 10\text{K} - T_{sink}}{T_i + 10\text{K}}\right) + W_{comp} + W_{add} \tag{7}$$

It uses a Carnot efficiency term that accounts for the increasing value of steam at high temperature. Additionally, 75% efficiency is applied to account for nonideal expansion in the steam turbines. T_i is the reboiler temperature (K); 10 K means the temperature of steam in the reboiler is 10 K higher than T_i; Q_{reb} is the reboiler duty (GJ/t CO_2); T_{sink} is the cold end temperature of Carnot engine, and set at 313 K here; W_{comp} is the compression work (GJ/t CO_2); W_{add} is the additional equipment work such as the compressor in LVC (GJ/t CO_2).

As for calculating the compression work, the simple following correlation can be used:

$$W_{comp} = 8.3673 + 22.216 \ln P_F - (27.118 + 0.0256\text{P}_F) \ln P_S \tag{8}$$

where W_{comp} is the compression work (kWh/t CO_2); P_F is the final delivery pressure, and P_F is set here as 110 bar; P_S is the initial pressure of compression, which equals to the stripper pressure.

The total equivalent work of each of the process modifications described previously is shown in Table 2 for MEA and Table 3 for DEA. All the process modifications apart from RSF and RSP exhibit lower energy consumption for MEA. As for DEA, only RSF has negative effect.

4.2 Effect of Amine Concentration and Lean Solvent Loading

The loading of the lean amine solution is a significant factor in reducing the energy consumption. More solvent is required to be circulated when the lean loading is high in order to capture the same amount of CO_2. The reboiler heat duty is rather sensitive to the solvent flow rate as the vaporization of water for CO_2 stripping

Table 2 Total equivalent work of process using MEA

Modifications	Rich Loading (mol CO_2/mol MEA)	Q_{reb} (GJ/t CO_2)	W_{add} (GJ/t CO_2)	W_{eq} (GJ/t CO_2)	Total Energy Savings (%)
Conventional	*0.465*	*3.460*	*0*	*0.911*	
ICA	*0.492*	*3.216*	*0*	*0.873*	*4.20*
FGP	*0.485*	*3.278*	*0*	*0.883*	*3.17*
RSS	*0.465*	*3.192*	*0*	*0.869*	*4.61*
RSP	*0.465*	*3.461*	*0*	*0.912*	*−0.02*
SSF	*0.463*	*3.196*	*0*	*0.870*	*4.54*
RSF	*0.465*	*3.634*	*0*	*0.939*	*−3.03*
SCB	*0.465*	*3.432*	*0*	*0.908*	*0.38*
SCH	*0.465*	*3.411*	*0*	*0.904*	*0.80*
LVC	*0.465*	*3.018*	*0.039*	*0.876*	*3.87*

Table 3 Total equivalent work of process using MEA

Modifications	Rich Loading (mol CO_2/mol DEA)	Q_{reb} (GJ/t CO_2)	W_{add} (GJ/t CO_2)	W_{eq} (GJ/t CO_2)	Total Energy Savings (%)
Conventional	*0.447*	*3.168*	*0*	*0.856*	
ICA	*0.468*	*3.078*	*0*	*0.842*	*1.64*
FGP	*0.468*	*3.080*	*0*	*0.842*	*1.60*
RSS	*0.447*	*2.945*	*0*	*0.821*	*4.06*
RSP	*0.447*	*3.153*	*0*	*0.854*	*0.27*
SSF	*0.440*	*2.921*	*0*	*0.818*	*4.50*
RSF	*0.447*	*3.302*	*0*	*0.877*	*−2.44*
SCB	*0.447*	*3.131*	*0*	*0.850*	*0.73*
SCH	*0.447*	*3.136*	*0*	*0.851*	*0.58*
LVC	*0.447*	*2.791*	*0.0368*	*0.833*	*2.70*

contributes most to the reboiler duty at low solvent flow rate values. If lean loading is extremely low, more heat is provided by the reboiler duty as the heat of reaction between amines and CO_2 accounts for the majority. As for amine concentration, it will affect the solvent capture capacity because low rich loading will be obtained if a more concentrated solution is used. And the proportion of water increases when diluted solution is implemented. These will all lead to a further reduction of reboiler duty. It can be observed in Figs. 11 and 12 that the optimal lean loading increases with MEA concentration rising. The minimum of reboiler duty occurs at approximately 0.17 mol CO_2/mol MEA in 30 wt% MEA. When DEA was used, it was noticed that irrespective of the concentration used, the optimal lean loading is obtained around 0.05 mol CO_2/mol DEA. It also can be concluded that the reboiler duty is more sensitive to lean loading in process using MEA. And these curves reveal furthermore that at higher amine concentrations, the flexibility of process increases because change in the lean loading will have a minor effect. Galindo et al. (2012) and Dinca (2013) also claimed the same point of view.

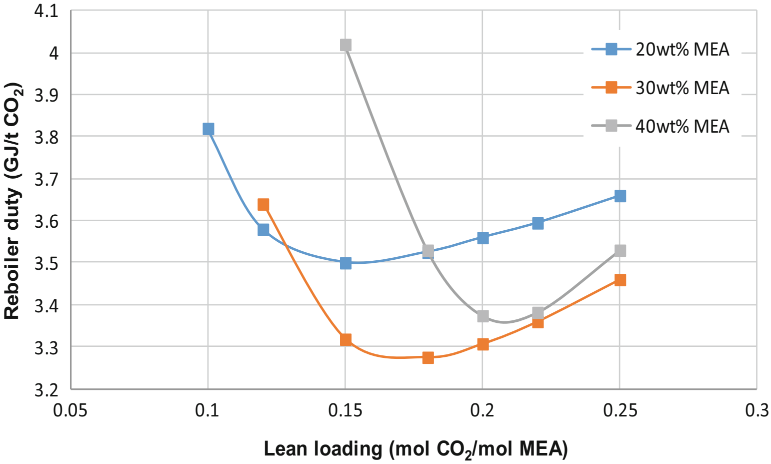

Fig. 11 Effect of MEA concentration and CO_2 loading of lean solvent on reboiler duty

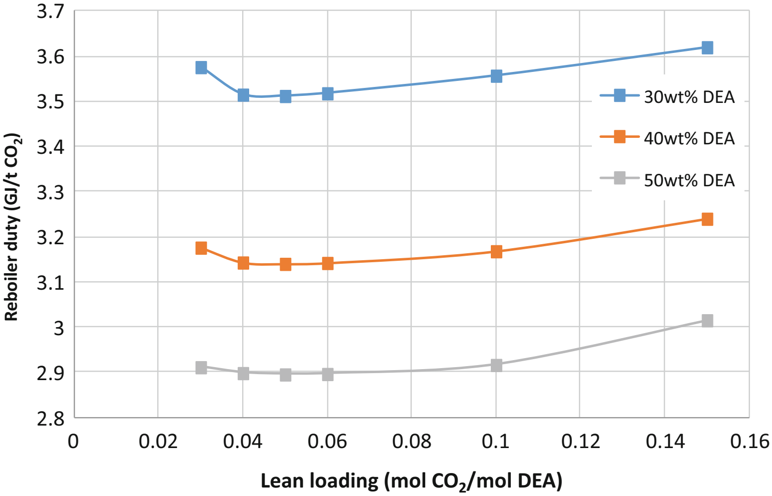

Fig. 12 Effect of DEA concentration and CO_2 loading of lean solvent on reboiler duty

4.3 Effect of CO_2 Concentration in the Flue Gas

The CO_2 content in the flue gas of typical coal-fired power plants lies in the range of 12–15 vol% (wet), and in natural gas combined cycle power plant, the CO_2 concentration will drop to below 5 vol%. It will be of value to explore how CO_2 concentration affects the energy consumption in two different amines. Figure 13

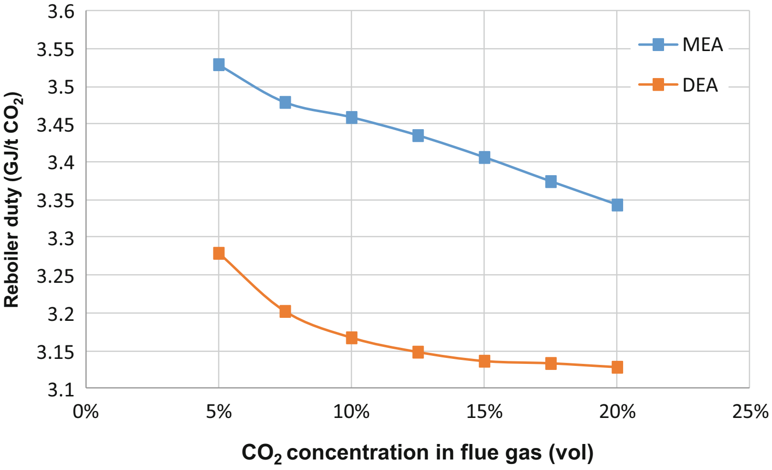

Fig. 13 Effect of CO_2 concentration in the flue gas on reboiler duty

illustrates the simulation results in conventional process when varying CO_2 concentration from 5 to 20 vol%. As expected, the reboiler duty decreases with CO_2 concentration rising up correspondingly both for MEA and DEA. A noticeable change in reboiler duty is observed when using MEA. In contrast, there is no significant difference for DEA when CO_2 is more than 10 vol%. These results are caused by the difference from the heat of reaction.

4.4 *Effect of Stripper Pressure*

It is a common view that the operating pressure of the stripper is a key parameter of reboiler duty reduction, which has been reported in many publications such as Oyenekan and Rochelle (2007). There is also a process modification proposed by Oyenekan and Rochelle (2006) and Le Moullec and Kanniche (2011a), which is to operate the stripper at vacuum/subambient pressure. CO_2 desorption becomes easier as stripper pressure is high. From another perspective, if stripper pressure is lower, lower pressure steam is required for solvent regeneration because the reboiler temperature goes down. Therefore, the influence of stripper pressure should be evaluated in total equivalent work to search for the optimal strategy. Figure 14 illustrates the results of simulation for conventional process, and it indicates that both for MEA and DEA, higher pressure is beneficial to reducing total energy consumption.

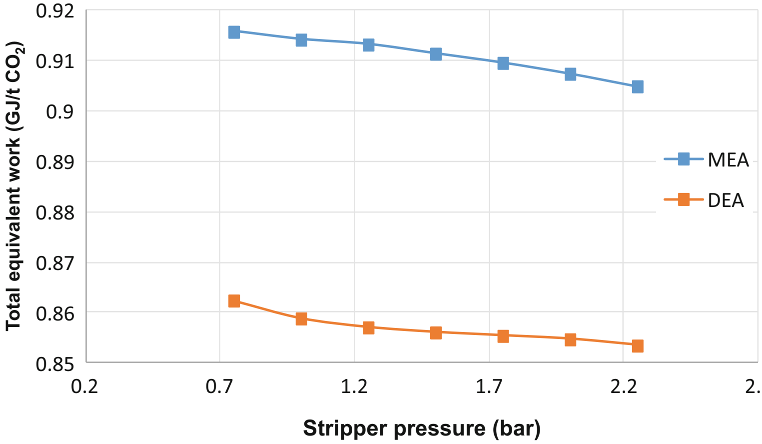

Fig. 14 Effect of stripper pressure on total equivalent work in the conventional process

4.5 *Effect of Lean Solvent Loading for Process Modifications*

As the principle and simulation results mentioned previously, the loading of the lean amine solution is of great significance in the reducing of energy consumption. And the simulation results indicated that the reboiler heat duty is rather sensitive to the lean loading. Thus, it is essential to simulate all the processes to come up with the optimal energy saving strategies. The process modification of ICA, RSS, SCH, LVC, SSF are selected to make comparison with the conventional process according to previous simulation results and discussion, as these configurations present better performance in terms of reducing energy consumption. The result shows in Fig. 15 for MEA and Fig. 16 for DEA. As for MEA, all the total equivalent work of these processes has a minimum point as the lean loading is increasing. In conventional process, ICA, and SCH, the minimums occur at approximately 0.18 mol CO_2/mol MEA, and it rises to 0.22 mol CO_2/mol MEA for RSS and SSF. In contrast, minimum of LVC appears at around 0.16 mol CO_2/mol MEA because the heat provided by compressed vapor is quite effective. At lower lean CO_2 loading, the total equivalent work of RSS, SCH, and SSF is higher than conventional process due to a larger amount of circulating solution. As a whole, the total equivalent works of these configurations for MEA are in the following order: LVC $<$ ICA $<$ RSS $<$ SSF $<$ SCH $<$ conventional process.

On the other hand, the results of processes using DEA appear somewhat different from MEA. The trends of conventional process, SCH, LVC are quite the same, as all of them have a minimum point at the lean CO_2 loading of about 0.15 mol CO_2/mol DEA. ICA raises this point to 0.2 mol CO_2/mol DEA while SSF lowers it to 0.1 mol CO_2/mol DEA. The total equivalent work of RSS has an obvious change as the minimum point occurs at 0.15 mol CO_2/mol DEA. The energy consumption of

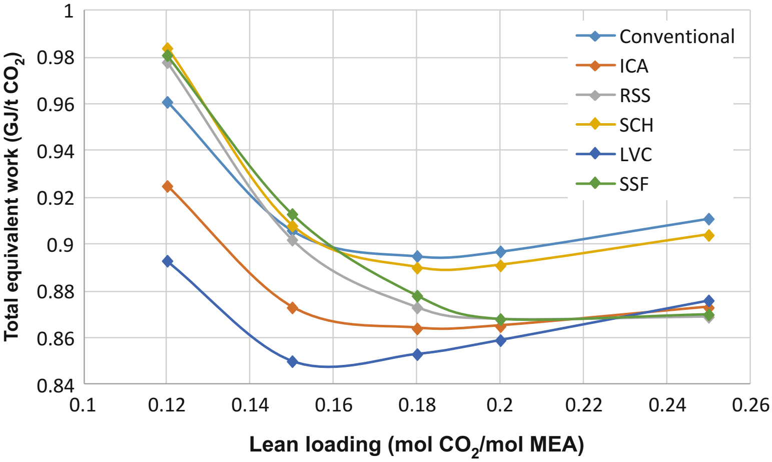

Fig. 15 Compare total equivalent work between different process configurations at different lean loading of MEA

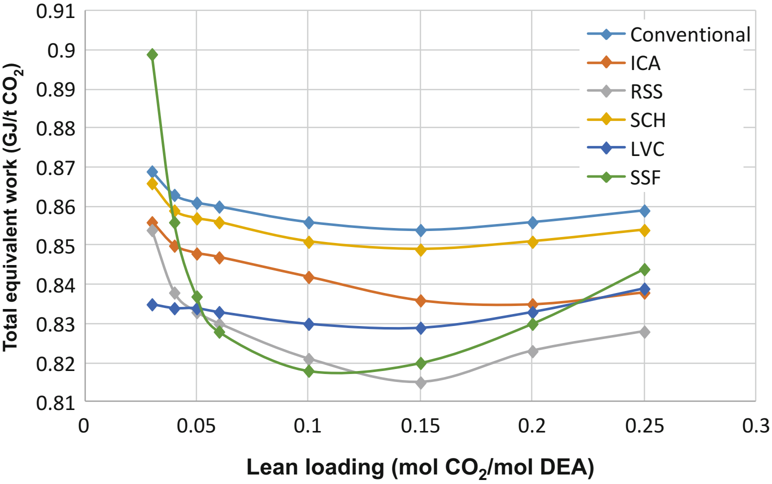

Fig. 16 Compare total equivalent work between different process configurations at different lean loading of DEA

SSF is higher than the conventional process due to lower working capacity of amine and a larger amount of circulating solution. In general, the total equivalent works of these configurations for DEA are in the following order: RSS ~ SSF < LVC < ICA < SCH < conventional process. Compared with MEA, it can be concluded that RSS or the variant of RSS is more efficient than DEA, because CO_2 is easier released from DEA solution in thermodynamics. And ICA is more favorable to MEA, which is also the reason from the difference of absorption heat. LVC gains benefits both for MEA and DEA.

5 Conclusions

In this work, including the conventional process, ten different process configurations have been simulated both for MEA and DEA on the same operating conditions, such as flue gas composition, CO_2 loading of lean amine solution, amine concentration, CO_2 capture ratio, and so on. To make a more valuable and comprehensive evaluation on energy consumption reduction, the performance is presented in terms of the total equivalent work as well as reboiler duty. It is worth mentioning that all simulations are restrained to maintain the temperature of amine solution below 120 °C, to avoid the degradation of MEA and DEA. As a result, process modifications are proved to be an efficient way to optimize the energy consumption in CO_2 capture process using MEA and DEA. It has been shown that ICA, RSS, SSF, LVC are favorable in MEA, as from 3.87% to 4.61% of total equivalent work is reduced, respectively, in preliminary simulation; and for DEA, RSS, SSF, LVC have better performance, as from 2.70% to 4.50% is reduced.

Meanwhile, this work presents the influence of four operating parameters in energy savings, namely, amine concentration, loading of lean amine solvent, CO_2 concentration in the flue gas, and stripper pressure. All these factors vary in both conventional process and process modifications with MEA and DEA. The study of amine concentration and lean loading shows that the optimal lean loading increases with MEA concentration rising, but it basically keeps constant in DEA. Moreover, reboiler duty is more sensitive to lean loading in process using MEA than DEA. When changing the CO_2 concentration in the flue gas, a more significant change in reboiler duty is observed when using MEA and less for DEA when CO_2 is more than 10 vol%. The effect of lean solvent loading on process modifications for MEA and DEA is quite different, and the minimum point of total equivalent work also depends on amine type and process. But, for both MEA and DEA, higher pressure is beneficial to reducing total energy consumption in all of the processes. Approximately, 10% reduction can be obtained in process modifications using MEA and 8% in DEA. LVC has the best performance when implement higher stripper pressure is used, and 9% of reboiler duty reduction is obtained in MEA and 8% in DEA.

The comparative study and evaluation of process modifications between MEA and DEA are proposed in this work, which present the influence of the interaction between solvent and process. It is essential in post-combustion process design to make optimization strategy. Further work will be continued with different solvents, such as MDEA, AMP, PZ, or amine blends in different processes.

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