

# A comparative study of MEA and DEA for post-combustion CO<sub>2</sub> capture with different process configurations

Boyang Xue<sup>1</sup> · Yanmei Yu<sup>1</sup> · Jian Chen<sup>1</sup>  · Xiaobo Luo<sup>2</sup> · Meihong Wang<sup>2</sup>

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**Abstract** This paper presented a comparative study of monoethanolamine (MEA) and diethanolamine (DEA) for post-combustion CO<sub>2</sub> capture (PCC) process with different process configurations to study the interaction effect between solvent and process. The steady state process model of the conventional MEA-based PCC process was developed in Pro/II<sup>®</sup> and was validated with the experimental data. Then ten different process configurations were simulated for both MEA and DEA. Their performances in energy consumption were compared in terms of reboiler duty and total equivalent work. The results show that DEA generally has better thermal performances than MEA for all these ten process configurations. Seven process configurations provide 0.38%–4.61% total energy saving compared with the conventional PCC process for MEA, and other two configurations are not favourable. For DEA, except one configuration, other process configurations have 0.27%–4.50% total energy saving. This work also analyzed the sensitivities of three key parameters (amine concentration, stripper pressure and lean solvent loading) in conventional process and five process modifications to show optimization strategy.

**Keywords** Process simulation · CO<sub>2</sub> capture · Energy consumption · MEA · DEA

## 1 Introduction

Carbon capture and storage (CCS) technology is considered to be the most effective technology to reduce greenhouse gas emissions in the future. The cost of CCS technology applied in fossil fuel power plant is about 40 to 60 \$/t CO<sub>2</sub> (Barry 2001), and the electricity price will increase by 45% when coupled with CCS (Le Moullec and Kanniche 2011b). Obviously, the application of CCS technology is limited by high energy consumption and high cost of capture process. Post-combustion CO<sub>2</sub> capture (PCC) technology is widely studied now because it can be

applied to most fossil fuel power plants, and among all kinds of PCC technology, chemical absorption with amine solutions is the most reliable and efficient method of CO<sub>2</sub> capture. Many studies are found in the literature that discuss the two main paths to reduce energy consumption in CO<sub>2</sub> capture process, developing new solvents and optimization of the process configurations (Oyeneke 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<sub>2</sub> 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. 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 solvent cost as

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✉ Jian Chen  
cj-dce@tsinghua.edu.cn

<sup>1</sup> State Key Laboratory of Chemical Engineering, Tsinghua University, Beijing 100084, China

<sup>2</sup> Process and Energy Systems Engineering Group, School of Engineering, University of Hull, Hull HU6 7RX, UK

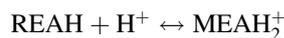
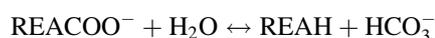
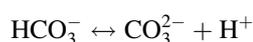
well as low regeneration heat requirement (Aaron and Tsouris 2005). Nevertheless, high energy requirement for stripping still exists due to the high heat of reaction with CO<sub>2</sub> using MEA (around 85 kJ/mol CO<sub>2</sub>). Consequently, DEA could be taken into account for low pressure operations and it has a lower heat of reaction with CO<sub>2</sub> (around 70 kJ/mol CO<sub>2</sub>). Secondary amines, like DEA, are much less reactive to sulphur components and their reaction products are not particularly corrosive. These all make DEA to be an attractive option for CO<sub>2</sub> 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).

However, most studies only evaluate process configurations for MEA solvent and the interaction between solvent and process is ignored (Le Moullec et al. 2014). 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<sub>2</sub> capture process configurations between MEA and DEA, which is aiming to compare the thermal performances between MEA and DEA, as well as investigating the energy saving by different process configurations using MEA and DEA. Including the conventional capture process, ten process configurations are evaluated, 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 configurations with a CO<sub>2</sub> compression process. For the 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 Model development and validation

### 2.1 Chemical equilibrium of amine system

The chemistry of aqueous primary and secondary amines scrubbing CO<sub>2</sub>, like MEA and DEA, behave similarly in thermodynamics. In aqueous solutions, CO<sub>2</sub> react 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):



where R represents an alkyl group, and REA equals to MEA, DEA here.

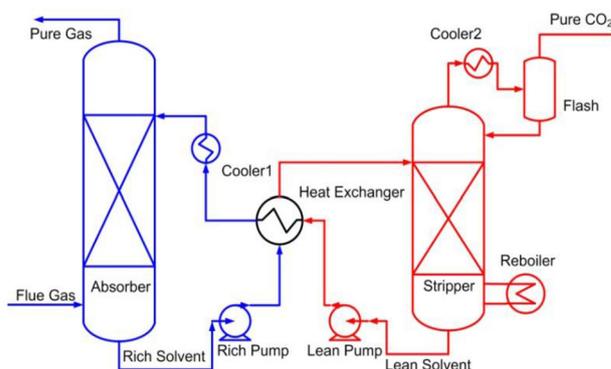
The chemical equilibrium constants for the dissociation reactions are represented by polynomials in temperature as follow:

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

### 2.2 Process model development in PRO/II

As many commercial simulation softwares perform well in process simulation such as Aspen Plus, ProMax, PRO/II, CO<sub>2</sub>SIM and so on, PRO/II (version 9.0) is selected to be the simulation software in this work due to the simplicity and usability which fulfil the purpose of this study. In the simulation, the property system uses the amine packages in PRO/II and electrolyte algorithm is for calculation. An equilibrium stage is assumed in absorber and stripper of all the processes. With the amine packages already implemented in PRO/II, results obtained for MEA and DEA are valuable as the parameters have been regressed from a large number of sources for MEA and DEA systems and resulting in good prediction of phase equilibrium. The accuracy of the simulations using PRO/II can also be validated in the following part. Although equilibrium models are known to give quantitatively different results from rate-based models, the first step of approximation is equilibrium models and proposed methodology will be then extended to rate-based models (Rodríguez et al. 2011).

A typical CO<sub>2</sub> capture process shows in Fig. 1, which is mainly consist of absorber, stripper and heat exchanger. The flue gas enters into the bottom of absorber and contacts with the counter-current lean CO<sub>2</sub> loading solvent flow introduced from the top of the column. After CO<sub>2</sub> absorption, the amine solvent becomes a rich CO<sub>2</sub> loading flow, which exits absorber from the bottom and is pumped to stripper. Before being injected, the cold rich solvent will be pre-heated by hot lean solvent exiting from the bottom



**Fig. 1** Conventional CO<sub>2</sub> capture process

of stripper. Heated rich solvent enters into stripper to release CO<sub>2</sub> and then becomes lean solvent again. Pure CO<sub>2</sub> flow can be collected from the top of stripper for further processing and the amine solvent is cycled in two columns to capture CO<sub>2</sub> continuously.

In process simulation of this work, the flue gas is made up of 10% CO<sub>2</sub>, 6% H<sub>2</sub>O, 84% N<sub>2</sub> 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 and stripper is 1 bar and 1.5 bar, respectively. A 0.1 bar pressure drop is set in two columns. 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<sub>2</sub> of flue gas, and the CO<sub>2</sub> lean loading is set at 0.25 mol CO<sub>2</sub>/mol MEA and 0.1 mol CO<sub>2</sub>/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 configurations in order to make effective comparison in energy consumption. For further study and discussion in this work, parameter changes will be highlighted individually.

### 2.3 Model validation

The work of Cousins et al. (2012) presented a lot of experimental data on CO<sub>2</sub> capture pilot plant. This work selected the pilot plant data of 1/02/2011, 11/02/2011, 31/03/2011 to validate the process model, and the comparison of simulation and experiment is shown in Table 1. As all the parameters were kept the same with literature, a good agreement between simulation and experiment is obtained. Because of the neglect of kinetics, deviations around 5% on rich loading and CO<sub>2</sub> capture ratio are acceptable. These results can validate the accuracy of process simulation.

## 3 Process simulation results

Many studies and reviews of process modification have already been published in open literatures (Oyekan and Rochelle 2007; Le Moulec and Kanniche 2011a; Cousins et al. 2011; Ahn et al. 2013; Le Moulec et al. 2014), which contain a variety of amine based capture process modifications for the purpose of energy consumption reduction. However, most studies only evaluate process configurations for MEA solvent and the interaction between solvent and process is ignored (Le Moulec et al. 2014). Therefore, it is worth investigating the energy consumption of different amine solvents in different processes. In this work, ten different process configurations are simulated using MEA and DEA solvent to compare their thermal performances and investigate the energy saving by different process configurations. To make a more valuable and comprehensive evaluation on energy consumption reduction, the performance is presented in terms of reboiler duty as well as the total equivalent work. The reboiler duty is calculated in process simulation with PROII based on given process parameters. And the calculation method for the total equivalent work will be described in the following Sect. 4.1. It is worth mentioning that all simulations are restrained to maintain the temperature of amine solution below 120 °C, which is to avoid the degradation of MEA and DEA. The lean solution temperature is 40 °C.

### 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<sub>2</sub> is an exothermic process which will lead to the temperature rise in the absorber. This has a negative effect on thermodynamic driving force for absorption and it results to lower 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

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

Date	Rich loading (mol CO <sub>2</sub> /mol MEA)	$T_{\text{reb}}$ (°C)	CO <sub>2</sub> capture (%)	$Q_{\text{reb}}$ (GJ/t CO <sub>2</sub> )
1/02/2011				
Literature	0.466	116.9	75.5	4.1
Simulation	0.478	116.1	79.9	4.0
11/02/2011				
Literature	0.471	117.6	74.4	4.0
Simulation	0.481	116.3	78.0	3.9
31/03/2011				
Literature	0.472	116.5	72.2	4.2
Simulation	0.477	115.2	70.1	4.0

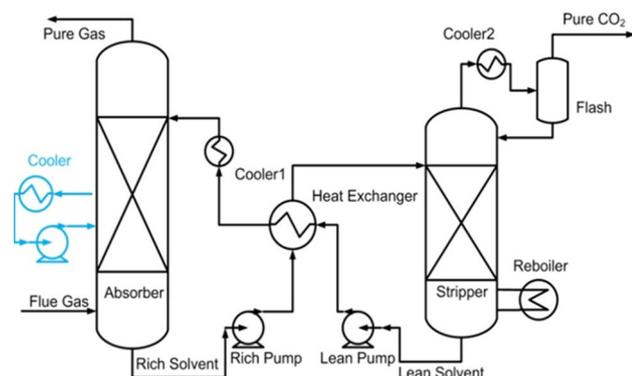
one of its stages, cooling it and then injecting it back at the same part. Intercooled absorber is efficient in control of the temperature in the absorber column which can increase the carrying capacity of the solvent and hence reduce the required amount of recycling solvent as well as the size of equipment. In simulation work, 5th stage temperature is cooled down to 45 °C for MEA and DEA. As a result, the rich CO<sub>2</sub> loading reaches 0.492 mol CO<sub>2</sub>/mol MEA, which is 0.465 mol CO<sub>2</sub>/mol MEA in conventional process. For DEA, 0.468 mol CO<sub>2</sub>/mol DEA is obtained as only 0.447 mol CO<sub>2</sub>/mol DEA before. It is found that the cycled 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 due to the heat of reaction with CO<sub>2</sub> is higher for MEA. In such favourable process in thermodynamics, MEA gains more benefits by cooling in absorber.

### 3.2 Flue gas pre-cooling (FGP)

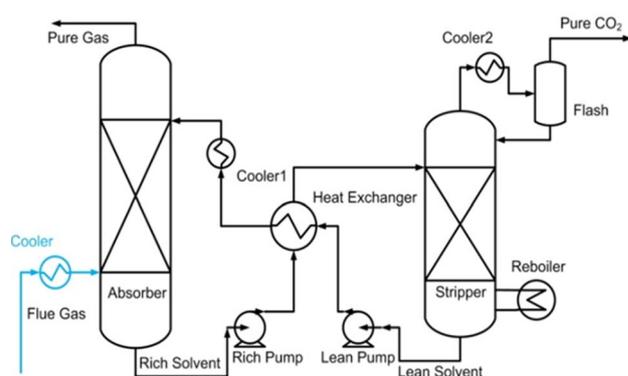
Flue gas pre-cooling is a simple modification discussed in the work of Tobiesen et al. (2007) and Le Moullec and Kanniche (2011a). As Fig. 3 shows, flue gas is cooled to a lower temperature before introduced to absorber. The principle of flue gas pre-cooled is similar with the intercooled absorber to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

### 3.3 Rich solvent split (RSS)

This process modification is suggested as long ago as Eisenberg and Johnson (1979). In Fig. 4, it splits the cold rich loaded solvent into two flows, and the split one



**Fig. 2** Intercooled absorber (ICA)

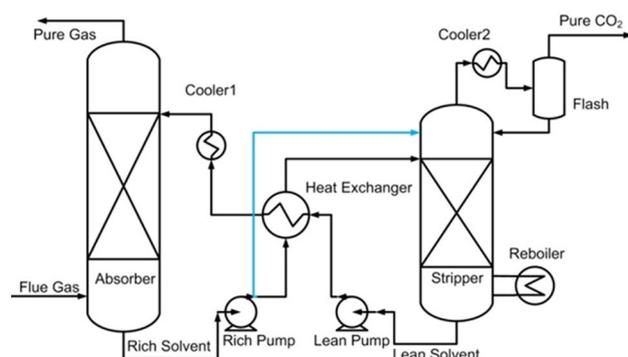


**Fig. 3** Flue gas pre-cooling (FGP)

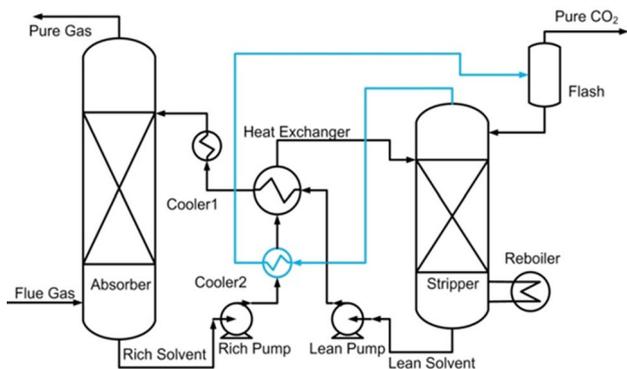
remains unheated when it enters the top of the stripper, while the other one is heated in the lean/rich heat exchanger and it is injected at lower stage. With the rich split modification, the heated rich solvent can reach a higher temperature at which CO<sub>2</sub> can desorb more easily. Meanwhile, the vapour released from the rich solvent meets with the cold solvent injected above, which is able to strip a little CO<sub>2</sub> from it. Thus, there is a reduction in reboiler duty. 10% of the rich solvent unheated is split to the top of stripper in simulation. There is a saving in reboiler duty of 7.7% by MEA and 7% by DEA. RSS has neutral effect on rich loading and solvent required as the absorption process remains the same. Different split rates larger than 10% are necessary to be used for further simulation.

### 3.4 Rich solvent pre-heating (RSP)

As Herrin (1989) proposed, the cold rich solvent can be heated by the hot vapour 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 vapour is exactly similar with the rich solvent temperature after



**Fig. 4** Rich solvent split (RSS)

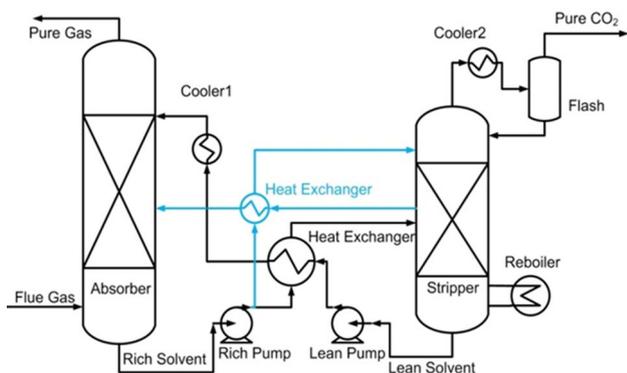


**Fig. 5** Rich solvent pre-heating (RSP)

heated by hot lean solvent, even a little lower, the heat transfer cannot exist if all rich solvent is heated. No energy reduction obtained in the simulation of MEA or DEA. But obvious benefits will be gained if combining rich solvent pre-heating with rich solvent split because heat can be transferred to the split cold rich solvent (Ahn et al. 2013). 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.

### 3.5 Solvent split flow (SSF)

The modification of split flow is first proposed by Shoeld (1934), which contains 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 then injected to the middle of absorber. Among all the variant of split flow modifications, the most common one is described by Leites et al. (2003) and Aroonwilas and Veawab (2007) as 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 inter-cooling. Many parameters needs to be taken into account to reach a minimal energy



**Fig. 6** Solvent split flow (SSF)

consumption, for example, the stages to draw off semi-lean solvent from stripper and inject into absorber, the split fraction of cold rich solvent and semi-lean solvent, and the introduced stage of hot rich solvent. In principle and simulation, the semi-lean stream is drawing 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, 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% of reduction on reboiler duty in MEA case, correspondingly with 7.8% in DEA.

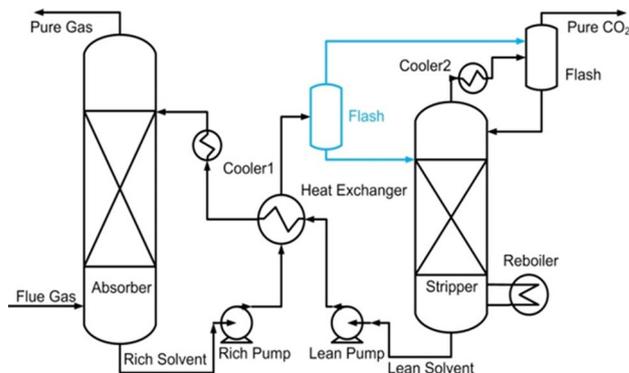
It is worth mentioning that the required amount of circulating solvent becomes larger in the solvent split flow modification than in the conventional process because the average solvent working capacity is lowered. Bigger equipments such as columns and pumps are required to match with the flow rate.

### 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<sub>2</sub> is gained whereas vaporization lowers the temperature of liquid phase. In fact, this flashing process is same with conducting separation process at an ideal stage. The phenomenon happens everywhere in stripper. As a result, this modification does not obviously reduce energy consumption except providing one more stripping stage. The simulation result in this work is the same with 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.



**Fig. 7** Rich solvent flashing (RSF)

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.

### 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 the vapour temperature is high at the top of stripper, 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 vapour compression (LVC)

Lean vapour 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), Woodhouse and Rushfeldt (2008). As Fig. 10 shows, the principle is to flash the hot lean solvent at a lower pressure, then compressing the hot vapour generated and re-injecting it into the bottom of stripper. As the vapour 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 the adiabatic efficiency of the compressor is 80%, and this modification shows significant savings in reboiler duty. With MEA, a 12.8% of reduction is obtained, and as for DEA, LVC

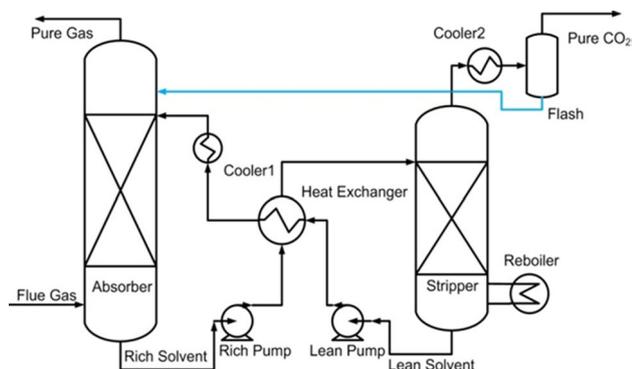


Fig. 8 Stripper condensate bypass (SCB)

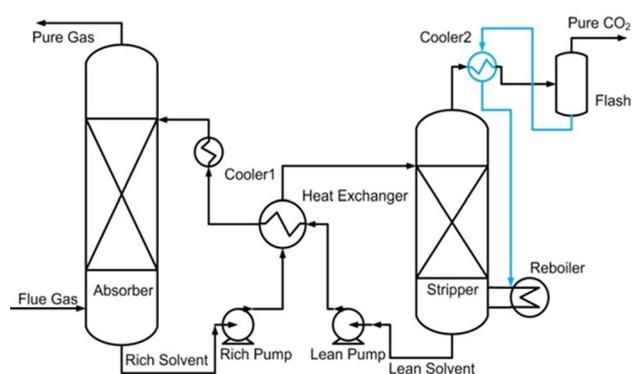


Fig. 9 Stripper condensate heating (SCH)

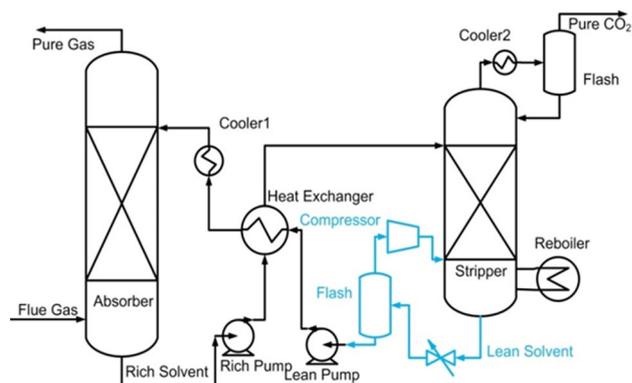


Fig. 10 Lean vapour compression (LVC)

allows a gain of 11.9% of reduction in reboiler duty. However, it should be noted that as a compressor introduced here, it leads to the additional electricity consumption that cannot be neglected, and the performance of total energy saving compared with the conventional process will be detailed discussed in the following part.

## 4 Discussions

Preliminary simulation results have been presented in previous part, and detailed simulation results and further discussion will be demonstrated in following paragraphs, including total energy consumption calculated for comparison and sensitivity analysis of three process operating key parameters.

### 4.1 Total work calculation

As mentioned before, the process modification of LVC introduces a compressor to generate vapour 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 is a variety of expressions in calculating the total equivalent work, such as Le Moulec and Kanniche (2011a), Ahn et al. (2013), Van Wagener et al. (2013). The following equation from the work of Van Wagener and Rochelle (2011) and Liang et al. (2015) is chosen.

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

It uses a Carnot efficiency term which accounted for the increasing value of steam at high temperature. Additionally, 75% efficiency was applied to account for non-ideal 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<sub>2</sub>);  $T_{sink}$  is the cold end temperature of Carnot engine, and set at 313 K here;  $W_{comp}$  is the compression work (GJ/t CO<sub>2</sub>);  $W_{add}$  is the additional equipment work such as the compressor in LVC (GJ/t CO<sub>2</sub>).

As for calculating the compression work, the following correlation from the work of Le Moulec and Kanniche (2011a) can be used.

$$W_{comp} = 8.3673 + 22.216 \ln P_F - (27.118 + 0.0256 P_F) \ln P_S$$

where  $W_{comp}$  is the compression work (kWh/t CO<sub>2</sub>);  $P_F$  is the final delivery pressure, and set  $P_F$  as 110 bar here;  $P_S$  is the initial pressure of compression, and stripper pressure is used, 1.5 bar here. So values of  $W_{comp}$  for all cases are 100.66 kWh/t CO<sub>2</sub>, i.e., 0.362 GJ/t CO<sub>2</sub>.

The total equivalent work of each process configurations 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 Sensitivity analysis of key parameters

### 4.2.1 Effect of amine concentration and lean solvent loading in conventional process

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<sub>2</sub>. The reboiler heat duty is rather sensitive to the solvent flow rate as the vaporization of water for CO<sub>2</sub> stripping contributes most to

the reboiler duty at low solvent flow rate values. In contrast, if lean loading is extremely low, more heat is provided by the reboiler duty as the heat of reaction between amines and CO<sub>2</sub> 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<sub>2</sub>/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<sub>2</sub>/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.

### 4.2.2 Effect of stripper pressure in conventional process

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 Moulec and Kanniche (2011a), which is to operate the stripper at vacuum/sub-ambient pressure. CO<sub>2</sub> desorption becomes easier as stripper pressure is high. From another perspective, if stripper pressure is higher, the temperature in the reboiler is also higher, and then higher pressure steam is required for solvent regeneration. Therefore, the influence of stripper pressure should be evaluated in total equivalent work to search for the optimal strategy. Figure 13 illustrates the results of simulation for conventional process, and it indicates that both for MEA and DEA, higher pressure is slightly beneficial to reducing total energy consumption because of higher reboiler temperature.

### 4.2.3 Effect of lean solvent loading for process modifications

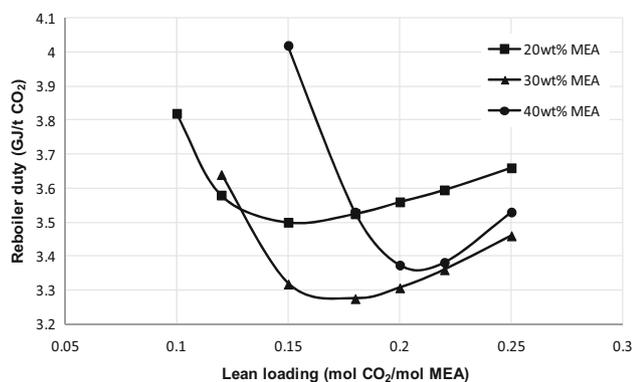
As the principle and simulation results mentioned previously, the loading of the lean amine solution is of great significance in the energy consumption reduction. And the simulation results indicated that the reboiler heat duty is rather sensitive to the lean loading. Thus, it is essential to simulate all the processes to come up with the optimal

**Table 2** Total equivalent work of process configurations using MEA

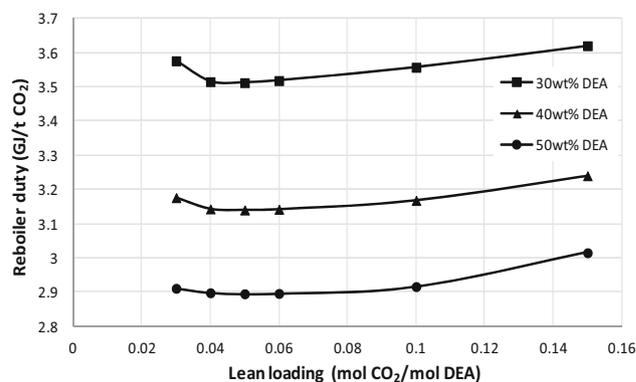
Modifications	Rich loading (mol CO <sub>2</sub> /mol MEA)	$Q_{reb}$ (GJ/t CO <sub>2</sub> )	$W_{add}$ (GJ/t CO <sub>2</sub> )	$W_{eq}$ (GJ/t CO <sub>2</sub> )	Total energy savings (%)
Conventional	0.465	3.460	0	0.911	–
ICA	0.492	3.216	0	0.873	4.20
FGP	0.485	3.278	0	0.883	3.17
RSS	0.465	3.192	0	0.869	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	0.0394	0.876	3.87

**Table 3** Total equivalent work of process configurations using DEA

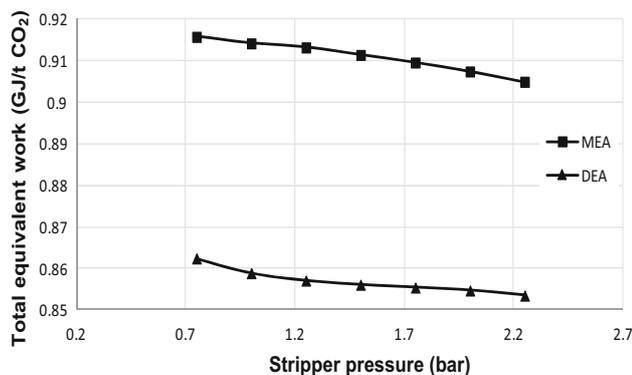
Modifications	Rich loading (mol CO <sub>2</sub> /mol DEA)	$Q_{reb}$ (GJ/t CO <sub>2</sub> )	$W_{add}$ (GJ/t CO <sub>2</sub> )	$W_{eq}$ (GJ/t CO <sub>2</sub> )	Total energy savings (%)
Conventional	0.447	3.168	0	0.856	–
ICA	0.468	3.078	0	0.842	1.64
FGP	0.468	3.080	0	0.842	1.60
RSS	0.447	2.945	0	0.821	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

**Fig. 11** Effect of MEA concentration and CO<sub>2</sub> loading of lean solvent on reboiler duty

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 term of reducing energy

**Fig. 12** Effect of DEA concentration and CO<sub>2</sub> loading of lean solvent on reboiler duty

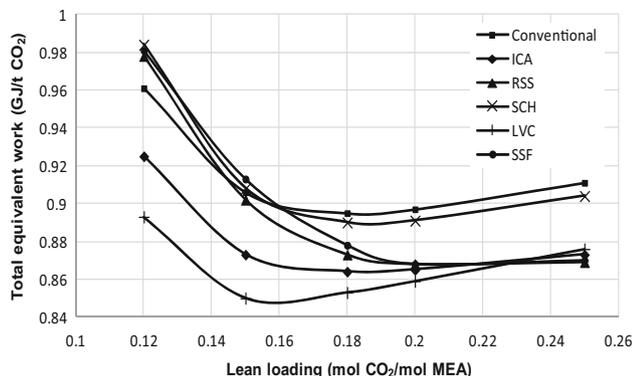
consumption. The result shows in Fig. 14 for MEA and Fig. 15 for DEA. As for MEA, all the total equivalent work of these processes have a minimum point as the lean loading is increasing. In conventional process, ICA, and SCH, the minimums occur at approximately 0.18 mol



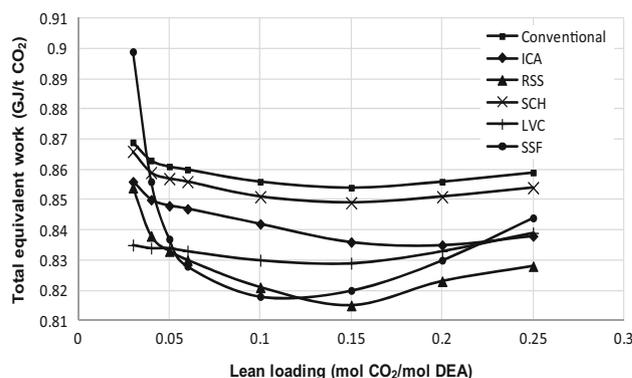
**Fig. 13** Effect of stripper pressure on total equivalent work in the conventional process

CO<sub>2</sub>/mol MEA, and it rises to 0.22 mol CO<sub>2</sub>/mol MEA for RSS and SSF. In contrast, minimum of LVC appears at around 0.16 mol CO<sub>2</sub>/mol MEA because the heat provided by compressed vapour is pretty effective. At lower lean CO<sub>2</sub> loading, the total equivalent work of RSS, SCH, 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.

Correspondingly, 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<sub>2</sub> loading of about 0.15 mol CO<sub>2</sub>/mol DEA. ICA raises this point to 0.2 mol CO<sub>2</sub>/mol DEA while SSF lowers it to 0.1 mol CO<sub>2</sub>/mol DEA. The total equivalent work of RSS has an obvious change as the minimum point occurs at 0.15 mol CO<sub>2</sub>/mol DEA. The energy consumption of 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



**Fig. 14** Compare total equivalent work between different process configurations at different lean loading of MEA



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

configurations for DEA are in the following order, RSS ~ SSF < LVC < ICA < SCH < conventional process. Compared with LVC is more favourable to MEA, it can be concluded that RSS or the variant of RSS are more efficient to DEA, because CO<sub>2</sub> is easier released from DEA solution in thermodynamics.

### 5 Conclusions

In this work, ten different process configurations have been simulated both for MEA and DEA on the same operating conditions. As a result, process modifications are proved to be an efficient way to optimize the energy consumption in CO<sub>2</sub> capture process using MEA and DEA. It has been shown that LVC, ICA, FGP, RSS, SSF are favourable in MEA, as from 3.17% to 4.61% of total equivalent work are reduced respectively in preliminary simulation. And for DEA, RSS, SSF, LVC have better performance, as from 2.70% to 4.50% are reduced.

Meanwhile, this work presents the influence of three operating parameters in energy savings, namely, amine concentration, stripper pressure, and loading of lean amine solvent. The study of amine concentration and lean loading in conventional process 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 processes using MEA than DEA. Both for MEA and DEA, higher pressure is beneficial to reducing total energy consumption in all of the processes. The effect of lean solvent loading on process modifications for MEA and DEA is pretty different, and the minimum point of total equivalent work also depends on amine type and process. RSS or the variant of RSS are more efficient to DEA, and LVC and ICA is more

favourable to MEA. Further work on blended amines will be carried out latter.

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