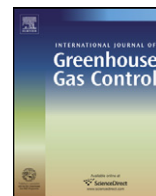




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Influence of stripper operating parameters on the performance of amine absorption systems for post-combustion carbon capture: Part II. Vacuum strippers

Sumedh S. Warudkar^{a,*}, Kenneth R. Cox^a, Michael S. Wong^{a,b}, George J. Hirasaki^a

^a Department of Chemical and Biomolecular Engineering, Rice University, 6100 Main St, MS 362, Houston, TX 77005, USA

^b Department of Chemistry, Rice University, 6100 Main St, MS 60, Houston, TX 77005, USA

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ABSTRACT

The alkanolamine absorption process is viewed favorably for use in the separation of carbon dioxide (CO₂) from point emission sources such as coal-fired power plants. At present, natural gas sweetening is the most important application for this technology. However, on a number of accounts such as the feed conditions of gas, its composition and process economics; natural gas sweetening and carbon capture are very different applications. Current technology is optimized toward providing a high performance for the former. As a part of this two-part study, we have used the process simulation software ProMax[®] to perform a detailed analysis on the effect of stripper operating pressure on factors like reboiler energy duty, absorber and stripper column sizing and parasitic power loss. We have examined the performance of monoethanolamine (MEA), diethanolamine (DEA) and diglycolamine (DGA) which are all commercial absorbents that can be reliably modeled in ProMax[®]. In part I of this study, we have analyzed the performance of strippers operated at pressures ranging from 150 kPa to 300 kPa. In this part of the study, we examine the performance of vacuum strippers operating under low vacuum at pressures of 30 kPa, 50 kPa and 75 kPa. Since vacuum strippers operate at lower temperatures than conventional stripper configurations, it is possible to use waste heat in the reboiler. In this study, we explore this possibility and consider 5 scenarios in which varying fractions of the reboiler steam are provided from waste heat sources located outside the turbine system. As with the cases presented in Part I, our comparisons of different configurations are based on energy consumption and column dimensions required for 90% CO₂ capture (separation + compression) from a 400 MW coal-fired power plant. CO₂ separated from the flue gas is compressed to a pressure of 16 MPa, typically maintained in the pipelines. On the basis of our findings, we report that vacuum stripping is an attractive alternative to conventional stripping. It is particularly attractive if significant sources of waste heat outside the turbine system can be located. We also conclude from our work that DEA and DGA have a superior performance than MEA when vacuum strippers are used. Use of vacuum strippers will certainly result in increased capital costs due to the need for larger equipment. However, in the view of potential savings in operating costs mainly by reduction in parasitic power loss; the increased capital expenditure may be justifiable.

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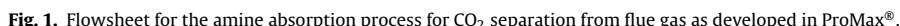
1. Introduction

Coal and natural gas fired power plants together produce more than 50% of the electricity worldwide and emit in excess of 40% of the carbon dioxide (CO₂) (Gielen and Podkafski, 2004). Flue gas from a typical pulverized coal fired power plant contains between 10% and 15% CO₂ by volume (Halmann and Steinberg, 1999). Carbon capture and storage (CCS) is widely considered as

a promising method for continuing our use of fossil fuels while also limiting the atmospheric CO₂ emissions (Metz, 2005). While several different methods of separating CO₂ from gas mixtures exist, the most mature and commercially accepted technique is the use of chemical absorbents such as alkanolamines. In the conventional alkanolamine absorption process adapted for CO₂ capture, the absorber is operated at slightly above atmospheric pressure (Kidnay et al., 2006). The stripper unit is generally operated between 150 kPa and 200 kPa and a temperature between 105 °C and 110 °C with steam at around 415 kPa providing the energy required for absorbent regeneration (Kohl and Nielsen, 1997). At a coal fired power plant retrofitted with this technology, the reboiler steam will likely be drawn from upstream of the low

* Corresponding author. Tel.: +1 713 501 8782.

E-mail addresses: sumedh@rice.edu, sumedh.warudkar@gmail.com (S.S. Warudkar).



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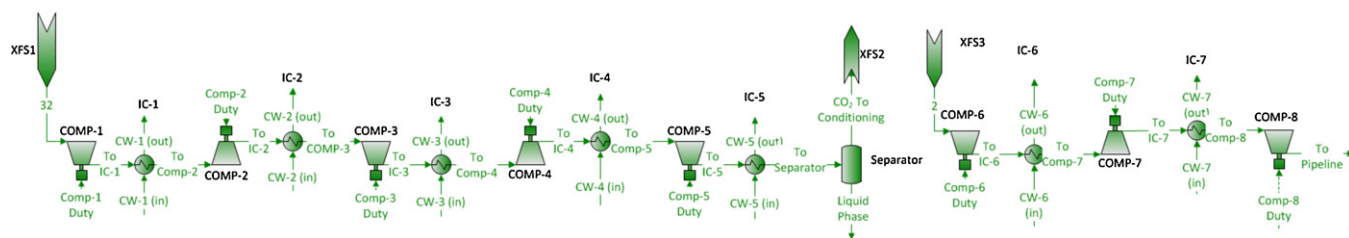


Fig. 2. Flowsheet for the CO₂ compression train as developed in ProMax®.

compressed CO₂ stream to below 150 ppm in order to minimize corrosion and hydrate formation problems in the pipeline (Thomas, 2005).

3. Vacuum stripping

Fig. 3 shows a plot of stripper pressure versus the average stripper operating temperature for 30 wt% diethanolamine (DEA) calculated using ProMax®. From the plot, it can be observed that as the stripper operating pressure is reduced, the average stripper operating temperature decreases. Thus, in a vacuum stripper the operating temperature is below 100 °C. In a stripper column operating at around 150 kPa, the temperature is around 110 °C and depending on the choice of absorbent; it increases by 10–20 °C when the pressure rises to 300 kPa. Hence, in a conventional stripper; reboiler steam, required at around 450 kPa must be drawn from the low pressure (LP) turbine. It is this withdrawal of steam that contributes significantly to the parasitic power loss. Since vacuum strippers operate at a lower temperature; it is possible to maintain the necessary conditions for stripping by using steam at much lower pressure such as near atmospheric steam. At a power plant, there are several potential sources of this very low pressure steam. The simplest, however the most inefficient way of extracting this steam is by drawing the steam toward the end of the low pressure (LP) turbine before it reaches the vacuum condenser. In addition, waste heat is available at power plants in varying quantities which could be used to provide a part of or the entire reboiler steam requirement.

Unlike in a conventional stripper, where the pressure can be increased or decreased by changing the pressure of the vapor and liquid phases; a vacuum stripper requires additional equipment to maintain the required degree of vacuum in the column. In large, commercial units; the most common means of generating vacuum

is the use of a steam jet ejector. A steam jet ejector employs the concept of a converging–diverging nozzle to convert the pressure energy of a motive fluid to kinetic energy which creates a low pressure zone which draws and entrains the suction fluid (Green, 2004). In industrial processes, steam jet ejectors are preferred to vacuum pumps due to their simplicity, absence of moving parts and the resulting reliability. Steam jet ejectors have a low isentropic efficiency of around 80% and require high pressure steam as a motive fluid.

4. Definition of process & simulation parameters

A typical 400 MW power plant emits approximately 3 million tons/year of CO₂ (Katzer, 2007). The flue gas flow rate is approximately 32 million sm³/day. The composition of the flue gas used in this study is provided in Table 1 (Singh et al., 2003). It can be seen that the concentration of CO₂ is close to 15% which corresponds to a low partial pressure of approximately 15 kPa. With an aim of estimating reasonable equipment sizes, we use a 3 absorber/stripper train configuration for all simulation cases in this study. As a part of this study, we compare three alkanolamines – monoethanolamine (MEA), a very fast reacting primary amine with a high heat of reaction, diethanolamine (DEA), a slow reacting secondary amine with a low heat of reaction and diglycolamine (DGA), another fast reacting primary amine, with low volatility. As explained in part I of this study, we employ certain guides and constraints to simulate the amine absorption process under conditions comparable to those held in commercial practice. These are:

1. At least 90% of the entering CO₂ must be captured and compressed (Spliethoff, 2010).
2. All amines are studied at the typical concentrations that they are used at in commercial practice. These are: MEA – 20 wt%, DEA – 40 wt% and DGA – 60 wt% (Kohl and Nielsen, 1997).
3. The maximum CO₂ loading of amine solutions is 0.4 moles-CO₂/mole-amine (Kohl and Nielsen, 1997).

It is quite well known that the use of aqueous alkanolamines is accompanied with corrosion problems. To minimize corrosion of carbon steel equipment, which is the common material of construction; the rich amine loadings are maintained below 0.4 mole-CO₂/mole-amine. This also forms the rationale behind the maximum working concentrations for the various amine

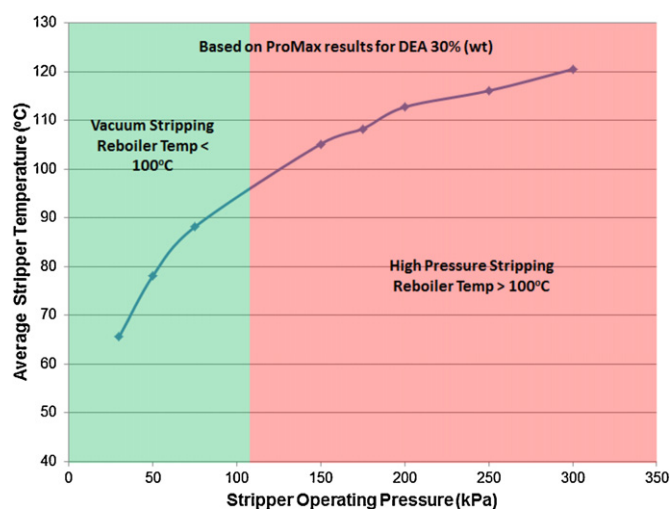


Fig. 3. Dependence of stripper operating temperature on operating pressure.

Table 1
Composition of flue gas.

Component	Value (mole %)
Water (H ₂ O)	11.69
Carbon dioxide (CO ₂)	14.59
Oxygen (O ₂)	2.85
Nitrogen (N ₂)	69.95
Sulfur dioxide (SO ₂)	0.01
Argon (Ar)	0.91

Table 2
System parameters used in ProMax simulations.

Simulation parameters	
Parameter	Value
Absorber/stripper flooding fraction	0.8
Tray type	Sieve trays
Absorber/stripper tray spacing	0.61 m
Absorber/stripper weir height	0.076 m
Condenser temperature	30 °C
Absorber # of trays (MEA, DGA)	2
Absorber # of trays (DEA)	10
Stripper # of trays (MEA, DGA)	15
Stripper # of trays (DEA)	10
Lean/rich exchanger end approach temperature	5 °C
Reboiler steam conditions	122 kPa, 140 °C
Jet ejector steam conditions	630 kPa, 284.6 °C
Jet ejector CO ₂ delivery pressure	101.33 kPa

solutions. Higher amine loadings can be maintained if stainless steel – a material much more resistant to corrosion is used for construction. However, replacing carbon steel with stainless steel results in a significant increase in costs and is considered an unattractive proposition.

Due to their faster reaction rates with CO₂, we have used a 2 ideal tray configuration in the absorbers for MEA and DGA. Since DEA reacts at a significantly slower rate; a 10 ideal tray configuration is required in the absorber. Due to the lower operating temperature in the stripper and thus, the relatively unfavorable conditions for CO₂ desorption; MEA and DGA require more stages than DEA, which is easier to regenerate. We use a 15 ideal tray configuration for MEA and DGA while using a 10 ideal tray configuration for DEA. We have listed all important simulation parameters in Table 2. Heat exchangers are critical pieces of equipment in amine absorption units since large quantities of heat are transferred between fluids in the Lean/Rich amine heat exchanger and the reboiler units. Thus, the amount of energy transferred in the heat exchanger and the end approach temperature are both important parameters to determine the heat exchanger sizing and thus its cost. In all the simulation cases, the end approach temperature for the Lean/Rich Heat exchanger is maintained at 5 °C. Table 3 is a summary of end approach temperature for the reboiler and the amount of heat transferred in the two heat exchangers.

Table 3
Heat exchanger parameters in the amine absorption system simulations.

Simulation case stripper pressure (kPa)	Lean/rich heat exchanger quantity of heat transferred (MW)			Reboiler minimum end approach temperature (°C)			Reboiler duty (MW)		
	MEA 20 wt%	DEA 40 wt%	DGA 60 wt%	MEA 20 wt%	DEA 40 wt%	DGA 60 wt%	MEA 20 wt%	DEA 40 wt%	DGA 60 wt%
Absorbent									
30	111.4	30.5	37.5	33.8	34.8	32.0	499.5	139.8	219.1
50	175.6	63.0	47.4	21.7	23.0	20.0	341.7	122.7	163.0
75	207.4	92.5	53.8	11.3	14.7	9.8	280.4	112.2	136.7

Table 4
Parameters for evaluation of equivalent work of reboiler steam.

Parameters for evaluation of equivalent power loss			
Parameter	Inlet to LP turbine		Inlet to steam jet ejector
Steam pressure (kPa)	630		630
Steam temperature (°C)	284.6		284.6
Mass enthalpy (kJ/kg)	3029.4		3029.4
Mass entropy (kJ/kg-°C)	7.3		7.3
Contribution of LP turbine to plant output			45%
Overall efficiency of turbine generator system			70%
Isentropic efficiency of steam jet ejector			80%
Steam flow to LP turbine (kg/s)			324.4

Table 5
Different waste heat availability scenarios considered in this study.

Simulation case #	Assumptions
0%	No suitable waste heat source available. All reboiler steam is tapped from downstream of LP turbine at 122 kPa
25%	25% of the reboiler steam provided by waste heat sources
50%	50% of the reboiler steam provided by waste heat sources
75%	75% of the reboiler steam provided by waste heat sources
100%	100% of the reboiler steam provided by waste heat sources

5. Evaluation of parasitic power loss

A power plant retrofitted with amine absorption technology for carbon capture has several energy sinks, which consume or reduce the power plant output. In case of the amine absorption process utilizing a vacuum stripper, these energy sinks involve auxiliary equipment like flue gas blower and pumps, the compression train, energy consumed by the reboiler and that by the steam jet ejector. While the energy consumed by the blower, pumps and compressors is direct in the form of generated electricity that by the reboiler and the jet ejector is in the form of steam which results in a loss of the power generation capacity. This loss of power generation capacity is estimated by calculating the equivalent work for the consumed steam. In part I, we have described the method that we follow to evaluate the equivalent power generated by the reboiler steam. We use the same methodology for estimating the contribution of reboiler and jet ejector steam to the parasitic power loss (Bartley et al., 2007; Spliethoff, 2010). As described in Section 3, varying amounts of waste heat – very low pressure steam, are available at power plants. Various parameters used to evaluate parasitic power losses are compiled in Table 4. In order to analyze the parasitic power loss under different conditions of waste heat availability, we consider 5 different scenarios in this study. These are discussed in Table 5 below. Considering the wide range of scenarios below gives a good estimate of the potential of vacuum stripping technology in reducing parasitic power loss.

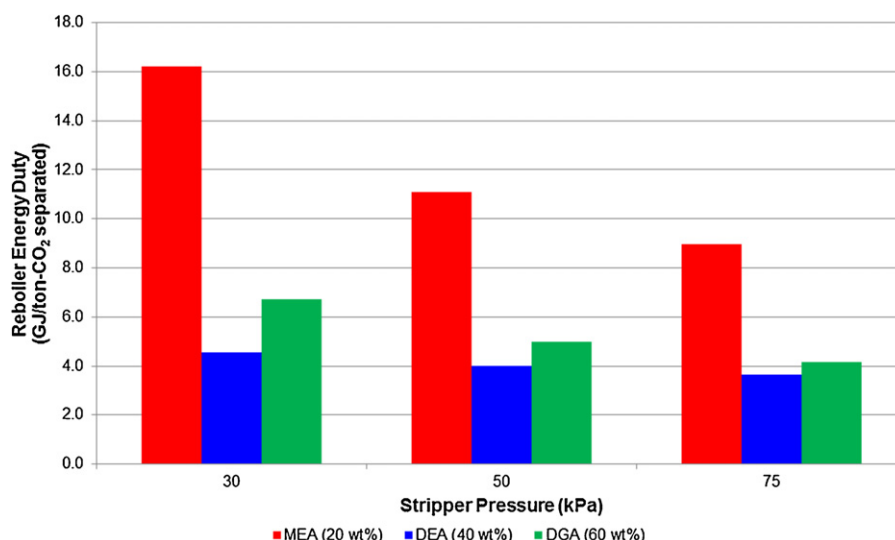


Fig. 4. Effect of change in stripper pressure on reboiler duty.

6. Results and discussion

6.1. Effect of amine concentration and stripper operating conditions

6.1.1. Effect on total energy consumption for CO₂ separation

Fig. 4 shows the effect of the stripper operating pressure on the reboiler energy duty for separation of CO₂. Amongst the three absorbents studied, MEA has the highest reboiler duty of approximately 16 GJ/ton-CO₂ at a stripper pressure of 30 kPa which decreases drastically to approximately 9 GJ/ton-CO₂ at 75 kPa. 40 wt% DEA has a slightly lower reboiler heat duty than 60 wt% DGA. DGA binds to CO₂ more strongly than DEA does. In a vacuum stripper which operates at a relatively low temperature, larger amounts of energy must be provided to release the regenerate the absorbent to the required degree. As the stripper pressure is raised from 30 kPa to 75 kPa, the reboiler energy duty reduces by roughly 50% for MEA, around 20% for DEA and by 38% for DGA. The lowest reboiler duty of 3.7 GJ/ton-CO₂ corresponds to a stripper pressure of 75 kPa with the use of 40% DEA as the absorbent.

A quick referral to part I of this study shows that the reboiler duty in all vacuum stripping cases is either higher or significantly higher than that for the high pressure strippers. A deeper examination of this data reveals that between the vacuum and high pressure strippers, the reboiler steam conditions have been modified which changes the equivalent work of the steam consumed. When we compare the parasitic power loss between different systems Section 6.4; we show that merely comparing the reboiler energy duty results in a misleading and incomplete analysis of stripper performance.

6.1.2. Effect on absorber and stripper sizing

At an amine absorption unit, reboiler energy duty is the primary contributor to the operating costs. A significant share of the capital costs is toward the construction of the large equipment such as absorber and stripper columns. Evaluating the effect of operating parameters on the equipment sizing is critical toward the economic feasibility of a novel configuration. Fig. 5 is a comparison of the absorber diameter for MEA, DEA and DGA at different stripper operating pressures. There is little to choose between the absorbents and the stripper operating conditions when it comes to absorber sizing. Column diameters are strongly dependent on the vapor flow-rates. Since, in the case of a CO₂ absorber; the vapor

phase is flue gas, the flow-rate of which is constant across all simulation cases. Hence, there is little variance in the absorber diameter with a change in absorbent or operating conditions. These absorber diameters are comparable in size to those for the high pressure stripper systems from Part I, suggesting that vacuum stripping will have minimal effects on the size of absorber column costs. Fig. 6 is a plot of the effect of stripper pressure on the stripper diameter. The most noticeable aspect of the plot is the size of the stripper columns required for MEA which are significantly bigger than the ones required for DEA and DGA. Comparing the size of vacuum strippers with high pressure strippers from Part I shows that the column diameter increases irrespective of the choice of absorbent. Comparing the stripper diameter for 75 kPa and 150 kPa system shows that the diameters for MEA, DEA and DGA increase by 36%, 13% and 32% respectively. Clearly, amine absorption systems with vacuum strippers require larger stripper columns and will result in increased equipment cost. The increase in column size can be explained on the basis of the change in the vapor flow rates in the stripper column. The specific volume of gases is inversely proportional to their pressure. Thus, when the operating pressure in a stripper column is reduced; the volumetric flow-rate of stripping vapors (steam) and CO₂ in the stripper increases. In addition, as the stripper pressure is reduced, so does the stripper temperature. This results in an increase in the ratio of p_{H_2O}/p_{CO_2} , which coupled with the increased volumetric flow rate of vapors results in larger diameter stripper columns (Oexmann and Kather, 2010).

6.2. Contribution of various constituent processes to reboiler heat duty

In Section 2, we have discussed that the energy provided to the reboiler is used for driving the CO₂ desorption reaction, for heating the rich amine entering the stripper and to generate stripping vapor (steam). In Part I of this study, we have discussed the approach that we have developed to estimate the contribution of each constituent processes toward the reboiler duty. We use the same approach to calculate the energy requirement for the 3 constituent processes in this work. Figs. 7 and 8 show the dependence of the energy consumed to generate the stripping vapor (steam) and the cumulative contribution of sensible heating and heat of reaction.

As we have remarked previously in Part I, the trend seen in Fig. 7 – the dependence of energy consumed for generation of stripping vapor (steam) has a similarity to the trends seen in the Fig. 4 which

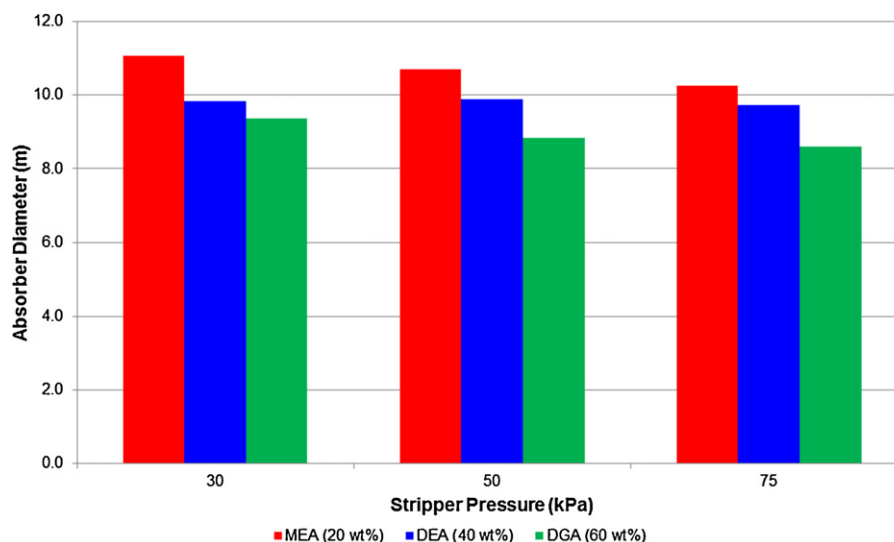


Fig. 5. Effect of stripper pressure on absorber diameter.

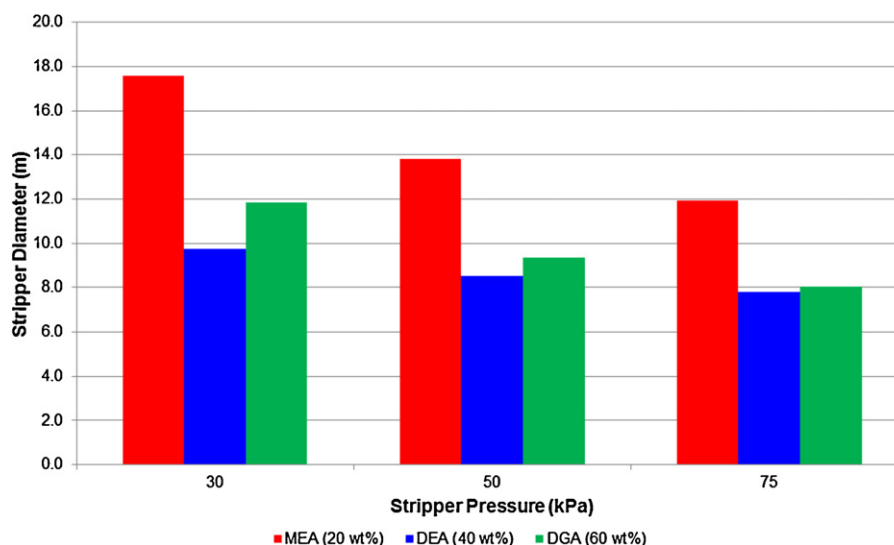


Fig. 6. Effect of stripper pressure on stripper diameter.

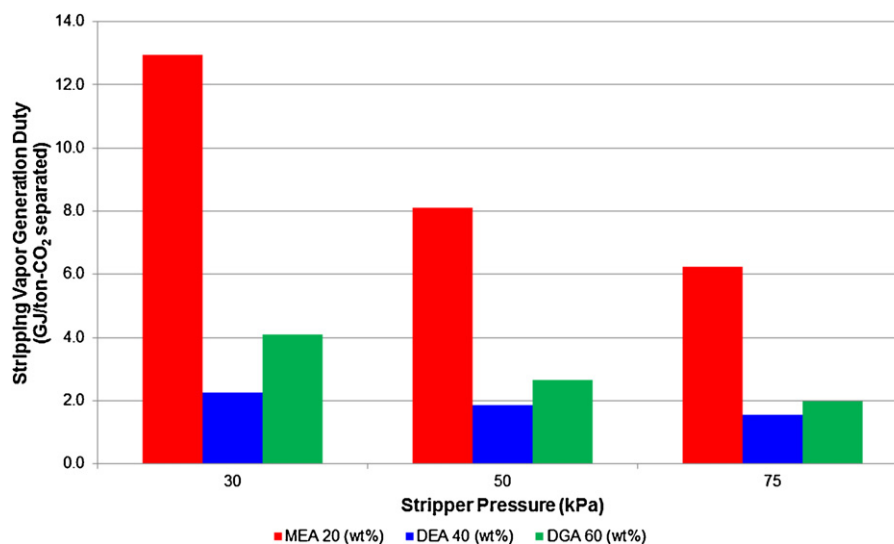


Fig. 7. Contribution of stripping vapor energy consumption to reboiler heat duty.

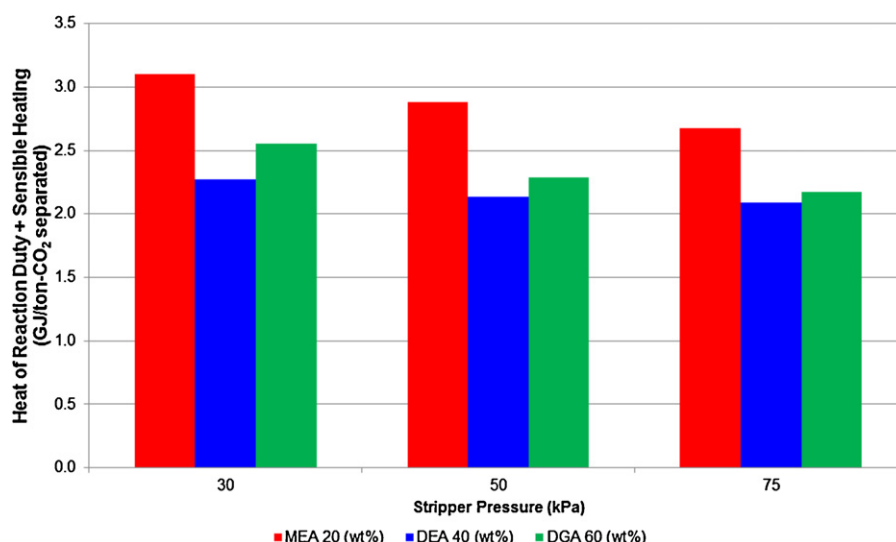


Fig. 8. Contribution of heat of absorption and sensible heating to reboiler heat duty.

Table 6

Effect of stripper pressure on average stripper and reboiler temperatures for various absorbents.

Simulation case stripper pressure (kPa)	Average stripper temperature (°C)			Reboiler temperature (°C)		
	MEA 20 wt%	DEA 40 wt%	DGA 60 wt%	MEA 20 wt%	DEA 40 wt%	DGA 60 wt%
30	67.5	63.7	66.6	71.6	70.4	73.6
50	80.2	74.6	78.5	83.8	82.3	85.8
75	90.7	82.8	87.8	94.3	92.0	96.5

is a plot of the dependence of reboiler duty on stripper pressure. This clearly suggests that irrespective of the change in operating pressure, the energy consumed for generating stripping vapor (steam) is a significant contributor to the reboiler energy duty. We also note that the stripping vapor (steam) requirement for MEA is enormous as compared to DEA and DGA. This could be attributed to the high stability of the MEA-CO₂ carbamate, which coupled with a low equilibrium vapor pressure for CO₂ at the lower stripper temperature results in a high $p_{\text{H}_2\text{O}}/p_{\text{CO}_2}$ and thus, high stripping vapor (steam) energy requirement. The stripping vapor (steam) energy requirements for DEA and DGA are consistent with the expected increase due to a decrease in the stripper temperature. As can be seen from Fig. 8, the cumulative contribution of sensible heating and heat of reaction has much less significant dependence on the stripper pressure and changes marginally as the stripper pressure increases from 30 kPa to 75 kPa.

6.3. Effect of stripper pressure on solvent losses

On the basis of the results discussed in Section 6, we recognize that vacuum strippers have disadvantages like a higher reboiler duty and the need for larger stripper columns. In particular, the need for larger equipment may result in an increased capital cost for the amine absorption plant. However, there are potential advantages such as the low stripper and heat exchanger operating temperature which reduces the risk of equipment corrosion as well as the absorbent losses. Most alkanolamines undergo thermal degradation at temperatures around 135 °C, thus the stripper and reboiler operating pressure is a particularly important parameter to predict the thermal degradation of amines. Table 6 is a compilation of the average stripper and reboiler temperatures for MEA, DEA and DGA. Table 6 clearly shows that the vacuum strippers operate at significantly lower temperatures than the high pressure strippers studied in Part I. This low operating temperature can be expected to result in a considerable reduction in the absorbent losses due

to thermal degradation. The lower temperatures will also reduce the severity of equipment corrosion problems which is an attractive feature of amine units with vacuum strippers. Table 7 is a compilation of the absorbent losses through absorber and stripper overhead vapors. In the simulations developed for this study, we do not include an amine wash section in the absorber. While this results in a higher amine slipping at the absorber overheads, it clearly indicates that the overheads losses with MEA and DGA as absorbents are more severe than with DEA.

6.4. Effect of carbon capture unit on parasitic power loss

6.4.1. Comparison of vacuum stripper and high pressure stripper systems

Fig. 4 shows that the reboiler energy duty for vacuum strippers, expressed in terms of energy required per unit mass of CO₂ separated is greater than that for high pressure strippers. However, reboilers for vacuum strippers are supplied with steam at 122 kPa which has significantly lower energy content than the 415 kPa steam required for high pressure strippers systems. Since the reboiler steam for vacuum and high pressure strippers is at different conditions, their resulting equivalent power loss is not the same – a larger quantity of the 122 kPa steam must be drawn to result in the same power loss as the 415 kPa. Fig. 9 is a plot of the parasitic power loss at the reference 400 MW power plant,

Table 7

Solvent losses through absorber and stripper overheads.

Simulation case stripper pressure (kPa)	Solvent losses absorber and stripper overheads (kg/day)		
	MEA 20 wt%	DEA 40 wt%	DGA 60 wt%
30	7136.2	59.5	4833.7
50	7875.9	62.7	7098.0
75	9091.2	62.7	10,181.6

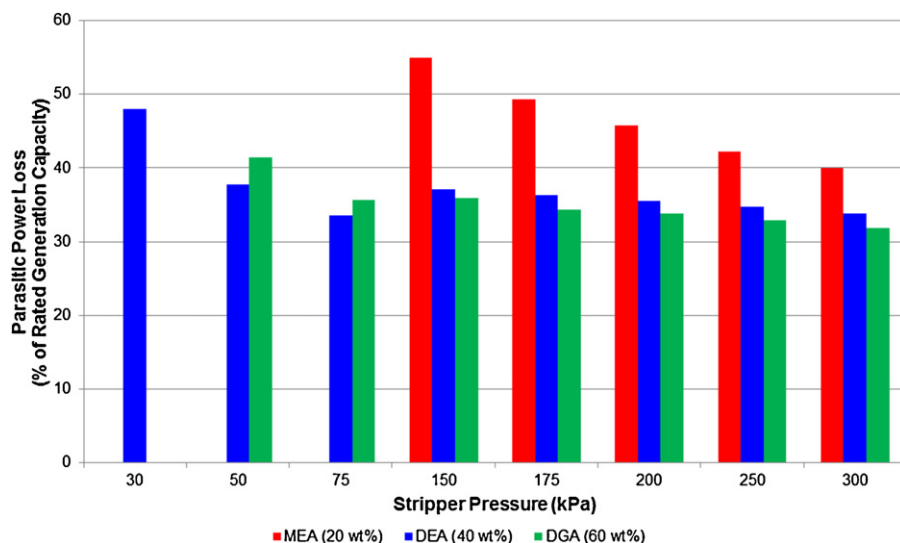


Fig. 9. Parasitic losses due to carbon capture system at reference power plant.

expressed as a % of the rated generation capacity. The most remarkable feature of this plot is the absence of any data points for MEA and DGA corresponding to a stripper pressure of 30 kPa and for MEA for 50 kPa and 75 kPa. In all cases, this is a result of the reboiler steam requirement being greater than the flow-rate of steam circulated to the low pressure (LP) turbine, suggesting that it is practically impossible to implement the 30 kPa configuration with any of the absorbents except DEA. We also note that with DEA and DGA as absorbents, the 75 kPa stripper has a competitive performance with all the high pressure strippers. DEA in particular has a lower parasitic power loss than all the high pressure stripper configurations. As was remarked in Part I of the study, MEA results in very high parasitic power losses under all system conditions studied.

6.4.2. Effect of waste heat availability on parasitic power loss

As described in Table 5, we have considered 5 different scenarios of waste heat availability at the power plant. At one end of the range is the case where no suitable sources of waste heat are available at the power plant, thus requiring that all steam be drawn from downstream of the low pressure (LP) turbine. At the other end of

the spectrum is a case where enough waste heat is available at the power plant to provide all reboiler steam and no steam is required from the turbine system. Three other intermediate scenarios where 25%, 50% and 75% of the reboiler steam is provided by waste heat are considered in this study. Fig. 10 shows the variance in parasitic power loss for DEA and DGA systems with vacuum strippers under different scenarios for waste heat availability. There is a gradual reduction in the parasitic power loss as the source of the reboiler steam is switched from the turbine system to the waste heat source. The parasitic power losses reduce by 15–17 percentage points for the 50 kPa strippers and by 12–15 percentage points for the 75 kPa strippers when the steam source is changed from the turbine system to waste heat. The main reason for the higher parasitic duty for the 50 kPa strippers is the greater jet ejector steam requirement and increased duty for the pumps due to larger absorbent flow-rates. As can be seen from Fig. 10; when all the reboiler steam is provided by waste heat sources, the parasitic power loss for DEA and DGA at 75 kPa is around 21% and the difference between them is less than 1%. This suggests that both these absorbents have a reasonably similar performance with the vacuum strippers.

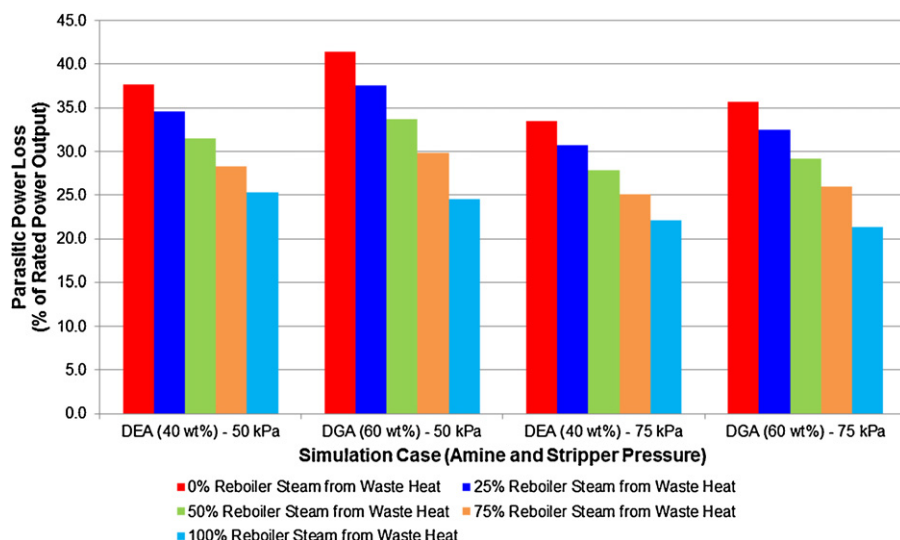


Fig. 10. Parasitic power loss with vacuum stripping under different waste heat availability scenarios.

7. Conclusions

This work has performed an in-depth exploration of the potential for using vacuum strippers in amine absorption systems for carbon capture. We have considered various relevant factors to analyze vacuum strippers and remark that they can be competitive with conventional and high pressure stripper systems when reboiler steam is drawn from the turbine system. In a scenario, where a large supply of waste heat is available at the power plant site; however, the 75 kPa vacuum strippers can result in parasitic power loss up to 15 percentage points lower than conventional and high pressure systems. We have found operational advantages, disadvantages and challenges toward the use vacuum strippers. We summarize these below:

- When all reboiler steam is tapped from the turbine system and the stripper pressure increased from 50 kPa to 75 kPa, the parasitic power loss for DEA decreases from 41.4% to 34.5% while that for DGA reduces from 41.3% to 32%. The parasitic power loss for both DEA and DGA is within 1% of that corresponding to a stripper pressure of 300 kPa.
- As remarked in Part I of this study, the absorber diameter is not sensitive to a change the stripper pressure. We find that the absorber columns for vacuum stripper systems are the same size as the high pressure systems of Part I.
- The stripper diameters are quite sensitive to the stripper operating pressures. When the stripper pressure is changed from 30 kPa to 75 kPa, the size of the MEA stripper decreases from 18 m to 11.8 m, the DEA stripper from 9.7 m to 7.8 m and DGA stripper from 11.8 m to 8 m. When compared to results from Part I of this work, we see that a 300 kPa stripper with MEA has a 6.5 m stripper column, DEA has 6.6 m and DGA has a 5.5 m column. Thus, vacuum strippers are significantly bigger in size than the conventional or high pressure strippers; which can be expected to increase the capital expenditure for the plant.
- Dissecting the reboiler energy duty reveals that MEA has an enormous requirement for stripping vapor (steam) which accounts for a majority of its reboiler duty. The high stripping vapor (steam) requirement also explains the large stripper columns, since their sizing is strongly dependent on vapor flow rates.
- Since vacuum strippers operate a low temperature (less than 100 °C), they may be expected to have a major operational advantage over conventional systems – low equipment corrosion. In addition, the lower temperature will have minimal (if any) thermal degradation problems as opposed to the 300 kPa configuration which may be expected to have solvent loss problems.
- Under the conditions used in this study, it is practically infeasible to use MEA as an absorbent for amine absorption systems with vacuum stripping due to the extremely large requirement for reboiler steam and impossibly large stripper columns. Similarly, the 30 kPa stripper configuration may be practically infeasible unless a very large source of waste heat is available at the power plant. Even then, this configuration will not be competitive with the 75 kPa system.

Based on the combined results of Part I and II of this work, we conclude that vacuum stripping at 75 kPa and high pressure at 300 kPa are more competitive configurations than conventional amine absorption. While high pressure stripping may be challenged by problems like thermal degradation, corrosion and absorbent losses, finding sources of waste heat is crucial to the cost effective deployment of vacuum strippers. The higher equipment cost of vacuum stripping maybe justified if a significant reduction in parasitic power loss can be achieved. Since the focus of this study is a technical analysis of amine absorption systems, it is not possible to make an accurate remark on the economics of the different process

configurations. Based on our investigations thus far, however, we recommend further detailed research into the economics of these novel configurations to determine their viability for commercial implementation.

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Appendix A. Supplementary data

Supplementary data associated with this article can be found, in the online version, at <http://dx.doi.org/10.1016/j.ijggc.2013.01.049>.

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