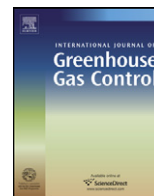




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Influence of stripper operating parameters on the performance of amine absorption systems for post-combustion carbon capture: Part I. High pressure strippers

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ABSTRACT

Amine absorption is a mature technology that is widely applied on a commercial scale for the removal of acid gases from gas mixtures. It is viewed as the most promising technology to separate carbon dioxide (CO₂) at point emission sources such as fossil fuel fired power plants as a part of a strategy called 'carbon capture and storage' being deployed to mitigate climate change. However, there are major challenges in advancing its use for this application; the most prominent one being – cost. It is estimated that the application of current technology to CO₂ capture will result in a 70–100% increase in the cost of electricity (COE). As a part of this two-part study, we have examined the influence of stripper operating parameters on carbon capture at a 400 MW pulverized coal-fired power plant retrofitted with amine absorption technology. We use the process simulation software, ProMax[®] to simulate the amine absorption process. Three commercial absorbents namely monoethanolamine (MEA), diethanolamine (DEA) and diglycolamine (DGA) are considered in this work. All the absorbents examined are studied at the typical working concentrations used in commercial operation. We also constrain the CO₂ loading of rich amine solution to 0.4 mol-CO₂/mol-amine to closely resemble the chemical environment in commercial operation. The main pointer used to compare system performance is the parasitic power loss due to the carbon capture. Parasitic power loss comprises of the loss of plant output due to withdrawal of reboiler steam from the low pressure turbine; the electricity required for CO₂ compression and by the plant auxiliaries such as blowers and pumps. We also examine the absorber and stripper column sizes and heat exchanger parameters which are key components of the system. On the basis of our findings in the first part of this study, we report that operating the stripper at higher pressures has significant advantages but also some disadvantages and operational challenges. We also report that DEA and DGA have a superior performance to MEA when used in high pressure strippers. We conclude that increasing the stripper pressure reduces the parasitic losses and the equipment sizing but may result in higher solvent losses and equipment corrosion.

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1. Introduction

The extensive burning of fossil fuels and the resulting release of carbon dioxide (CO₂) has been linked to global climate change. While there's no universal consensus on the causes of climate change yet, steps must be taken to mitigate the disastrous consequences of climate change. Coal and natural gas fired power plants together account for more than 50% of electricity produced in world and more than 40% of the CO₂ emissions (Gielen and Podkański, 2004). At these power plants, feedstock burns in the

presence of excess air to release energy which is converted to electricity. Flue gas leaving a power plant contains between 4 and 8% CO₂ by volume for natural gas feedstock and 10 and 15% CO₂ for coal (Halmann and Steinberg, 1999). Carbon capture and storage (CCS) involves the separation of CO₂ from flue gas, followed by compression and long term storage in suitable geological formations. Carbon capture and storage is viewed favorably as a strategy for reducing CO₂ emissions in the short and medium term (Metz, 2005). Several methods such as adsorption, absorption and membrane separation are available for the separation of CO₂ from gas mixtures (Davison and Thambimuthu, 2004). Of these, chemical absorption of CO₂ using aqueous alkanolamines is the most commercially mature and widely practiced technique. Several alkanolamines such as monoethanolamine (MEA), diethanolamine (DEA), methyl-diethanolamine (MDEA), diglycolamine (DGA) and

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2-amino-2-methyl-1-propanol (AMP) with different physiochemical properties are produced and used commercially (Kohl and Nielsen, 1997). A major challenge in the wide-scale application of amine absorption technology for carbon capture is the high energy consumption for absorbent regeneration. It is estimated that even with the application of state-of-the-art amine absorption technology, there will be a 70–100% increase in the cost of electricity (Toth, 2011).

Scientific researchers are taking different approaches to tackle the problem of high energy consumption in amine absorption. The more common approach adopted to reduce the energy consumption has been to develop alternatives to existing amines. Cullinane and Rochelle have explored the use of aqueous potassium carbonate (K_2CO_3) promoted by piperazine (PZ) – a cyclic diamine with very fast reaction kinetics. The activated K_2CO_3 solution has a lower heat of reaction than MEA (Cullinane and Rochelle, 2004). Freeman et al. have explored the use of concentrated, aqueous piperazine (PZ) as a fast reacting absorbent for CO_2 (Freeman et al., 2010). Aronu screened several novel absorbents for their potential for high absorption capacity, fast reaction kinetics which were estimated based on their molecular structures and their pK_a (Aronu et al., 2011). In contrast to the conventional, amine based absorbents; ionic liquids – salts that exist as liquids at room temperatures have been suggested as absorbents for carbon capture processes (Zhang et al., 2012). An alternative and relatively less common theme in ongoing research is the modification of the amine absorption process itself with an aim of optimizing it for carbon capture. Oyekan and Rochelle modeled various innovative stripper configurations in Aspen Plus[®]. This included a simple stripper operating at 160 kPa, a multipressure stripper with stages operating at 330, 230 and 160 kPa and a vacuum stripper operating at 30 kPa. They found that the vacuum stripper was the most attractive option and had the least equivalent power loss due to reboiler steam consumption when an absorbent with a low heat of reaction was selected (e.g. K_2CO_3 activated with PZ) (Oyekan and Rochelle, 2006). A subsequent study by the same authors found that the vacuum stripper, operating at 30 kPa was disadvantaged by the high cost of CO_2 compression (Oyekan and Rochelle, 2009). Reviews of several innovative configurations of the amine absorption process involving heat integration schemes, split flow arrangements, etc. have been authored by Cousins et al. (2011) and Le Moullec and Kanniche (2011).

While Oyekan and Rochelle have explored some novel stripper configurations, they have limited these cases to the multipressure stripper and the 30 kPa vacuum stripper. It is however, possible to operate a simple stripper at a range of pressures – elevated above the conventional pressure of 150 kPa and at varying degrees of vacuum. Most equipment used in industrial amine absorption systems is constructed using carbon steel which is susceptible to corrosion with acid gas containing alkanolamine solutions. Even with the use of corrosion inhibitors, the practical limit of acid gas loading of absorbent solutions is not more than 0.4 mol- CO_2 /mol-amine. Higher gas loadings can be maintained in routine operation with the use of stainless steel equipment. However, stainless steel is 2–3× more expensive than carbon steel. While it can be argued that lower parasitic power loss resulting from the use of stainless steel equipment will offset the higher capital costs, there are very few commercial CO_2 capture facilities in operation. Thus, carbon steel equipment with a maximum CO_2 loading of 0.4 represents the limit of reliable operation for current technology. Most current research efforts to model innovative stripper configurations do not discuss or account for this constraint on the CO_2 loading of amine solutions, which may result in lower than actual parasitic loss. We attempt to address these gaps in published research through our study.

In this two-part study, we have used the process simulator ProMax[®] to explore the operating parameter space of the amine absorption process. We have analyzed the effect of process parameters like stripper pressure, temperature and reboiler steam conditions (and source) on equipment sizing and parasitic power loss for CO_2 capture. We have considered 3 commercial alkanolamines – MEA, DEA and DGA as the absorbents for this study. In part I of this research, we evaluate the performance of amine units with strippers operating between 150 kPa and 300 kPa and utilizing 413 kPa (absolute) steam drawn from the low-pressure (LP) turbine. In addition, we evaluate the distribution of energy consumption between the various physical processes constituting absorbent regeneration. In part II, we evaluate the performance of the amine absorption process with the strippers operating under vacuum between 30 kPa and 75 kPa utilizing steam at 122 kPa respectively. Finally, we examine the performance of cases in parts I and II to comment on the most favorable stripper configurations for carbon capture.

2. Process description

We have designed the flow sheet for the amine absorption process in ProMax[®] – a process simulator used frequently in the oil and gas industry for modeling amine sweetening facilities. The flowsheet shown in Fig. 1 is based on a simplified scheme for alkanolamine acid gas removal process. Flue gas enters the carbon capture system at atmospheric pressure. A blower raises the pressure of flue gas in order to overcome pressure drop encountered in the absorber column. Before entering the carbon capture unit, flue gas passes through the Flue Gas Desulfurization (FGD) unit which reduces the sulfur oxides (SO_x) concentration to around 100 ppmv (Sajwan, 2006). SO_x compounds form Heat Stable Salts (HSS) with alkanolamines which result in absorbent losses. To minimize these losses, the flue gas is passed through a polishing scrubber where the SO_x concentration is reduced to 10 ppmv by contacting it with a dilute sodium hydroxide solution (Wu et al., 2010). After passing through the polishing scrubber, flue gas enters the absorber. Lean aqueous amine solution flows in at the top of the absorber at a temperature of around 45 °C. As the amine solution flows down, CO_2 from flue gas dissolves in the absorbent and reacts reversibly with the alkanolamine to form a soluble intermediate. Amine- CO_2 reaction is exothermic and results in a moderate increase in the solution temperature. The decarbonized flue gas leaves the absorber column at the top along with some absorbent vapors. For some volatile absorbents like MEA, these losses can be significant and a water wash section at the top of the absorber is required to recover lost absorbent. Rich amine solution exits the absorber at the bottom, where a booster pump provides the pressure head to overcome the head losses downstream. Rich amine solution from the absorber is contacted with hot lean amine exiting the reboiler in a heat exchanger to transfer sensible heat. After passing through the heat exchanger, lean amine enters a make-up/blow-down unit where water and/or amine is added or removed to maintain the required absorbent concentration and flow rate. A booster pump provides pressure head to the lean amine. At this point, lean amine temperature is around 65 °C which is brought down to 45 °C in a cooler. Rich amine solution at a temperature between 90 °C and 100 °C enters the stripper which operates between 150 and 200 kPa pressure and at approximately 110 °C. Steam at a pressure of 415 kPa is supplied to a reboiler located at the bottom of the stripper. The steam provides the energy required for sensible heating – to raise the temperature of the entering rich amine to that in the stripper, drives the endothermic CO_2 desorption reaction and generates the stripping vapor (steam). At the high temperature (110–120 °C) maintained in the stripper, the alkanolamine – CO_2 intermediate

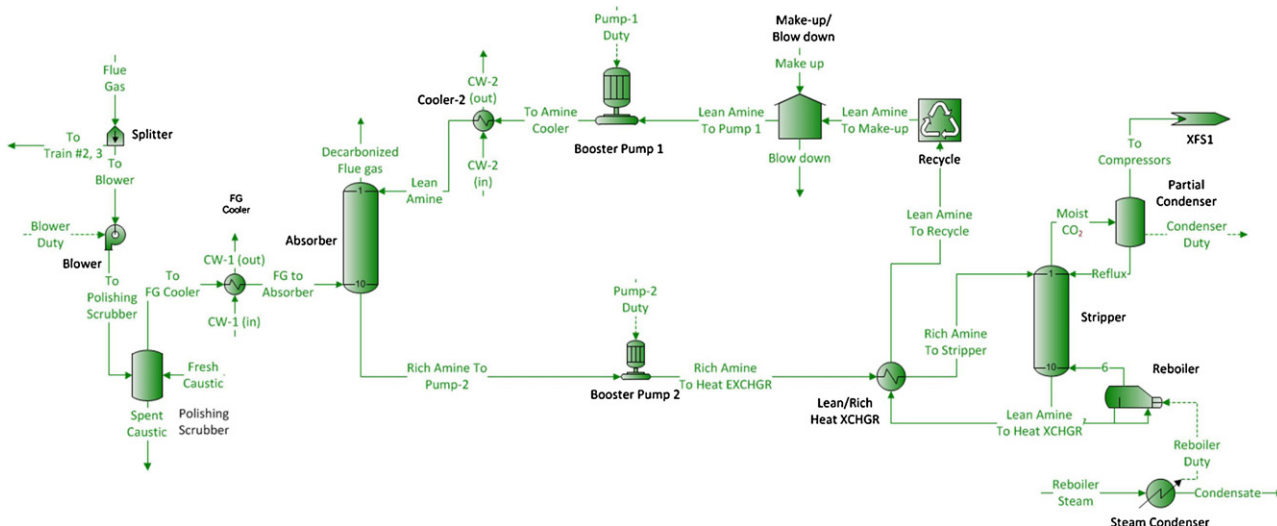


Fig. 1. Flowsheet for the amine absorption process for CO₂ separation from flue gas as developed in ProMax®.

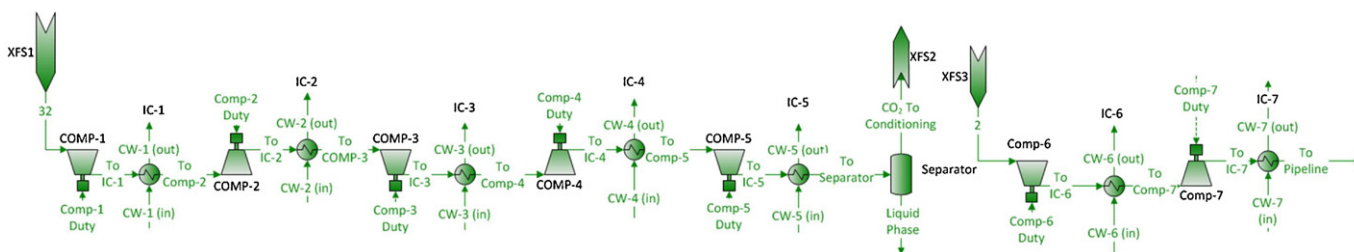
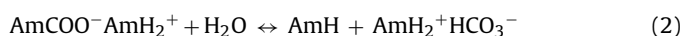
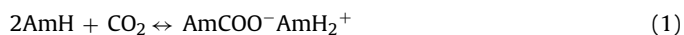


Fig. 2. Flowsheet for the CO₂ compression train as developed in ProMax®.

dissociates to release the CO₂ molecule. CO₂ along with water vapor exits the stripper at the top. It passes through a partial condenser where most of the water vapor condenses and is refluxed back into the stripper. CO₂ leaving the condenser enters a compressor – intercooler train for compression to a pipeline pressure of 16 MPa. A conditioning unit is required for water removal which is critical to the prevention of the formation of gas hydrates in the pipeline. Fig. 2 is the flowsheet used to model the compressor – intercooler train.

3. Amine absorption modeling

When a gas is physically absorbed into a solvent, the molecules merely transfer from one phase to another. Absorption of CO₂ into an alkanolamine solution involves chemical reactions between the acidic CO₂ and basic amine. In the first reaction, represented by Reaction (1) below; one CO₂ molecule reacts with 2 amine molecules to form one carbamate molecule. In the presence of water, the carbamate molecule dissociates to form one bicarbonate molecule and releases one amine molecule as depicted in Reaction (2).



In case of primary and secondary amines, the carbamate molecule formed is relatively stable and the formation of bicarbonate molecules is slow and insignificant (Sartori and Savage, 1983). At the typical temperature in an absorber (40–60 °C), the amine–CO₂ reaction does not take place instantaneously. Hence, when modeling CO₂ absorption by aqueous alkanolamines; reaction kinetics must be considered along with mass transfer to avoid

erroneous results (Vaidya and Kenig, 2007). Table 1 is a compilation of the heat of reaction and rate constants for the amines considered in this study (Alper, 1990a,b; Aresta, 2003; Vaidya and Kenig, 2007).

For the purpose of this study, we use the process simulation software ProMax® developed by Bryan Research & Engineering. ProMax® has a built-in package, TSWEET which accounts for the amine–CO₂ reaction kinetics in the absorber. The actual mechanism which limits the rate of CO₂ absorption in amines is its ionization in water. This is handled in ProMax® by the TSWEET kinetics model which is analogous to a liquid-side resistance in a mass transfer model. The TSWEET kinetics model calculates simultaneous mass transfer (physical process) and chemical reaction to account for the relatively slow CO₂–amine reaction rate. The total contacting time between the sour gas and the absorbent streams is a major factor in determining the amount of CO₂ picked up. In ProMax®, columns are often simulated using ideal trays rather than real trays. The liquid residence time on an individual tray is evaluated based on the tray geometry and further used to evaluate the total residence time in the column by multiplying it with the ratio of ideal to real stages. ProMax® has been demonstrated to be reliable in predicting the performance of amine gas sweetening

Table 1
Physiochemical properties of amines studied in this work.

Absorbent	Heat of reaction (kJ/kg)	Reaction rate constant (m ³ kmol ⁻¹ s ⁻¹)
Monoethanolamine (MEA)	1919	5545
Diethanolamine (DEA)	1519	2375
Diglycolamine (DGA)	1977	4517

units under a variety of system operating conditions (Alfadala and Al-Musleh, 2009; Luo et al., 2009; Polasek and Bullin, 1984).

4. Simplified analysis of overall regeneration heat duty

In Section 2, we discussed that steam supplied to the reboiler provides the energy required to heat the rich amine solution entering the stripper, to drive the endothermic CO₂ desorption reaction and to generate stripping vapor (steam). Historically, absorbents for CO₂ separation have been compared almost exclusively on the basis of their heat of reaction. However, dissecting the reboiler energy consumption into the above constituent processes helps to develop a deeper insight into absorbent performance. Energy required for absorbent regeneration in the stripper can be described as the sum of contributions of three processes, as shown below in Eq. (3) (Oexmann and Kather, 2010).

$$q_{\text{reboiler}} = q_{\text{reaction}} + q_{\text{stripping}} + q_{\text{sensible}} \quad (3)$$

In Eq. (3), q_{reboiler} is the reboiler duty, q_{reaction} is the heat of reaction for desorption of CO₂, $q_{\text{stripping}}$ is the energy consumed to generate stripping vapor (steam) and q_{sensible} is the energy required for sensible heating of the incoming rich amine solution to the stripper operating temperature. To evaluate these three component terms, we have to make the following simplifying assumptions.

All stripping vapor (steam) is condensed in the partial condenser and returned to the reboiler. Thus, the condenser duty corresponds to the energy carried by the stripping vapor (steam).

Heat of reaction for the amine–CO₂ desorption reaction is independent of temperature, pressure and amine concentration.

q_{reboiler} is the net energy provided to reboiler by condensing steam. As represented in Eq. (4), it is the difference between the enthalpy of entering vapor and exiting condensed liquid.

$$q_{\text{reboiler}} = \dot{m}_{\text{vapor}} H_{\text{vapor}} - \dot{m}_{\text{condensed liquid}} H_{\text{condensed liquid}} \quad (4)$$

Based on assumption #1, $q_{\text{stripping}}$ is the condenser duty, which can be evaluated by performing an energy balance on the partial condenser.

$$q_{\text{stripping}} = \dot{m}_{\text{moist CO}_2} H_{\text{moist CO}_2} - \dot{m}_{\text{reflux}} H_{\text{reflux}} - \dot{m}_{\text{dry CO}_2} H_{\text{dry CO}_2} \quad (5)$$

Based on assumption #2, q_{reaction} is the energy required to drive the endothermic amine–CO₂ desorption reaction.

$$q_{\text{reaction}} = \dot{m}_{\text{separated-CO}_2} \Delta H_{\text{reaction}} \quad (6)$$

Using the above equations, we can calculate q_{sensible} which remains as the unknown from Eq. (3). With the value of q_{sensible} , $q_{\text{stripping}}$ and q_{reaction} known; the effect of various system parameters on their contribution to q_{reboiler} can be studied.

5. Definition of process and simulation parameters

A typical 400 MW coal-fired power plant emits approximately 3 million tons/year of CO₂ (Katzner, 2007). The corresponding flue gas flow rate is approximately 32 million Sm³/day, which is the flow rate used in this study. The molar composition of flue gas used in this study is provided in Table 2 (Singh et al., 2003). As can be seen, the CO₂ concentration in flue gas is low which results in a low partial pressure. These are unfavorable conditions for the absorption of CO₂ into the absorbent. Coal combustion occurs in the presence of excess air, leading to the presence of unused oxygen in flue gas. Oxygen reacts with the alkanolamines and results in absorbent degradation problems. A single absorber – stripper train configuration results in an impractically large equipment design on account of the large flue gas, absorbent and stripping vapor flow

Table 2
Composition of flue gas.

Component	Value (mol%)
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

rates to be handled. Hence, we use a 3 absorber – stripper train configuration for this study. We examine the performance of three alkanolamines; MEA – a fast reacting primary amine with high heat of regeneration, DEA – a slow reacting secondary amine with a low heat of regeneration and DGA – another fast reacting primary amine with a high heat of reaction but lower vapor pressure than MEA. In order to simulate the amine absorption process under conditions similar to those maintained in commercial practice, we introduce the following constraints and guides.

1. At least 90% of the entering CO₂ must be captured and compressed (NETL, 2012).
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 mol-CO₂/mol-amine (Kohl and Nielsen, 1997).

Corrosion problems in carbon steel equipment with alkanolamines, especially at high concentrations and high solution loadings are quite well known. It is possible to achieve significantly higher CO₂ loadings by switching to stainless steel equipment; however it comes at a significantly higher cost. Stainless steel is 2–3× more expensive than carbon steel. While it can be argued that the higher cost maybe justified by lowered parasitic power losses, carbon steel equipment represents current technology that has been found to be reliable and economically feasible. This forms the scientific rationale behind the constraints imposed above.

MEA and DGA react with CO₂ much faster than DEA; thus requiring fewer trays/stages in the absorber column. In this study, we used absorber columns with 2 ideal stages for all MEA and DGA simulations and 10 ideal stages for all DEA cases. Various simulation parameters used in this study are compiled in Table 3. Heat exchangers are important pieces of equipment at amine absorption facilities. Two of the most critical heat exchangers in the amine absorption flowsheet shown in Fig. 1 are the lean/rich amine heat exchanger and the reboiler. Two key parameters which decide the sizing of the heat exchanger units are the minimum

Table 3
System parameters used in ProMax simulations.

Simulation parameters	
Parameter	Value
Absorber flooding fraction	0.8
Tray type	Sieve trays
Absorber/stripper tray spacing	0.61 m
Absorber/stripper weir height	0.076 m
Stripper flooding fraction	0.8
Condenser temperature	30 °C
Absorber # of trays (MEA, DGA)	2
Absorber # of trays (DEA)	10
Stripper # of trays (MEA, DEA, DGA)	10
Reboiler steam	413 kPa, 145 °C
Lean/rich heat exchanger	5 °C
Minimum approach temperature	

Table 4
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)			
	Absorbent	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%
150		218.2	137.1	89.1	32.4	34.3	34.0	215.1	104.7	99.9
175		187.2	149.2	96.2	27.6	30.3	31.7	181.3	102.2	91.9
200		190.3	168.9	101.9	23.5	28.1	28.9	161.1	99.5	89.8
250		189.2	187.8	111.6	16.2	22.3	24.3	141.0	97.0	87.1
300		197.6	203.5	119.1	10.1	17.5	20.8	131.3	94.8	84.7

Table 5
Parameters for evaluation of equivalent work of reboiler steam.

Parameters for evaluation of equivalent work			
Parameter	Inlet to LP turbine	Steam to reboiler	Steam to downstream condenser
Steam pressure (kPa)	630	413.7	5.52
Steam temperature (°C)	284.6	144	34.5
Mass enthalpy (kJ/kg)	3029.4	2928.0	2236.2
Mass entropy (kJ/kg·°C)	7.3	7.3	7.3
Contribution of LP turbine to plant output		45%	
Overall efficiency of turbine generator system		70%	

approach temperature and the energy duty (quantity of heat transferred). Both these parameters for the two heat exchangers for the different simulation cases are tabulated in Table 4. In this entire study, we have chosen against varying the reboiler steam pressure and temperature in the interest of simplicity. To account for the effect of this selection, we have summarized the reboiler approach temperatures in Table 4. For the same energy duty, a heat exchanger with a lower approach temperature will require a greater surface area to achieve the necessary heat transfer.

6. Evaluation of parasitic power loss

At a power plant retrofitted with amine absorption technology for carbon capture, there are several energy sinks that lead to a reduction in the net power output. This includes plant auxiliaries like the flue gas blowers and pumps, the compressors and energy consumed by the reboiler. The cumulative loss in plant output due to these energy sinks is termed as the parasitic power loss. Energy consumption by the reboiler is typically reported as energy consumed per unit mass of CO₂ separated. However, at an amine absorption facility; energy is provided to the reboiler unit by steam. In order to estimate the parasitic power loss at the plant, it is required to calculate the equivalent power generation capacity of the consumed steam. The parameters used to evaluate this are compiled in Table 5 (Bartley et al., 2007; Spliethoff, 2010).

Assuming the expansion of steam through the turbine to isentropic, the amount of steam required to generate one unit of electricity depends only on the inlet and outlet conditions of steam and the efficiency of the turbine. The relation between these is described in Eq. (7) where the units of the steam flow rate are kg/kWh, η_t is the turbine efficiency, H_1 is the mass enthalpy of the steam at its inlet pressure and temperature into the low pressure (LP) turbine and H_{2S} is the mass enthalpy at exit pressure and the entropy corresponding to the inlet conditions (Ganapathy, 1994).

$$\text{Stream rate} = \frac{154.81}{\eta_t(H_1 - H_{2S})} \quad (7)$$

It is known that the low pressure (LP) turbine contributes around 45% of the electric output of the power plant (Bartley et al., 2007). This allows for the calculation of the steam circulation rate to the low pressure (LP) turbine which is 324.4 kg/s. With the plant setup

remaining unaltered, this is the maximum amount of steam supply available to the reboiler. For each simulation case, the steam flow rate to the reboiler is known. This allows for evaluation of the power loss due to the reboiler steam consumption and thus, its contribution to the parasitic power loss.

7. Results and discussion

7.1. Effect of stripper operating conditions

7.1.1. Effect on total energy consumption for CO₂ separation

As discussed in Section 2, energy provided to the reboiler is employed to heat the rich amine solution entering the stripper, to drive the amine–CO₂ desorption reaction and to vaporize water to generate stripping vapor (steam). A conventional stripper column is operated between 150 kPa and 200 kPa, at a temperature of around 105–110 °C depending on the choice of absorbent. As the stripper pressure is increased, so does the operating temperature. As the stripper operating temperature increases, (p_{H_2O}/p_{CO_2}); the ratio of the partial pressure of water to that of CO₂ in equilibrium with the amine solution decreases and results in a decreased stripping vapor (steam) requirement. Since, in this case the stripping vapor is steam; an increased stripper pressure results in a decrease in the total reboiler duty. Fig. 3 shows the reboiler energy

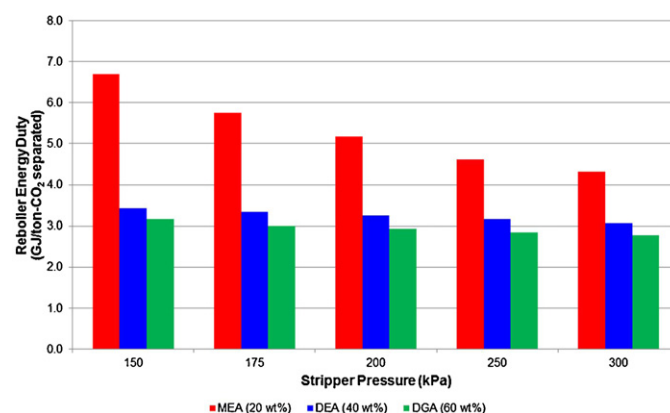


Fig. 3. Effect of change in stripper pressure on reboiler duty.

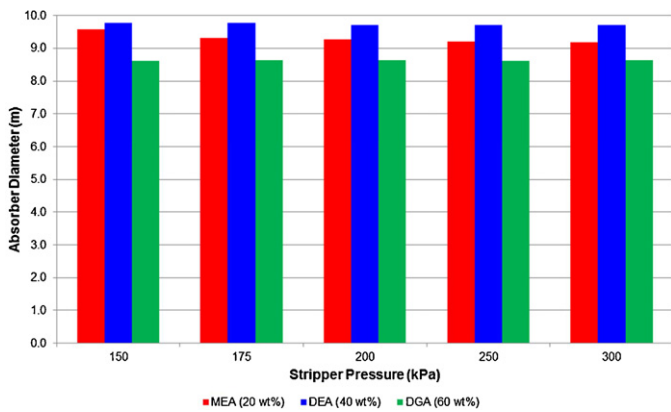


Fig. 4. Effect of change in stripper pressure on absorber column size.

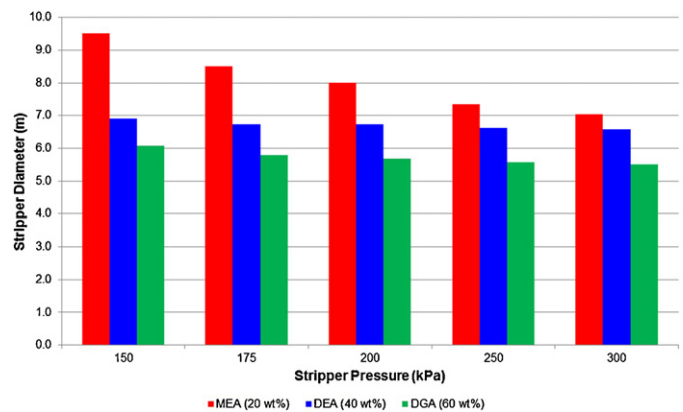


Fig. 5. Effect of change in stripper pressure on stripper column size.

duty for separation of CO₂ for 20 wt% MEA, 40 wt% DEA and 60 wt% DGA. The most noticeable feature of the plot is the significantly higher reboiler duty of MEA throughout, but especially under conventional (150 kPa) stripper operating conditions. At a stripper pressure of 150 kPa; MEA requires almost 6.75 GJ/ton-CO₂ whereas, DEA and DGA require only about 3.25 and 3.1 GJ/ton-CO₂ respectively. Where DGA has an advantage over DEA is in that DGA solutions can be used at a significantly higher concentration without serious corrosion or solvent loss problems. Thus, at almost all the stripper pressures studied in this work; 60 wt% DGA has a reboiler duty which is almost 10% lower than the 40 wt% DEA.

The reboiler duty values for MEA in published literature when the CO₂ concentration of flue gas is around 15% lie between 3.57 and 4.01 GJ/ton-CO₂. The reboiler duty value we report for MEA with the 150 kPa stripper is around 6.5 GJ/ton-CO₂ which is close to twice the previously reported value. To understand this, we must refer back to Section 5 where we discussed that rich amine loading must be maintained below 0.4 mol-CO₂/mol-amine to avoid excessive corrosion. The carbamate formed by the reaction between MEA and CO₂ is reasonably stable. In order to constrain the amine loadings, higher amine circulation and steam flow rate must be maintained. With other parameters unchanged, an increase in the amine circulation rate or the reboiler steam flow results in an increased CO₂ pickup in the absorber and consequently a higher rich amine loading. Consequently, significantly larger amounts of energy must be provided to the reboiler to maintain lower rich amine loadings. The reboiler duty for MEA and DGA when the rich amine loading is constrained to 0.4 mol-CO₂/mol-amine is similar to the value of 5.6 GJ/ton-CO₂, reported elsewhere in published literature (Salkuyeh and Mofarahi, 2012). As the stripper pressure is increased from 150 kPa to 300 kPa; the energy duty for MEA reduces from 6.5 GJ/ton-CO₂ to 4.25 GJ/ton-CO₂, DGA reduces from 3.1 GJ/ton-CO₂ to 2.8 GJ/ton-CO₂ and DEA reduces from 3.25 GJ/ton-CO₂ to 3 GJ/ton-CO₂. This decrease is primarily a result of reduced stripping vapor (steam) requirement. A detailed analysis of the contributions of the 3 constituent processes of absorbent stripping is given in Section 7.2.

7.1.2. Effect on absorber and stripper sizing

In the design of an amine absorption unit, energy consumption is the most important contributor to the operating costs. One of the significant contributions to the capital investment is the absorber and stripper column size, which is also strongly influenced by amine selection. ProMax[®] uses the degree of fractional flooding as an input parameter to calculate column diameter. As disclosed in Table 3, all columns in this study are set to operate at 80% flooding. Fig. 4 shows the effect of amine selection and stripper operating pressure on absorber diameter. It can be seen that the

absorber diameter varies only marginally depending on the amine selected and the stripper pressure. This happens because the most important factor affecting column diameters is the vapor flow-rate. Since, in all cases flue gas inflow rate of flue gas remains constant, there's only a small change in the absorber diameter.

Fig. 5 shows the effect of the stripper pressure and amine choice on the stripper diameter. Unlike the absorber; the stripper column shows a strong dependence on both the choice of amines and the stripper pressure. As before, this can be explained by observing the vapor inflow rates to the stripper. As the stripper pressure increases, the stripping vapor (steam) requirement reduces. Due to an increase in the stripper pressure, there is also a reduction in the volume of the stripping vapor (steam) and the CO₂ flowing inside the column. These two factors combine to reduce the overall vapor flow rate in the stripper column which consequently reduces the stripper size. Since MEA has the highest stripping vapor (steam) requirement, it requires the largest stripper column for all amines compared, followed by DEA and DGA.

7.2. Contribution of various constituent processes to reboiler heat duty

Fig. 6 shows the contribution of the energy consumption for generating stripping vapor (steam) to the reboiler heat duty. Energy supplied to the reboiler is used for heating the amine solution, driving the CO₂ desorption reaction and to generate stripping vapor (steam). Using the equations developed in Section 4, we calculate the contribution of each of these components to the reboiler heat duty. It can be seen from Fig. 6 that stripping vapor (steam) consumption decreases as the stripper pressure is increased from 150 kPa to 300 kPa. This results from an increase in

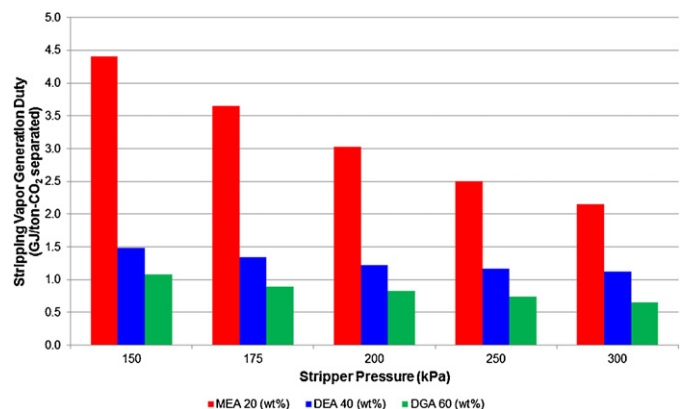


Fig. 6. Contribution of stripping vapor energy consumption to reboiler heat duty.

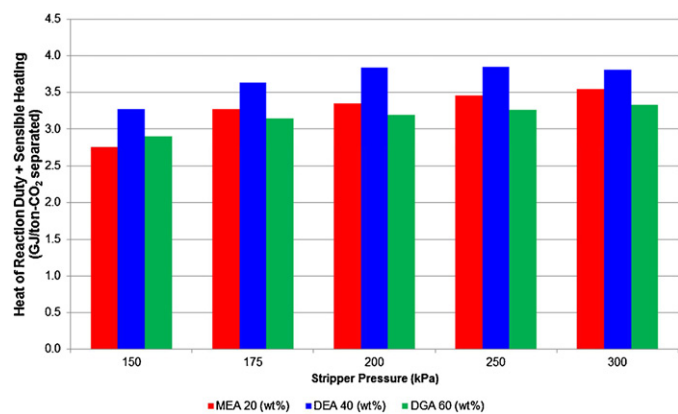


Fig. 7. Combined contribution of sensible heating and heat of reaction to reboiler heat duty.

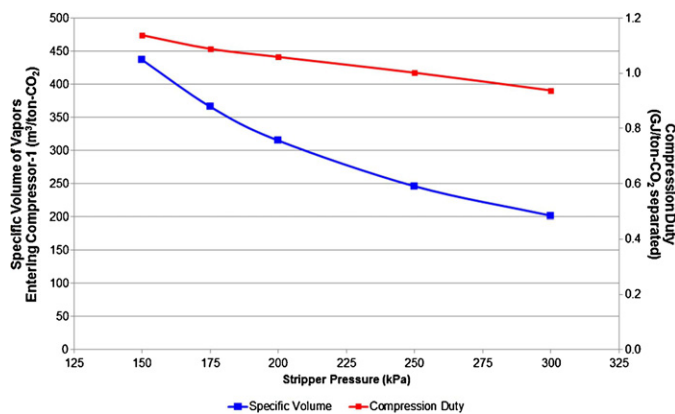


Fig. 8. Effect of stripper pressure on compression duty.

the partial pressure of CO₂ and H₂O in equilibrium with the amine solution in the stripper unit. As the CO₂ partial pressure increases, the stripping vapor (steam) requirement reduces.

From Fig. 6, it can be clearly identified that MEA has a significantly higher stripping vapor (steam) requirement than DEA and DGA, which suggests a lower equilibrium CO₂ partial pressure. We hypothesize that this is a result of the relatively stable carbamate intermediate formed by the reaction between MEA and CO₂. One of the most noticeable features of Fig. 6 is its close similarity to trend in Fig. 3, which presents the effect of stripper pressure on the reboiler duty. This suggests that the stripping vapor (steam) energy requirement is more sensitive to a change in stripper pressure than sensible heating and heat of reaction. Fig. 7 is a plot of the lumped contribution of sensible heating and heat of reaction to reboiler heat duty. As seen in the plot, the combined contribution of these two processes is near constant and is only weakly sensitive to changes in stripper pressure. The remarkable steadiness in the plot of the lumped contributions of sensible heating and heat of reaction is partly a result of the method we have used to evaluate the contributions, where the heat of absorption is the difference between the reboiler duty and the sum of sensible heating and condenser duty. The dependence of stripping vapor (steam) requirement on stripper pressure can be explained by analyzing the ratio of partial pressure of water to that of CO₂ in equilibrium with the rich amine solution. Using a modified form of the Clausius–Clayperon equation, the ratio of the partial pressure of water to CO₂ can be expressed as shown below (Oexmann and Kather, 2010).

$$\frac{p_{\text{H}_2\text{O}}^{\text{sat}}}{p_{\text{CO}_2}^*} = \frac{p_{\text{H}_2\text{O},T_{\text{ref}}}^{\text{sat}}}{p_{\text{CO}_2,T_{\text{ref}}}^*} \exp\left(\left[\frac{T - T_{\text{ref}}}{RTT_{\text{ref}}}\right] [\Delta H_{\text{vap,H}_2\text{O}} - \Delta H_{\text{abs,CO}_2}]\right)$$

As the stripper pressure increases, so does the operating temperature which results in a decrease in this ratio and a resultant reduction in the stripping vapor (steam) requirement.

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%
150	108.8	98.5	97	114.1	110.9	110.9
175	113	102.1	98.8	118.8	114.9	113.3
200	116.8	104.4	101.4	123.0	117.1	116.0
250	122.5	109.7	105.8	130.0	122.8	120.7
300	127.5	114.1	108.9	136.0	127.6	124.1

7.3. Effect of stripper pressure on compression duty

CO₂ capture is the separation of CO₂ from flue gas followed by compression to a pipeline pressure of 16 MPa. Compression duty accounts for nearly 10% of the total energy consumption. An increase in the stripper pressure results in a reduction in the reboiler steam requirement. However, it also reflects in the compression duty because at higher stripper pressures, separated CO₂ is at a higher pressure and consequently at a lower volume. Thus, the energy duty, especially of the first two stages of the compression train is greatly reduced. From Fig. 8, we can observe that as the stripper pressure is increased from 150 kPa to 300 kPa; compression duty falls by nearly 60%. This suggests that a higher stripper operating pressure has advantages in reducing the energy duty both during separation and compression of CO₂.

7.4. Effect of stripper pressure on solvent losses

As shown in Section 7.1.1, the reboiler duty reduces as the stripper pressure is increased which results in energy savings at the carbon capture plant. There is a reduction in the compression duty because the CO₂ is delivered to the compression train at a higher pressure. However, as explained previously; an increase in the stripper pressure results in an elevated stripper temperature. Most alkanolamines undergo degradation at a temperature of around 135 °C, and thus it may be expected that as the stripper pressure increases, solvent losses due to the thermal degradation of absorbents increase as well. Table 6 is the compilation of the average stripper temperature (including the reboiler) and the reboiler temperature itself. It can be clearly seen from this data that the average stripper temperature approaches 130 °C in case of MEA and 115 °C and 110 °C for DEA and DGA respectively. The reboiler itself operates at a much higher temperature – close to 135 °C for MEA and around 125 °C for DEA and DGA.

Scientific views in published literature about the rate of solvent loss due to thermal degradation are divided. Davis reports that MEA

Table 7
Solvent losses through absorber and stripper overheads.

Simulation case stripper pressure (kPa)	Solvent losses absorber and stripper overheads (kg/day)		
Absorbent	MEA 20 wt%	DEA 40 wt%	DGA 60 wt%
150	4252.4	30.9	2798.1
175	4594.5	30.9	2493.9
200	4978.4	24.7	2504.9
250	5369.7	24.7	2541.7
300	5893.2	24.7	2501.2

(30 wt%) exhibits a decrease in activity of around 37% due thermal degradation, when maintained at 135 °C at a CO₂ loading of 0.4 for 4 weeks. Under the same conditions, DGA (42 wt%) activity reduces by only by 9% (Davis, 2009). On the contrary, Stewart and Lanning recommend that the stripper be operated below a bulk temperature of 127 °C and the reboiler below 175 °C to avoid excessive thermal degradation (Stewart and Lanning, 1994). Thus, it is difficult to conclusively predict the dependence of solvent loss due to thermal degradation on stripper pressure. However, it may be expected that higher stripper pressures lead to greater losses.

Table 7 is a compilation of the amounts of solvent escaping with the absorber and stripper overheads for the various simulation cases. In the current work, we do not consider a water wash section in any of the simulation cases; meaning that these solvent losses are much higher than found in commercial practice. However, this data indicates that solvent loss problems are quite significant for MEA as well as DGA than for DEA. Based on the information in Tables 6 and 7, DEA and DGA appear to have significant advantages in system performance than MEA.

7.5. Effect of carbon capture on electrical output

A typical sub-critical pulverized coal-fired power plant generates electricity in 3 stages. High pressure steam generated in the boiler first passes through high pressure (HP) turbine, followed by an intermediate pressure (IP) turbine and lastly, a series of low pressure (LP) turbines. The low pressure (LP) turbines are crucial to the power generation since they produce around 45% of the total electricity output (Bartley et al., 2007). Reboilers for conventional amine absorption process require steam at 415 kPa which must be drawn from the low pressure (LP) turbines. This results in a reduction in the electricity output of the low pressure (LP) turbines and the overall output of the power plant. Fig. 9 is a plot of the parasitic power loss at the reference power plant represented as a % of its rated capacity of 400 MW. As can be seen, for all cases considered in this study; the parasitic power loss is a significant fraction of the

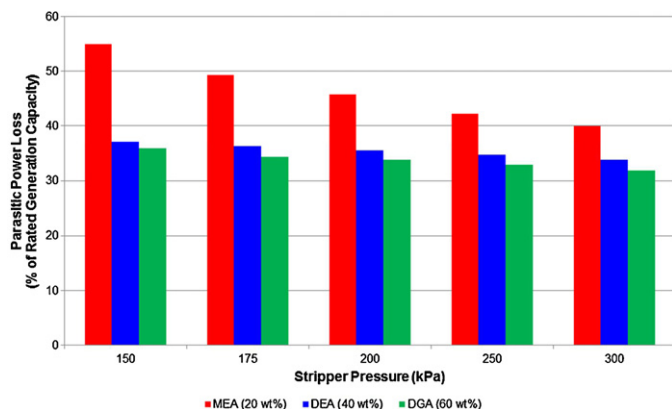


Fig. 9. Parasitic power loss at reference power plant due to carbon capture.

plant's output. As the stripper pressure increases, there is a reduction of approximately 15% points in the parasitic power loss when CO₂ capture takes place using MEA. For DEA and DGA, the decrease is around 4.5% points. We note that increasing the stripper pressure makes MEA a more competitive absorbent with DEA and DGA.

8. Conclusions

Based on this study, we conclude that there are significant advantages to operating the stripper columns of amine absorption units for carbon capture at higher pressures. These advantages are tied up with some disadvantages and operational challenges. We comment on these below:

- An increase in the stripper pressure from 150 kPa to 300 kPa results in a drop in the parasitic power loss for MEA from 54% to 38.7%. For DEA, the parasitic power loss reduces from 38.3% to 33.6% while for DGA it reduces from 36% to 31.6%.
- The absorber sizing is not sensitive to the stripper pressure. However, the stripper column diameter decreases significantly with an increase in pressure and is sensitive to the choice of absorbent. Between 150 kPa and 300 kPa, the MEA stripper diameter reduces by 25%, the DEA stripper by 5% and the DGA stripper by 9%.
- A dissection of the reboiler heat duty reveals that sensible heating and heat of reaction are almost independent of the stripper pressure. The energy required for generation of stripping vapor, one of the largest contributors to the reboiler heat duty is sensitive to stripper pressure. When the stripper pressure is increased from 150 kPa to 300 kPa, it decreases by 50% for MEA, 25% for DEA and 40% for DGA.
- Increasing the stripper operating pressure and keeping the reboiler steam conditions unchanged may result in a requirement for a larger reboiler. The two factors affecting the size of the reboiler are its energy duty and the minimum end approach temperature.
- Increasing the stripper pressure may result in increased absorbent losses due to thermal degradation. It is however, difficult to conclusively predict the disadvantage of increasing the average stripper temperature by 15–25 °C. Opinions in published literature offer varying recommendations.
- On the basis of a comparison of the parasitic power loss, it can be concluded that DEA and DGA have better performance than MEA. Both DEA and DGA result in stripper columns smaller than MEA, which results in a reduced capital expense. On the other hand, of the three amines compared; MEA has the lowest cost, followed by DEA and lastly, DGA.

Based on the first part of our work, we remark that operating the stripper at higher pressures with DEA and DGA as the absorbents has advantages over using MEA in the conventional stripper configurations. Since this study takes into account only the technical aspects of amine absorption, it is difficult to comment upon the economics of using the various amines and the novel stripper configurations. However, the results presented here strongly suggest that further investigation must be conducted to understand the pros and cons of high pressure stripper operation.

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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.2013.01.050>.

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