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Energy performance of advanced reboiled and flash stripper configurations for CO_2 capture using monoethanolamine

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Abstract: CO_2 capture by absorption using amine solvents has the potential to significantly reduce the CO_2 emissions from fossil-fuel power plants. One of the major costs of this technology is the energy required for solvent regeneration. Complex process configurations claim to have promising potential to reduce the energy required for solvent regeneration. In this work, the effect of flow-sheet complexity is explored by studying two advanced stripping flow-sheets, an advanced flash stripper and an advanced reboiled stripper. Both advanced configurations recover the stripping steam heat by means of a heat integration comprised of cold and warm rich solvent bypasses. The advanced configurations are simulated and optimised in Aspen Plus® V.8.4 using 7 m monoethanolamine (MEA) with lean loading from 0.15 to 0.38 (mol CO_2/mol MEA). The rich loading associated with each lean loading is determined by simulating the absorber providing 90% capture from flue gas with 4 mole%

CO_2, typical of a natural gas-fired turbine. The results are compared to a simple stripper in terms of total equivalent work. Both the advanced reboiled stripper and the advanced flash stripper require 12% less equivalent work than a simple stripper. The associated cold rich and warm rich bypasses for the optimum cases are respectively 20% and 50% for the advanced reboiled stripper and 15% and 35% for the advanced flash stripper.

1. Introduction

The implementation of post-combustion CO_2 capture by absorption/desorption with a chemical solvent such as monoethanolamine (MEA) is the most promising process for near term deployment according to IEA[1]. However, one major disadvantage of this process is its large energy requirement. CO_2 capture by chemical absorption/desorption is based on a reversible reaction between CO_2 and a suitable solvent. Regenerating the solvent in a stripper column accounts for more than 60% of the total energy required in a post-combustion CO_2 capture unit[2]. This energy is usually provided as low pressure steam from the power plant steam cycle[3]. The conventional solvent regeneration technology is a simple stripper, with a significant loss of exergy as water is condensed from the CO_2[4,5]. Studies have shown that the addition of an amine-based CO_2 capture plant to a natural gas combined cycle power plant leads to a net power plant efficiency penalty of 7-11%[6].

Previously, a number of research studies have explored various alternative process configurations and optimisation of CO_2 capture processes[5,7-19]. One of the best configurations proposed earlier with PZ will be evaluated in this paper with MEA. The potential for energy saving therefore exists and design and operation of energy efficient amine based CO_2 capture will have a substantial effect on the overall plant energy consumption and operating costs.

Fundamental research has shown the benefit of reduced driving forces in chemical processes. In a chemical process, driving forces for heat transfer (temperature), mass transfer, and chemical reaction[4,5] generally result in thermodynamic irreversibility, by which the process consumes more energy than ideally required[5,11]. However, a chemical process with reasonable capital cost must have finite driving forces to expend some thermodynamic availability (exergy) and consume more energy compared to an ideal process.

Although it is not possible to have a thermodynamically reversible process because of excessive equipment sizes, by proper design and operation it is possible to minimise the system exergy losses [11,17]. Reducing excess driving forces will induce energy savings to the process.

Complex configurations had previously been proposed to improve the energy efficiency of stripping columns. For example, Leites et al.[11] proposed several complex configurations that incorporate a combination of stripper column inter-heating and split-flow and a multi-feed arrangement at varying temperature. The original idea of the rich solvent split flow was suggested and patented by Johnson and Eisenberg[20,21]. They modified the stripping process by splitting the rich solvent into two streams downstream of the absorber. One is passed without further heating to the top of the stripper column while the other is passed to an intermediate point in the stripper column after being pre-heated in the lean/rich cross heat exchanger. Their suggested scheme however showed some energy deficiency where a portion of the rich solvent enters the column top with no prior preheating. Preheating the rich solvent to a temperature close to the stripper operating temperature is crucial to avoid the condensation of vapour water that would otherwise take place at the condenser, which causes an increase in the energy requirement[17].

Van Wagener and Rochelle[22] evaluated the benefits of increasing process reversibility by introducing more complexity to the system using multi-stage flash and inter-heated stripping. They showed using the inter-heated configuration improves the performance of the stripper column by approximately 8% based on total equivalent work.

Furthermore, their study confirmed that increasing pressure will typically yield better performance in terms of energy consumption due to more reversible operation. Madan[8] showed stripper columns with various complex configurations perform better than a conventional one. His results showed that an advanced flash stripper with rich solvent split flow entering the column at different temperature levels offers the best performance. Later, Lin et al.[7] developed advanced configurations incorporating thermal integration based on excess regeneration heat and rich solvent split flow and studied the improvement brought by these modifications for 8 m piperazine (PZ) and 9 m MEA. They showed that the proposed configurations provide 10% less equivalent work for 8 m PZ and 6% for 9 m MEA when compared to a simple stripper. The advanced flash stripper proposed by Lin was performed as expected when tested with 5 m PZ in a 0.1 MW pilot plant[23].

Recent work in industry has shown interest in the development of more complex configurations with higher efficiency. MHI examined more efficient heat recovery from the stripper column and studied an interheated stripper column[24]. The MHI configuration attained a more reversible process in the stripper column by recycling a portion of the heat available in the lean solvent back into the column. Previously, Barchas and Davis[25] also claimed substantial saving in steam requirement for solvent regeneration when the total rich solvent is preheated to the stripper temperature before entering the column, with only a minor increase in equipment costs. However, the temperature of rich solvent before and after the proposed modification was not disclosed.

The present study aims to evaluate energy improvements offered by the two advanced configurations proposed earlier by Lin et al.[7, 30] for use with 8 m PZ at 12% CO_2. The present study uses 7 m MEA (approximately 30 wt %) and NGCC conditions (approximately 4 mole% CO_2), and quantifies the optimum operating conditions for lean loading from 0.15 to 0.38 (mol CO_2/mol MEA). The advanced reboiled stripper uses a conventional reboiler to provide the heat required for solvent regeneration. The advanced flash stripper replaces the reboiler with a convective steam heater. Both configurations incorporate a system including cold and warm rich solvent bypass to recover heat from the product vapour and employ an additional cold rich bypass heat exchanger. Splitting the rich solvent into two streams before entering the stripper column further increases the complexity of the system. Process modelling of these two advanced configurations was performed using Aspen Plus V.8.4. The optimum fraction of the cold rich solvent and warm rich solvent bypasses was determined over a range of lean loading. The associated rich loading for each lean loading is obtained by simultaneously modelling the absorber column providing 90% capture from flue gas with 4 mole% CO_2, typical of a natural gas-fired turbine.

2. Methodology

The analysis started with the simulation of a standard CO_2 absorption/desorption process for a range of lean loading from 0.15 to 0.38 (mol CO_2/mol MEA) with a fixed flue gas flow rate and compositions, as a baseline for comparison against the advanced configurations. The flue gas used in the simulation represents the exhaust gases of natural gas combustion with typically 4% mole CO_2[26]. The standard process consists of a simple absorber column and simple stripper column with no process optimisation and designed for 90% CO_2 capture efficiency. The details of the column design and their specifications are provided in[27].

2.1. Process simulation

The standard 7 m MEA (7 mol MEA/kg water, about 30 wt. %) was chosen for this study As this solvent has long been the industry standard for removal of acid gases such as CO_2 due to its low cost per mole of amine, high heat of absorption, high absorption capacity, and high rates of reaction. This is the highly anticipated solvent to be used in the first generation of large-scale CO_2 capture plant.

The Aspen Plus V.8.4 RateSep model, a rigorous framework for the modelling of rate-based separations, was used to simulate the absorber and stripper. The model used in Aspen Plus for the thermodynamic properties is based on the work by Zhang et al.[28]. The model uses the asymmetric electrolyte Non-Random-Two-Liquid (e-NRTL) property method to describe the CO_2-MEA-H_2O chemistry in liquid phase, and the Redlich-Kwong (RK) equation of state for the vapour phase. The model has been validated by Zhang et al.[28] against experimental data available in open literature. The absorber model comprises both equilibrium and kinetic rate-based controlled reactions, while the stripper model comprises equilibrium rate-based controlled reactions. In the absorber column, the reactions that involve CO_2 were described with a kinetic model. The equilibrium reactions describing the solution chemistry of CO_2 absorption with MEA, which are integral components of the thermodynamic model, are expressed as[28]:

$$2H_2O \leftrightarrow H_3O^+ + OH^- \tag{1}$$

$$CO_2 + 2H_2O \leftrightarrow HCO_3^- + H_3O^+ \tag{2}$$

$$HCO_3^- + H_2O \leftrightarrow H_3O^+ + CO_3^{2-} \tag{3}$$

$$MEAH^+ + H_2O \leftrightarrow MEA + H_3O^+ \tag{4}$$

$$MEACOO^- + H_2O \leftrightarrow MEA + HCO_3^- \tag{5}$$

The formation of carbamate and bicarbonate are kinetically limited, and the forward and reverse reactions are expressed as follows[29]:

$$CO_2 + OH^- \rightarrow HCO_3^- \tag{6}$$

$$HCO_3^- \rightarrow CO_2 + OH^- \tag{7}$$

$$MEA + CO_2 + H_2O \rightarrow MEACOO^- + H_3O^+ \tag{8}$$

$$MEACOO^- + H_3O^+ \rightarrow MEA + CO_2 + H_2O \tag{9}$$

The reboiler section and the flash vessel used in the advanced flash stripper were modelled as equilibrium stages. Figure 1 shows the flow-sheet developed in Aspen Plus to simulate the base case with the simple stripper.

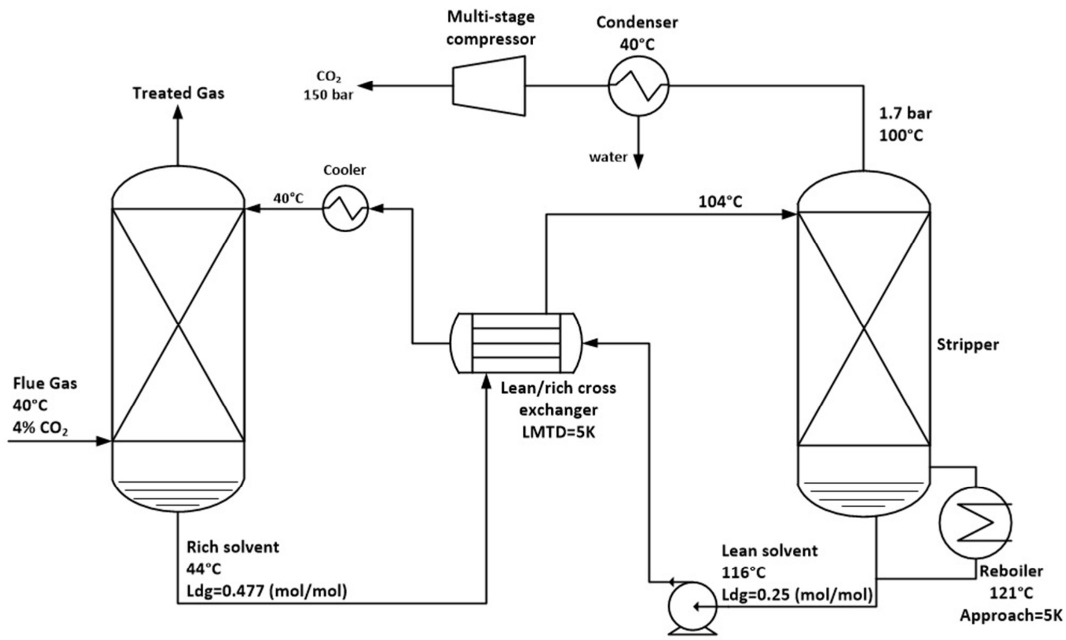

Figure 1. The CO_2 capture process with simple stripper for the solvent lean loading of 0.25 (mol CO_2/mol MEA)

2.2. Process specification

For the range of lean loading from 0.15 to 0.38 (mol CO2/mol MEA), the CO_2 capture process consisting of one absorber column and one stripper column in a closed loop arrangement, as shown in Figure 1, was simulated in Aspen RateSep^{TM}.

Packed columns were defined with Sulzer Mellapak 250Y structured packing. The column diameters were specified to give a 75% approach to flooding. The height of packing was specified as 20 m for both the stripper and absorber columns, resulting in a pinch for all cases. This excess packing height should provide an accurate estimate of the relative energy use, but will underestimate the actual energy requirement. For a given lean loading, the solvent flow rate was determined to provide 90% CO_2 removal rate with respect to the flue gas condition and the absorber packed column specified height. The liquid to gas ratio (L/G) and associated rich loading at the absorber discharge for the range of lean loading are presented in Table 1.

The stripper pressure was kept constant at 170 kPa (1.7 bar) resulting in variable solvent temperatures at the stripper bottom to achieve the desired lean loading. For each lean loading, the regenerated solvent temperature at the stripper discharge is also presented in Table 1.

Table 1. Predicted absorber and stripper performance (90% CO_2 removal, with 20 m Sulzer Mellapak 250Y structured packing, 7 m MEA fed to the absorber at 40°C, flue gas fed to the absorber at 40°C with 4 mole% CO_2, 170 kPa stripper P)

Lean loading (mol CO_2/mol MEA)	**L/G (kg/kg)**	**Rich loading (mol CO_2/mol MEA)**	**Lean solvent temperature at stripper discharge (°C)**
0.15	0.80	0.475	118.5
0.18	0.89	0.476	118
0.20	0.96	0.476	117.5
0.21	1.00	0.476	117
0.25	1.18	0.477	116
0.30	1.53	0.477	114
0.32	1.79	0.471	113
0.36	3.87	0.431	110
0.38	4.14	0.446	109

For all cases, the overall log mean temperature difference (LMTD) of the lean/rich cross heat exchanger, was specified as 5°C. The LMTD of the rich solvent bypass heat exchanger was set at 20°C. A 5°C hot side approach was specified on the steam reboiler, and a 5°C LMTD was specified for the convective steam heater. The process specifications used to simulate various flow sheets are summarised in Table 2.

Table 2. Process design specifications used in process simulations

Design specifications	Advanced reboiled stripper	Advanced flash stripper
Process simulation tool	Aspen Plus V8.4	
Thermodynamic model	e-NRTL-RK	
Packing type	Sulzer Mellapak 250Y	
Absorber column packed height (m)	20	
Stripper column packed height (m)	20	
Lean/rich cross heat exchanger LMTD (°C)	5	
Cold rich bypass heat exchanger LMTD (°C)	20	
Reboiler approach temperature (°C)	5	-
Steam heater LMTD (°C)	-	5
Stripper pressure (kPa)	170	

2.3. Process Evaluation

Equivalent work was used to evaluate the energy requirement of the advanced configurations at various lean loading. This result estimates the total electrical work penalty from the power plant by operating the stripper, compressors and pumps. Eq. (1) shows the three main contributors to the overall equivalent work: regeneration heat, compression work, and pump work.

$$W_{eq} = W_{heat} + W_{comp} + W_{pump} \tag{1}$$

The compression and pump work would draw electricity directly from the power plant output, therefore their respective work values are added directly into the equivalent work. The regeneration heat, on the other hand, would draw steam from the steam turbine of the power plant that would be otherwise expanded in low pressure steam turbines to generate electricity[22]. Oyenekan suggested the use of an equivalent work term to evaluate the heat duty of a similar basis as the pump and compression work[4]. The pump work includes only the required head at the efficiency of the pump, e.g. 75%, to move and circulate the solvent from absorber to the pressure of stripper and vice versa. The flue gas blower work is excluded from this calculation, assuming the flue gas pressure at the absorber inlet is sufficiently high to overcome the passage and packing pressure drops. The Aspen Plus pump block is used to calculate the pump work. The compression work is the work to compress the product vapour from the stripper pressure (P_{in}), to the storage pressure of 15 MPa (150 bar), that can be calculated using Eq. (2)[30].

$$W_{comp} = -3.48\ln(P_{in}) + 14.85, \qquad 1 < P_{in}\,(bar) < 20 \tag{2}$$

The equivalent electrical penalty associated with solvent regeneration, called the heat equivalent work, is calculated using the Carnot efficiency method, as represented by Eq. (3).

$$W_{heat} = \eta_{turbine}\left(\frac{T_{reb} + \Delta T - T_{sink}}{T_{reb} + \Delta T}\right) Q_{reb} \tag{3}$$

The assumptions made for Eq. (2) include a compression ratio of 2 or less for each compression stage, a compressor polytropic efficiency of 86%, inter-stage cooling to 40°C with knocked out water between stages with zero pressure drop[30]. Assumptions made for Eq.

(3) include a 90% efficiency to account for non-ideal expansion in steam turbines[31], an approach temperature of 5°C for the steam side, and a sink temperature of 40°C.

3. Proposed configurations

This study evaluates and quantifies the energy savings that advanced stripper configurations will offer when using 7 m MEA. The proposed configurations include a reboiler-based stripper and flash-based stripper. In both configurations, the stripping steam heat is recovered by incorporating a heat integration system comprised of cold and warm rich solvent bypasses. This system resulted in employing an additional heat exchanger, called the cold rich bypass exchanger, to the advanced configuration flowsheets. The details of each process configuration are described in the following paragraphs and their associated flow sheets are shown in Figures 1-3.

3.1. Simple stripper

The base case of this study uses a simple stripper as shown in Figure 1. The rich solvent enters the stripper at the top after being pre-heated in the lean/rich cross heat exchanger by the hot lean solvent leaving the stripper column at the bottom. The heat exchanger was modelled with rich side flashing and 5°C LMTD rather than using a back pressure valve with flashing at the top of the stripper. In the stripper column, the energy required for the solvent regeneration is provided by the reboiler. The regenerated lean solvent returns to the absorber column through the lean/rich cross heat exchanger. The product vapour leaves the column from the top and after being cooled to 40°C in the overhead condenser is fed to a multi-stage compressor train. The product vapour cooling at the overhead condenser is associated with a loss of latent heat of its excess water vapour.

3.2. Advanced reboiled stripper

Figure 2 shows the advanced reboiled stripper with cold rich bypass and warm rich bypass. This configuration is an advanced version of a simple stripper that includes a heat recovery of the latent heat available in the product vapour by the cold rich solvent. In this configuration, the cold rich solvent splits into two streams downstream the absorber column. One split bypasses the lean/rich cross heat exchanger and enters the cold rich bypass heat exchanger, to partially recover the latent heat available in the product vapour exiting the system. The product vapour usually contains more than 50% water vapour.

The second stream enters the lean/rich cross heat exchanger and recovers the heat available in the lean solvent leaving the stripper column. Subsequently, a portion of this stream splits further into two streams, and one stream is drawn from the cross heat exchanger at its bubble point (bp) and mixed with the preheated, bypassed rich solvent before entering the stripper column at the top. The remainder of the warm rich solvent heats further up in the cross heat exchanger before entering the stripper column in the middle. The temperature of this stream is usually higher than the bubble point. Using this arrangement is expected to be more efficient than the conventional practice since it avoids inevitable flashing of the rich solvent at the top of the stripper column due to recovering all the heat available in the hot lean solvent at once at the lean/rich cross exchanger. Using the additional heat exchanger will therefore balance the heat transfer more efficiently and reversibly by making smaller heat transfer driving force between the rich solvent and the product vapour at the top of the column[7].

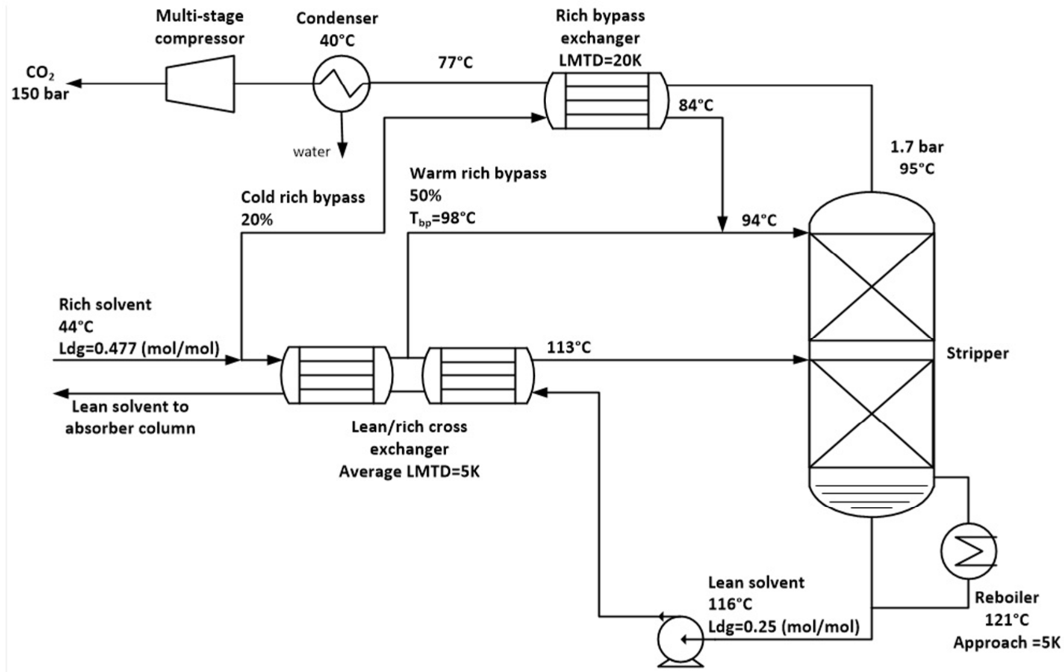

Figure 2. The advanced reboiled stripper for the lean solvent loading of 0.25 mol CO_2/mol MEA

3.3. Advanced Flash stripper

Figure 3 shows the flowsheet of the advanced flash stripper. This configuration is similar to that of the advanced reboiled stripper, except the reboiler is replaced by a convective steam heater and a flash in the sump of the stripper. In this configuration, one split of the rich solvent downstream of the absorber column bypasses the lean/rich cross heat exchanger and preheats by the hot product vapour exiting the stripper column at the top. The rest of the rich solvent preheats in the cross heat exchanger, where a portion of it, at its bubble point, is drawn to mix with the preheated cold rich bypass, prior entering the stripper at the top. The rest of the boiling rich solvent is further heated in the cross heat exchanger before entering the steam heater. The hot flashing rich solvent is then fed into a flash vessel from the bottom where the flashed vapour counter-currently contacts the rich solvent. Since the convective steam heater has less solvent hold-up and residence time at elevated temperature, compared to a reboiler, it will minimise the solvent thermal degradation[7,9].

With respect to process specifications described earlier, the proportion of the cold rich and warm rich solvent flow rates at various lean loadings is subject to optimisation to quantify the highest energy savings offered by each advanced configuration.

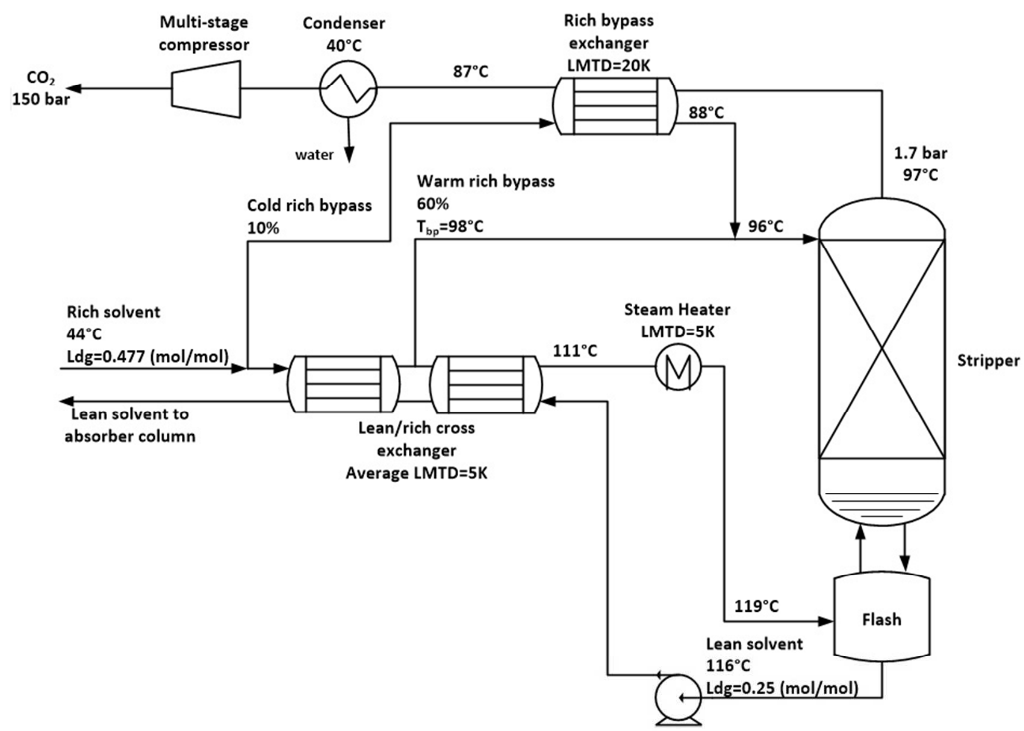

Figure 3. The advanced flash stripper for the lean solvent loading of 0.25 (mol CO_2/mol MEA)

4. Results and discussion

4.1. Total equivalent work

Total equivalent work is an appropriate indicator to evaluate and compare the advanced configurations against each other and the base case. The calculated overall equivalent work was normalised by the moles of CO_2 removed. For a given lean loading, the optimum equivalent work was quantified by varying the cold and warm rich bypass flow rates. The optimum flow rates are given as their fraction of the total rich solvent flow for a given lean loading. Also, for each advanced configuration, there was an optimum lean loading at which the reduction in the total equivalent work is highest when compared to their respective base case. The total equivalent work of the simple stripper for the range of lean loading from 0.15 to 0.38 is summarised in Table 2. These values are the baseline values against which the advanced configurations are compared. For advanced reboiled and advanced flash configurations, the results of optimum cases with their cold and warm rich bypass flow fractions are summarised in Table 3 and 4, respectively. For each lean loading, the reported optimum cold rich and warm rich bypass fractions are the relative proportion of their flow rates to the total rich solvent flow rate in percentage.

Table 2. Performance of the simple stripper for 90% capture for various lean loading

Lean Loading (mol CO_2/mol MEA)	**Regeneration heat duty (kJ/mol CO_2)**	W_{heat} **(kJ/mol CO_2)**	W_{comp} **(kJ/mol CO_2)**	W_{eq} **(kJ/mol CO_2)**
0.15	183.8	34.8	13.0	48.2
0.18	169.6	32	13.0	45.4
0.20	166.6	31.2	13.0	44.7
0.21	166	31	13.0	44.6
0.25	164.4	30.4	13.0	44.1
0.30	164	29.7	13.0	43.6
0.32	167.2	30	13.0	44.0
0.36	205.3	35.8	13.0	51.1

0.38	197.3	33.8	13.0	49.3

Table 3. Optimum results for the advanced reboiled stripper for 90% capture for various lean loading

Lean Loading (mol CO_2/mol MEA)	Cold rich bypass (%)	Warm rich bypass (%)	Regeneration specific heat duty (kJ/mol CO_2)	W_{heat} (kJ/mol CO_2)	W_{comp} (kJ/mol CO_2)	W_{eq} (kJ/mol CO_2)
0.15	35	50	170.6	32.3	13.0	45.7
0.18	30	55	148.3	27.9	13.0	41.4
0.20	30	50	143.3	26.9	13.0	40.4
0.21	30	50	140.3	26.2	13.0	39.8
0.25	20	50	136.6	25.3	13.0	39.0
0.30	20	35	139.7	25.3	13.0	39.2
0.32	20	30	147.7	26.5	13.0	40.5
0.36	15	10	190.0	33.1	13.0	48.5
0.38	13	12	185.5	31.8	13.0	47.3

Table 4. Optimum results for the advanced flash stripper for 90% capture rate for various lean loading

Lean Loading (mol	Cold rich bypass	Warm rich bypass	Regeneration specific heat duty	W_{heat} (kJ/mol CO_2)	W_{comp} (kJ/mol CO_2)	W_{eq} (kJ/mol CO_2)

CO_2/mol MEA)	**(%)**	**(%)**	**(kJ/mol CO_2)**			
0.18	10	75	160.1	34.4	13.0	47.9
0.20	20	60	151.9	29.2	13.0	42.7
0.21	30	50	143.4	27.4	13.0	40.9
0.25	10	60	138.0	25.5	13.0	39.2
0.30	15	35	136.1	24.3	13.0	38.2
0.32	10	35	140.9	24.9	13.0	39.0
0.36	10	15	182.0	31.3	13.0	46.6
0.38	10	10	178.1	30.1	13.0	45.6

As shown in Table 4, the results for the advanced flash stripper at the lean loading of 0.15 (mol CO_2/mol MEA) could not be obtained because the optimum theoretically occurs when the total bypass exceeds 85% of the total rich solvent flow. This means that the total heat required for the solvent regeneration should be provided by the remaining rich solvent flow (i.e. less than 15% of the total rich solvent flow), resulting in a significant rise on the rich solvent temperature after the convective steam heater (i.e. more than 180°C). This temperature is excessive and would result in thermal degradation of the amine. In principle, for the convective steam heater, the highest acceptable operating temperature with respect to the solvent thermal degradation is 135-140°C, while, for the reboiler application this limit is 120-125°C. The calculated results show that the lean loading of 0.18 (mol CO_2/mol MEA) is the limit for the advanced flash stripper, as at this loading the rich solvent temperature after the steam heater is 138°C. Although at this loading the regeneration specific heat duty of the advanced flash stripper is smaller than that of the simple stripper, however, from the total equivalent work point of view, at this loading the advanced flash stripper offers no energy savings. In fact, the total equivalent work of the advanced flash stripper is nearly 6% higher than that of the simple stripper. This finding offers another limit than the solvent thermal degradation for the applicability of the advanced flash stripper. From the total equivalent work viewpoint, the lowest lean loading at which the advanced flash stripper is capable of providing energy savings in terms of overall equivalent work is 0.20 (mol CO_2/mol MEA).

Figure 4 and 5 present graphically the regeneration specific heat duty and the total equivalent work of the advanced configurations and a comparison to the simple stripper.

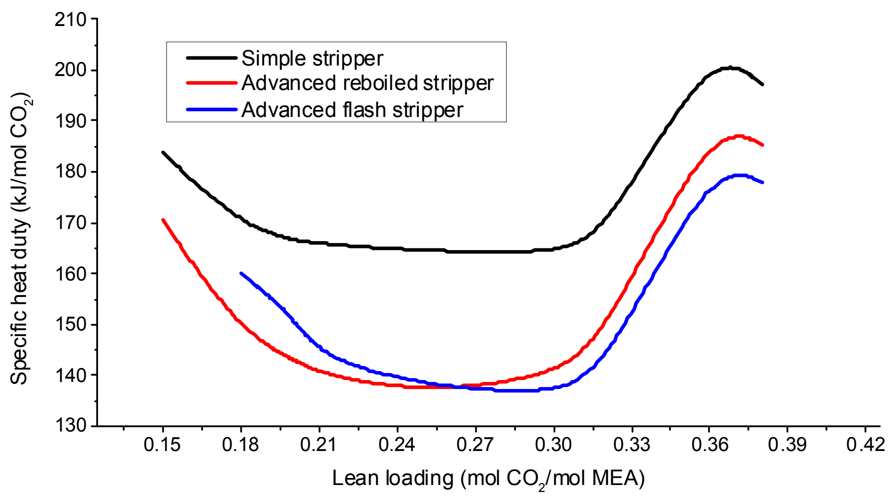

Figure 4. Comparison of the regeneration specific heat duty of advanced configurations for a range of lean loading.

Adding complexity improves the stripper energy requirements. The advanced reboiled stripper requires 6 to 16.9% less heat duty than the simple stripper, which is 4.1 to 11.7% less total equivalent work, where the lean loading associated with the highest and lowest improvements is 0.25 and 0.38 (mol CO_2/mol MEA), respectively. Likewise, for the advanced flash stripper, the improvement in specific heat duty varies from 8.8 to 17 %, and in total equivalent work varies from 4.4 to 12.4% at the lean loading of 0.30 and 0.20, respectively.

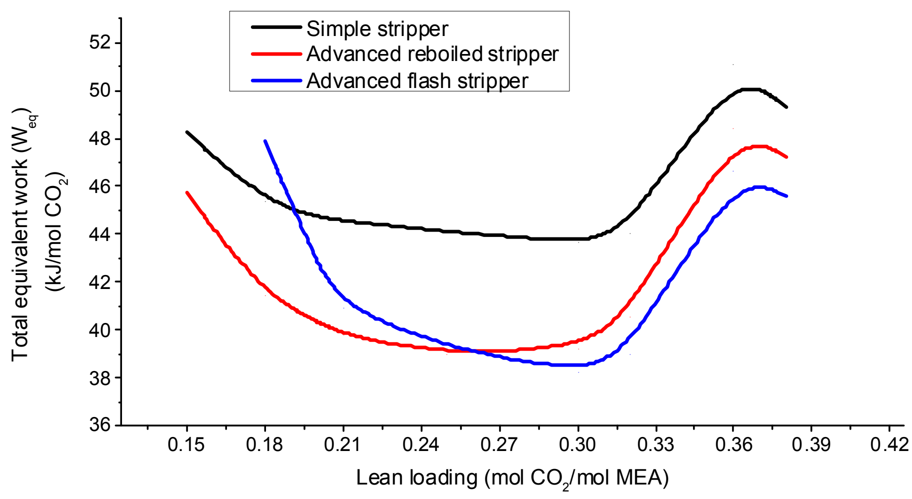

Figure 5. Comparison of the total equivalent work (W_{eq}) of advanced configurations for a range of lean loading, CO_2 compression to 15 MPa.

At low lean loading, i.e. below 0.25 (mol CO_2/mol MEA), the performance of the advanced reboiled stripper is better than the advanced flash stripper. However, at higher lean loading, the trend reverses and the advanced flash stripper provides greater improvement, to the point that at the lean loading of 0.38, the improvement provided by the advanced flash stripper is almost double than that of the advanced reboiled stripper.

One reason for this change might be correlated with the steam temperature. For the advanced reboiled stripper, the temperature of steam is identical to that of the simple stripper as both configurations employ the reboiler to provide the heat required for solvent regeneration with a 5°C steam side approach temperature. However in the advanced flash stripper, the reboiler is replaced by a convective steam heater, by which the heat required for regeneration is provided by steam using a 5°C LMTD. This difference resulted in different steam temperature used at each configuration. Figure 6 shows the temperature of steam used in the advanced reboiled and advanced flash strippers at optimum cases, and the relation to the solvent temperature at the bottom of stripper column. As shown, at the lean loading of 0.25 (mol CO_2/mol MEA), the temperature of steam used at both advanced configurations is similar. At loading below 0.25, the steam temperature used at the advanced flash stripper is higher than that of the advanced reboiled stripper, whereas this trend reverses for lean loading higher than 0.25.

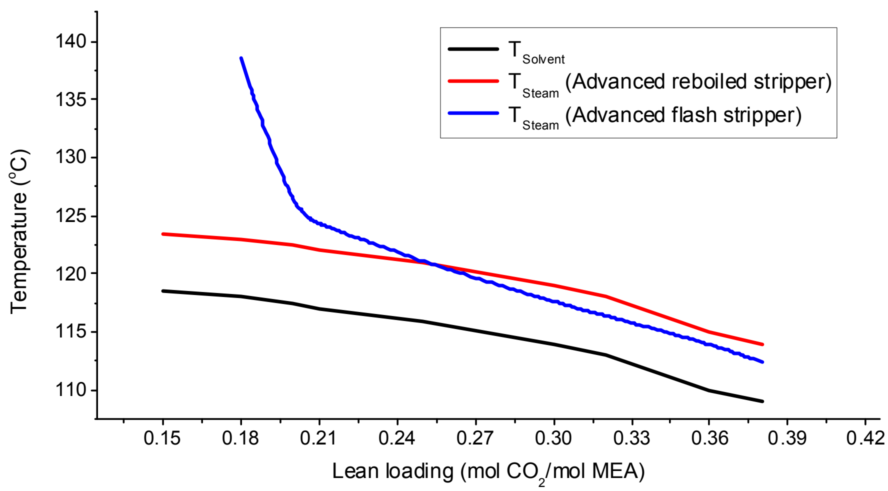

Figure 6. Comparison of steam and solvent temperatures for advanced configurations.

4.2. Temperature pinch

The stripper operation is frequently determined by a rich end pinch because of larger liquid to gas ratio at the top of the column relative to that at the bottom. In a simple stripper, when the pinch occurs at the rich end, the driving force at the lean end is excessively large with a loss of available work[4]. This condition is more pronounced at higher lean loading. In general, the stripping process is more reversible at lower lean loading since driving forces are to be relatively smaller at the lean end. Advanced reboiled and advanced flash stripper configurations were suggested to develop an equally distributed driving force through the column to reduce the energy required for regeneration and thus the total equivalent work.

To study the effectiveness of the advanced configurations on the stripper driving force, liquid and vapour temperature gradients through the stripper column at various lean loading were analysed and compared. The stripper column is comprised of 20 identical stages, followed by a reboiler in the simple stripper and advanced reboiled stripper, or by a flash drum in the advanced flash stripper, as the stage 21. The temperature driving force is calculated by the difference between the temperature of the liquid stream leaving a stage (stage "n") and the temperature of the vapour stream entering that stage, i.e. the temperature of the vapour stream leaving one stage below that stage (stage "n+1").

Figures 7 to 9 show the temperature driving at each stage of the stripper packed column for simple, advanced reboiled and advanced flash stripper configurations at lean loading of 0.21, 0.25, and 0.30 (mol CO_2/mol MEA), respectively. For the simple stripper, the temperature driving force is more consistent at lean loading of 0.21 than that of 0.30 (mol CO_2/mol MEA). This confirms the stripping process in the simple stripper configuration is more reversible at low lean loading compared to higher lean loading. In this configuration, the pinch was observed at the rich end at various lean loading. As lean loading increases the area of pinch expands through the column height followed by extensively increasing temperature driving force at the lean end. For instance, the magnitude of the temperature driving force at stage 20 for lean loading of 0.30 is nearly three times higher than that of 0.21 (mol CO_2/mol MEA) causing excess energy requirement for solvent regeneration.

For the advanced reboiled stripper, regardless of lean loading, the column is pinched at the middle of the column where the second rich solvent feed enters. From this point, as the solvent flows downward the temperature driving force increases. Although in the advanced reboiled stripper, the magnitude of the temperature driving force at the lean end is similar to that of the simple stripper, the difference between the temperature driving force of the top and the bottom of the column is lesser than that of the simple stripper. The effect of the advanced reboiled stripper in terms of column driving force is shifting the pinch from the top of the column to the middle of the column.

As shown in Figures 7 to 9, the advanced flash stripper has the smallest temperature driving force at lean ends among the three configurations. This configuration shows a tendency to be also pinched at lean ends which is more evident at higher lean loading. For instance, at 0.30 lean loading, the temperature driving force at the lean end is 0.6°C, compared to 4.5°C and 4.1°C of the simple stripper and advanced reboiled stripper, respectively. Results suggest the effect of the advanced flash stripper on the column driving force is to form a pinch at the lean end, which contributes to improve the thermodynamic efficiency and lower the energy requirement for solvent regeneration. This finding is aligned with what was shown earlier that at 0.30 lean loading, the advanced flash stripper provides the highest improvement in terms of total equivalent work.

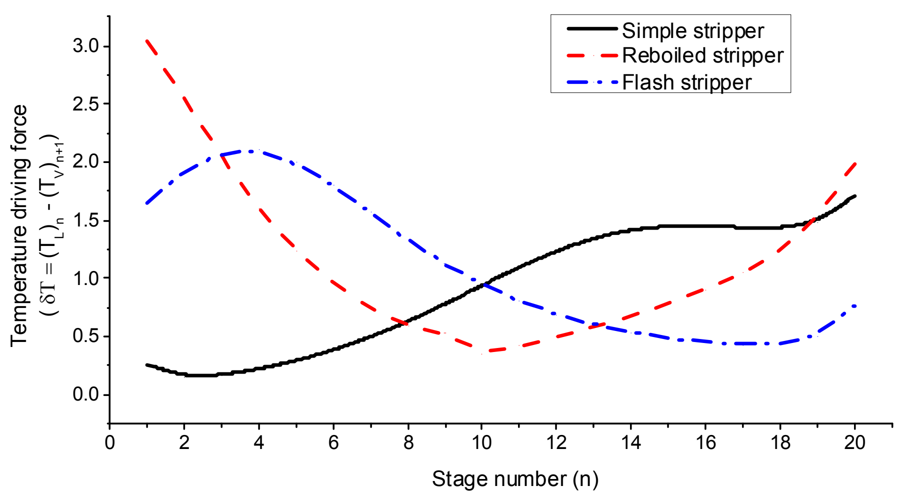

Figure 7. Temperature driving force at each stage for simple, advanced reboiled and advanced flash stripper at lean loading of 0.21 (mol CO_2/mol MEA), (stripper packed column = 20 stages, stage 1 at the top of the column, stage 20 at the bottom of the column, ΔT= liquid temperature leaving stage (n) –vapour temperature leaving stage (n+1))

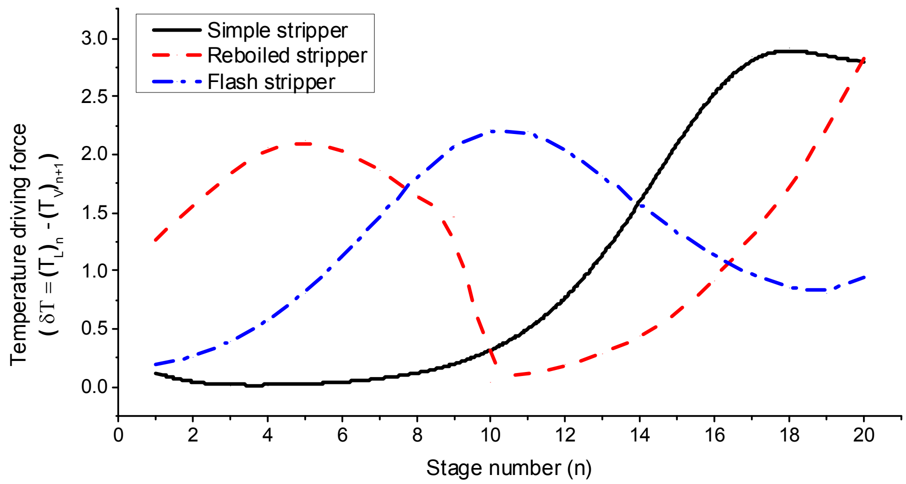

Figure 8. Temperature driving force at each stage for simple, advanced reboiled and advanced flash stripper at lean loading of 0.25 (mol CO_2/mol MEA), (stripper packed column = 20 stages, stage 1 at the top of the column, stage 20 at the bottom of the column, ΔT= liquid temperature leaving stage (n) –vapour temperature leaving stage (n+1))

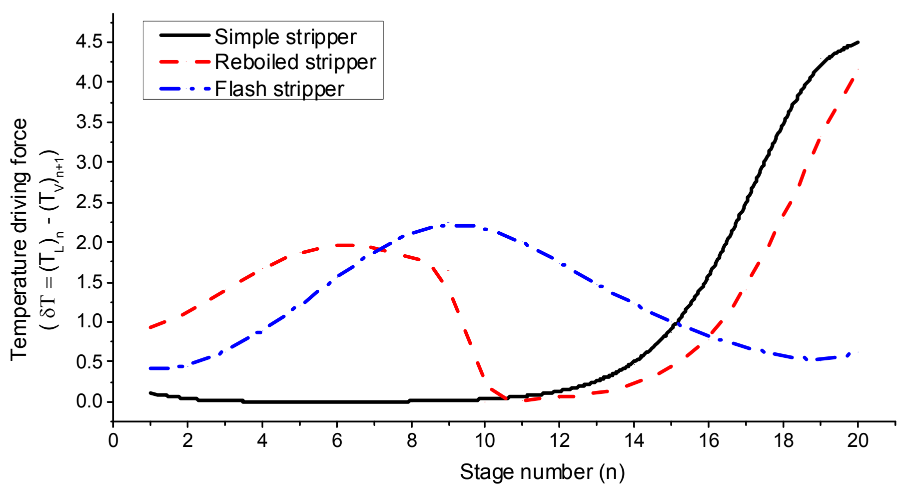

Figure 9. Liquid and vapour temperature driving forces at each stage for simple, advanced reboiled and advanced flash stripper at lean loading of 0.30 (mol CO_2/mol MEA), (stripper

packed column = 20 stages, stage 1 at the top of the column, stage 20 at the bottom of the column, ΔT= liquid temperature leaving stage (n) –vapour temperature leaving stage (n+1))

4.3. Heat recovery at the rich bypass heat exchanger

In a simple stripper, the product vapour typically leaves the column containing 40-60% water vapour (mole basis). This stream is cooled in a where the latent heat of the water vapour is lost. In the advanced reboiled and flash stripper configurations, the latent heat of the water vapour is partially recovered by the cold rich bypass stream at the rich bypass heat exchanger contributing to improve the energy efficiency of the system. In fact, the rich bypass heat exchanger acts as a part of the overhead condenser where the cooling water is replaced by the cold rich bypass stream recovering the heat dissipated from the product vapour which would be otherwise wasted. Figure 10 shows the water vapour in the product vapour before and after the rich bypass heat exchanger of optimum cases of advanced reboiled and advanced flash strippers for a range of lean loading from 0.20 to 0.32 (mol CO_2/mol MEA). For comparison, the water vapour in the product vapour of the simple stripper before entering the overhead condenser is also shown.

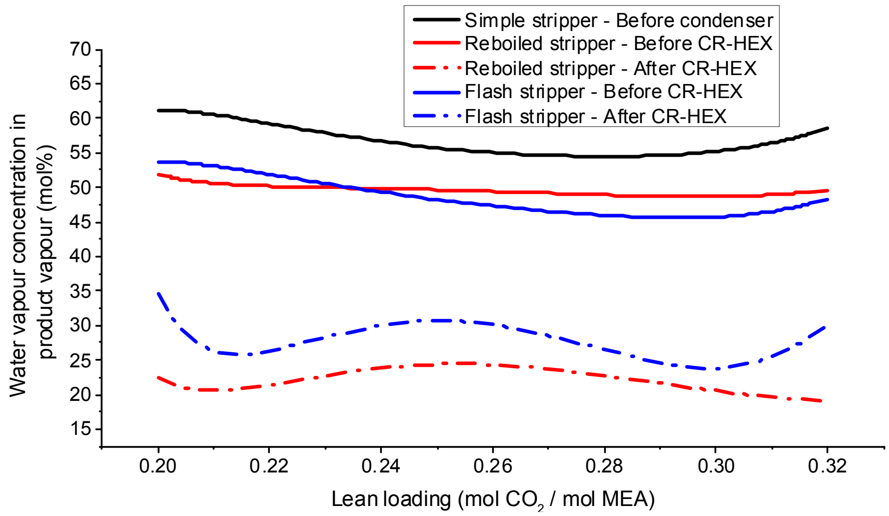

Figure 10. Water vapour concentration in the product vapour before and after cold rich heat exchanger (CR-HEX) of the optimum cases for simple, advanced reboiled and advanced flash strippers

The heat required for stripping is approximately the summation of three terms: the heat required to desorb the CO_2, the heat required to generate the water vapour at the top of the column, and the sensible heat required to increase the solvent temperature to the column temperature. According to Figure 10, in advanced configurations the water vapour content in the product vapour is 9 to 18% smaller than that of the simple stripper. This shows one of the positive contributions of advanced configurations on lowering the total heat requirement. In addition, the study showed the advanced stripper configurations contribute in lowering the plant total cooling water requirements. Table 5 summarises the reduction in the cooling water requirements (cooling water for the overhead condenser and the trim cooler) when using advanced reboiled and advanced flash strippers in relation to the simple stripper configuration.

Table 5. Reduction in cooling water consumption in percentage when using advanced strippers in relation to the simple stripper configuration

Lean loading (mol CO2/mol MEA)	**Advanced reboiled stripper**	**Advanced flash stripper**
0.18	19.2	3.8
0.2	23.1	15.4
0.21	30.4	26.8
0.25	17.4	17.4
0.3	20.0	20.0
0.32	16.7	23.3
0.36	1.6	4.8
0.38	3.2	6.3

The highest latent heat recovery in terms of the difference between the water vapour content before and after the rich bypass heat exchanger was observed at lean loading 0.25 (mol CO_2/mol MEA)for the advanced reboiled stripper and at 0.30 (mol CO_2/mol MEA) for the advanced flash stripper. These are the lean loading at which the corresponding advanced configurations offer the highest energy improvements in terms of total equivalent work. Furthermore, at lean loading 0.30 (mol CO_2/mol MEA) that the advanced flash stripper offers the highest energy savings, the water vapour content in the product vapour leaving the stripper column is the minimum amongst all optimum cases.

A comprehensive economic evaluation of the advanced configurations is outside the scope of this paper. The incremental capital cost to implement the advanced configurations should be small, so the energy saving should more than justify use of one of the advanced configurations. The reboiler or steam heater will cost less because it will have a reduced heat duty. The condenser is mostly replaced by cold rich exchanger, which will have significantly less heat duty than the condenser with the simple stripper. The cross exchanger will require two heat exchangers, but the total area will be about the same. The trim cooler will be larger. Additional piping and instrumentation will be required for the bypasses. Frailie[32] showed the purchase equipment cost of the advanced flash stripper with Piperazine (PZ) is smaller than the conventional stripper working with PZ and this is almost entirely due to the decrease in capital expenditure from using steam heater rather than reboiler.

The steam required for the advanced reboiled and advanced flash strippers will be extracted from the IP/LP cross over pipe at conditions similar to that of the simple stripper configuration, as both advanced configurations require steam at temperatures of 115-135°C (with saturated pressures of 170-312 kPa) compared to 115-125°C (with saturated pressure of 170-232 kPa) of the simple stripper configuration.

5. Conclusions

The advanced reboiled and advanced flash stripper were evaluated with 7 m MEA to remove 90% mole CO_2 from flue gases with 4% CO_2, typical of a natural gas fired application, for a range of lean loading from 0.15 to 0.38 (mol CO_2/mol MEA). The energy efficiency improvement offered by the advanced configurations was evaluated and compared with that of a simple stripper configuration using the total equivalent work.

Simulation results confirmed both advanced configurations work equally well over the specified range of lean loading, except the advanced flash stripper fails to operate at lean loading below 0.18 (mol CO_2/mol MEA), as the solvent temperature at the steam heater outlet exceeds the solvent thermal degradation limit.

With lean loading from 0.21 to 0.32 (mol CO_2/mol MEA), the advanced reboiled stripper and flash stripper require an equivalent work of only 38 to 41 kJ/mol of CO_2 recovered, compared to 44-45 kJ/mol with the simple stripper. The regeneration heat duty was reduced 11 to 18% to 136-148 kJ/mol of CO_2 recovered compared to 166-167 kJ/mol with the simple stripper. At lean loading of 0.30 (mol CO_2/mol MEA), the advanced flash stripper offers the highest reduction in the total equivalent work of 12.4%, and the highest reduction offered by the advanced reboiled stripper is 11.7% at the lean loading of 0.25 (mol CO_2/mol MEA).

Simulations showed that the advanced flash stripper requires more equivalent work than the advanced reboiled stripper at lean loading less than 0.26 (mol CO_2/mol MEA) and more than the simple stripper at a lean loading less than 0.2 (mol CO_2/mol MEA), mainly due to the higher steam temperature required at those lean loading.

The variation of temperature driving force through the column showed that the advanced flash stripper tends to pinch at the lean end, opposed to the simple stripper which usually pinches at the rich end, contributing to enhance the thermodynamic efficiency of the stripping process and reducing the loss of work.

In both advanced reboiled and advanced flash stripper configurations, one contributor to improve the energy efficiency is less water vapour at the top of the column. In addition both configurations contribute in lowering the plant cooling water requirement when compared with the plant with a simple stripper configuration.

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