Performance of Innovative Stripper Options for CO$_2$ Capture

By

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Abstract

The state-of-the-art technology for the capture of CO$_2$ from coal-fired power plants is aqueous absorption / stripping. The technology has been applied in small plants but requires some innovations for application in large power plants because of its energy -intensive nature. It is estimated that the operating cost of running this system constitutes a 15 – 30% energy hit to the power plant. In this paper, we present results of simulating different process alternatives including (1) using a more reactive solvent (2) operating the cross exchanger at a lower temperature approach (3) optimizing the stripper operation (4) using innovative configurations such as vacuum and multipressure strippers. The model was developed in the Aspen Custom Modeler (ACM) Simulation Environment using three stripper configurations – a simple, vacuum and multipressure stripper for two solvents 7m monoethanolamine (MEA) and 5m K$^+$ / 2.5m PZ. With some approximations, the use of solvents with varying heats of desorption are investigated and an optimum heat of desorption that minimizes reboiler duty and equivalent work is proposed. The vacuum stripper is favored for solvents with $\Delta H_{des} \leq 25$ kcal/gmol CO$_2$ while the multipressure configuration is attractive for solvents with $\Delta H_{des} \geq 25$ kcal/gmol CO$_2$ at a rich P$_{CO2}$* = 2.5 kPa and rich absorber temperature of 40°C.

Stripper Configurations

Simple Stripper

The simple reboiled stripper (Figure 1a) is run at 160 kPa. The vapor leaving the top of the stripper is cooled and the condensed water is refluxed. The CO$_2$ is compressed in five stages (intercooled to 40°C) to 1000 kPa. The reboiler runs at 110 – 120°C in this configuration.

Multipressure Stripper

The multipressure stripper, (Figure 1b), is divided into three sections, each operating at a different pressure. The CO$_2$ compressor is integrated with the stripper. The vapor from a lower pressure stage is compressed and subsequently

used as stripping vapor in a higher-pressure section. Water vapor condenses with the increased pressure and the latent heat of water is recovered. This leads to lower reboiler duties and CO_2 is produced at a greater pressure than with the simple (isobaric) stripper. However the compression work is greater than that of the simple stripper because some water vapor is compressed with the CO_2. The pressure levels are 160 kPa, 230 kPa and 330 kPa from the bottom to the top of the stripper. The vapor exiting the stripper is cooled and water is refluxed. The CO_2 is further compressed in three stages (intercooled to $40^{\circ}C$) to 1000 kPa. Therefore, the five compression stages include two integrated with the stripper.

(a) Typical Absorber/Stripper Configuration for 7m MEA (Rich $[CO_2]_T = 3.68$ m, lean $[CO_2]_T=2.39$ m at $40^{\circ}C$)

(b) Multipressure Stripper (Rich $[CO_2]_T = 5.33$ m, Lean $[CO_2]_T=4.51$ m at $40^{\circ}C$)
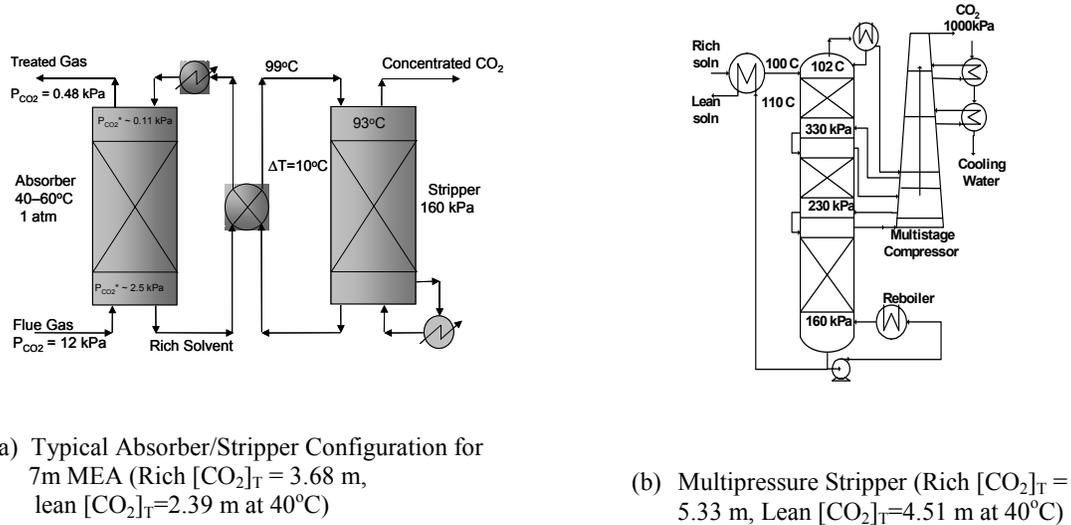
Figure 1: Stripper Configurations

Multipressure stripping has the following features:

1. The latent heat of water is recovered at the rich end.
2. It makes use of the high temperature preheat in the high pressure flash thereby rewarding a closer approach temperature in the cross exchanger.
3. CO_2 can be recovered at a greater concentration and pressure. This leads to less compression work downstream of the stripper.
4. This configuration should be best with high ΔH_{des} solvents such as 7m MEA.

Vacuum Stripper

This configuration is identical to the simple stripper. The stripper is operated at 30 kPa and the reboiler runs at 60 – 80°C. The CO_2 is compressed in five intercooled stages to 1000 kPa.

Vacuum stripping has the following features:

1. Lower temperature (less valuable) steam is used to run the reboiler so more electricity can be extracted before the steam is used in the stripper.
2. Additional compression is required for the CO_2.
3. The mass transfer is not as fast as that of the simple stripper because the lower temperature results in slower kinetics.

Model Development

A stripper model for aqueous solutions of 7m monoethanolamine and of 5m K^+/ 2.5m PZ was developed in Aspen Custom Modeler (ACM). This model divides the stripper into ten sections with Murphree efficiencies assigned to CO_2, water and temperature. In the multipressure configuration, four sections are at 160 kPa, four at 230 kPa and two at 160 kPa. An empirical expression with six adjustable constants was used to represent the vapor-liquid equilibrium (VLE) and heat of absorption/desorption. The three stripper configurations were modeled with variations of the rich and lean CO_2 loading, temperature approach (5-10°C) and stripper operating pressure. The equivalent work consumed by the process was calculated.

Modeling Assumptions

(a) The sections are well mixed in the liquid and vapor phases.
(b) The reaction takes place in the liquid phase.
(c) The reboiler is assumed to be in equilibrium.
(d) There is negligible vaporization of the amine.
(e) The top flash was 40% efficient.

The CO_2 vapor pressure (kPa) under stripper conditions was represented by the empirical expression in Table 1.

Table 1: Adjustable constants in VLE expression

$$\ln P_{CO2}* = a + b*[CO_2]_T + \frac{c}{T} + d\frac{[CO_2]_T^{\,2}}{T^2} + e\frac{[CO_2]_T}{T^2} + f\frac{[CO_2]_T}{T}$$

	7m MEA	5m K$^+$/2.5m PZ
a	35.12	-0.263
b	-6.43	0.148
c	-14281	-5306
d	-11148.5	-16995.5
e	-485777	-469758
f	4667.14	2808

The total CO_2 concentration, $[CO_2]_T$, is the concentration of CO_2 in all forms including free CO_2, bicarbonate, carbonate and carbamate. The adjustable constants in Table 1 were obtained by regressing the points from the rigorous model for 5m K$^+$/2.5m PZ by Cullinane [1] and for 30 wt% MEA using equilibrium flashes in AspenPlus based on the rigorous model developed by Freguia [2] from data of Jou et al. [3].

The heat of absorption/desorption is calculated by differentiating the equation in Table 1 with respect to 1/T:

$$-\frac{\Delta H}{R} = c + 2d\frac{[CO_2]_T^{\,2}}{T} + 2e\frac{[CO_2]_T}{T} + f[CO_2]_T \qquad (1)$$

The heat of vaporization of water, partial pressure of water, and heat capacities of steam, CO_2 and the solvent (essentially water) were calculated with equations from the DIPPR database [4].

The partial pressure of CO_2 and water in each section was calculated by:

$$P_n = E_{mv}(P_n* - P_{n-1}) + P_{n-1} \qquad (2)$$

A Murphree efficiency (E_{mv}) of 40% and 100% was assigned to CO_2 and water. The model assumed a 100% efficiency with respect to heat transfer.

The model inputs were the rich $[CO_2]_T$ and liquid rate (1 kg/s), the temperature approach in the cross exchanger (difference between the temperature of the rich stripper feed and the lean solution leaving the bottom of the stripper) and column pressure. Initial guesses of the lean $[CO_2]_T$, section temperatures, partial pressures and concentrations were provided. The model solves equations for calculating VLE and for material and energy balances. It calculates temperature and composition profiles, reboiler duty and equivalent work.

The total energy required by the stripper is given as total equivalent work:

$$W = 0.75Q \left[\frac{(T_{reb} + 10) - 313}{(T_{reb} + 10)} \right] + W_{comp} \qquad (3)$$

W_{comp} constitutes the isentropic work of compression of the gas exiting the top of the stripper to 1000 kPa. An efficiency of 75% was assumed for the compressor.

The first term in (3) accounts for the electricity generation lost by extracting steam from a turbine while the second is the compressor work. The condensing temperature of the steam is assumed to be 10K higher than the reboiler fluid. The turbine assumes condensing steam at 313K.

Results and Discussion

Predicted Stripper Performance
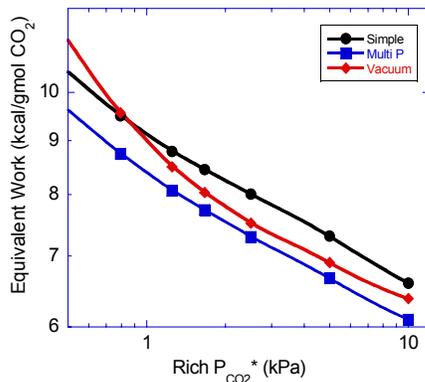
Figure 2: Total Equivalent Work for Different Configurations with 7m MEA (ΔT=10°C, Abs. Rich T = 40°C)

Figure 3: Total Equivalent Work for Different Configurations with 5m K+/2.5m PZ (ΔT=10°C, Abs. Rich T = 40°C)

Figure 2 shows the minimum equivalent work for 7m MEA using the three configurations. The multipressure stripper gives the least equivalent work over the entire rich P_{CO2}* range. The simple stripper is the least attractive configuration with the highest work over most of the rich P_{CO2}* range. The multipressure stripper offers 8% energy savings when compared to the simple stripper. The vacuum stripper requires 6% less energy at high rich P_{CO2}*.

Figure 3 shows the minimum equivalent work for 5m K$^+$/2.5m PZ using the three configurations. The vacuum stripper gives the least equivalent work over most of

the rich $P_{CO2}*$ range with the multipressure stripper competitive at higher rich $P_{CO2}*$. The simple stripper is the least attractive configuration at high rich $P_{CO2}*$. In comparison to the simple stripper, the vacuum stripper requires 18% less energy at lower rich $P_{CO2}*$ and offers savings up to 8% at higher $P_{CO2}*$.

Operating the stripper using 5m K^+/2.5m PZ with a closer temperature approach, 5^oC instead of 10^oC, offers 2 to 6% savings over the practical range of rich $P_{CO2}*$ as evident from Figure 4. An economic analysis is desirable before additional investment in heat transfer area is made.

Generic Solvent Modeling

The vapor-liquid equilibrium for the generic solvents was approximated by the expression:

$$\ln P = a + b*[CO_2]_T \cdot \frac{\Delta H}{RT} \qquad (4)$$

The constant b was set to 3.07 while the constant, a , was varied. The value of constant, a , used in equation (4) for the generic solvents is from our previous work [5].
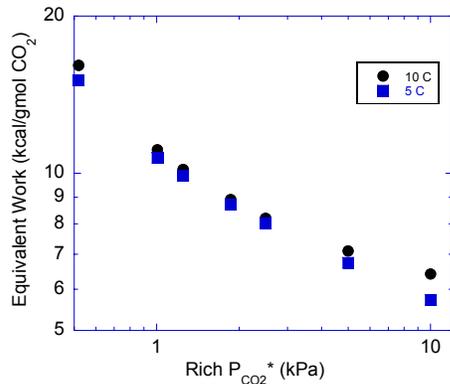
Figure 4: Total Equivalent Work for Different ΔT for 5m K^+/2.5m PZ for a Simple Stripper (Abs. Rich T = 40^oC)
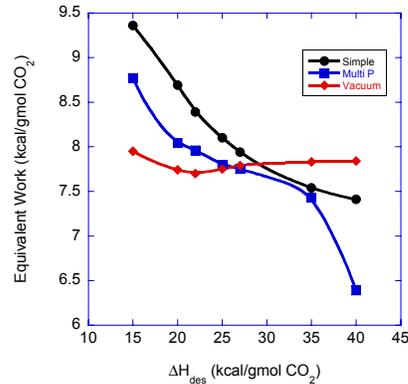
Figure 5: Total Equivalent Work for Generic Solvents (Rich P_{CO2}= 2.5 kPa at 40^oC, $\Delta T = 10^oC$)

Figure 5 shows the minimum total equivalent work for the generic solvents using the three configurations. The vacuum stripper requires the least equivalent work with solvents with $\Delta H_{des} \leq 25$ kcal/gmol CO_2 while the multipressure stripper requires the least equivalent work for solvents with $\Delta H_{des} \geq 25$ kcal/gmol CO_2.

Conclusions

1. The multipressure configuration gives the least equivalent work at a fixed absorber rich P_{CO2}* for 7m MEA.
2. For 5m K$^+$/2.5m PZ, the vacuum stripper is the most attractive option. Lower temperatures, 60-80°C, will help reduce corrosion and allow for alternative materials of construction like fiber-reinforced plastic (FRP).
3. With 5m K$^+$/2.5m PZ, operating the cross exchanger with a 5°C approach offers 2-6% energy savings over one with a 10°C approach.
4. For generic solvents, the optimum ΔH_{des} for the solvent that will minimize equivalent work is a function of the stripper configuration used. The vacuum stripper is favored for solvents with $\Delta H_{des} \leq 25$ kcal/gmol CO_2 while the multipressure configuration is attractive for solvents with $\Delta H_{des} \geq 25$ kcal/gmol CO_2 at a rich P_{CO2}* = 2.5 kPa when the rich absorber temperature is 40°C.

Acknowledgements

Nomenclature

$[CO_2]_T$ = total concentration of CO_2 [molal]
E_{mv} = Murphree section efficiency defined in terms of partial pressures [-]
ΔH = heat of absorption/desorption [kcal/gmol CO_2]
P_{CO2} = partial pressure of CO_2 in the bulk gas [kPa]
P_{CO2}* = equilibrium partial pressure of CO_2 [kPa]
P_n = partial pressures on sections n [kPa]
P_{n-1} = partial pressures sections n-1 [kPa]
P_n* = equilibrium partial pressure leaving section n [kPa]
Q_{reb} = reboiler duty [kcal/gmol CO_2]
R = universal gas constant [cal/K-mol]
T = temperature [K]
W_{comp} = isentropic work of compression [kcal/gmol CO_2]
W_{eq} = equivalent work [kcal/gmol CO_2]

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