Optimization of Advanced Flash Stripper for CO₂ Capture using Piperazine

Yu-Jeng Lin, a Gary T. Rochelle"a*

aMcKetta Department of Chemical Engineering, The University of Texas at Austin, 200 E. Dean Keeton St., C0400, Austin, TX 78712-1589

Abstract

CO₂ capture from coal-fired power plants using amine scrubbing typically incurs a 20–30% energy penalty. The major energy requirements are the reboiler duty for solvent regeneration and the compression work for CO₂ sequestration. The advanced flash stripper using 8 m PZ was proposed, which recovers the stripping steam heat by employing cold and warm rich bypasses. The objective of this work is to quantify the benefits of the advanced regeneration system as energy and capital cost. The advanced flash stripper saves 10% total equivalent work that includes heat duty work, compression work, and pumping work compared to the simple stripper. From the economic analysis, it was found that the compressor and the cross exchangers are the two major capital costs of the regeneration system. By considering the cost of the main heat exchangers, the temperature approaches and pressure drops of the cross exchanger and the steam heater were optimized. The optimum LMTD of the cross exchanger is around 10 K, which saves 8% regeneration cost compared to the base case with 5 K. The annualized regeneration cost of the advanced flash stripper is $31/tonne CO₂ at the optimum lean loading 0.20, providing 15% cost savings compared to the base case simple stripper (5 K cross exchanger LMTD).

© 2014 The Authors. Published by Elsevier Ltd.
Selection and peer-review under responsibility of GHGT.

Keywords: Process design; Aspen Plus®; amine scrubbing

1. Introduction

In post-combustion CO₂ capture using amine scrubbing (Figure 1), steam usage for solvent regeneration and CO₂ compression work are the main contributors to the energy requirement. Implementing CO₂ capture incurs a 20–30% penalty on electricity output for a typical coal-fired power plant [1].
Alternative stripper configurations could improve energy efficiency significantly compared to a simple stripper. Van Wagener [2,3] emphasized the importance of increasing process reversibility by introducing more complex configurations including multi-stage flash, cold rich bypass, and an interheated column. Van Wagener showed that the interheated stripper offers the best energy savings. Piperazine (PZ) is a new standard solvent [4] that provides higher reaction rate, capacity, and thermal stability than the conventional solvent, monoethanolamine (MEA). The energy performance compared to MEA has been demonstrated [3].

Loss of stripping steam vapor from the regenerator is one of the reasons that the simple stripper is inefficient. When the stripper is operated at 120–150 °C, water in the rich solvent is vaporized and emitted with CO₂ from the top of stripper. The proposed advanced flash stripper [5] applies cold and rich bypasses to recover the stripping steam heat, providing better energy savings than the interheated stripper. The CO₂ lean loading (mol CO₂/mol alkalinity) is the most important process operating parameter that determines the vapor-liquid equilibrium and regeneration capacity. The optimum lean loading has been investigated for MEA and PZ [5], however, only energy performance was considered. Besides reboiler duty, compression work and pump work should be considered. Capital cost should be included to determine optimum design conditions.

In this work, the advanced flash stripper using 8 m PZ has been modeled using Aspen Plus®. Total equivalent work is used as an indicator of energy performance that includes reboiler duty work, pump work, and compression work. The capital cost will be estimated for the regeneration system. The regeneration cost, including energy and capital cost, will be used to determine the optimum lean loading.

2. Process descriptions

2.1. Simple stripper

The simple stripper shown in Figure 1 is the base case. The cold rich solvent is heated by the hot lean solvent in the cross exchanger and then sent to the top of stripper. The reboiler provides the sensible heat, the heat of CO₂ desorption, and the vaporization heat of water. The hot lean solvent from the reboiler is returned to the absorber through the cross exchanger. The hot CO₂ vapor from the top of the stripper is cooled to 40 °C in the overhead condenser with loss of the latent heat of the excess water vapor.

![Figure 1. Amine scrubbing with simple stripper](image)
2.2. Advanced flash stripper

Figure 2 shows the advanced flash stripper with warm rich bypass and cold rich bypass. In this configuration, the rich exchanger is used to preheat the cold rich solvent by hot CO₂ vapor coming out of the stripper. A portion of the cold rich solvent obtains latent heat of steam from the stripped vapor. Warm rich bypass is extracted between two cross exchangers and fed to the top of the stripper after mixing with cold rich bypass. The temperature was selected as the bubble point temperature at the stripper operating pressure. The rest of the rich solvent is heated by a steam heater and fed into the bottom of the flash stripper. The reboiler in a typical stripper is replaced by a convective steam heater and a flash vessel. Only part of the rich solvent countercurrently contacts vapor in the flash stripper. Since the convective steam heater has less solvent hold-up and residence time, it will minimize thermal degradation.

![Diagram of Advanced Flash Stripper](image)

Figure 2. Advanced flash stripper

2.3. Multi-stage compressor

The inlet pressure of the compressor train is determined by the stripper pressure, which is dependent on the system lean loading and reboiler temperature. To sequester the CO₂ underground, the target pressure of the compressor has to be at least above its supercritical pressure, 74 bar. To replace the pressure loss during transportation, 0.4-0.5 bar/km is required. In this work, the target pressure is 150 bar. When the CO₂ is in the supercritical phase such that its density is similar to liquid, the difference between a pump and a compressor for the compression task disappears and becomes a question of density rather than phase. The supercritical pump is suggested for the last stage when the density is above 500 kg/m³ [6].

To decrease the pipeline diameter, aftercooling can be applied to increase CO₂ density and reduce volume flow rate [7]. In this work, the aftercooler that cools CO₂ is employed before the supercritical pump to attain a density suitable for pumping and to reduce the pump work from reduced volume flow rate. The density of CO₂ increases dramatically with temperature around the critical region. The aftercooling temperature is specified as 30 °C to obtain the major density increase.

For a coal-fired power plant with 593 MW gross output, the CO₂ volume flow rate is around 140,000 ft³/min at 1 bar [8]. A centrifugal compressor with intermediate pressure ratio and large capacity has been suggested for CO₂ capture [9].
3. Methods

3.1. Process specifications

The process specifications used in the simulations are shown in Table 1. Since the absorber is not included, a split flow sheet was used, with typical rich solvent conditions such as loading and temperature fixed as inputs.

This work simulates the solvent regeneration process, with the exception of the absorber, as shown in Figures 1 and 2. Simulation results were obtained from Aspen Plus® version 7.3. The Electrolyte Non-Random Two-Liquid (e-NRTL) property method was used. For the gas-liquid contacting separator unit, Aspen Plus® RateSep™ provides a rigorous rate-based model for heat and mass transfer using a non-equilibrium approach, applying two-film theory. The thermodynamic and kinetics model used for PZ in this work was “Independence” [10]. The model has been regressed in Aspen Plus®, with experimental data including amine volatility, heat capacity, CO₂ solubility, and amine pKa over a range of amine concentration and CO₂ loading.

The multi-stage compressor was also modeled in Aspen Plus®. Since only CO₂ and water are presented, the Benedict-Webb-Rubin-Starling (BWRS) thermodynamic model was used. The multi-stage compressor will compress the CO₂ from the stripper pressure to 76 bar that includes 2 bar of net positive suction head (NPSH), and the supercritical pump will pressurize the supercritical CO₂ to 150 bar. The maximum pressure ratio of the compressor at each stage is specified as 2 due to high molecular weight.

Table 1. Stripper specifications

<table>
<thead>
<tr>
<th>Solvent</th>
<th>8 m PZ</th>
</tr>
</thead>
<tbody>
<tr>
<td>Process modeling tool</td>
<td>Aspen Plus® v7.3</td>
</tr>
<tr>
<td>Thermodynamic model</td>
<td>Independence</td>
</tr>
<tr>
<td>Packing type</td>
<td>2 m Mellapak standard 250X</td>
</tr>
<tr>
<td>Regeneration temperature (°C)</td>
<td>150</td>
</tr>
<tr>
<td>Rich loading (mol CO₂/mol alk)</td>
<td>0.4</td>
</tr>
<tr>
<td>Rich solvent temperature (°C)</td>
<td>46</td>
</tr>
<tr>
<td>Cross exchanger LMTD (°C)</td>
<td>5</td>
</tr>
<tr>
<td>Rich exchanger LMTD (°C)</td>
<td>20</td>
</tr>
</tbody>
</table>

Table 2. Multi-stage compressor specifications

<table>
<thead>
<tr>
<th>Process modeling tool</th>
<th>Aspen Plus® v7.3</th>
</tr>
</thead>
<tbody>
<tr>
<td>Thermodynamic model</td>
<td>BWRS</td>
</tr>
<tr>
<td>Maximum pressure ratio/stage</td>
<td>2</td>
</tr>
</tbody>
</table>
### 3.2. Energy cost

Total equivalent work is used to account for the electricity penalty due to the steam extraction from the power plant, which is a more useful metric than reboiler duty alone. As Equation 1 shows, the equivalent work consists of pump work, compression work, and heat duty work. The pump is required to move the solvent from the absorber to the pressure of the stripper. The heat duty work is obtained from reboiler duty by multiplying by turbine efficiency and Carnot cycle efficiency (Equation 2). The turbine efficiency is set to a typical value of 90%. The compression work shown in Equation 3 is regressed using the configuration in Figure 3 from Aspen Plus® in the range of inlet pressure from 1 to 20 bar.

\[
W_{eq} \left(\frac{kJ}{\text{mol CO}_2}\right) = W_{Heat} + W_{pump} + W_{comp}
\]

\[
W_{Heat} = 90\% \left(\frac{T_{steam} - T_{sink}}{T_{steam}}\right) Q_{reb}
\]

\[
W_{comp} \left(\frac{kJ}{\text{mol CO}_2}\right) = -3.48 \ln(P_{in}) + 14.85, \quad 1 < P_{in}(\text{bar}) < 20
\]

### 3.3. Capital cost

The purchased equipment cost (PEC) of unit operations were obtained either from vendor quotes or empirical correlations, and then scaled to 2014 cost level. Techno-economic analysis for CO₂ capture process have been studied by Frailie [10], who concluded that the total annualized capital cost can be calculated from PEC by Equation 4. The scaling factor \(\alpha\) converts PEC to total capital cost and the annualizing factor \(\beta\) annualizes cost. The \(\alpha\) includes direct cost, indirect cost, and working capital. The \(\beta\) takes into account return on investment (10%), taxes (35% of return on investment), depreciation, and maintenance (2–3%). The \(\alpha\) and \(\beta\) are recommended as 5 and 0.2, respectively.

\[
\text{Total annualized capital cost} \left(\frac{\$}{\text{yr}}\right) = \alpha \times \beta \times \text{PEC}($)
\]

### 3.4. Optimization of main heat exchangers

In the amine scrubbing process, the capital cost of the cross exchanger is one of the cost centers since the heat transferred from the hot lean solvent to the cold rich solvent is about 4–5 times the heat input to the reboiler/steam heater. The temperature approach and the pressure drop are the most important design parameters for the heat exchangers.

The pressure drop through the heat exchangers can be used to enhance the heat transfer coefficient, which reduces heat exchanger area but increases pumping cost. A pump for the lean solvent is not necessary when the stripper pressure is sufficient to get the solvent through the cross exchangers, the trim cooler and the absorber. The pressure drop of the trim cooler and the absorber is assumed as 1 bar and 3 bar, respectively. With the optimization, a lean solvent pump is possibly added when it is worth to save the capital cost of the cross exchanger by increasing pressure drop. To determine the temperature approach of the cross exchanger, the tradeoffs involve the capital cost...
and the steam cost of the reboiler/steam heater. The costs associated with the optimization of the main heat exchangers are summarized in Table 3. The decision variables include the pressure drop and the temperature approach of the cross exchanger and reboiler/steam heater.

Table 3. Capital and energy cost associated with the optimization of main heat exchangers.

<table>
<thead>
<tr>
<th>Main heat exchanger cost</th>
<th>Capital cost</th>
<th>Energy cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cross exchanger</td>
<td>Plate and frame exchanger</td>
<td>Pump</td>
</tr>
<tr>
<td>Steam heater/reboiler</td>
<td>Shell and tube exchanger</td>
<td>Pump</td>
</tr>
<tr>
<td>Pump</td>
<td>Pumping cost</td>
<td>Pumping cost</td>
</tr>
<tr>
<td>Steam cost</td>
<td>Steam cost</td>
<td></td>
</tr>
</tbody>
</table>

4. Results and discussions

4.1. Optimization of heat exchangers

The pressure drop and the temperature approach of the cross exchangers and the steam heater/reboiler were optimized. Figure 4 shows the main exchanger cost of the advanced flash stripper with varied log mean temperature difference (LMTD), which is the parameter that has the greatest effect on the cost. The pressure drop and the steam temperature for each point are optimized. Three representative lean loadings are shown. The optimum LMTD is around 10 K in the range of optimum lean loading. From lean loading 0.32 to 0.26, the cost savings come from the improved capacity, but diminishing returns can be seen when the lean loading decreases further. The cost is less sensitive to LMTD at lower loading since the size of the cross exchanger is relatively small.

![Fig. 4. Annualized cost of main heat exchangers of the advanced flash stripper with varied cross exchanger LMTD; optimum pressure drops; optimum steam temperature.](image-url)
4.2. Total equivalent work and energy cost

The total equivalent work of the simple stripper and the advanced flash stripper is shown in Figure 5. Compared to the simple stripper, the energy saving of the advanced flash stripper is 10%, which is mostly derived from the recovery of stripping steam heat. The improvement is more significant at low loading because more stripping steam comes out with the CO₂ vapor. Figure 5 also shows the total equivalent work with 5 and 10 K cross exchanger LMTD. The energy performance is always better when a tighter temperature approach is employed. Less reboiler duty is required to heat the solvent to 150 °C when the temperature of the rich solvent coming into the stripper is higher by using a small LMTD, however, the capital cost of the cross exchanger is prohibitive compared to the energy savings.

The total equivalent work is converted to energy cost by using $100/MWh levelized cost of electricity (LCOE). Figure 6 shows the distribution of the energy cost with varied lean loading. The steam usage accounts for about 2/3 of total energy consumption. Both pumping work and compression work are driven by the stripper pressure. The pumping work increases faster when the lean loading is above 0.30 due to the low capacity, which offsets a part of benefits from the compression work when the stripper is operated at high pressure.

![Fig. 5. Total equivalent work of simple stripper and advanced flash stripper at varied 5 and 10 K cross exchanger LMTD.](image-url)
4.3. Capital cost

The capital cost of the advanced flash stripper is presented with varied lean loading (Figure 7). The compressor and the cross exchange are the two major costs. The compressor cost is mainly affected by the stripper pressure. A discontinuity of increasing compressor cost is observed between lean loading 0.18 and 0.20, where the compressor train needs an extra stage to attain the target pressure of 150 bar. The capital cost of the cross exchanger is driven by the capacity and is most sensitive when the lean loading changes. The capital cost of the rich exchanger reflects the amount of the stripping steam recovered. Generally, at high lean loading range, the capital cost is dominated by the cross exchanger, but with decreasing lean loading, the compressor cost starts to take over.
4.4. Lean loading optimization

The regeneration cost includes the energy cost and the capital cost shown in Figures 6 and 7. Figure 8 compares the simple stripper with and without the heat exchanger optimization. A typical cross exchanger LMTD, 5 K is arbitrarily chosen as the base case without optimization. The optimum case saves $3/tonne CO₂, an 8% improvement compared to the base case. The results confirm the importance of heat exchanger design in the amine scrubbing process.

By applying the advanced flash stripper, the total regeneration cost improves 7% further, which mainly comes from the lower energy requirement. The optimum lean loadings for both the simple stripper and the advanced flash stripper are around 0.20, lower than the optimum that only the energy performance is considered (Figure 5), but is close to the optimum of the capital cost (Figure 7). Even though the energy cost accounts for 2/3 of the regeneration cost, the capital cost is more sensitive with varied lean loading. The low capacity at high lean loading reduces both energy and capital cost, so the optimum is forced to lower lean loading until the compression cost and the heat loss of stripping steam dominate. The results demonstrate that the advanced flash stripper provides the lowest regeneration cost even though process complexity increases.

With 8 m PZ a lean loading of 0.20 may not be operationally attractive because of the potential precipitation of PZ.6H₂O solid at upset conditions or during shutdown when the solvent cools below 40°C. This possibility may require additional capital cost to heat trace lines and provide other means to recover from an upset. Alternate solvents may also be considered that provide the same performance as 8 m PZ without precipitation at the optimum lean loading.
5. Conclusions

The advanced flash stripper that recovers the stripping steam heat by the cold and warm bypasses was proposed, which provides 30.4 kJ/mol CO$_2$ total equivalent work, giving 10% energy savings compared to the simple stripper. Besides the energy performance, the capital cost of regeneration was also evaluated. From the optimization of the main heat exchangers, the optimum LMTD of the cross exchanger is around 10 K. The compressor and the cross exchanger account for over 50% of the capital cost of regeneration. Combining the energy cost and the capital cost, the regeneration cost of the advanced flash stripper is $31/tonne CO$_2$ at the optimum lean loading 0.20, providing 15% and 7% cost savings over the base case simple stripper (5 K cross exchanger LMTD) and optimized simple stripper, respectively.

References