CO₂ Capture by Absorption with Potassium Carbonate

Quarterly Progress Report

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Abstract

The objective of this work is to improve the process for CO₂ capture by alkanolamine absorption/stripping by developing an alternative solvent, aqueous K₂CO₃ promoted by piperazine. The stripper model with Aspen Custom Modeler and careful optimization of solvent rate suggests that 7 m MEA and 5 m K₂CO₃/2.5 m PZ will be practically equivalent in energy requirement and optimum solution capacity. The multipressure stripper reduces energy consumption by 15% with a maximum pressure of 5 atm. The use of vanadium as a corrosion inhibitor will carry little risk of long-term environmental or health effects liability, but the disposal of solvent with vanadium will be subject to regulation, probably as a hazardous waste. Analysis of the pilot plant data from Campaign 1 has given values of the mass transfer coefficient consistent with the rate data from the wetted wall column. With a rich end pinch, 30% MEA should provide a capacity of 1.3-1.4 mole CO₂/kg solvent.

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Introduction

The objective of this work is to improve the process for CO₂ capture by alkanolamine absorption/stripping by developing an alternative solvent, aqueous K₂CO₃ promoted by piperazine. This work expands on parallel bench-scale work with system modeling and pilot plant measurements to demonstrate and quantify the solvent process concepts.

The bench-scale and modeling work is supervised by Gary Rochelle. Frank Seibert is supervising the pilot plant. Three graduate students (Babatunde Oyekan, Ross Dugas, and Jennifer Lu) have received support during this quarter for direct effort on the scope of this contract. Three students supported by other funding have made contributions this quarter to the scope of this project (Eric Chen – EPA Star Fellowship; Tim Cullinane and Marcus Hilliard – Various Industrial Sponsors).

Experimental

Subtask 2.1 describes the test plan to be used in the pilot plant for the second campaign.

Results and Discussion

Progress has been made on three subtasks in this quarter:

Subtask 1.3 – Develop Integrated Absorber/Stripper Model

Babatunde Oyekan has an initial version of a stripper model using Aspen Custom Modeler. The model has been used to calculate the performance with 7 m MEA and 5 m K⁺ / 2.5 m PZ in both simple and multipressure stripper.

Subtask 1.5 – Simulate Base Case Pilot

Material balances and rate analyses were completed with the data from Campaign 1.

Subtask 1.9 – Economic and Applicability Analysis

A literature review is presented on the environmental hazards of vanadium.

Subtask 1.10 – Simulate MEA Baseline Testing

The Freguia model of the MEA solvent has been used to predict absorber performance of the pilot plant with 30% MEA.

Subtask 2.1 – Pilot Plant Test Plan – Campaign 2

A detailed test plan was developed for Campaign 2 to obtain a more complete data set on absorber performance with Flexipac 1Y and IMTP#40 packing and to obtain performance data for the stripper.

Conclusions

1. The stripper model with Aspen Custom Modeler and careful optimization of solvent rate suggests that 7 m MEA and 5 m K+/2.5 m PZ will be practically equivalent in energy requirement and optimum solution capacity.
2. The multipressure stripper reduces energy consumption by 15% with a maximum pressure of 5 atm.
3. The use of vanadium as a corrosion inhibitor will carry little risk of long-term environmental or health effects liability, but the disposal of solvent with vanadium will be subject to regulation, probably as a hazardous waste.
4. Analysis of the pilot plant data from Campaign 1 has given values of the mass transfer coefficient consistent with the rate data from the wetted wall column.
5. With a rich end pinch, 30% MEA should provide a capacity of 1.3-1.4 mole CO₂/kg solvent.

Future Work

We expect the following accomplishments in the next quarter:

Subtask 1.1 – Modify Vapor-Liquid Equilibrium (VLE) Model

The AspenPlus™ model will be regressed to represent the CO₂ solubility in 5 m K⁺/2.5 m PZ.

Subtask 1.2 – Modify Point Rate Model

Parameters will be developed to represent measured absorption rates with the rigorous model of mass transfer by Bishnoi.

Subtask 1.3 – Develop Integrated Absorber/Stripper Model

Initial results will be obtained to simulate the stripper with Aspen Custom Modeler. The existing absorber model for MEA will be modified to accept piperazine species.

Subtask 1.5 – Simulate Base Case Pilot

The absorber data from Campaign 1 will be simulated with a rigorous differential equation model.

Subtask 1.10 – Simulate MEA Baseline Testing

The stripper performance will be simulated with the Freguia model.

Subtask 2.1 – Pilot Plant Test Plan

A detailed test plan will be developed for the third pilot plant campaign, the MEA base case.

Subtask 2.5 – Campaign 2

The second campaign in the pilot plant is scheduled to begin in mid-October 2005. The pilot plant will be operated for four weeks with 5 m K⁺/ 2.5 m PZ. Performance data will be obtained for both the stripper and the absorber.

Task 1 – Modeling Performance of Absorption/Stripping of CO₂ with Aqueous K₂CO₃ Promoted by Piperazine

Subtask 1.3 – Develop Integrated Absorber/Stripper Model – Aspen Custom Modeler for Stripper

By Babatunde Oyenekan
(Supported by this contract)

Introduction

We have continued to develop the stripper submodel in Aspen Custom Modeler™ (ACM) for the overall model of CO₂ absorption/stripping by aqueous solutions of 7m monoethanolamine and of potassium carbonate and piperazine. This model divides the stripper into sections with Murphree efficiencies assigned to CO₂, water and temperature. A new expression with six adjustable constants is used to represent the VLE and heat of absorption/desorption. Two process configurations (a simple stripper and a multipressure stripper) are simulated and the effect of varying the rich and lean [CO₂]_T, approach temperature (5-10°C) and stripper operating pressure on the equivalent work consumed by the process is calculated by this model. The results show that when the 5m K⁺ / 2.5m PZ solvent is run at a closer approach to saturation, it requires a comparable energy input as 7m monoethanolamine. Both solvents have comparable capacities. The multipressure stripper reduces equivalent work by ~ 15% when compared to a simple stripper at a fixed temperature approach.

Experimental (Model Formulation)

Aspen Custom Modeler™ Model

A model has been developed in Aspen Custom Modeler™ to simulate the stripper operation. The model was designed for a wide variety of solvents but has currently been applied to a 7m MEA and a 5m K⁺ / 2.5m PZ solution.

Modeling Assumptions

- (a) The sections were assumed to be well mixed in the liquid and vapor phases.
- (b) The reboiler was assumed to be in equilibrium.

The CO₂ vapor pressure (kPa) under stripper conditions is represented by the empirical expression

$$\ln P = a + b * [\text{CO}_2]_T + \frac{c}{T} + d \frac{[\text{CO}_2]_T^2}{T^2} + e \frac{[\text{CO}_2]_T}{T^2} + f \frac{[\text{CO}_2]_T}{T} \quad (1)$$

P = the equilibrium partial pressure of CO₂ in kPa

T = temperature in Kelvin

[CO₂]_T = total CO₂ concentration (m)

The adjustable constants (Table 1) were obtained by regressing the points from the rigorous model for 5m K⁺ / 2.5m PZ by Cullinane and Rochelle (2004) and for MEA equilibrium flashes in AspenPlus™ using the rigorous model developed by Freguia (2002) from data of Jou and Mather (1995).

Table 1. Adjustable Constants in VLE Expression

	7m MEA	5m K ⁺ / 2.5m PZ
a	35.12	18.61
b	-6.43	2.49
c	-14280	-14210
d	-11150	-70614
e	-485800	468800
f	4670	589

The heat of absorption/desorption is calculated by differentiating Equation (1) with respect to 1/T. This is given by the following:

$$-\frac{\Delta H}{R} = c + 2d \frac{[\text{CO}_2]_T^2}{T} + 2e \frac{[\text{CO}_2]_T}{T} + f[\text{CO}_2]_T \quad (2)$$

where ΔH represents the heat of absorption/desorption [=] kcal/gmol CO₂, and

R is the Universal gas constant [=] cal/K-mol

The rich [CO₂]_T at specified rich P_{CO2} (kPa) leaving the absorber at 40°C and 60°C for both solvents is shown in Table 2.

The heat of vaporisation of water, partial pressure of water, heat capacities of steam, CO₂ and the solvent (essentially water) were calculated from equation derived from the DIPPR database.

The partial pressure of CO₂ and water leaving each section were calculated from Equation (3),

$$E_{mv} = \frac{P_n - P_{n-1}}{P_n^* - P_{n-1}} \quad (3)$$

where E_{mv} is the Murphree plate efficiency defined in terms of partial pressures

P_n, P_{n-1} are the partial pressures of the component on sections n and n-1

P_n^{*} is the equilibrium partial pressure of the component leaving section n.

Efficiencies of 40% and 100% were assigned to CO₂ and water. The model assumed 100% efficiency with respect to heat transfer.

Table 2. Rich [CO₂]_T at Absorber Conditions

Solvent	Rich P _{CO₂} (kPa)	[CO ₂] _T (m)	
		60°C	40°C
7m MEA	1.25	2.79	3.49
	2.5	3.08	3.76
	5	3.37	4.03
	10	3.67	4.30
5m K ⁺ /2.5m PZ	1.25	4.19	4.81
	2.5	4.46	5.07
	5	4.74	5.33
	10	5.02	5.58

The equivalent work is a convenient way to quantify the heat requirement of the process. It constitutes the work lost from the turbine upstream of the power plant since the condensing steam used to run the reboiler is no longer available to generate electric power. Assuming the enthalpy difference between the feed and products is negligible compared to the heat input and cooling water at 313 K is used to remove heat in the condenser, the equivalent work, W_{eq} , consumed by the process is given by

$$W \text{ (kcal/gmolCO}_2\text{)} = Q \left[\frac{T_{cond} - T_o}{T_{cond}} \right] + W_{comp} \quad (4)$$

where Q is the reboiler duty in kcal/gmol CO₂, T_{cond} is the temperature of the condensing steam (temperature of reboiler plus 10K) in the shell of the reboiler and T_o is the temperature of the cooling water (313K). W_{comp} constitutes the adiabatic work of compression of the gas exiting the top of the stripper to 300 kPa (an arbitrary pressure selected). For this analysis isentropic efficiency of the compressor was assumed to be 75%.

The model can be run as a stand-alone model in ACM or it can be run via Microsoft Excel when sensitivity analyses are performed. The ACM model has the rich [CO₂]_T, temperature approach and column operating pressures as inputs. Initial guesses of lean [CO₂]_T, section temperatures, pressures and concentrations are made. The model solves the VLE, material balance, and energy balances to give the gas rates exiting each section. It then returns the concentration, temperature, and composition profiles, reboiler duty and equivalent work as outputs. In order to minimize energy requirement a sensitivity analysis is carried out in Microsoft Excel using a Visual Basic code that links Excel to ACM and solves the same equations as the stand-alone ACM code but for a range of lean concentrations. This reports the reboiler duty and equivalent work for each lean concentration. Using the minimum function in Excel, the minimum reboiler duty and equivalent work are identified. The lean concentration that corresponds to minimum reboiler duty or equivalent work is not necessarily the same for a specified rich concentration. This means that the system can be optimised in two ways either to minimize reboiler duty or equivalent work. For our application, the steam will be withdrawn from the power plant so it is important to minimize the total equivalent work. The optimum capacity is defined as the difference between the rich [CO₂]_T and the lean [CO₂]_T that will minimize reboiler duty or equivalent work.

Results and Discussion

Predicted Stripper Performance

The stripper performance for a simple stripper operating at 160 kPa (Figure 1) with five compression stages to compress the gas to 1000kPa with intercooling to 313K between the stages downstream of the stripper was simulated.

Sensitivity analyses were carried out on 5m K⁺/2.5m PZ when the stripper operates at 160 kPa and the gas exiting the stripper is sent to a condenser that knocks out the water and the exiting CO₂ is further compressed to 1000 kPa. Figure 2 shows the results when the W_{eq} is minimized for the range of solvent concentrations given in Table 2. The results show that the reboiler duty and the equivalent work can be reduced by a factor of two over the entire range. The optimum capacity increases increasing rich [CO₂]_T. This is because at higher solvent concentrations CO₂ is easier to strip because it has a higher propensity to move into the vapor phase due to flashing that might occur. This reduces the amount of reboiler duty required and hence the equivalent work.

Figure 1. Simple Stripper

Figure 2. Minimizing W_{eq} for 5m K+/2.5m PZ, $P_T = 160$ kPa, $P_{final} = 1000$ kPa

The effect of varying the pressure at which the stripper operates was also studied. We expect that higher pressures in the stripper will lead to higher temperatures. Since the vapor pressure of CO₂ increases with pressure, increasing the pressure increases the amount of CO₂ that will go into the vapor phase. This should have some effect on the reboiler duty and equivalent work. The operating pressure should be one that will not raise the temperature to a level that will cause degradation of the solvent. For 7m MEA the maximum operating temperature is about 120°C. Operating the 5m K+/2.5m PZ at 300 kPa increased the CO₂ capacity over the entire range of rich solution concentrations by 0.2 m. The reboiler duty decreased by 18% at a rich solution concentration of 4.19 m to 0% at 5.58 m. No significant change was observed in the equivalent work over this range.

The effect of varying the temperature approach in the cross heat exchanger was investigated. Changing this approach from 10°C to 5°C constitutes using a larger heat exchanger. The results for the two solvents when the equivalent work is minimized are shown in Figure 3.

Figure 3. Effect of temperature approach, $P_T = 160\text{kPa}$, $P_{\text{final}} = 1000\text{kPa}$

At a fixed rich partial pressure of CO_2 in the rich solution, the 7m MEA system requires less energy than 5m K^+ /2.5m PZ. Using a larger heat exchanger i.e. operating the cross exchanger at 5°C instead of 10°C gives some benefit in terms of equivalent work. However it is not so significant. 5m K^+ /2.5m PZ can be run at a closer approach to saturation than 7m MEA. If the 5m K^+ /2.5m PZ is run at twice the pressure of 7m MEA the equivalent work requirements are comparable. This is shown in Table 3.

Table 3. Work Requirement for Solvents

Solvent	Rich P_{CO_2} (kPa) at 60°C	Total Weq (kcal /gmol CO_2)
7m MEA	2.5	13.4
5m K^+ /2.5m PZ	5	13.1
7m MEA	1.25	16.6
5m K^+ /2.5m PZ	2.5	16.7

Figure 4. McCabe-Thiele Plot ,Rich $[\text{CO}_2]_T = 4.46\text{m}$ using $5\text{m K}^+/2.5\text{m PZ}$

In order to determine driving forces, McCabe-Thiele plots were generated. Figures 4 and 5 show such plots for $5\text{m K}^+/2.5\text{m PZ}$ when a rich $[\text{CO}_2]_T$ of 4.46m , 5.07m respectively. In figure 4, a subcooled feed at 372K is fed to the stripper. The temperature at the top section of the stripper is 376K . The plot shows that there is a rather evenly distributed driving force in the stripper.

In Figure 5, the feed enters at 373K and flashes at the inlet of the stripper. This reduces the temperature to 368K after which the temperature gradually rises until the exit temperature of 383K is achieved. A rich end pinch (region of zero driving force) is experienced for this run. This shows that the height of contactor in this region achieves little or no removal since the partial pressure of CO_2 in those sections is essentially at equilibrium. The units of solution concentration are given in terms of loading.

Figure 5. McCabe-Thiele Plot, Rich $[\text{CO}_2]_T = 4.46\text{m}$ using $5\text{m K}^+/2.5\text{m PZ}$

Multipressure Stripping Results

In this configuration (Figure 6), the stripper is divided into stages with each operating at a different pressure. CO_2 compressors are integrated with the stripper. The vapor from a lower pressure stage is compressed and subsequently used as stripping vapor in a higher-pressure stage. The water vapor condenses as it passes through the higher pressure stages 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 both water vapor and CO_2 must be compressed.

In the current model, three pressure levels were selected for the stripper. The lowest pressure level was in the bottom of the stripper while the highest was at the top. The middle pressure was calculated by distributing the pressures geometrically. The vapor exiting the stripper is sent to a condenser to condense the water. The CO_2 is further compressed to 1000 kPa. The compression is carried out in five stages: two to raise the pressure in the stripper and three downstream of the stripper. Cooling water is used to intercool the gas downstream of the stripper. Figure 7 shows the effect of varying the maximum stripper pressure on equivalent work for a 7m MEA solvent with a rich $[\text{CO}_2]_T = 3.37\text{m}$. The result shows that increasing the pressure

from 160kPa to 500kPa reduces the equivalent work from 11.5 kcal/gmol CO₂ to 9.9 kcal/gmol CO₂. Further increasing the pressure to 600 kPa increases the equivalent work.

Figure 6. Multipressure Stripper

Figure 7. Maximum Pressure Effect for Multipressure Stripper using 7m MEA

The performance of the two process configurations were compared for a 7m MEA solution with a rich $[\text{CO}_2]_T = 3.37\text{m}$ with varying temperature approaches. The results are presented in Table 4.

Table 4. Performance of Stripper Configurations with 7m MEA, Rich $[\text{CO}_2]_T = 3.37\text{m}$

	kcal/gmol CO ₂					
Stripper	T _{app} (°C)	Q	W _{comp,int}	W _{comp,ext}	W _{comp,tot}	Total W _{eq}
Simple	10	38.4	0.0	1.7	1.7	11.5
(160kPa)	5	37.3	0.0	1.7	1.7	10.7
Multipressure	10	25.3	3.1	0.6	3.7	9.9
(500/280/160 kPa)	5	21.1	3.0	0.6	3.6	8.9

From the table, we see that the reboiler duty decreases by 35% when operating the multipressure stripper at 10°C and it decreases by 44% when operating at 5°C. The compression work however doubles. The total equivalent work is the sum of the equivalent work of the reboiler duty and the work of compression. The total equivalent work reduces by about 15% with the multipressure configuration when operating the stripper at fixed temperature approach.

Conclusions and Future Work

In this quarter, the ACM model has been developed. The work requirement has been minimized and applied to 7m MEA and 5m K⁺ / 2.5m PZ. The results show that at a closer approach to saturation, 7m MEA and 5m K⁺ / 2.5m PZ require comparable energy. The capacities of the two solvents are also comparable. The equivalent work is minimized by a lean solution $P_{\text{CO}_2^*} \sim 10\text{kPa}$ at 110°C for a simple stripper. The multipressure stripper reduces equivalent work by $\sim 15\%$ when compared to the simple stripper at a fixed temperature approach. For most practical situations, the solvent is over stripped in order to minimize energy requirement. In the next quarter, a model for a generic solvent will be developed and initial packing height estimations will be calculated. This will involve some rigorous rate based modeling. The second campaign of the pilot plant will be carried out.

Subtask 1.3b – Develop Integrated Absorber/Stripper Model

By Jennifer Lu
(Supported by this contract)

Introduction

This report presents the continuing development of an absorber model in Aspen Plus™ for use in modeling the pilot plant. A previous model was based on the existing 30% MEA-system model by Freguia (2002) and altered in order to make the MEA system emulate a system that uses potassium carbonate promoted by piperazine. For a more rigorous approach, the model is being converted to a piperazine system. This report details these continuing efforts.

Experimental

In the previous model, the rate constant of the interface pseudo first order approximation is increased by a factor of two and the temperature dependence of the carbamate equilibrium constant was set to zero, in order to decrease the heat of absorption of CO₂ by approximately 5 kcal/mol. This was done in order to make the MEA system behave like a 2.5m PZ/5m K⁺ solvent.

In this work, the MEA system was converted to a piperazine system. The MEA species were changed to their corresponding PZ species, from MEA to PZ, from MEA⁺ to PZ⁺, and so on. The dicarbamate, PZ(COO⁻)₂, and the protonated carbamate, HPZCOO, were also added. The MEA reactions were replaced by the following PZ reactions:

Potassium carbonate, potassium bicarbonate, and the potassium ion were also added to the component list of the simulation. The dissociation reactions for potassium carbonate and potassium bicarbonate were also added.

The model uses a FORTRAN subroutine to calculate reaction rates. The reactions for the MEA model are as follows:

A PZ model would have three reactions, one for the carbamate species, the dicarbamate species, and the bicarbonate.

Equilibrium constants for reactions (13) and (14) may be obtained by combining reactions (6) and (7) with the following reaction for bicarbonate:

The equilibrium constant for reaction (15) is obtained by combining reaction (16) with the water dissociation reaction.

In the Freguia (2002) FORTRAN subroutine, the CO_2 flux is given by the following equation:

$$N_{\text{CO}_2} = \sqrt{(k_{2,\text{MEA}} [\text{MEA}]_i + k_{2,\text{OH}^-} [\text{OH}^-]_i) \frac{D_{\text{CO}_2}}{\gamma_{\text{CO}_2}} (a_{\text{CO}_2,i} - a_{\text{CO}_2,i}^*)} \quad (17)$$

where

$$a_{\text{CO}_2,i}^* = \frac{\frac{k_{2,\text{MEA}}}{K_{\text{CO}_2-\text{MEA}}} \frac{a_{\text{MEACOO}^-} a_{\text{H}_3\text{O}^+}}{a_{\text{H}_2\text{O}} \gamma_{\text{MEA}}} + \frac{k_{2,\text{OH}^-}}{K_{\text{CO}_2-\text{OH}^-}} \frac{a_{\text{HCO}_3^-}}{\gamma_{\text{OH}^-}}}{k_{2,\text{MEA}} \gamma_{\text{MEA}} + k_{2,\text{OH}^-} \gamma_{\text{OH}^-}} \quad (18)$$

For the PZ system, with its three kinetic reactions, the flux will have three terms, as follows:

$$N_{\text{CO}_2} = \sqrt{(k_{2,\text{PZ}} [\text{PZ}]_i + k_{2,\text{PZCOO}^-} [\text{PZCOO}^-] + k_{2,\text{OH}^-} [\text{OH}^-]_i) \frac{D_{\text{CO}_2}}{\gamma_{\text{CO}_2}} (a_{\text{CO}_2,i} - a_{\text{CO}_2,i}^*)} \quad (19)$$

where

$$a_{\text{CO}_2, i}^* = \frac{\frac{k_{2, \text{PZ}}}{K_{\text{CO}_2 - \text{PZ}}} \frac{a_{\text{PZCOO}^-} a_{\text{H}_3\text{O}^+}}{a_{\text{H}_2\text{O}} \gamma_{\text{PZ}}} + \frac{k_{2, \text{PZCOO}^-}}{K_{\text{CO}_2 - \text{PZCOO}^-}} \frac{a_{\text{PZ}(\text{COO}^-)_2} a_{\text{H}_3\text{O}^+}}{a_{\text{H}_2\text{O}} \gamma_{\text{PZ}(\text{COO}^-)_2}} + \frac{k_{2, \text{OH}^-}}{K_{\text{CO}_2 - \text{OH}^-}} \frac{a_{\text{HCO}_3^-}}{\gamma_{\text{OH}^-}}}{k_{2, \text{PZ}} \gamma_{\text{PZ}} + k_{2, \text{PZCOO}^-} \gamma_{\text{PZCOO}^-} + k_{2, \text{OH}^-} \gamma_{\text{OH}^-}} \quad (20)$$

The subroutine must be rewritten to account for the third reaction and the change in the equilibrium interfacial activity coefficient for CO_2 .

The Freguia (2002) FORTRAN subroutine uses the Onda correlation to calculate the wetted area in order to calculate the film thickness. However, the Onda correlation only applies to random packings that are larger than 3 inches. Additionally, the correlation does not account for the dependence on the gas rate. Wilson (2004) has shown that the wetted area does indeed depend on both the gas rate and the liquid rate. Thus, a new subroutine must be used to calculate the wetted area.

Discussion

Five tasks have been identified to move the model to a more rigorous approach. These are summarized in Table 5.

Table 5. Identified Tasks for Making the Current Absorber Model More Rigorous

Task	
(1) Replace MEA species/add PZ species	✓
(2) Replace MEA reactions/add PZ reactions	✓
(3) Develop interfacial area subroutine	✓
(4) Regress interfacial area data	
(5) Rewrite kinetics subroutine	

The first three tasks have been completed. The MEA species have been replaced and the additional PZ species have been added. Regressed interaction parameters have been taken from results of Task 1.1. The MEA reactions have been replaced with PZ reactions and the additional PZ reactions added. Equilibrium constants have also been taken from Task 1.1.

The interfacial area subroutine is overly simplistic. Currently, the subroutine allows the user to input the specific wetted area directly. (The kinetic subroutine has also been modified to allow the user to input the specific wetted area. This number must be inputted into both places.) The liquid and gas rates are calculated and the wetted area can be found from the graphs developed by Wilson (2004). The wetted area is then input into the subroutine and new liquid and gas rates are calculated. The new wetted area is then found from the graphs. The process is iterative and can become tedious at times. Thus, the data should be regressed in order to code a subroutine that will calculate the wetted area for a number of different packings.

Freguia's (2002) subroutine begins by calculating the density and molecular weight of the liquid, in order to find a total concentration. The viscosity and diffusion coefficients for CO₂ are then defined for the solvent. The rate constants and equilibrium constants are then given. Variables are defined for the activity coefficient of CO₂, the equilibrium activity of CO₂ for the carbamate reaction, and the equilibrium activity of CO₂ for the bicarbonate reaction. The liquid holdup and the wetted area are calculated to give a film thickness. Then the rates are calculated using the film thickness and the variables are returned to Aspen.

The main task is to rewrite the kinetics subroutine. Work has already begun, but is still in the preliminary stages.

Future Work

As the kinetics and interfacial area subroutines are completed, pilot plant performance will be modeled. The equilibrium constants, rate constants, and interaction parameters will be updated, as refinements become available from detailed thermodynamic (Task 1.1) and rate (Task 1.2) modeling.

Subtask 1.5 – Simulate Base Case Pilot

By Eric Chen

(Supported by EPA Star Fellowship)

Results and Discussion

Material balance

Figure 8 gives results for a material balance around the absorber. The capacity (moles CO₂ removed/kg solvent) can be calculated from either the gas phase balance or the liquid phase balance. The gas phase balance uses the inlet and outlet CO₂ concentrations measured by the IR analyzers, the gas flow rate, and the liquid flow rate. The liquid phase balance uses the liquid flow rate and the CO₂ concentration in the rich and lean solution determined by the inorganic carbon analyzer.

The gas capacity is consistently 30 to 40% greater than the liquid capacity. We believe that the analysis of the rich solution underreported the actual CO₂ concentration because CO₂ was lost from the solution during sampling and sample dilution. The samples were collected hot, directly from the process into an open beaker. The samples were diluted by taking 1 ml of sample into an open beaker, then adding 1 liter of water. These procedures will be modified in Campaign 2 to minimize loss of CO₂. For Campaign 1 the rich solution loading will be calculated from the gas capacity and the lean solution loading.

We also found inconsistencies in measurements of lean solution CO₂ loading. Duplicate analyses of sample at the inorganic carbon analyzer on campus produced values that were 15% lower than the values determined by the carbon analyzer at the Pickle Research Center. We have adjusted the measured values to be consistent with the campus analyzer, because we have been using it successfully in all of our rate and thermodynamic studies and because its results are more consistent in the analyses that follow.

Absorber Performance

Figure 9 gives a revised analysis of the 17 runs from Campaign 1. CO₂ penetration is given as a function of the solution capacity (mol CO₂/kg solvent) calculated from the gas phase balance. The log of penetration is linearly correlated with the capacity. With 3% CO₂ the penetration varied from 0.2 to 12% as the L/G varied from 4.5 to 0.9 kg/kg. With 12 % CO₂ the penetration varied from 3 to 12% as the L/G varied from 5.6 to 3.4 kg/kg. In the practical range of absorber operation at the high capacity required to minimize the steam rate for the stripper, the penetration is about 12%. The maximum capacity observed was about 1.15 mol CO₂/kg solvent. It was limited by a rich end pinch (by the equilibrium of the rich solution with the inlet gas).

Figure 8. Absorber Material Balance

Figure 9. Absorber Performance - 17 Runs, Lean Loading = 0.4-0.5 (mol CO₂/(K+2PZ))

McCabe–Thiele Analysis

Figure 10 illustrates a typical rich end pinch with a McCabe-Thiele diagram. Sets of points are given for the top, middle, and bottom of the absorber. The measured CO_2 loading is used for the lean solution. The solution loading at the middle and bottom of the column is calculated from the measured lean loading and the material balance with the measured gas compositions, gas rate, and liquid rate. The equilibrium points are calculated at the respective solution loading and measured temperature from an empirical correlation of Cullinane's (2004) equilibrium data. With this low L/G the gas leaving the absorber is practically in equilibrium with the rich solution, so it is not possible to calculate a mass transfer coefficient from the bottom half of the column. However, the two data points from the top half of the column are useful for estimating a mass transfer coefficient.

Figure 10. McCabe-Thiele – Rich End Pinch

Figure 11 illustrates a lean end pinch. With a greater L/G , the gas exiting the absorber is practically in equilibrium with the inlet lean solution. In this case it is only possible to calculate a mass transfer coefficient around the bottom of the column.

Figure 11. McCabe-Thiele – Lean End Pinch

Figure 12 shows all data points collected in Campaign 1 that correspond to pinch. The gas concentration is plotted as a function of its respective loading and compared to CO₂ vapor pressure calculated from Cullinane's (2004) equilibrium model. There are 12 of the 32 possible data sets that are at the pinched condition. The other 20 data sets should be useful for estimating mass transfer coefficients.

Figure 12. Absorber McCabe Pinch Points, 12/32 Possible Data Pairs

Figure 13 illustrates a run with a distributed driving force. There are no apparent pinches, so it should be possible to calculate mass transfer coefficients for both the top and bottom of the column.

Figure 13. McCabe Thiele Plot – Run 1.7, $G = 0.65 \text{ kg/m}^2\text{-s}$, $L/G = 4.1$, $PZ = 1.4 \text{ mol/kg}$ Solvent

The overall gas phase mass transfer coefficient was calculated for each pair of points that did not have a pinch. Each run has the potential for two pairs of points, the top of the column and the bottom of the column. The driving force was estimated as the logarithmic mean of the driving force at either end of the column section and the K_gA is given by:

$$N_G = \frac{\ln \left[\frac{CO_{2,IN} - CO_{2,IN}^*}{CO_{2,OUT} - CO_{2,OUT}^*} \right]}{\frac{((CO_{2,IN} - CO_{2,IN}^*) - (CO_{2,OUT} - CO_{2,OUT}^*))}{(CO_{2,IN} - CO_{2,OUT})}} = \frac{K_G a}{G}$$

The effective wetted area of the packing, A , was estimated from the KOH experiments presented in the last quarterly progress report. Figure 14 shows the K_G calculated for 16 data pairs as well as values of the gas phase mass transfer coefficient, k_g' taken directly from Cullinane's experimental work with the wetted wall column. The line is a fit of the data from the wetted wall column. We would expect K_g and k_g' to be equal in the absence of gas film resistance. Most of

the pilot plant results with 3% CO₂ are consistent with the data from the wetted wall column. However the data with 12% CO₂ are less consistent and generally lower than expected. This analysis does not account accurately for the effects of the temperature bulge, which is larger with 12% CO₂. The data with 12% CO₂ are also mostly at greater liquid rates, which may result in poor gas/liquid distribution in the structure packing.

Figure 14. Gas Phase Mass Transfer from the Pilot Plant (16 of 32 Possible Data Pairs) and the Wetted Wall Column (Cullinane) Reference???

Absorber Temperature Profile

Figure 15 gives the surface temperature profile of the absorber for three runs, all with a lean end pinch. The maximum temperature was observed with 12% CO₂ near the top of the absorber. No data are reported in the five-foot section in the middle of the column which contains a collection plate and redistributor.

Figure 15. Absorber Temperature Profile, $L = 2.6$ & $2.7 \text{ kg/m}^2\text{-s}$ (3 & 12%), $P = 0.2\%$ (3% CO_2), 5% (12% CO_2), Lean-End Pinch All Three Cases

Conclusions and Future Work

The bench-scale rate data have been confirmed by experiments in the absorber with 3% CO_2 . A more non-isothermal analysis will be performed to validate rates with the data for 12% CO_2 .

The CO_2 penetration through the absorber increases at lower liquid rates, producing greater solution capacity. The maximum solution capacity was about 1 gmol CO_2/kg solvent.

The absorber material balance suggests that CO_2 losses occurred in the sampling and analysis of the rich solution. Therefore the sampling procedure will be modified for Campaign 2.

Subtask 1.9 – Economic and Applicability Analysis

By Gary T. Rochelle (supported by the State of Texas)

The applicability of the K_2CO_3 /piperazine solvent will depend on avoidance of corrosion, solvent makeup costs, and acceptable disposal of the spent solvent. These issues will be the topics of a proposal for a contract to be completed by the end of November. The discussion that follows addresses the potential environmental impact of vanadium, the most probable corrosion inhibitor for this system.

V^{+5} is a known effective corrosion inhibitor for the hot potassium carbonate process at concentration from 500 to 5000 ppm. Bienstock et al. (1961) showed that sodium metavanadate at 0.2% (830 ppm V) eliminated corrosion of carbon steel in boiling solutions of 40 wt% potassium carbonate saturated to CO_2 . It was ineffective at 0.05%. Its use in processes for CO_2 capture may complicate the disposal or reuse of spent solvent. As a “heavy metal” it will immediately raise flags because of potential environmental impact, whether warranted or not. There appears to be no evidence that vanadium has effects like heavy metals such as mercury or lead. It may rather be in the same category as iron or manganese.

Vanadium is listed as a hazardous substance. The §268.48 Universal treatment standards set a value of 4.3 mg/liter as a standard for wastewater (www.setonresourcecenter.com/cfr/40CFR/P268_035.htm) and 1.6 mg/liter TCLP (www.environmental.usace.army.mil/info/technical/comply/complguide/complyfs/98-04/98-04.html).

Although vanadium is listed as a hazardous substance by ATSDR and EPA, there appears to be no significant health risk from oral consumption. Most of the quantified health effects result from inhalation of vanadium pentoxide dust. Because vanadium is nonvolatile and we expect no particulate emissions, the primary risk associated with use of vanadium is contamination of drinking water. Because “the gastrointestinal absorption [of vanadium] is so low, the health implications for people drinking water [contaminated with vanadium] are not readily apparent” (USPHS, 1992). “EPA has decided that if you eat less than 9 micrograms (μg) of vanadium pentoxide per kilogram (kg) of your body weight, your health is protected” (USPHS, 1992). There are no reports of death in humans following ...oral ... exposure to vanadium” (USPHS, 1992). The LD_{50} for rats and mice is 41 and 31 mg V/kg/day, respectively (Llobet and Domingo, 1984). Human volunteers given 1.3 mg V/kg/day for 45 days had intestinal cramping and diarrhea (Dimond et al., 1963). This dosage is equivalent to about 45 ppm in drinking water. Workers who have been exposed to vanadium dusts did not show an increased number of cancer deaths (USPHS, 1992). When animals are exposed to vanadium, “prolonged retention of vanadium occurs only in the skeleton” (USPHS, 1992).

There appears to be no significant risk of vanadium to plants. Rock phosphate and superphosphate fertilizers contain 10 to 2000 ppm vanadium (Van Zinderen Bakker and Jaworski, 1980). Vanadium is a micronutrient for plant growth. Commonly applied fertilizers with phosphate contain 14 – 949 ppm vanadium (Shaffer, 2001).

Therefore, the use of vanadium as a corrosion inhibitor will carry little risk of long term environmental or health effects liability, but the disposal of solvent with vanadium will be subject to regulation, probably as a hazardous waste. It will be easier to deal with waste solvent if it has less than 4.3 ppm vanadium.

Subtask 1.10 – Simulate MEA Baseline Testing

By Ross Dugas
(Supported by this contract)

Introduction

This report contains simulated results from the MEA absorption/stripping model developed by Freguia (2002). The Aspen Plus model contains a rate-based absorber and an equilibrium-based stripper. Both the absorber and the stripper were modeled via Aspen's Ratefrac block. Freguia's MEA model was adjusted to better represent conditions in the pilot plant instead of industrial systems, its original aim. The system was then run with a wide variety of operating conditions to quantify absorber and stripper performance.

Experimental

The kinetics of Freguia's MEA absorption/stripping model were modified slightly to increase HCO_3^- formation in the absorber. It was found that the original code underestimated the bicarbonate formation.

Listed below are some of the common model specifications that are applicable to all simulation cases:

- Flue Gas – 55°C, 0.1 psig, $\approx 10\%$ H_2O
- Lean Liquid - 40°C, 30 wt% (7 molal) MEA
- Absorber – 1 atm, 24.6' Height
- Stripper – 20' Height
- Diameter – 168" for absorber, stripper
- CMR #2 packing in absorber, stripper
- Temp Approach of Cross Exchanger - 10°C

Although the absorber and stripper column diameters are 168" in the model, the columns are only 16.8" in the pilot plant. Convergence issues were encountered when reducing the diameter from an industrial scale to a pilot plant scale. 168" was a convenient size since all flow rates for the pilot plant would be 1/100 of those inputs into the model. Therefore, all liquid and gas flow rates reported are not those in the model but the proportional flow rates of a 16.8" column.

Listed below are the varying operating conditions that were modified in the simulation:

- 3%, 12% CO_2
- 0.2, 0.3 Lean Loadings
- 0.5, 1 atm Stripper Pressures
- 250, 500 acfm
- 0.875 – 40 gpm

Results and Discussion

Freguia's original code and the new code give similar absorption results at very low liquid holdups, $\approx 0.01\%$, but different results at higher holdups. However, this new code did not resolve the difficulty in converging cases with liquid holdups greater than 0.2% . A maximum liquid holdup of 0.42% was obtained but was very time consuming. Therefore, all of the cases shown in this report are for liquid holdups of 0.01% . However it can be seen that a holdup greater than $.15\%$ will effectively model the system. It is important to remember that these are not actual holdups but simulated parameters to simulate actual conditions. Results showing the sensitivity of this parameter can be seen below in Figure 16.

Figure 16. Influence of Liquid Holdup on the Performance of the Absorber

The absorber performance for all simulated cases can be seen below in Figure 17. It can be seen that the absorber performance increases with increased liquid rate, decreased gas rate, and decreased lean loading. This is all intuitional but the graph also quantifies these effects as well as shows conditions which will yield rich end pinches. These rich end pinches are located where the penetration approaches 1 and the performance curves start to merge for varying gas rates.

Figure 17. Absorber Performance of Simulated Results

Future Work

The next step of this work is to analyze the stripper performance of the model. Other column internals also need to be analyzed in the simulation. The current simulations only include CMR #2 but a metal structured packing and a random plastic packing will be analyzed. Once these factors are sufficiently understood, a matrix of operating conditions for the MEA pilot plant campaign will be constructed and the MEA campaign will be ready to commence.

Task 2 – Pilot Plant Testing

Subtask 2.1 – Pilot Plant Test Plan – Campaign 2

by Eric Chen

(Supported by EPA Star Fellowship)

Objectives

The first objective will be to obtain a more complete data set on absorber performance as a function of gas rate, liquid rate, CO₂ concentration, and CO₂ lean loading with Flexipac™ 1Y and IMTP#40 packing, including the data required to optimize the absorber gas velocity. Due to the problems encountered during Campaign 1, a portion of Campaign 2 will be devoted to obtaining additional data with Flexipac™ 1Y.

The second objective will be to obtain performance data for the stripper over a range of solvent rates, rich loadings, and stripper pressures. As originally planned, the data will be used to optimize stripper performance.

Post Campaign 1 Issues

Before the commencement of Campaign 2, a number of issues will need to be resolved. The two online Rosemount pH meters that failed have been replaced. The original probes and transmitters were sent to Rosemount Analytical for examination, but it appeared that nothing was wrong. The probes and transmitters were not designed for outdoor use. Therefore, it was suspected that the heavy rains had shorted the connection between the probe and transmitter, causing the pH meters to malfunction. Rosemount Analytical sent two new pH meters with one of the probes hardwired to the transmitter. A shelter will be constructed to protect the non-hardwired pH probe and transmitter.

At 12% CO₂, the 0-5% Horiba analyzer was over-ranged when taking measurements at the middle of the absorber. An additional Horiba CO₂ analyzer has been procured for analysis in the range of 0-20% CO₂ to expand the measuring range at the middle and the outlet of the absorber.

At the conclusion of the first campaign, it was discovered that the polyethylene filter bag for the reboiler had completely dissolved. The filters will need to be replaced with materials that are compatible with the potassium carbonate and piperazine solvent. The filter had previously been used in a C6/C7 system with no adverse effects at the approximately the same temperatures (~245 °F). Therefore, temperature should not have been an issue.

It was suspected that there was CO₂ loss during liquid sampling and analysis preparation. A new sampling procedure involving sample bombs have been developed and will be tested.

The steam traps on the stripper feed heater were undersized in the first campaign and as a result the solvent was not adequately heated. New steam traps will be installed on the stripper feed heater to rectify the problem. In addition, a larger orifice will be installed for steam flow measurement to the reboiler. In the first campaign, the steam flow rates exceeded the measuring

range of the orifice. Finally, cooling water to the air cooler will be blinded off to permit operations at higher gas temperatures.

Experimental Setup

In the second campaign, the absorber will first be packed with 20 ft of Flexipac™ 1Y. In the third week of the campaign, the absorber packing will be replaced with 20 ft of IMTP #40 random packing. The new packing has less surface area than the Flexipac™ 1Y, which will allow us to run at lower lean loading and still maintain modest CO₂ removal rates. The main focus of the second campaign will be to obtain data for the stripper. The sieve trays in stripper will be replaced with 20 feet of IMTP #40 random packing, divided into two beds. A collector place plate and redistributor will be placed in between the two beds. Additional RTD's will be installed in the absorber column to measure the temperature profile.

Sampling/Analytical Procedures

There will be a total of five liquid sample points: absorber lean, absorber middle, absorber rich, stripper middle, and stripper lean. All of the samples will be withdrawn using 10 ml sample bombs with Swagelok® quick-connects. The samples will then be brought inside the laboratory. A syringe will be used to extract the liquid samples from the sample bomb. The samples will be injected and diluted into vial containing chilled deionized (DI) water. Two sample sets will be taken for each operating condition.

The liquid samples will be analyzed for CO₂ loading, piperazine and potassium concentration, total alkalinity, total iron, vanadium, and possibly some degradation products. CO₂ loading measurements will be conducted with the Total Organic Carbon (TOC) analyzer, using the total inorganic carbon (TIC) analysis portion. Standards will be placed in between every ten samples to monitor analyzer drift. In addition, the samples will be split every five samples and analyzed with the on campus IC analyzer to validate the results from the TIC.

Piperazine and potassium concentration and total alkalinity will be determined by titration with 0.5 N NaOH and 0.5 N HCl. The TOC will also be used to determine the piperazine concentration. Total iron, vanadium, and potassium will be determined by Ion Coupled Plasma (ICP) analysis.

Online gas phase concentration of CO₂ will be measured at the inlet, middle, and outlet of the absorber. The inlet and outlet concentration will be measured in-situ with Vaisala CO₂ analyzers. The middle CO₂ gas concentration will be measured with Horiba CO₂ analyzers via continuous sample extraction. Two Horiba CO₂ analyzers will be used to measure the middle CO₂ gas concentration. The maximum measurement ranges of the analyzers are 0-5% and 0-20%, respectively.

Test Schedule

The first week of Campaign 2 will be dedicated to the setup and re-configuration of the pilot plant (Table 6). The system will be loaded with diluted solvent stored from the first campaign. Water will be evaporated and removed from the system to give solvent with a density of 1220 g/liter, corresponding to 5 m K⁺ /2.5 m piperazine.

In the second week, operations will begin with Flexipac™ 1Y in the absorber and will last for approximately four days, with two 12-hour and two 24-hour runs (Table 7). For each operating condition, the experimental run will last approximately four hours. Therefore, for a 12-hour run, a total of three experiments will be conducted. Operations with the Flexipac™ 1Y will duplicate points from Campaign 1 and additional runs will be conducted at higher gas rates.

In week 3, the structured packing in the absorber will be replaced with IMTP#40 random packing. Pilot operations will resume with the IMTP#40 packing on day 4 with two days of 12-hour operations (Table 8). Week 4 will be dedicated to 24-hour operations. In week 5, two additional days of 24-hour operations will be conducted and clean up will begin on day 3. On day 4, the effective area of the absorber will be determined by the addition of sodium hydroxide (0.1N NaOH) to the clean rinse water. The NaOH solution will be analyzed for residual piperazine and potassium carbonate as a way of validating the rinse process.

In each of the experiments, the gas rate will be set to give a pressure drop of approximately 0.5 or 1.0 inches of H₂O per foot of packing. The liquid rate will be adjusted to meet one of three criteria: 90% CO₂ removal, absorber distributed, and stripper optimized. In order to satisfy the absorber distributed criteria, the absorber must be operated without any pinches. This will be accomplished by adjusting the liquid flow rate until the CO₂ gas concentration at the middle of the absorber is approximately the geometric mean of the inlet and outlet. Stripper optimized means that the stripper is operating without any pinches. The temperature profile of the stripper will be monitored and the liquid flow rate will be adjusted until the temperature gradient is distributed across the column.

During the first week of operations with the Flexipac™ 1Y packing, pilot plant operations once the CO₂ loading has reached 0.5 total alkalinity (TAlk). The system will be operated without the addition of CO₂ if possible. The loading will be allowed to drift and CO₂ losses will be monitored.

For the IMTP#40 runs, three sets of CO₂ loadings will be conducted. The IMTP#40 campaign will begin at a loading of 0.5 total alkalinity and will be allowed to drift downward. The focus of Campaign 2 will be on the loading range of 0.4-0.45. Therefore, CO₂ will be periodically added to the system if the loading drifts below 0.4 TAlk.

The stripper will be operated at a pressure of 1.6 atm for most of the campaign. However, attempts will be made to operate at 4 atm during the course Campaign 2. Vacuum operations require mechanical change-outs and thus will be attempted towards the end of the 0.4-0.45 loading campaign segment. The final and lowest set of loading experiments will be attempted at the 1.6-4 atm range without switching back to the original plant configuration.

Table 6. Campaign 2 Schedule

Week	Day 1	Day 2	Day 3	Day 4	Day 5
Oct. 11	Absorber/Stripper Setup				
Oct. 18	Troubleshoot	Flexipac 1Y 12hrs		Flexipac 1Y 24hrs	
Oct. 25	Packing Change Out			IMTP#40 12hrs	
Nov. 1	IMTP#40 24hrs				
Nov. 8	IMTP#40 24hrs		Rinse	NaOH	Rinse

Table 7. Campaign 2 Operating Matrix for Flexipac™ 1Y

CO2 In %	Gas Rate acfm	Ldg mol/TAlk	Liq Rate ^{1,2,3} gpm	T Gas C	T Liq C	P Strip atm
12	160	0.5	5	40	40	1.6
12	250	0.5	7.75	40	40	1.6
12	300	0.5	7.75	40	40	1.6
12	700	0.5	90 % Removal	40	40	1.6
12	700	0.5	Distributed	40	40	1.6
12	700	0.5	Stripper Optimum	40	40	1.6
12	900	0.5	90% Removal	40	40	1.6
12	900	0.5	Distributed	40	40	1.6
12	900	0.5	Stripper Optimum	40	40	1.6

Table 8. Campaign 2 Operating Matrix for IMTP#40

CO2 In %	Gas Rate acfm	Ldg mol/TAlk	Liq Rate ^{1,2,3}	T Gas C	T Liq C	P Strip atm
3	700	0.5	90% Removal	40	40	1.6
3	700	0.5	Distributed	40	40	1.6
3	700	0.5	Stripper Optimum	40	40	1.6
3	1000	0.5	90% Removal	40	40	1.6
3	1000	0.5	Distributed	40	40	1.6
3	1000	0.5	Stripper Optimum	40	40	1.6
12	700	0.5	90% Removal	40	40	1.6
12	700	0.5	Distributed	40	40	1.6
12	700	0.5	Stripper Optimum	40	40	1.6
12	1000	0.5	90% Removal	40	40	1.6
12	1000	0.5	Distributed	40	40	1.6
12	1000	0.5	Stripper Optimum	40	40	1.6
Loading Change						
3	700	0.4-0.45	90% Removal	40	40	1.6
3	700	0.4-0.45	Distributed	40	40	1.6
3	700	0.4-0.45	Stripper Optimum	40	40	1 - 4
3	1000	0.4-0.45	90% Removal	40	40	1.6
3	1000	0.4-0.45	Distributed	40	40	1.6
3	1000	0.4-0.45	Stripper Optimum	40	40	1 - 4
12	700	0.4-0.45	90% Removal	40	40	1.6
12	700	0.4-0.45	Distributed	40	40	1.6
12	700	0.4-0.45	Stripper Optimum	40	40	1 - 4
12	1000	0.4-0.45	90% Removal	40	40	1.6
12	1000	0.4-0.45	Distributed	40	40	1.6

CO2 In %	Gas Rate acfm	Ldg mol/Talk	Liq Rate ^{1,2,3}	T Gas C	T Liq C	P Strip atm
12	1000	0.4-0.45	Stripper Optimum	40	40	Min
Loading Change						
3	700	0.35	90% Removal	40	40	0.2-4.0
3	700	0.35	Stripper Optimum	40	40	0.2
3	1000	0.35	90% Removal	40	40	1.6
3	1000	0.35	Distributed	40	40	1.6
3	1000	0.35	Stripper Optimum	40	40	1.6
12	700	0.35	90% Removal	40	40	1.6
12	700	0.35	Distributed	40	40	1.6
12	700	0.35	Stripper Optimum	40	40	1.6
12	1000	0.35	90% Removal	40	40	1.6
12	1000	0.35	Distributed	40	40	1.6
12	1000	0.35	Stripper Optimum	40	40	1.6

Notes:

1. 90% Removal – 90% CO₂ removal across absorber
2. Distributed – Middle CO₂ gas concentration in between inlet & outlet
3. Stripper Optimum – Temperature profile is distributed across stripper

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