

# Dynamic Operations of the Stripping Column of a CO<sub>2</sub> Capture Plant

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## Abstract

Post combustion CO<sub>2</sub> capturing holds an important position in the area of carbon capture and sequestration (CCS). Research is operating in this area ranging from experimental work to modelling work. Dynamic models are interesting since these describe the plant operation during variations, up-stream or down-stream. A model for the stripping column of a capture plant is developed following the rate based approach to represent the heat and mass transfer. Sensitivity of the model for different physiochemical property correlations is analyzed. The predictions of the dynamic model for the stripping column of the capture plant under varying operating conditions in the re-boiler are presented. Predictions of the transient behaviour of the developed stripper model appear realistic.

**Keywords:** CO<sub>2</sub> capture; Dynamic model; Stripping column; MATLAB

## 1. Introduction

Achieving a sound solution for the CO<sub>2</sub> emission reduction is of interest and a challenge for the researchers who are actively involved in dealing with the climate issue.

Power generation via fossil fuel-fired power plants is known to be the largest single source of CO<sub>2</sub> emission [1]. The development of capture technologies targeting such sources therefore is important for achieving the goals in CO<sub>2</sub> emission reduction. Post-combustion capture, pre-combustion capture and oxy-fuel combustion are the three main technologies available at present [2], and much research is done with the prospect of developing those techniques further. Post combustion capture is still the best known technique, possibly due to the large number of existing power plants, and the promising developments that are available. CO<sub>2</sub> capture by amine absorption and stripping is currently considered to be the most feasible option for the removal of carbon dioxide from the power plants' exhaust gases [3].

Modelling work related to CO<sub>2</sub> capture technologies plays an important role with respect to the design, control and optimization of the capture process. Several steady state models are already in use for design and optimization purposes, but dynamic simulation models are scarce. Development of dynamic models is important since there is a demand for information related to the dynamics of a plant, such as the transient conditions during the start-up and shut-down and the operation of the plant under varying loads.

A CO<sub>2</sub> capture plant with solvent absorption (absorber/ stripper process) consists of several units which will interact and eventually influence the control and optimization of the process. The operation of a power plant always depends on the power demand and availability of other utilities which can lead to variations in its load. The load variations in the up-stream power plant cause varying exhaust gas rates which may cause operating challenges for the CO<sub>2</sub> capture plant. A dynamic simulation model should predict the influence of the components of the capture plant on each other. Further, the model should predict the influence from the up-stream power plant on the operation of the capture plant, when the power plant is running under varying load conditions.

In recent years, the popularity of the non-equilibrium rate based approach (NEQ) for column modelling has surpassed that of the equilibrium stage approach (EQ), where the liquid and vapour phases are assumed to be in equilibrium. The EQ stage and efficiency based approaches are not very accurate because columns rarely, if ever, operate at equilibrium in actual operation. Further, heat and mass transfer are actually rate based processes that are driven by the gradients of temperature and chemical potentials. In the NEQ approach, the finite mass transfer rates across the liquid-vapour interface are considered. The NEQ approach has been introduced in steady state simulators like Aspen Plus [4] and is even more appropriate for dynamic models.

The dynamic models developed for representing industrial stripping processes have to be accurate and rigorous in order to give insight in the complex transient conditions, and simple to ensure the feasibility of the process simulations [5]. Simplicity will ensure higher execution speed, which is an important factor since simulation of transient conditions is much more complex than steady state simulations. Therefore, it is important to make the model as simple as possible yet good enough to be used in a dynamic simulator. Use of simple thermodynamics can be considered as a wise strategy for developing a dynamic model to ensure simplicity. Inclusion of simple equilibrium models such as the Kent-Eisenberg model [6] instead of much more complex models like e-NRTL [7] in a dynamic model is a good example where the proposed simple equilibrium model will introduce simplicity to the model as well.

Validation of the predictions of the dynamic model still remains a challenge due to the absence of appropriate experimental data. Comparing the predicted steady states from such a dynamic model with the experimental data available in the literature is a possible first step in validating the model.

Little work exists on the development of dynamic models of the stripping tower of a post combustion CO<sub>2</sub> capture plant [8]-[11]. The majority of the existing models have used a complex thermodynamic model which increases the complexity of the models.

In the present work a dynamic model is developed for a stripping tower of a MEA (Mono-Ethanol-Amine) based CO<sub>2</sub> capture plant from the flue gas, following the NEQ approach, and implemented in MATLAB. The structure of the stripping tower considered for modelling is given in Figure 1.

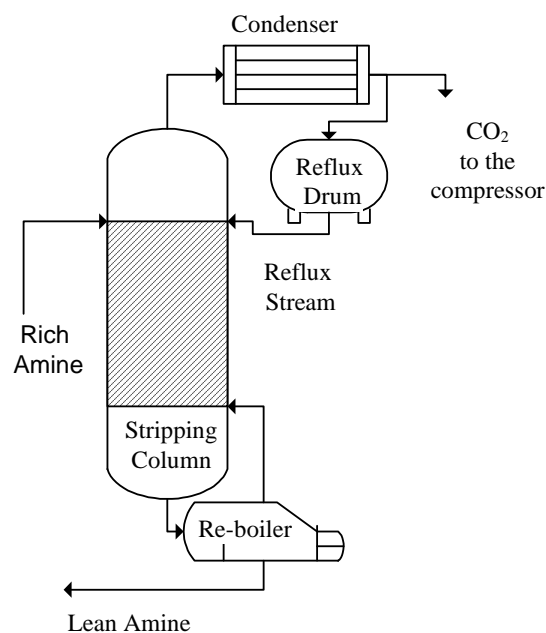


Figure 1. Diagram of a Stripping column

Some steady state results and dynamic predictions are found and recorded. The results are compared with experimental data found in the literature [12]. The sensitivity of the model for different physiochemical correlations is analyzed. Effect on the predictions of the model from the inclusion of minor reactions is also analyzed.

## 2. Dynamic model

A dynamic model for the stripping column of a CO<sub>2</sub> capturing system is developed and implemented in MATLAB. The stripping column is modelled as a combination of the main tower, the re-boiler and the condenser. The main column is discretized along the height and a set of time dependant equations are developed. Each control volume consists of a separate liquid and vapour phase. Physics and thermodynamics of each phase and interfacial heat and mass transfer are considered with assumptions for developing the set of equations.

The important model assumptions are summarized below.

1. Each phase in a control volume behaves as a continuous stirred tank (CST)
2. Ideal gas phase and ideal liquid phase
3. Interfacial mass transfer of only H<sub>2</sub>O, CO<sub>2</sub> and MEA are considered
4. Only the reactions in the liquid phase are of importance
5. Linear pressure drop along the column
6. The packing height of the column is considered
7. Constant volumetric flows of vapour and liquid are considered
8. Heat loss to the surroundings is neglected

### 1.1. Main model equations

The main model equations consist of the molar (component) and energy balances for the liquid and vapour phases. The component balances for the gas and liquid phases are given by eqs. 1 and 2, respectively,

$$\frac{dc_i^l}{dt} = \frac{u^l}{\lambda} \frac{dc_i^l}{dz} + \frac{1}{\lambda} \dot{n}_{i,trans}^m + \dot{n}_{i,gen}^m \quad (1)$$

$$\frac{dc_i^v}{dt} = -\frac{u^v}{(1-\lambda)} \frac{dc_i^v}{dz} - \frac{1}{(1-\lambda)} \dot{n}_{i,trans}^m \quad (2)$$

where  $c_i$  is the concentration of component  $i$ ,  $t$  is the time,  $u$  is the velocity,  $\lambda$  is the liquid hold up of the column,  $dz$  is the height of a control volume and  $\dot{n}_i^m$  is the volumetric molar flow or generation. The superscripts “l” and “v” and the subscripts “trans” and “gen” stand for the liquid and vapour phases, and the interfacial transfer and the rate of generation, respectively.

Only the main reaction between CO<sub>2</sub> and MEA, which is given by reaction R1, is considered for computing the rate of generation of species.



Here MEAH<sup>+</sup> is the protonated MEA and MEACOO<sup>-</sup> is the carbamate ion formed by the reaction.

The energy balances for the liquid and vapour phases are given by the eqs. 3 and 4, respectively.

$$\frac{dT^l}{dt} = \frac{u^l}{\lambda} \frac{dT^l}{dz} + \frac{h_{ov} A_w}{\lambda \tilde{C}_p \Sigma c_i^l} (T^v - T^l) + \frac{\dot{n}_{CO_2,trans}'''}{\lambda \tilde{C}_p \Sigma c_i^l} (-\Delta \tilde{H}_{ab}) + \frac{1}{\lambda \tilde{C}_p \Sigma c_i^l} \sum \dot{n}_{i,trans}'''' (-\Delta \tilde{H}_{i,vap}) \quad (3)$$

$$\frac{dT^v}{dt} = -\frac{u^v}{(1-\lambda)} \frac{dT^v}{dz} - \frac{h_{ov} A_w}{(1-\lambda) \Sigma (c_i^v \tilde{C}_{p,i}^v)} (T^v - T^l) \quad (4)$$

Here  $T$  is the temperature,  $h_{ov}$  is the overall heat transfer coefficient between the two phases,  $A_w$  is the effective contact area between the phases and  $C_p$  is the specific heat capacity. Decoration “~” denotes molar basis, while  $(-\Delta H_{ab})$  and  $(-\Delta H_{i,vap})$  represent heat of absorption of  $CO_2$  and heat of condensation of specie  $i$ .

The reflux flow, which enters the column in the first control volume from the top of the stripping tower, introduces the following changes (given by eqs. 5 and 6) into eqs. 1 and 3.

$$\frac{u^l}{\lambda} \frac{dc_i^l}{dz} = \frac{u_1^l}{\lambda} \left( \frac{c_{i,1}^l - c_i^l}{\Delta z} \right) + \frac{u_2^l}{\lambda} \left( \frac{c_{i,2}^l - c_i^l}{\Delta z} \right) \quad (5)$$

$$\frac{u^l}{\lambda} \frac{dT^l}{dz} = \frac{u_1^l}{\lambda} \left( \frac{T_1^l - T^l}{\Delta z} \right) + \frac{u_2^l}{\lambda} \left( \frac{T_2^l - T^l}{\Delta z} \right) \quad (6)$$

Here  $u_1^l$  and  $u_2^l$  are the velocities of the amine flow from the absorption tower and the reflux flow from the reflux drum. Concentrations  $c_{i,1}^l$  and  $c_{i,2}^l$  represent the concentrations of the rich amine leaving the absorption tower and the reflux stream, respectively. Temperatures  $T_1^l$  and  $T_2^l$  are correspondent to the rich amine flow and the reflux flow at the inlet to the stripping tower.

The MEA solvent system is considered for analysis, and the thermodynamic and physical parameters are given accordingly. The interfacial mass transfer, reaction kinetics, and phase equilibrium formulations used are the same as presented by Jayarathna et al. [13].

The model is implemented in MATLAB and solver ODE15s is used to solve the set of differential and algebraic equations.

### 1.2. Condenser and re-boiler models

The condenser of the stripping column is modelled with a reflux drum to hold the liquid until it is refluxed. A flash calculation is performed (the feed stream to the condenser is flashed at constant pressure and temperature) in the condenser to find the liquid and vapour fractions leaving the condenser and their compositions. The liquid flow leaving the condenser enters the reflux drum, and the gas flow leaves for the  $CO_2$  compressor.

The reflux drum is assumed to have constant cross sectional area. The temperature inside the drum and the liquid phase density (which is dominated by water) are also assumed to be constants. An overall mass balance and specie balances are performed for the liquid phase inside the tank in order to find the rate of change of the liquid height and the liquid phase composition (given by eqs. 7 and 8).

$$\frac{dH_{RD}^l}{dt} = \frac{1}{A_{RD}\rho^l} \left( \dot{m}_{RD,in}^l - \rho^l \dot{V}_{RD}^l \right) \quad (7)$$

$$\frac{dc_{i,RD}^l}{dt} = \frac{1}{A_{RD}H_{RD}^l} \left( \dot{n}_{i,RD,in}^l - \frac{c_{i,RD}^l}{\rho^l} \dot{m}_{RD,in}^l \right) \quad (8)$$

Here  $\dot{m}_{RD,in}^l$  is the mass flow rate of liquid into the reflux drum, which is found from the molar flow into the reflux drum:  $\dot{m}_{RD,in}^l = \sum (\dot{n}_{i,RD,in}^l \cdot M_i)$  where  $M_i$  is the molar mass of specie  $i$ . Liquid height, specie concentration, liquid density and cross sectional area of the reflux drum are given by  $H_{RD}^l$ ,  $c_{i,RD}^l$ ,  $\rho^l$  and  $A_{RD}$ . Liquid height of the reflux drum is controlled between an upper and a lower boundary by manipulating the flow rate of the liquid from the reflux drum ( $\dot{V}_{RD}^l$ ). When the liquid phase composition and the mass flow rate from the reflux drum are known, the conditions of the reflux stream are known.

The re-boiler is modelled using a fixed vapour to feed fraction. Ideal temperature and pressure control in the re-boiler is assumed. The re-boiler heat duty ( $\dot{Q}_{RB}$ ) is calculated using:

- the energy required to heat up the feed ( $\dot{m}_{RB,in}^l \hat{c}_p^l \Delta T_{RB}$ ),
- energy required to vaporize the pre-defined fraction of the feed ( $\sum (\dot{n}_{i,RB}^v \Delta \tilde{H}_{vap,i})$ ), and
- heat of desorption of CO<sub>2</sub> ( $\Delta H_{ab}$ ), which leads to eq. 9,

$$\dot{Q}_{RB} = \dot{m}_{RB,in}^l \hat{c}_p^l \Delta T_{RB} + \sum (\dot{n}_{i,RB}^v \Delta \tilde{H}_{vap,i}) + \Delta H_{ab} \quad (9)$$

where  $\Delta T_{RB} = T_{RB} - T_{RB,in}^l$ . The feed flow into the re-boiler, specific heat capacity of the liquid and the molar rate of vapour leaving the re-boiler are given by  $\dot{m}_{RB,in}^l$ ,  $\hat{c}_p^l$  and  $\dot{n}_{i,RB}^v$ .

### 1.3. Physical properties and other parameters

Physical properties and other parameters are introduced to the model either as correlations or constant values found in the literature, or else using well known calculation methods. Some of the important physical properties and other parameters are given in Table 1 with their literature sources.

### 1.4. Numerical method

The model is implemented in MATLAB and the solver ODE15s is used to solve the set of differential and algebraic equations. Each tower model is discretized into 50 control volumes using the method of lines. The 50 control volumes are of uniform size.

Table 1. Physical properties and other parameters used in the MATLAB stripper model.

Property	Source	Comments
Enhancement factor	Hoff et al. [14]	
Forward reaction rate coefficient	Jamal et al. [15]	
Heat of absorption of CO <sub>2</sub>	Kohl et al. [16]	
Heat of vaporization (H <sub>2</sub> O and MEA)	Gáspár et al. [10]	
Liquid density	Weiland et al. [26]	
Liquid diffusivity of CO <sub>2</sub>	Versteeg et al.[18]	N <sub>2</sub> O analogy is used.
Liquid hold-up	Billet et al. [19]	
Liquid viscosity	Cheng et al. [17]	
Mass transfer coefficients	Hanley et al. [20]	Has developed for Flexipac1Y packing.
Overall heat transfer coefficient between phases	Cussler et al.[21]	Chilton-Colburn analogy is used.
Phase and reaction equilibrium	Kent et al. [6]	Simple and easy to apply.
Saturation pressure of water	Hoff et al. [14]	Clausius Clapeyron model is used
Thermal conductivity of N <sub>2</sub>	Incropera et al.[22]	Value of N <sub>2</sub> is used for the gas mixture.
Vapour diffusivity	Poling et al. [23]	Fuller equation is used together with Blanc's law.
Vapour viscosity	Perry et al. [24]	Mixture viscosity is found by combining the individual component values.
Wetted area of packing	R.E.Dugas. [12]	Constant value, specific for Flexipac1Y packing.

### 1.5. Model predictions

The MEA solvent system is considered for the analysis, and the thermodynamic and physical parameters are given accordingly. Several pilot plant cases from the Separations Research Program at the University of Texas at Austin are used for the validation of the dynamic model [12]. The stripping column in the pilot plant is a packed column with Flexipac 1Y type of structured packing. This column consists of two 3.05m packing sections with a collector plate and redistributors in between. Applied flow rates and packing properties are taken from the literature [12], [25]. The model prediction is satisfactory, and details are given in Table 2 and Figure 2.

Table 2. Inputs and predictions for the Texas cases 25, 28, 30 and 32.

Case no:	Inlet liquid temp: [K]	Inlet liquid rate [L/min]	Re-boiler temp: [K]	Inlet rich CO <sub>2</sub> loading	Lean CO <sub>2</sub> loading		Re-boiler duty [kW]	
					Pilot plant data	Simulated	Pilot plant data	Simulated
25	342	104.1	390	0.386	0.277	0.280	469	433
28	345	82.1	393	0.412	0.282	0.284	366	387
30	349	54.9	394	0.453	0.280	0.276	255	259
32	359	40.7	400	0.428	0.272	0.268	152	164

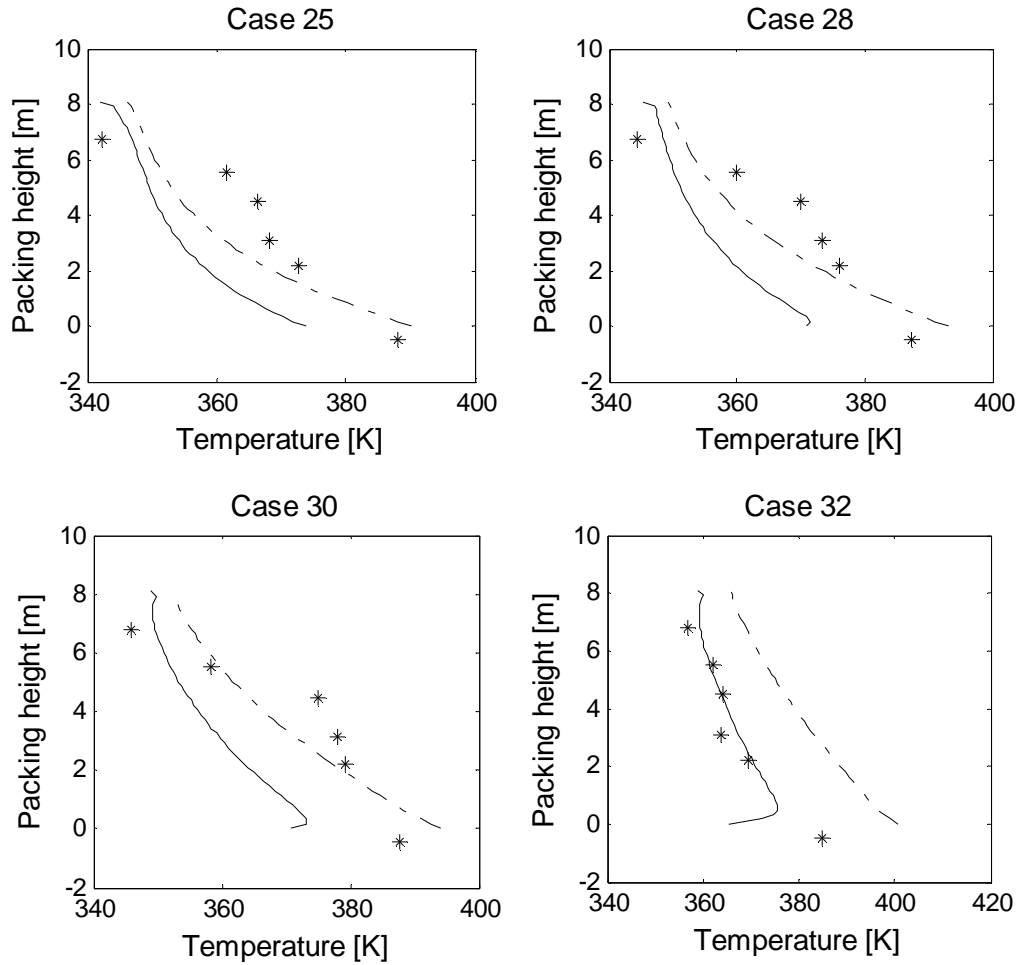


Figure 2. Temperature profiles inside the stripping column at the steady state. — is the liquid phase temperature, -.- is the vapour phase temperature, \* is the experimental data.

### 3. Sensitivity analysis

Sensitivity of the model for different physiochemical correlations and inclusion or exclusion of the minor reactions is analyzed. Eleven different cases are simulated and the predictions of the steady state are compared with the steady state results of the base case. Pilot plant case 32 is selected for the analysis. The combination of the physical properties and other parameters given in Table 1 are used for the base case simulation. Each case analyze the effect on the re-boiler heat duty, lean  $\text{CO}_2$  loading and prediction of the temperature profiles inside the stripping tower.

Details of the cases with the alterations made, the predicted re-boiler heat duties and lean  $\text{CO}_2$  loadings are given in Table 3.

Table 3. Details of the simulations performed for the sensitivity analysis.

Case number	Amendment	Re-boiler hear duty [kW]	Lean CO <sub>2</sub> loading
Pilot plant 32	-	152.2	0.272
Base case	-	164	0.268
1	Density model from Cheng et al. [17] is used	164	0.268
2	Viscosity model from Poling et al. [23] is used	164	0.268
3	Viscosity model from Weiland et al. [26] is used	164	0.268
4	Specie generation is presented with additional set of minor equations	166	0.259
5	Model for wetted specific area is taken from Billet et al. [19]	178	0.343
6	Model for wetted specific area is taken from Onda et al. [27]	169	0.293
7	Model for wetted specific area is taken from Hanley et al. [20]	169	0.294
8	Correlation for local mass transfer coefficients is taken from Onda et al. [27]	151	0.281
9	Correlation for Henry's law is taken from Jiru et al. [28]	159	0.313
10	Saturation pressure of H <sub>2</sub> O is calculated from Antoine equation	145	0.316
11	Saturation pressure of H <sub>2</sub> O is calculated from the Clausius Clapeyron relation	147	0.319

Cases 1 – 3 analyse the effect on the prediction of the stripper model by the correlations for computing the liquid phase density presented by Cheng et al. [17], liquid phase viscosity presented by Poling et al. [23] and liquid phase viscosity presented by Weiland et al. [26], respectively. Case 4 considers the effect by including the minor reactions into the model. The term minor reactions refer to the set of reaction presented by Liu et al. [29] for the system of CO<sub>2</sub>-MEA-H<sub>2</sub>O. Cases 5 to 7 analyse the effect from different correlations available for calculating the wetted specific area of the packing that are presented by Billet et al. [19], Onda et al. [27] and Hanley et al. [20], respectively. Case 8 considers the effect on the model predictions by the inclusion of the correlation presented by Onda et al. [27] to predict the local mass transfer coefficients of the species. Changes of the model predictions when the phase equilibrium of CO<sub>2</sub> is calculated using Henry's law constant presented by Jiru et al. [28] is examined by case 9. Cases 10 and 11 show the effect on the model predictions by inclusion of the different methods, such as the Antoine equation and the Clausius Clapeyron relation, to calculate the saturation vapour pressure of H<sub>2</sub>O.

The predictions of the temperature profiles inside the stripping tower, re-boiler heat duties and lean CO<sub>2</sub> loadings for the cases 1 to 4 are very similar to the predictions by the base case simulation, showing that there is no significant effect from the alterations made in the cases 1 to 4 on the model predictions.



Comparison between the experimental data (from Pilot plant case 32), temperature profiles from the base case simulation and the temperature profiles from the simulations of the cases 5 to 7 and 8 to 11 are given in Figure 3 and Figure 4, respectively.

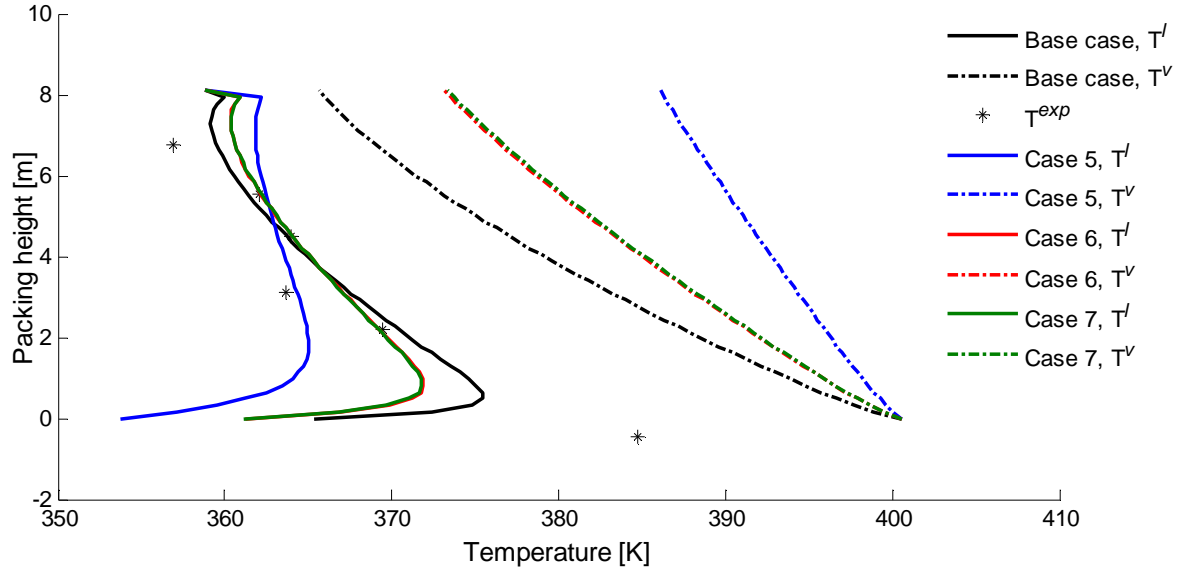


Figure 3. Temperature profiles predicted by the simulations of the base case and the cases 5 to 7, and the experimental data.

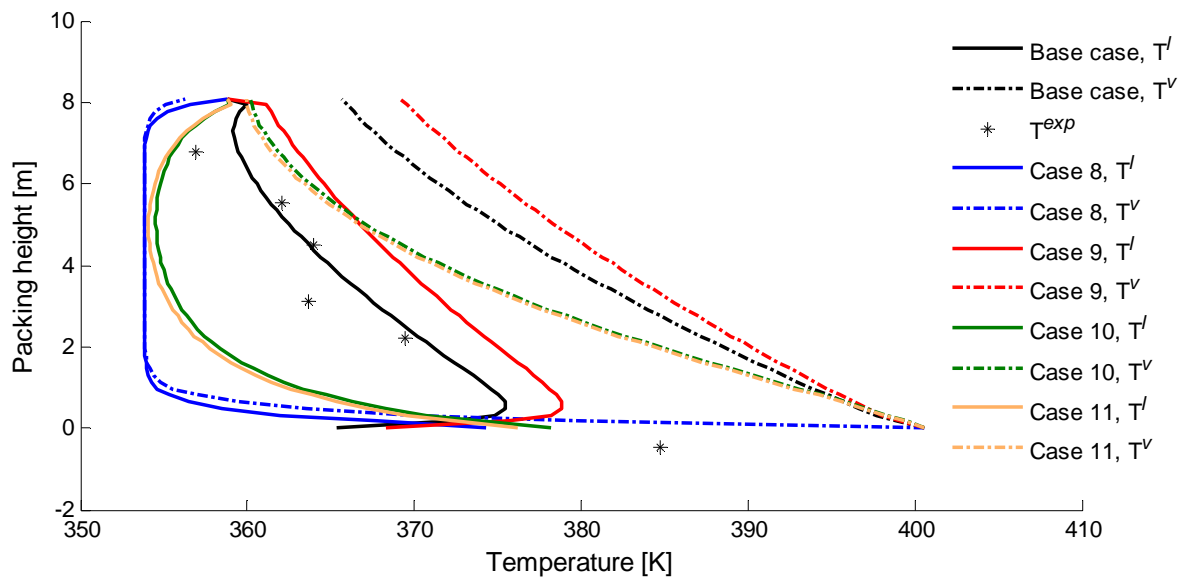


Figure 4. Temperature profiles predicted by the simulations of the base case and the cases 8 to 11, and the experimental data.

In Figure 3, the temperature curves predicted by the simulations of the base case, case 5, case 6 and case 7 for the liquid and the vapour phases, and the experimental data points are presented. Predicted curves from the cases 6 and 7 are very much the same for each phase along the tower height. Liquid

phase temperature profiles predicted by the cases 6 and 7 lie close to the liquid phase temperature profile from the base case. Vapour phase temperature profiles predicted by the cases 6 and 7 show an increasing deviation which starts from the bottom of the tower, from the vapour phase temperature profile of the base case, with a maximum deviation of about 7 K at the top of the tower. The temperature profile predicted for the liquid phase by case 5 has a considerable deviation from that of the base case up to about 4 m along the column height. The vapour phase temperature profiles from the base case simulation and the case 5 simulation have an increasing deviation with a maximum of about 12 K at the top of the tower.

Wetted specific area predictions from the correlations used in the cases 5 to 7 are lower than that of the base case, resulting increased re-boiler duties and lean loadings. As a result of the reduced transfer area followed by the reduced heat and mass transfer, deviation between the liquid and vapour phase temperature profiles in each case has also increased.

In Figure 4, the temperature curves predicted by the simulations of the base case, case 8, case 9, case 10 and case 11 for the liquid and the vapour phases, and the experimental data points are presented. Liquid and vapour temperature profiles predicted by case 8 lie very close to each other and consist of a steep curve (almost constant temperatures) from about 1 m to the top of the tower, while being very much different from the results of the base case and the experimental data. Curves predicted by case 9 have shifted towards higher temperatures than in the base case, but lie closer to the curves from the base case. The predictions of case 10 are very similar to those of case 11. The liquid phase profiles from cases 10 and 11 have predicted lower temperatures than the base case simulation all the way to the top of the tower from about 0.5 m and the predicted temperature is higher than the base case up to about 0.5 m from the bottom of the tower.

The correlation by Onda et al. [27] used in the case 8, has predicted higher values for the local mass transfer coefficients compared to the base case. The effect on the overall mass transfer coefficient from the over-prediction of the local mass transfer coefficients is dampening by the effect on the enhancement factor from the same alteration. Temperature profiles inside the tower have been predicted very similar to each other as a result of the over-prediction of the gas side local mass transfer coefficient. The gas side local mass transfer coefficient has a direct effect on the overall heat transfer coefficient according to the Chilton-Colburn analogy [21]. The re-boiler heat duty has reduced as a result of the increased temperature of the liquid leaving the stripping columns. Lower temperature values inside the tower compared to the base case may have increased lean CO<sub>2</sub> loading value due to the reduced reversed reaction rate.

The correlation by Jiru et al. [28] have predicted higher solubility of CO<sub>2</sub> in amines compared to the base case, which in return have predicted higher lean CO<sub>2</sub> loading value. As a result of the reduced CO<sub>2</sub> transfer the H<sub>2</sub>O transfer may also have reduced causing higher temperature values inside the tower compared to the base case. The increased temperature of the liquid entering the re-boiler has reduced the re-boiler heat duty.

The correlations used in the cases 10 and 11 for predicting the saturation pressure of water have predicted higher values compared to the correlation by Hoff et al. [14], which is used in the base case. Higher saturation pressure values have resulted in increased vaporisation of water (or reduced condensation), which can be considered as the reason for reduced temperature values inside the tower. The reduced temperature values have resulted in increased lean CO<sub>2</sub> loading. The re-boiler duty has reduced as a result of the increased temperature of the liquid leaving the stripping tower.

#### 4. Dynamic analysis

The predictions of the transient conditions from the dynamic model are analysed via a simulation with varying re-boiler temperature.

The steady state results from the base case simulation are used as the initial conditions for this simulation. Firstly, the re-boiler temperature can be increased gradually as an attempt to introduce a varying heating load condition. Secondly, the re-boiler temperature can be reduced gradually in order to achieve the temperature set-point of the re-boiler, which represents the action of a temperature controller to reassure the stability of the process.

The simulation procedure is given below.

- The base case is simulated for 10 minutes
- The re-boiler temperature is increased from 400.5 K to 405.5 K in 5 minutes (until  $t = 15$  min)
- Then, the re-boiler temperature is reduced from 405.5 K to 400.5 K in 5 minutes (until  $t = 20$  min)
- The simulation keeps on for another 10 minutes ( until  $t = 30$  min)

The temperatures are changed linearly for the simulation rather than introducing a step change, which represents a more sudden change and more challenging operation.

Figure 5-(a) and 5-(b) demonstrate the change of the temperature profiles of the liquid and vapour leaving the stripping column with respect to the time. Figure 5-(c) and 5-(d) show the change of the re-boiler heat duty and the  $\text{CO}_2$  loading in the liquid leaving the stripping column during the transient situation.

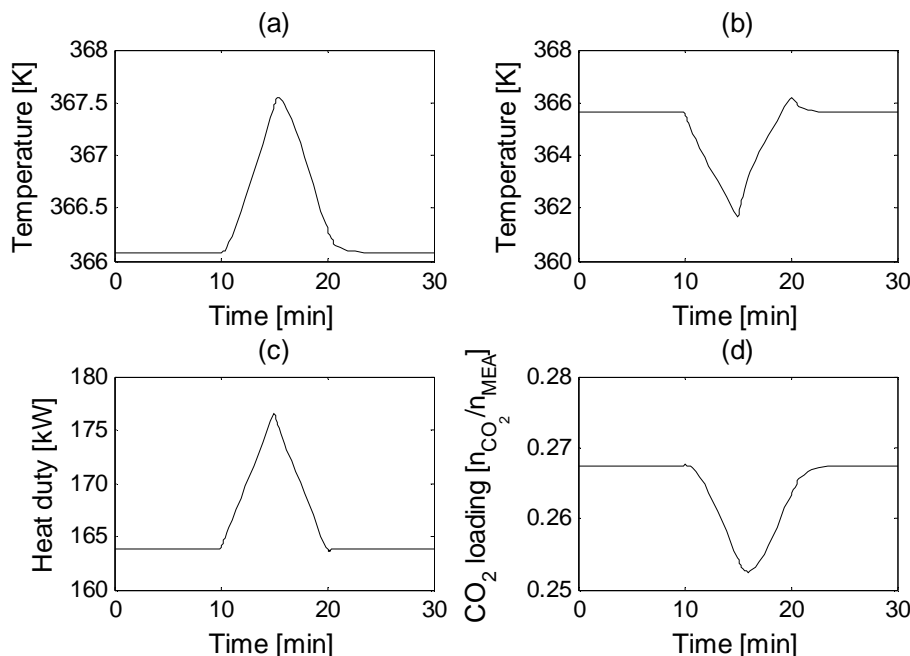


Figure 5. Predictions from the dynamic simulation. (a). Temperature of the vapour leaving the stripping column to the condenser, (b). Temperature of the liquid phase leaving the stripper to the re-boiler, (c). Re-boiler heat duty and (d).  $\text{CO}_2$  loading in the amine leaving the stripping column.

The heat duty shows an increase as the re-boiler temperature increases (Figure 5-(c)), obviously due to the increased heat demand to raise the temperature. The vapour temperature appears to be increasing as the re-boiler temperature increases (Figure 5-(a)), which is a direct effect from the increasing temperature of the inlet vapour to the stripping column. The liquid phase temperature decreases as the re-boiler temperature increases (Figure 5-(b)), even though one may expect an increase as the inlet temperature of the vapour increases. A possible reason for this observation is the increased vapour rate as a result of the increased temperature in the re-boiler, which may then limit the heat transfer from the vapour to the liquid. Increased vapour rate has increased the mass transfer between phases and increase the rate of the reversed reactions in the liquid phase (this can be seen from the figure 5-(d)), which leads to a reduction in the liquid phase temperature as well.

During the attempt to bring the temperature of the re-boiler back to the temperature set-point, the curves show a reversed effect as expected.

## 5. Conclusions

A good dynamic model provides the possibility to study the effect of various disturbances on the operating conditions of a plant and to apply improvements. Further, a dynamic model is useful for implementing a control system for the plant and to perform optimization. Development of dynamic models is vital for the understanding and improvement of the CO<sub>2</sub> capture process.

A dynamic simulation model for the stripper of a CO<sub>2</sub> capture plant has been developed in order to predict the transient conditions during various operating scenarios. The simplicity of the model has been maintained by use of simple models such as the Kent-Eisenberg model.

The steady state results from the simulation of the stripping column have shown acceptable accordance with the pilot plant data from the Separations Research Program at the University of Texas at Austin, which is taken as a primary validation of the model.

Simulations are performed to analyze sensitivity of the model predictions to different physical property correlations and inclusion of the minor reactions. The model predictions are not effected by the inclusion of the minor reactions, different density correlation for the liquid phase or different viscosity model for the liquid phase. Significant effect could be seen by the selected correlation for calculating the wetted specific area of the packing, Henry's law coefficient or saturation vapour pressure of water. A simulation is performed to analyse the model predictions under varying operating conditions. The model predictions of the transient conditions seem reasonable.

Further validation of the model will be performed by comparing it with the behaviour of a real plant. This is not included in the current work due to the limitation in available data. The model will be expanded to cover the whole CO<sub>2</sub> capture plant, which can be used for developing a control system for the capture process.

## Acknowledgement

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