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Dynamic Modelling and Analysis of Post-Combustion CO_2 Chemical Absorption Process for Coal-fired Power Plants

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Abstract
Post-combustion capture by chemical absorption using MEA solvent remains one of the most popular technologies for CO_2 emission mitigation for coal-fired power plants. This paper presents a study of the dynamic responses of a post-combustion CO_2 capture plant by modelling and simulation. Such a plant consists mainly of the absorber (where CO_2 is chemically absorbed) and the regenerator (where the chemical solvent is regenerated). Model development and validation are described followed by dynamic analysis of the absorber and regenerator columns linked together with recycle. The gPROMS (Process Systems Enterprise Ltd.) advanced process modelling environment has been used to implement the proposed work. The study gives insights into the operation of the absorber-regenerator combination with possible disturbances arising from integrated operation with a power generation plant. It is shown that the performance of the absorber is more sensitive to the molar L/G ratio than the actual flow rates of the liquid solvent and flue gas. In addition, the importance of appropriate water balance in the absorber column is shown. A step change of the reboiler duty indicates a slow response. A case involving the combination of two fundamental CO_2 capture technologies (the partial oxyfuel mode in the furnace and the post-combustion solvent scrubbing) is studied. The flue gas composition was altered to mimic that observed with the combination. There was an initial sharp decrease in CO_2 absorption level which may not be observed in steady state simulations.

Keywords: CO_2 Capture, Post-combustion, Chemical absorption, Dynamic modelling, Coal-fired power plant

1. Introduction

1.1 Background

Post-combustion capture by chemical absorption using MEA solvent remains one of the most popular technologies available for significant reduction in CO_2 emissions. It offers advantages over other CO_2 capture approaches in effecting significant CO_2 emission reductions as it is a suitable technology for retrofitting of existing power generation plants. It is also well suited for treating flue gas streams with low CO_2 partial pressures typical of coal-fired power plants.

Amongst stationary anthropogenic sources of CO_2 emissions, fossil-fuelled power plants release the largest amounts of CO_2. Coal-fired power plants release twice as much CO_2 per unit of electricity generated as their natural gas counterparts and are generally less efficient. However, these plants offer some advantages to operators due to the relatively high availability of coal compared to other fuels. Projections from the Energy Information Administration (EIA) suggest the continued use of coal especially in light of increasing electricity demand [1]. In addition, coal-fired power plants can be operated flexibly in response to changes in supply and demand [2] which provides great advantage to the operator.

This study investigates the operation of coal-fired power plants with post-combustion capture using dynamic modelling and simulation. Several authors have studied the reduction in the overall efficiency of the power generation process due to the large energy requirements of chemical absorption of CO_2 on a steady-state basis [3-6]. However, these studies thus assume that the power plant operates continuously at base load conditions. Such studies cannot provide insight into plant performance during certain operations such as start-up, load variations and other disturbances. Accurate dynamic models are required to study this behaviour.

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1.2 Novel contributions and outline of the paper

[7] and [8] present the dynamic model development and simulation of the absorber only. [8] also discusses different types of models used for modelling reactive absorption and the developments made in this regard. [9] describes the dynamic model development and simulation of the regenerator only while [10] extends this to the two stand-alone columns. Analysis on stand-alone columns may be inaccurate due to the inevitable coupling of the two columns linked with a recycle loop. This study presents dynamic modelling and simulation of the CO_2 chemical absorption plant that is the absorber and regenerator linked together with recycle. As a result the dynamic interaction between the absorber and regenerator could be studied more accurately through dynamic analysis. This work presents dynamic analyses

a. The effect of reducing flue gas flow rate to the absorption process
b. The effect of reducing reboiler heat duty
c. The effect of increasing CO_2 concentration in the flue gas to absorber

2. Operation of the CO_2 chemical absorption plant and Grid Code Requirements

Chemical absorption is a reactive absorption process and involves the combination of mass transfer and chemical reaction of CO_2 with a chemical solvent to form a weakly bonded intermediate compound which may be regenerated with the application of heat producing the original solvent and a CO_2 stream [11,12]. Monoethanolamine (MEA) being a primary amine reacts with CO_2 to produce carbamate via sets of liquid phase reactions [5,13].

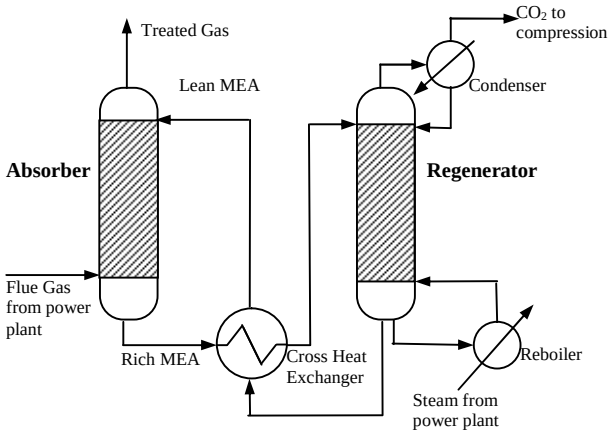


Figure 1 Simplified process flow diagram of Chemical Absorption process for post-combustion capture from [7]

Figure 1 describes a typical chemical absorption plant. The facility consists of two main units – the absorber and regenerator columns which could be packed or tray columns. Flue gas from the power plant is contacted counter-currently with lean MEA solution in the absorber. MEA chemically absorbs CO_2 in the flue gas. This leaves a treated gas stream of much lower CO_2 content. The solvent solution (now Rich MEA) is regenerated in the regenerator column using steam derived from the power generation process. This steam is proposed to be extracted between the intermediate pressure and low pressure turbines. CO_2 from the top of the column is compressed and transported away while the lean (regenerated) MEA solution is returned to the absorber column completing the cycle.

System frequency in a Grid is determined by a balance in system demand and total power generation. When the demand rises above the power generated, the frequency drops and vice versa. The UK Grid Code demands relatively strict minimum requirements for primary frequency response in power generation. Power plants are required to increase their output by 10% over a 10-second period. Secondary response refers to the requirement of output increase between 30s and 30 minutes [14]. One reason why subcritical power plants can meet the UK Grid requirements is that they have boiler drums and as such have a large

amount of stored energy. Higher efficiency supercritical boilers do not have drums and therefore it may be harder for them to meet Grid requirements. There are therefore concerns that these supercritical plants may not be able to meet the primary response requirement of the UK Grid Code [14]. In such a situation, some benefits may accrue from addition of a post-combustion plant downstream the power plant as the steam supplied to the reboiler could be reduced to make more steam available for power generation thus aiding primary response. It is, therefore, important to understand the behaviour of the absorption process with such a reduction. Storage of rich amine solvent could enable this to be done without a drop in CO_2 capture rates. In addition, the flue gas supplied to the absorption process may vary in flow rate or composition as the power plant operates in varying load conditions. This paper therefore studies the dynamic response of the CO_2 chemical absorption plant to investigate its behaviour with disturbances from the upstream power plant.

In Section 5, three dynamic cases describe how disturbances from the power generation process affect the operation of the CO_2 chemical absorption plant – the first deals with increased flue gas flow rate to the absorber. This mimics a secondary response scenario. The second mimics a possible primary response scenario where there is reduced steam flow available for regeneration. The third looks at a less conventional case where there is an increased concentration of CO_2 in the flue gas from the power generation process. This would be the outcome of a partial O_2-fired coal power plant with post-combustion capture [15].

3. Model development

This section describes development of the whole plant model. Common to both absorber and regenerator columns is the packed column section. These two columns were modelled based on the Gas-Liquid Contactor model from Process Systems Enterprise's Advanced Model Library. Two phenomena determine the rates of CO_2 absorption and solvent regeneration – mass transfer and chemical reaction. Mass transfer rates were modelled based on the two-film theory (Figure 2) using the Maxwell-Stefan formulation hence the models are referred to as "rate-based models". Heat and mass transfer resistances are modelled in the liquid and vapour films. Mass transfer coefficients in the liquid and vapour films were determined by correlations given by Onda et al [16]. The diffusivity (χ) of CO_2 in the liquid phase was based on expressions provided by Vaidya et al [17]. The diffusivity (χ) of CO_2 and other components in the vapour phase was estimated using the Fuller's equation [18]. Expressions for the heat of absorption were obtained from literature [19].

The regenerator model consists of the Gas-Liquid Contactor model (for the packed column section) linked with two flash drum models which represent the condenser and reboiler of the unit.

In the absorber, it is assumed that there is no heat lost to the surroundings. Due to its relatively higher operating temperature, the regenerator energy balance includes a heat loss term. Heat loss was assumed constant along the height of the column. The total heat loss from the column was estimated using relations given by [20].

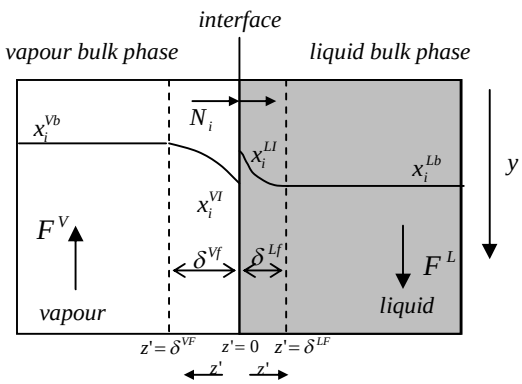


Figure 2 Liquid and vapour bulks, films and interface

3.1Model Assumptions

The following assumptions were used in developing this dynamic model:

1. All reactions are assumed to attain equilibrium
2. Plug flow regime and linear pressure drop along the column
3. Phase equilibrium at interface between liquid and vapour films
4. Negligible holdup in the vapour bulk
5. Negligible solvent degradation
6. Negligible heat loss in the absorber column
7. Chemical reactions are assumed to reach equilibrium in the condenser and reboiler.

3.2 Model Equations

More details of the absorber dynamic model are described in [8] while regenerator dynamic model is further described in [10]. In the packed column section, the main models include the vapour and liquid bulk models, the vapour and liquid film models and the interface (Figure 2). The main equations are listed for the convenience of readers:

3.2.1 Liquid Bulk Model

Mass Balance: $$\frac{dM_i}{dt} = \frac{-1}{L \cdot A}\frac{\partial F_i^L}{\partial y} + N_i \cdot Sp \cdot MW_i \cdot \omega \quad (1)$$

Energy Balance: $$\frac{dU}{dt} = \frac{-1}{L \cdot A}\frac{\partial F_H^L}{\partial y} + Sp \cdot \omega\left(H_{liq}^{cond} + H_{liq}^{conv} + H_{abs}\right) + HL \quad (2)$$

Heat Loss (HL) = $$\frac{342.88 \times (T_{regen\_bottom} - T_{ambient})}{L \times A \times NAE} \quad (3)$$

The Heat Loss in the regenerator is calculated based on the temperature difference between the regenerator bottoms temperature (T_{regen_bottom}) and the ambient temperature. It is calculated per unit volume and distributed evenly along the axial length of the column.

Heat of absorption (or desorption): $$H_{abs} = N_{CO_2} \times h_{abs} \quad (4)$$

$$h_{abs} = R \times \left(-14281 - \left(\frac{1092554 \times \gamma^2}{T}\right) - \left(\frac{6800882 \times \gamma}{T}\right) + 32670.01 \times \gamma\right) \quad (5)$$

3.2.2 Vapour Bulk Model

Mass Balance: $$0 = \frac{-1}{L \cdot A}\frac{\partial F_i^V}{\partial y} - N_i \cdot Sp \cdot MW_i \cdot \omega \quad (6)$$

Energy Balance: $$0 = \frac{-1}{L \cdot A}\frac{\partial F_H^V}{\partial y} + Sp \cdot \omega\left(H_{vap}^{cond} + H_{vap}^{conv}\right) \quad (7)$$

3.2.3 Vapour and Liquid Film Models

Maxwell-Stefan Formulation:

$$\frac{1}{\delta}\frac{\partial x_i^M}{\partial z'} = \frac{1}{c_t}\sum_{k=1}^{n}\left(\frac{x_i^M N_k - x_k^M N_i}{\chi_{i,k}}\frac{\mu^R}{\mu}\frac{T}{298.15}\right) \quad (8)$$

3.2.4 Interface model

$$f_i^L \times x_i^{M,L} = f_i^V \times x_i^{M,V} \quad (9)$$

3.2.5 Physical property estimation

Physical property estimations for viscosities, enthalpies and densities were determined using Multiflash while the fugacity coefficients and flash calculations in the reboiler and condenser were determined using the Electrolyte Non-random-two-liquid (NRTL) model in Aspen Properties and accessed through the CAPE-OPEN Thermo interface.

3.3 Controllers for the process

Control schemes are illustrated in Figure 3. PI controllers are used for temperature control of the regenerator column. The temperatures of the top and bottom of the regenerator are manipulated using the condenser and reboiler heat duties. The temperature of the regenerated solvent determines its CO_2 loading and thus affects its absorption capacity. Thus the temperature of the solvent from the reboiler bottoms must be maintained at desired levels.

The reboiler level was controlled using a P controller by manipulating the reboiler bottoms flow rate.

Water balance is achieved by controlling the water mass fraction in the lean solvent with the water makeup flow rate used as manipulated variable. The Lean MEA stream consists of 30.48wt% MEA, 6.18wt% CO_2 which leaves a 63.34 wt% H_2O. The set point for the mass fraction of lean solvent was therefore set as 0.6334.

The cross heat exchanger was sized to yield the reported regenerator feed temperature and a temperature approach of over 15°C. The cooler upstream the Lean MEA tank is specified to have a constant outlet temperature. The pressure of the partial condenser is also assigned to maintain the column operating pressure. The Rich MEA from absorber pump is specified to deliver the differential pressure between the absorber and regenerator including the pressure drops of the equipment in between.

Table 1 Process Controllers

	Controlled Variable	Manipulated Variable	Set point (K)
Condenser Temperature Controller	Condenser temperature	Condenser heat duty	320
Reboiler Temperature Controller	Reboiler temperature	Reboiler heat duty	387
Reboiler Level Controller	Reboiler level	Reboiler bottoms flow rate	0.50
Water Makeup Controller	Water mass fraction in lean solvent	Water Makeup flow	0.6334

3.4 Model Topology

Assumptions for Model Topology

1. No water wash section in the absorber
2. To achieve water balance, lean amine solvent from the regenerator is fed into the Lean MEA tank where water is made up before being pumped to the absorber

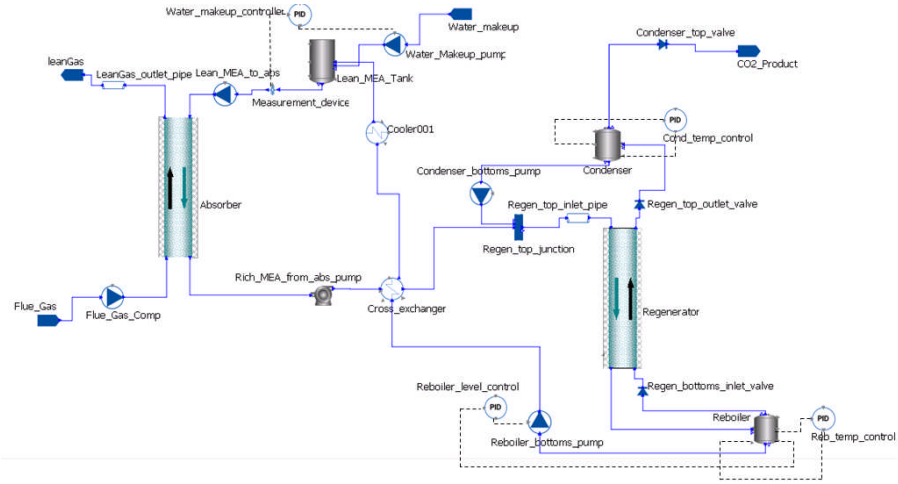


Figure 3 Model topology in gPROMS

4. Model Validation

The validation of the chemical absorption plant model was carried out using data from the Separations Research Program at the University of Texas at Austin [20]. The absorber and regenerator columns of the pilot plant are both packed columns with diameters of 0.427m and total packing height of 6.1m. Each column consists of two 3.05m packed bed sections with a collector plate and redistributor between the beds.

Out of the 48 experimental cases carried out in the research program, two cases (Cases 32 and 47) were selected for steady state validation purposes. These two cases were selected because of their relatively high and low liquid to gas (L/G) molar flow ratios for the absorber respectively. This further implies relatively high and low circulation rates of solvent respectively.
Table 2 shows the process conditions for the lean MEA and flue gas streams to the absorber while Table 3 shows some absorber column and packing specifications.
Simulation results were validated using the temperature profile of the absorber column measured in the pilot plant [20]. In addition the measured CO_2 loading of the amine solvent taken at different positions was compared with values obtained from simulation.

Table 2 Process conditions for Cases 32 and 47

	Case 32			Case47		
Stream ID	**FLUE GAS**	**LEAN MEA TO ABSORBER**	**RICH MEA FROM ABSORBER**	**FLUE GAS**	**LEAN MEA TO ABSORBER**	**RICH MEA FROM ABSORBER**
Temperature (K)	320	314	358	332	313	356
Pressure (10^5 kPa)	1.035	1.703	1.630	1.033	1.703	0.690
Total flow (kg/s)	0.130	0.720	0.745	0.158	0.642	0.746
Mass Fractions						
H_2O	0.0148	0.6334	0.6122	0.0193	0.6334	0.6085
CO_2	0.252	0.0618	0.2901	0.2415	0.0618	0.2943
MEA	0	0.3048	0.0971	0	0.3048	0.0966
N_2	0.7332	0	0.0006	0.7392	0	0.0006
L/G Ratio		6.5			4.6	

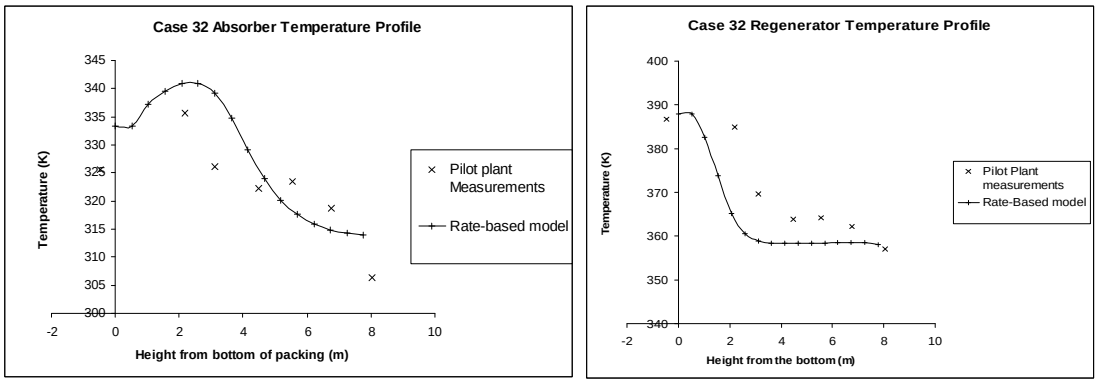
Table 3 Equipment Specifications

Description	Value
Column internal diameter (m)	0.427
Height of packing (m)	6.1
Nominal packing size (m)	0.0381
Packing specific area (absorber) (m^2/m^3)	145
Packing Specific area (regenerator) (m^2/m^3)	420
Cross Heat Exchanger Heat transfer area (m^2)	7
Reboiler volume (m^3)	1
Condenser volume (m^3)	2

4.1 Validation Results of stand-alone columns

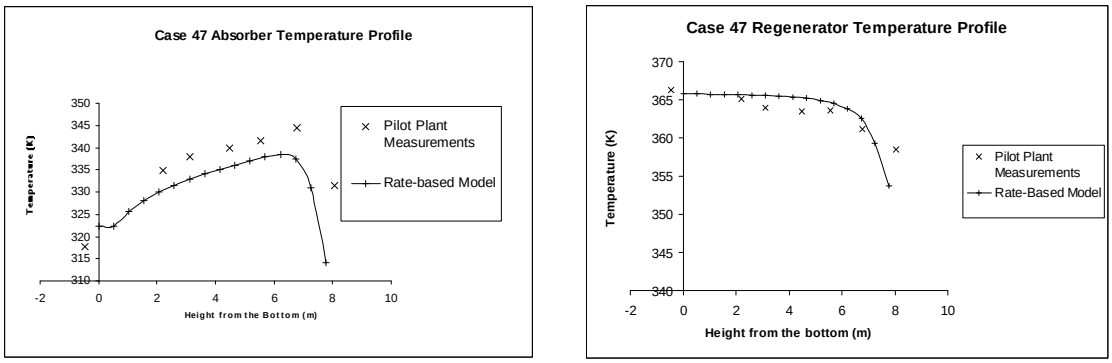
Figures 4 and 5 show the validation results for the stand-alone column based on the temperature profile measured in the pilot plant for the two selected cases. Feed conditions to the absorber were inputted from the information available in [20] there was limited information available for the feed conditions to the regenerator. Therefore, for the stand-alone regenerator, the feed conditions were obtained from the rich solvent conditions estimated by the absorber model.

In general, the stand-alone models predict the general shape of the column temperature profiles. The deviations appear higher in the regenerator profile, however, this may be because (as explained earlier) any inaccuracies in the absorber model results would be propagated to the regenerator simulation.



(a) (b)

Figure 4 (a) Absorber and (b) Regenerator temperature profile of columns for Case 32 (Stand-alone)

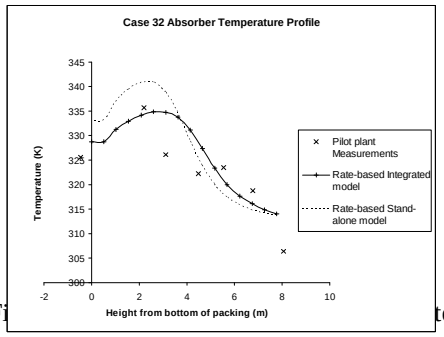


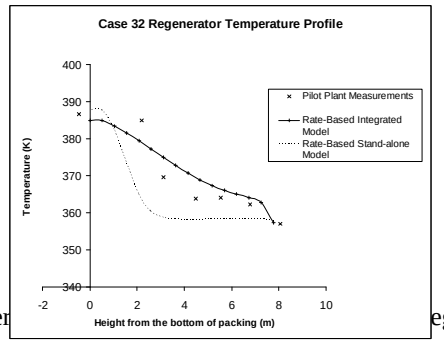
(a) (b)

Figure 5 (a) Absorber and (b) Regenerator temperature profile for Case 47 (Stand-alone)

4.2 Validation Results of integrated model

Figure 6 shows the validation results for the absorber and regenerator columns when they are linked together with recycle as shown in Figure 3. The process conditions used are those of Case 32. The results show a different prediction of the temperature profile in the integrated model especially in the regenerator because of the limitations discussed in Section 4.1. In general, the integrated model gives a better prediction of the observed temperature profile. This is expected as the pilot plant consists of columns linked together with recycle and not stand-alone structure.





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5. Dynamic analysis

Post-combustion capture offers a wider range of flexibility compared to pre- and oxy-combustion capture with the possibility of turning the regeneration process on, off or reducing the flow of rich solvent to the regenerator [21]. Electricity demand levels vary throughout the day. The regeneration process could be "turned off" or reduced during peak hours to meet increased demand. It is also likely that as the proportion of renewable energy sources increase, CCS plants would be forced to operate flexibly to support the intermittency of such renewable sources [21].

Dynamic analysis investigates the behaviour of the integrated system with disturbances from the power generation process. Three cases were investigated:

a. The effect of increasing flue gas flow to the absorber.
b. The effect of reducing reboiler heat duty on plant performance.
c. The effect of increasing CO_2 concentration of flue gas to the absorber.

It was important to keep control of the water balance in the process because the phenomenon would introduce additional disturbances to the system. Kvamsdal et al explained that if the flue gas stream to the absorber is not saturated with water, there is a tendency that some water would be evaporated in the column yielding a saturated stream at the top of the column and concentrating the solvent. This could also occur if the temperature of the flue gas significantly increases as it passes through the column. On the other hand, if the temperature of the flue gas reduces significantly as it passes through the column, some water vapour would condense and thus dilute the solvent [22]. Water balance is therefore a key control issue in the process.

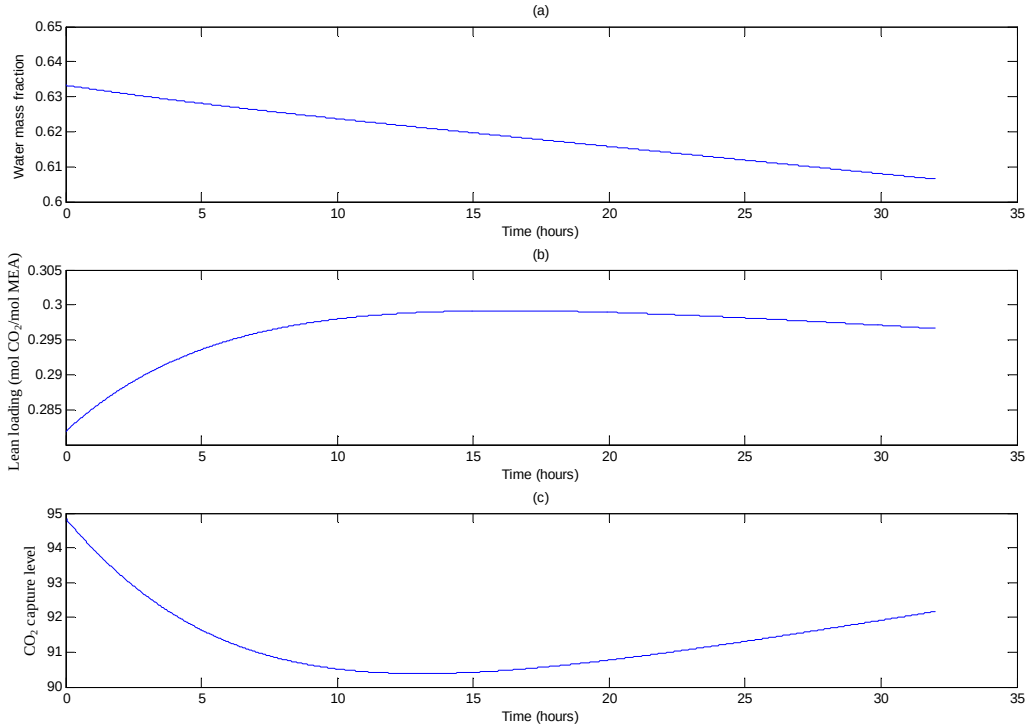

Figure 7 Effects of poor water balance in CO_2 absorption process

Figure 7 above illustrates some of the effects the phenomenon could have. The same conditions for Case 32 were maintained for the integrated model as described in Section 4.2. Without any disturbances introduced, the CO_2 capture level varies significantly (Figure 7c). As the water mass fraction in the lean solvent reduced fairly steadily (Figure 7a), the effect was an initial increase in the CO_2 loading of the lean solvent (Figure 7b). A maximum loading is reached followed by a subsequent reduction with a resulting increase in CO_2 capture levels (Figure 7c). This process could lead to severe operational problems especially with regards corrosion due to the high concentrations of MEA and CO_2 in the solvent.

5.1 Increasing flue gas flow rate to the absorber

This case simulates the effect of an increase in power plant output over a period of 10 minutes. It is assumed that this leads to a corresponding increase in flow rate of flue gas to the absorber and its composition is maintained. The process was simulated with the base-load conditions (Case 32) for two hours and a 10% increase in flue gas flow rate was carried out over a period of 10 minutes after which the simulation was left to run for just less than eight hours. In Case 1, the solvent circulation rate was kept constant while in Case 2, the circulation rate was correspondingly increased to maintain the molar L/G ratio (Figure 8).

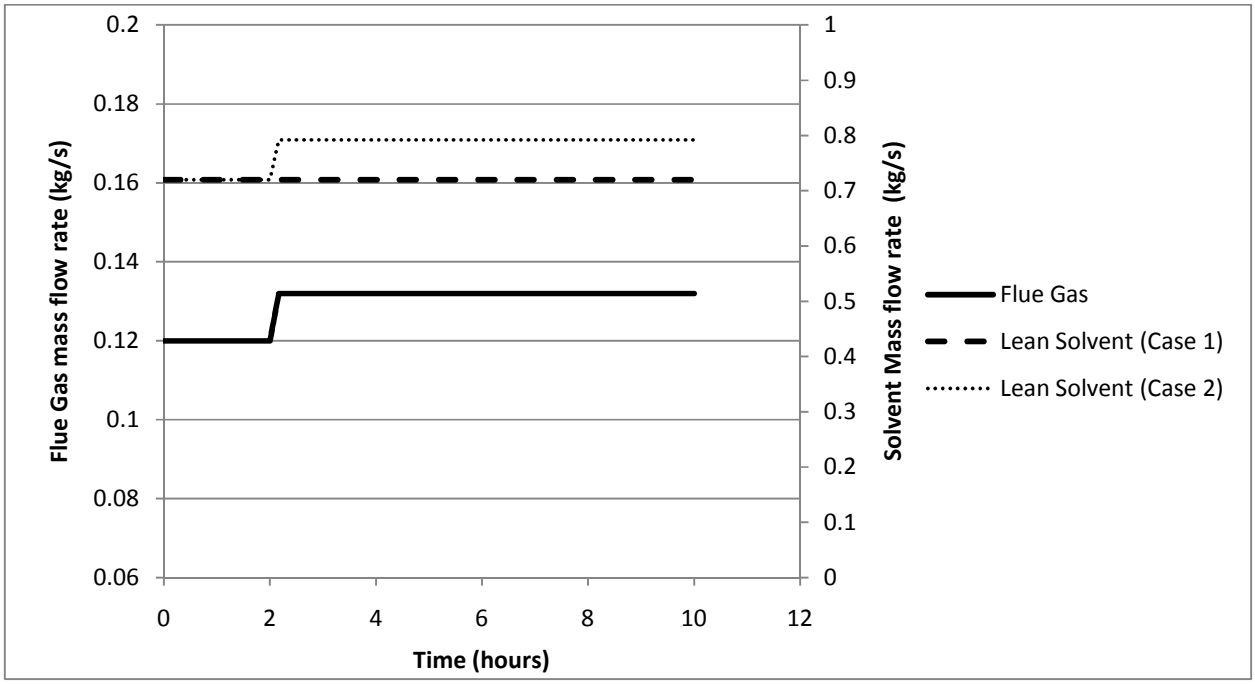

Figure 8 Flue gas and Lean MEA flow rate while increasing power plant load

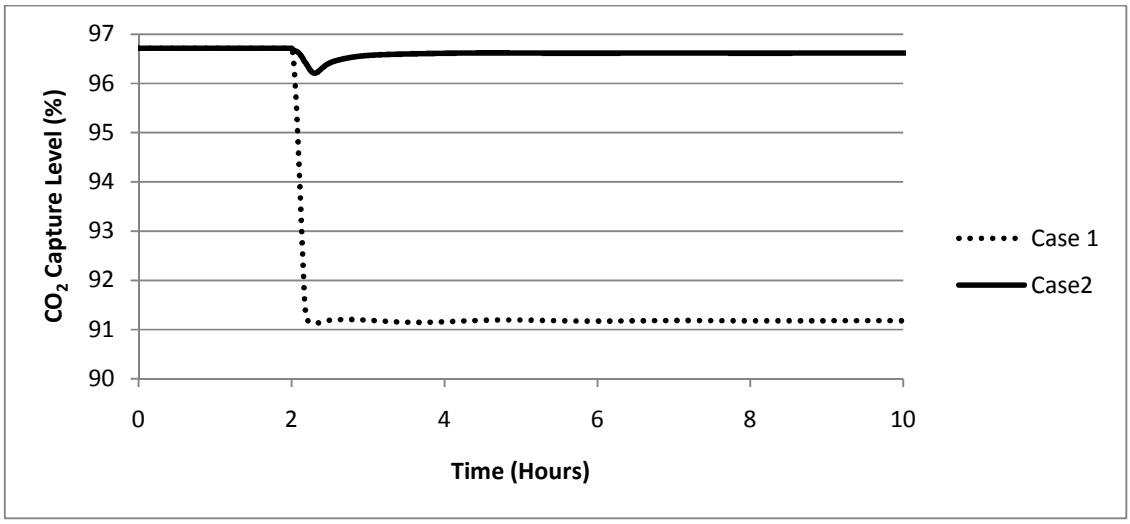

Figure 9 CO_2 capture level with increasing power plant load (Case1 and Case 2)

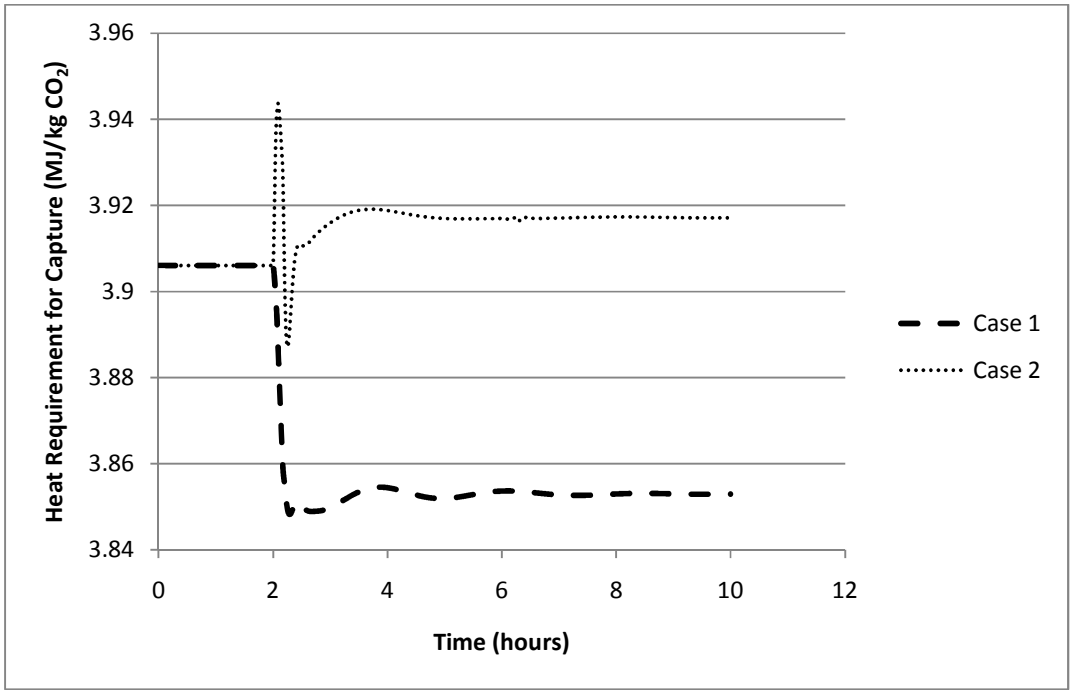

Figure 10 Heat requirement with increasing power plant load (Case1 and Case2)

Figure 9 shows a decrease in CO_2 absorption levels at the onset of the disturbance for Case 1 and virtually the same capture level maintained for Case 2 after a slight drop. This is however at the cost of increased heat requirement for capture (Figure 10). The heat required to capture a kg of CO_2 in Case 1 drops possibly because the lean solvent had the capacity to capture more CO_2 at the same circulation rate. The heat duty requirement for capture for Case 2 is higher than that of Case 1 but is maintained at almost the same level as before the disturbance. The oscillations observed are likely due to controller action.

5.2 Reducing reboiler heat duty

This case investigates the effect of a decrease in reboiler heat duty. This required heat duty for regeneration is supplied from the power generation process. Storage of rich amine solvent from the absorber could enable this to be done without a drop in CO_2 capture rates. This scenario explores the effect of a drop in reboiler duty and would not correspond to normal operation in an actual post combustion capture plant. Base conditions were maintained for a period of two hours. A 10% reduction in reboiler duty was implemented over ten seconds (Figure 11). Conditions were maintained for just less than eight hours.

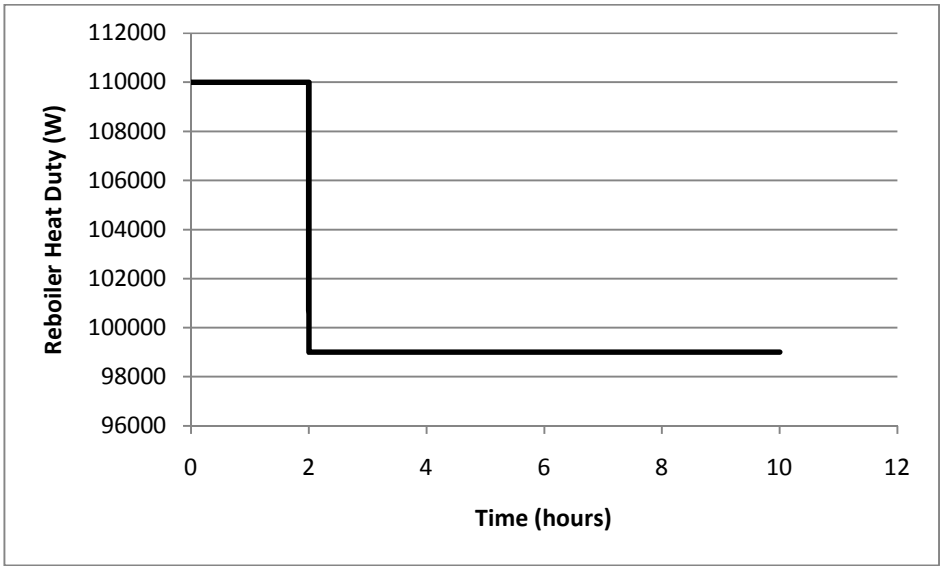

Figure 11 Reducing reboiler heat duty

A significant increase in CO_2 loading is observed (Figure 12) resulting from reduced reboiler temperatures. This also had a significant effect on the absorption levels in the absorber column (Figure 13).

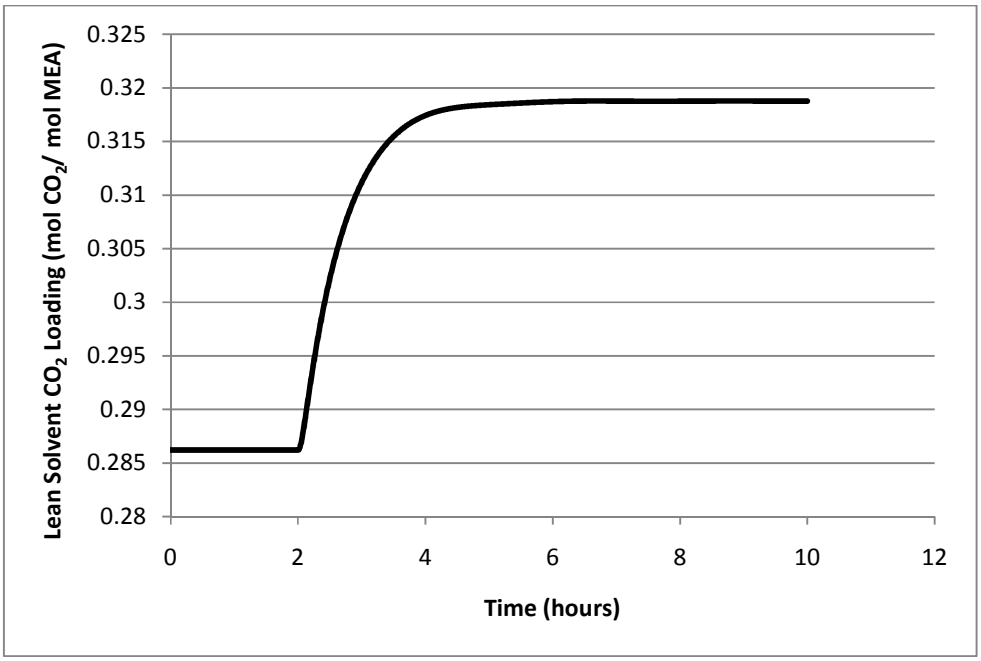

Figure 12 CO_2 loading change with reduction in reboiler duty

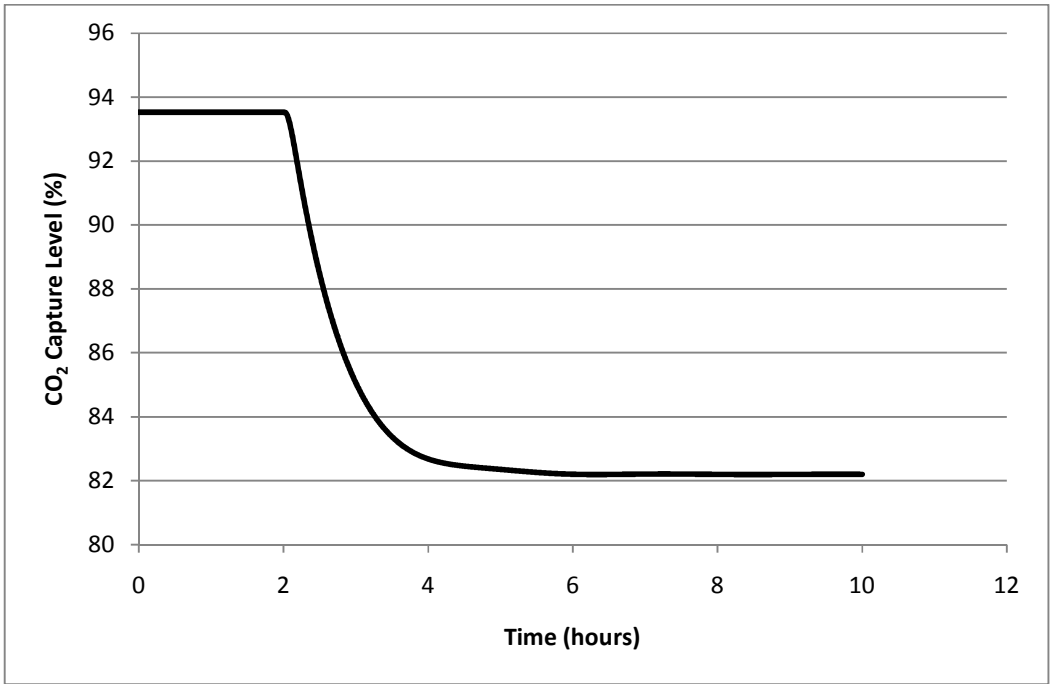

Figure 13 CO_2 capture level with reduction in reboiler duty

Since the change is carried out over a relatively short period of time, a time constant was estimated. The time constant estimated for the process was 2800s or 46.67 minutes. It should be noted that the lean MEA storage tank in the system would contribute to such a big time constant. Based on the volume of liquid in the tank as well as the flow rate of solvent from the tank, this contribution was estimated as 1388s or 23.14 minutes. It should be noted that in this model, transport delay has not been accounted for.

5.3 Increasing CO_2 concentration of flue gas to the absorber column

This case investigates the effect of increasing CO_2 concentration in the flue gas to the absorber. Some studies show that there could be benefits in a combination of two fundamental CO_2 capture technologies (the partial oxyfuel mode in the furnace and the post-combustion solvent scrubbing) [15]. It is estimated that the enrichment of oxygen content in the combustion air supplied to the furnace from atmospheric 21% by volume (or 100% air) to 50% by volume (or 60% air) would result in a new composition of flue gas to the absorber as shown in Table 4. The CO_2 mass flow to the absorber was kept constant while the total mass flow rate of the flue gas was correspondingly reduced (Figure 14). Figure 15 shows the changes in component compositions as this happens. This is done over a period of 30 minutes. Conditions were maintained for about 7.5 hours.

Table 4 New composition of flue gas with increasing CO_2 flue gas concentration

	Mass Fractions	
	100% air	60% air
Water	0.0148	0.1014
Carbon Dioxide	0.2520	0.4885
Nitrogen	0.7332	0.4101
Total Mass flow (kg/s)	0.1200	0.0620

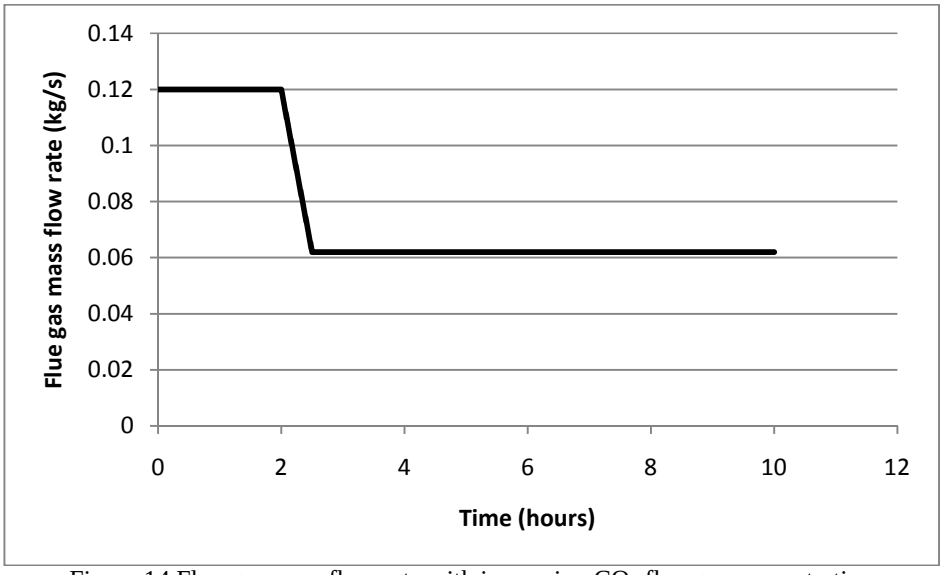

Figure 14 Flue gas mass flow rate with increasing CO_2 flue gas concentration

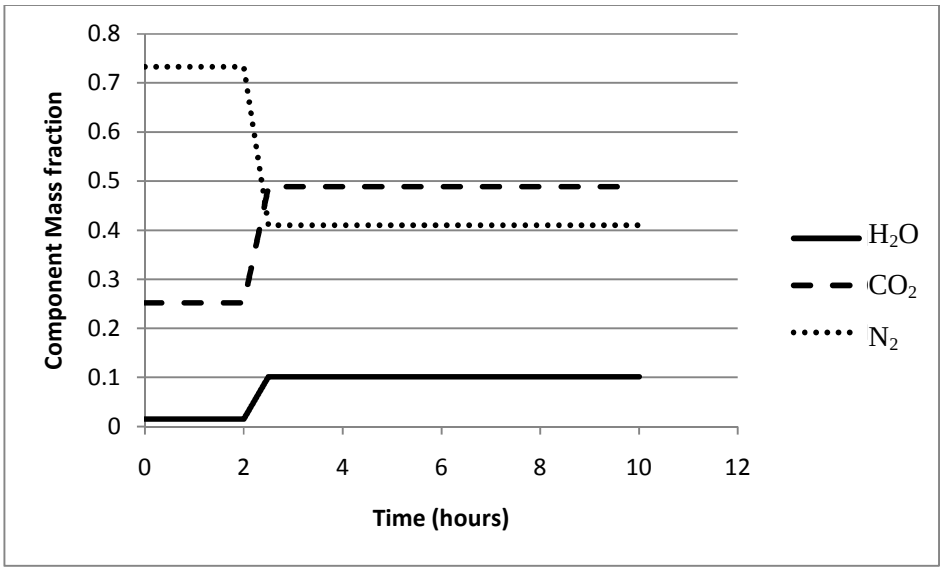

Figure 15 Flue gas composition with increasing CO_2 flue gas concentration

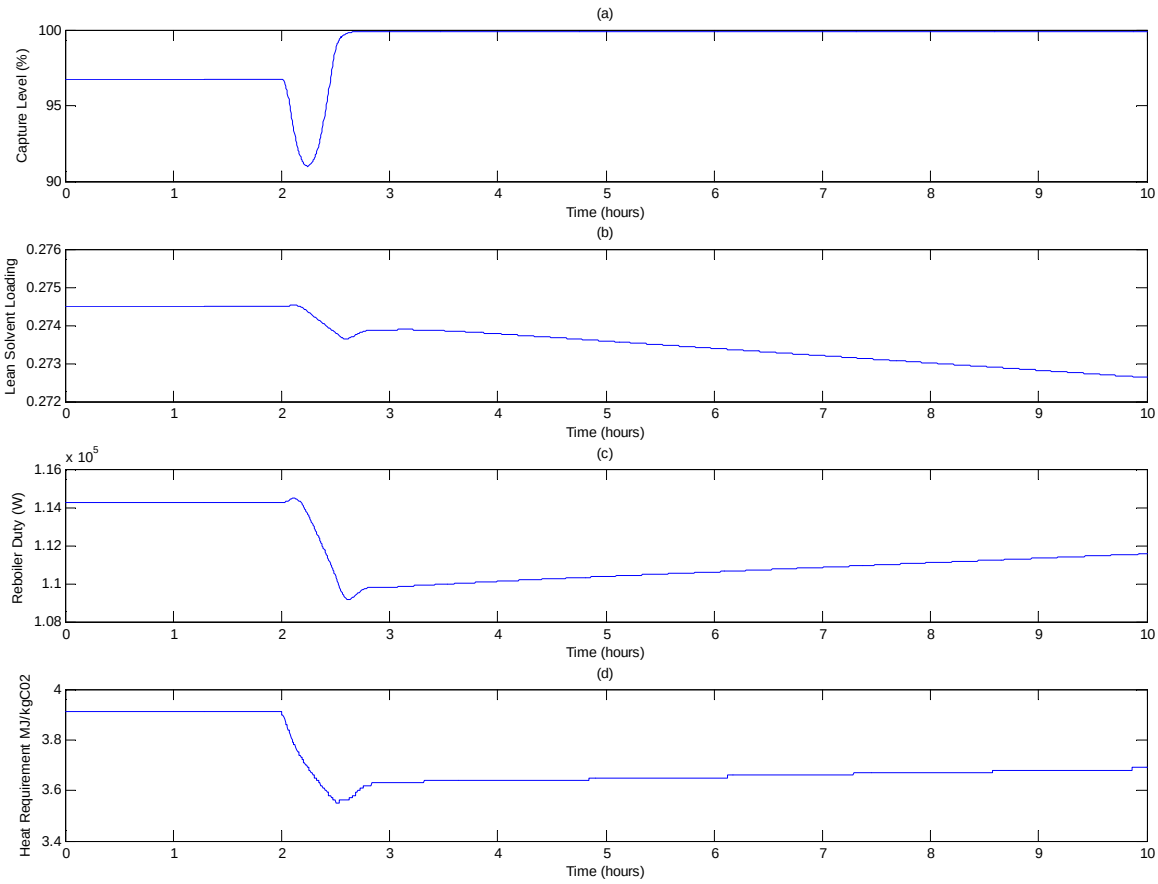

Figure 16 Effects of increasing CO_2 flue gas concentration

Figure 16a show a significant increase in CO_2 capture level after a sharp drop starting from the onset of the disturbance. The initial sharp drop is probably due to the initial increase in absorber temperature observed in the column (Figure 17). The solid line in Figure 17 indicates the onset of the disturbance. Afterwards, these temperatures drop sharply and the temperature bulge shifts towards the top of the column (Figure 17). This indicates an offset in water balance in the absorber where more water vapour in the gas stream condenses and dilutes the solvent in the column. Since this water imbalance cannot be remedied by the control system employed in the simulation, the dilution cannot be avoided. On dilution, the amount of MEA in the rich and subsequently the lean solvent reduces. Since the reboiler is temperature controlled, the equilibrium CO_2 concentration in the liquid phase reduces and so does the CO_2 loading in the lean solvent (Figure 16b). Figures 16c and 16d show the reboiler duty and heat requirement for capture respectively variation with the disturbance. The increased absorber temperatures are propagated to the regenerator and thus less steam is required in the reboiler. The heat requirement (Figure 16d) for capture also reduces since the same mass flow rate of solvent absorbs significantly more CO_2.

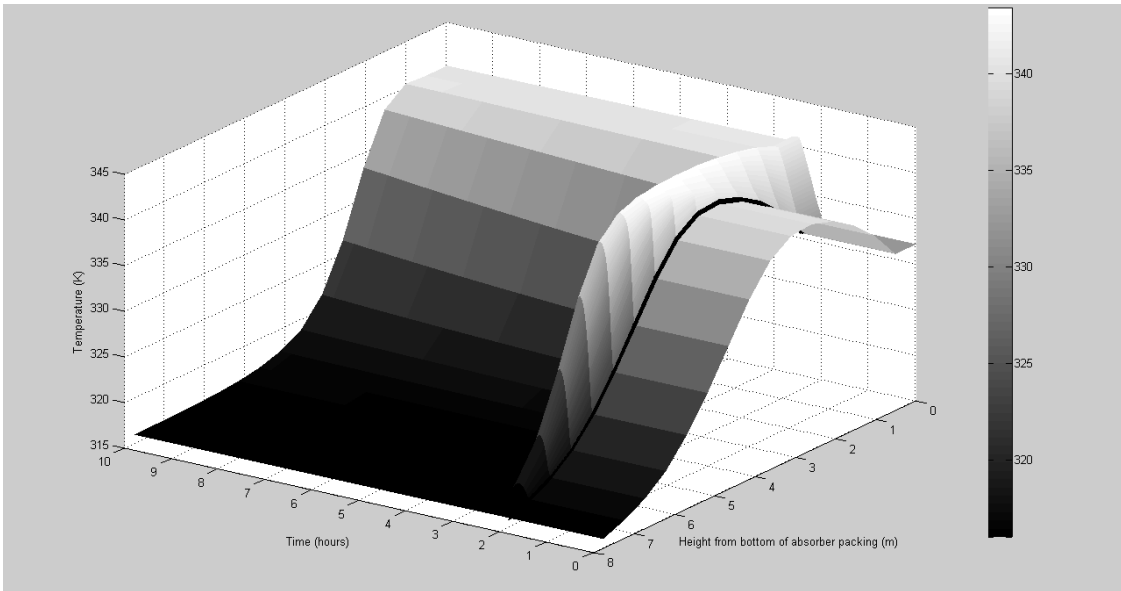

Figure 17 Absorber Column Temperature Profile with increasing CO_2 flue gas concentration

6. Conclusions

This paper presents a study of CO_2 capture using chemical absorption based on the dynamic modelling of the absorber and regenerator columns linked together. Validation results show that the model predicts the absorber and regenerator temperature profiles and CO_2 loadings fairly well. It is shown that water balance in the absorption process is very important. Dynamic analyses show that the absorber performance is more sensitive to the molar L/G ratio than the actual flow rates of the solvent and flue gas. The performance of the regenerator is affected by the reboiler duty and a time constant for the system was estimated as 46.67minutes. Some effects of increasing CO_2 flue gas concentration observed in the dynamic simulations such as sharp drops in capture levels and declining performance due to water balance, may not be observed with steady state simulations.

7. Acknowledgements

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Nomenclature

A	Cross sectional area (m^2)
c_t	Total molar concentration (mol/m^3)
f	Fugacity coefficient
F_i	Component mass flow rate (kg/s)
F_H	Enthalpy flow rate (J/s)
H	Heat flux (J/m^2)
h	Specific Enthalpy (J/kg)
HL	Heat loss to surroundings (J/m^3)
L	Length of column section (m)
L/G	Liquid to gas
M	Mass Holdup (kg/m^3)
MW	Molecular weight (kg/mol)
N	Molar flux ($mol/m^2.s$)
NAE	Number of Axial Elements

Greek Symbols

δ	Film thickness (m)
μ	Viscosity (Pa.s)
ω	Wetted area ratio
χ	Diffusivity (m^2/s)
γ	CO_2 Loading (mol CO_2/mol MEA)

Subscripts

abs	Absorption
H	Enthalpy
i	Component number
Liq	Liquid
Vap	Vapour

Superscripts

Cond	Conduction

n	Number of components
R	Universal Gas Constant (J/mol.K)
Sp	Specific area (m^2/m^3)
U	Energy Holdup (J/m^3)
x	Mass fraction
x_i^M	Molar fraction
y	Axial position
z'	Film position

$Conv$	Convection
I	Interface
L	Liquid
Lb	Liquid bulk
Lf	Liquid film
R	Reference
V	Vapour
Vb	Vapour bulk
Vf	Vapour film

References

[1] Energy Information Administration (EIA). International Energy Annual. 2006. Available at: http://www.eia.doe.gov/iea/ Accessed August 20, 2009.

[2] Chalmers H, Gibbins J. Initial evaluation of the impact of post-combustion capture of carbon dioxide on supercritical pulverised coal power plant part load performance. Fuel, 2007;86(14):2109-23.

[3] Abu-Zahra MRM, Schneiders LHJ, Niederer JPM, Feron PHM, Versteeg GF. CO_2 capture from power plants: Part I. A parametric study of the technical performance based on monoethanolamine. International Journal of Greenhouse Gas Control, 2007;1(1):37-46.

[4] Aroonwilas A, Chakma A, Tontiwachwuthikul P, Veawab A. Mathematical modelling of mass-transfer and hydrodynamics in CO_2 absorbers packed with structured packings. Chemical Engineering Science, 2003;58(17):4037-53.

[5] Davidson R. Post-combustion carbon capture from coal fired plants - solvent scrubbing. 2007;CCC/125.

[6] Davison J. Performance and costs of power plants with capture and storage of CO_2. Energy, 2007;32(7):1163-76.

[7] Kvamsdal HM, Jakobsen JP, Hoff KA. Dynamic modeling and simulation of a CO_2 absorber column for post-combustion CO_2 capture. Chemical Engineering and Processing: Process Intensification 2009;48(1):135-44.

[8] Lawal A, Wang M, Stephenson P, Yeung H. Dynamic modelling and simulation of CO_2 absorption for post combustion capture in coal-fired power plants. Fuel 2009;88(12):2455-62.

[9] Ziaii S, Rochelle GT, Edgar TF. Dynamic modeling to minimize energy use for CO_2 capture in power plants by aqueous monoethanolamine. Ind. Eng. Chem. Res. 2009;DOI: 10.1021/ie801385q.

[10] Lawal A, Wang M, Stephenson P and Yeung H. Dynamic Modeling and simulation of CO_2 chemical absorption process for coal-fired power plants. 10th International Symposium on Process Systems Engineering – PSE '09; 16-20 August 2009; Brazil; 2009.

[11] IEA GHG. The capture of carbon dioxide from fossil fuel fired power stations. 1993;IEA GHG/SR2.

[12] IPCC. IPCC special report on carbon dioxide capture and storage. 2005; Available at: http://www1.ipcc.ch/ipccreports/srccs.htm. Accessed September 24, 2009.

[13] Kucka L, Müller I, Kenig EY, Górak A. On the modelling and simulation of sour gas absorption by aqueous amine solutions. Chemical Engineering Science, 2003;58(16):3571-8.

[14] EON UK. Supercritical coal fired plant requirements and the Grid Code. 2008; Available at: http://www.nationalgrid.com/NR/rdonlyres/E3196A90-C091-463B-9113-440126B02A5E/25336/pp08_26_SuperCriticalcoalfiredplant.pdf. Accessed September 24, 2009.

[15] Doukelis A, Vorrias I, Grammelis P, Kakaras E, Whitehouse M, Riley G. Partial O_2-fired coal power plant with post-combustion CO_2 capture: A retrofitting option for CO_2 capture ready plants. Fuel 2009;88(12):2428-36.

[16] Onda K, Takeuchi H, Okumoto Y. Mass transfer coefficients between gas and liquid phases in packed columns. Journal of Chemical Engineering of Japan 1968;1(1):56-62.

[17] Vaidya PD, Mahajani VV. Kinetics of the Reaction of CO_2 with Aqueous Formulated Solution Containing Monoethanolamine, N-Methyl-2-pyrrolidone and Diethylene Glycol. Ind. Eng. Chem. Res. 2005;44(6):1868-73.

[18] Reid RC, Prausnitz JM, Sherwood TK. The properties of gases and liquids. Third ed. New York: McGraw-Hill; 1977.

[19] Oyenakan BA. Modeling of Strippers for CO_2 capture by Aqueous Amines. University of Texas at Austin; 2007.

[20] Dugas ER. Pilot plant study of carbon dioxide capture by aqueous monoethanolamine. University of Texas at Austin; 2006.

[21] Haines MR and Davison JE. Designing carbon capture power plants to assist in meeting peak power demand. 9th International Conference on Greenhouse Gas Control Technologies; 16 - 20 November 2008; Washington DC: Energy Procedia; 2008.

[22] Kvamsdal HM, Rochelle GT. Effects of the temperature bulge in CO_2 absorption from flue gas by aqueous monoethanolamine. Ind. Eng. Chem. Res. 2008;47(3):867-75.