



(51) International Patent Classification:

B01D 53/14 (2006.01) **B01D 53/78** (2006.01)
B01D 53/62 (2006.01) **C01B 31/20** (2006.01)

(21) International Application Number:

PCT/AU2013/000978

(22) International Filing Date:

30 August 2013 (30.08.2013)

(25) Filing Language:

English

(26) Publication Language:

English

(30) Priority Data:

2012903801 31 August 2012 (31.08.2012) AU

(71) Applicant: THE UNIVERSITY OF SYDNEY [AU/AU];
Parramatta Road, Sydney, New South Wales 2006 (AU).(72) Inventors: **KHALILPOUR, Rajab**; 10 Saint George
Road, Bexley, New South Wales 2207 (AU). **ABBAS, Ali**;
Unit 5, 5-7 Preddys Road, Bexley, New South Wales 2207
(AU).(74) Agent: **DAVIES COLLISON CAVE**; Level 14, 255
Elizabeth Street, Sydney, New South Wales 2000 (AU).

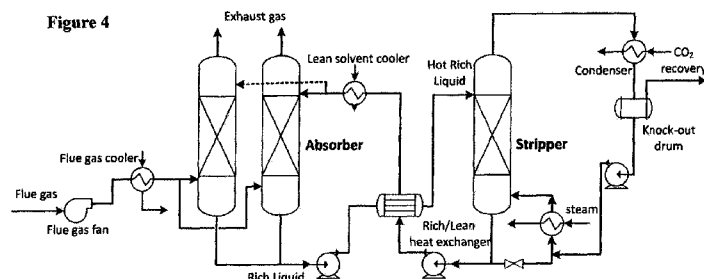
(81) Designated States (unless otherwise indicated, for every kind of national protection available): AE, AG, AL, AM, AO, AT, AU, AZ, BA, BB, BG, BH, BN, BR, BW, BY, BZ, CA, CH, CL, CN, CO, CR, CU, CZ, DE, DK, DM, DO, DZ, EC, EE, EG, ES, FI, GB, GD, GE, GH, GM, GT, HN, HR, HU, ID, IL, IN, IS, JP, KE, KG, KN, KP, KR, KZ, LA, LC, LK, LR, LS, LT, LU, LY, MA, MD, ME, MG, MK, MN, MW, MX, MY, MZ, NA, NG, NI, NO, NZ, OM, PA, PE, PG, PH, PL, PT, QA, RO, RS, RU, RW, SA, SC, SD, SE, SG, SK, SL, SM, ST, SV, SY, TH, TJ, TM, TN, TR, TT, TZ, UA, UG, US, UZ, VC, VN, ZA, ZM, ZW.

(84) Designated States (unless otherwise indicated, for every kind of regional protection available): ARIPO (BW, GH, GM, KE, LR, LS, MW, MZ, NA, RW, SD, SL, SZ, TZ, UG, ZM, ZW), Eurasian (AM, AZ, BY, KG, KZ, RU, TJ, TM), European (AL, AT, BE, BG, CH, CY, CZ, DE, DK, EE, ES, FI, FR, GB, GR, HR, HU, IE, IS, IT, LT, LU, LV, MC, MK, MT, NL, NO, PL, PT, RO, RS, SE, SI, SK, SM, TR), OAPI (BF, BJ, CF, CG, CI, CM, GA, GN, GQ, GW, KM, ML, MR, NE, SN, TD, TG).

Published:

— with international search report (Art. 21(3))

(54) Title: A SOLVENT BASED CARBON CAPTURE PROCESS AND PLANT AND A METHOD OF SIZING AND/OR CONFIGURING SAME



(57) Abstract: A method of determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, the method including the following steps: providing information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and, determining the size and/or configuration of the solvent based carbon capture process, wherein, step b includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant. The method also includes implementing a solvent based carbon capture plant according to the determined size and configuration of the solvent based carbon capture process and/or plant.

- 1 -

A SOLVENT BASED CARBON CAPTURE PROCESS AND PLANT AND A METHOD OF SIZING AND/OR CONFIGURING SAME

The present invention relates to a solvent based carbon capture process and plant and a
5 method of sizing and/or configuring same.

Background

Solvent based post-combustion carbon capture (PCC) has been evaluated and described as
10 the most mature technology for commercial scale industrial carbon capture from power plant flue gas. The technology is currently used in different small-scale industrial applications. Indeed, PCC began in the 1970s not out of concern for climate change but as a potential economic source of CO₂ for enhanced oil recovery (EOR) operations.

15 Figure 1 shows a schematic of a solvent-based PCC process. The flue gas, with a temperature range of 40-60 °C passes through the absorber column (packed or tray) where the lean solvent enters from the top of the absorber in a countercurrent process. In the absorber, the solvent removes CO₂ from the flue gas through physico-chemical interaction; the rich solvent then exits from the bottom of the absorber while the cleaned flue gas
20 leaves the absorber overhead towards the stack. In the stripper column, the rich solvent is stripped of CO₂ by thermal treatment at 100-120 °C. The lean solvent is recycled to the absorber while the CO₂ is sent from overhead to the compression unit.

Despite notable developments in solvent-based PCC, the implementation of this
25 technology in power plants still incurs a notable energy penalty, mainly due to solvent regeneration. Although it varies according to the type of power plant, the energy penalty is estimated to be above 20%. This results in a serious reduction in the power plant load. Therefore, if this technology is to become economically viable there is a need to develop methods to make significant reductions to the capital cost and operating cost of solvent-
30 based PCC processes.

- 2 -

Summary

According to one aspect there is provided a method of determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns
5 and one or more desorber columns, the method including the following steps:

- a) providing information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,
 - b) determining the size and/or configuration of the solvent based carbon capture process,
- 10 wherein, step b includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

In certain embodiments the method includes implementing a solvent based carbon capture plant according to the determined size and configuration of the solvent based carbon capture process and/or plant.

- 15 In certain embodiments the one or more design parameters of the solvent based carbon capture process and/or plant are selected from the following: the number of absorber columns, the number of desorber columns, the diameter of the absorber column(s), the diameter of the desorber column(s), the height of the absorber column(s), the height of the desorber column(s), the type of packing of the absorber column(s), and the type of packing
20 of the desorber column(s).

- In certain embodiments the one or more operational parameters of the solvent based carbon capture process and/or plant are selected from the following: the ratio of gas to liquid flow rates entering the absorber and/or the desorber, the ratio of the source of gas including carbon dioxide to solvent, the carbon dioxide capture rate, the temperature of the gas
25 including carbon dioxide, the temperature of the inlet gas to the desorber, the solvent temperature, the lean and rich loadings of the solvent, the operating pressure of the

- 3 -

absorber column(s), the operating pressure of the desorber column(s), the duty of the boiler associated with the desorber column(s).

In certain embodiments step b includes the following steps:

- i. providing a mathematical model of the solvent based carbon capture
5 process and/or plant dependent upon the one or more design parameters and the one or more operational parameters;
- ii. determining one or more objective functions associated with the mathematical model; and,
- iii. determining the value of the one or more design parameters and the one or
10 more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.

In certain embodiments the mathematical model of the solvent based carbon capture process and/or plant includes the following:

- 15
 - calculation of the number of absorber columns;
 - calculation of the size of the absorber column(s);
 - calculation of the number of desorber columns required;
 - calculation of the size of the desorber column(s); and
 - optionally calculation of the size and/or configuration of the auxiliaries
20 required.

In certain embodiments the one or more objective functions associated with the mathematical model includes the minimisation of the capital installation cost and/or the operating cost of the solvent based carbon capture process and/or plant. In one form the

- 4 -

objective function associated with the mathematical model includes one or more of the following: economic objective functions, techno-economic objective functions, and/or technical objective functions. In another form objective function is selected from: minimising the capital cost of the solvent based carbon capture process, minimising the
5 operating cost of the solvent based carbon capture process, the net present value and/or combinations thereof.

In certain embodiments the source of the gas is derived from a fossil fuel combustion process.

According to another aspect there is provided a solvent based carbon capture process
10 and/or plant that has been sized and/or configured by the method as herein described.

According to another aspect there is provided a method of determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, wherein the method includes:

15 a. a processing system obtaining information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,

 b. the processing system determining the size and/or configuration of the solvent based carbon capture process,

20 wherein, step b includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

In certain embodiments the method includes the processing system:

 i. providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or
25 more operational parameters;

- 5 -

ii. determining one or more objective functions associated with the mathematical model; and

iii. determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which corresponds to the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.

In certain embodiments the method includes the processing system iteratively manipulating the one or more design parameters and the one or more operational parameters within the mathematical model to determine the value of the one or more design parameters and the one or more operational parameters which corresponds to the one or more objective functions.

According to another aspect there is provided a processing system for determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, wherein the processing system is configured to:

a. obtain information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,

b. determine the size and/or configuration of the solvent based carbon capture process;

wherein the determination by the processing system includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

In certain embodiments the method includes the processing system:

i. providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters;

- 6 -

ii. determining one or more objective functions associated with the mathematical; and

iii. determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.

In certain embodiments the processing system is configured to iteratively manipulate the one or more design parameters and the one or more operational parameters within the mathematical model to determine the value of the one or more design parameters and the one or more operational parameters which corresponds to the one or more objective functions.

According to another aspect there is provided a computer readable medium for configuring a processing system to determine the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, wherein the computer readable medium includes executable instructions for configuring the processing system to:

- a. obtain information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,
- b. determine the size and/or configuration of the solvent based carbon capture process;

wherein the determination by the processing system includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

In certain embodiments the executable instructions configure the processing system to:

- 7 -

- i. provide a mathematical model of the solvent based carbon capture process dependent upon the one or more design parameters and the one or more operational parameters;
 - ii. determine one or more objective functions associated with the mathematical
 - 5 model; and
 - iii. determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond with the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.
- 10 In certain embodiments the executable instructions configure the processing system to iteratively manipulate the one or more design parameters and the one or more operational parameters within the mathematical model until the one or more objective functions are optimised.
- According to another aspect there is provided an absorber associated with a solvent based
- 15 carbon capture process and/or plant wherein the absorber has been sized by providing information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.
- 20 In certain embodiments determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant includes providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters, determining one or more objective functions associated with the
- 25 mathematical model; and determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size of the absorber.

- 8 -

In certain embodiments the absorber is sized by providing one or more of the following parameters: the height of the absorber, the diameter of the absorber, the packing height of the absorber and/or the packing type.

5 According to another aspect there is provided a desorber associated with a solvent based carbon capture process and/or plant wherein the desorber has been sized by providing information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

10 In certain embodiments determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant includes providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters, determining one or more objective functions associated with the
15 mathematical model; and determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size of the absorber.

In certain embodiments the desorber is sized by providing one or more of the following parameters: the height of the desorber, the diameter of the desorber, the packing height of
20 the desorber and/or the packing type.

According to another aspect there is provided a solvent based carbon capture plant including an absorber as herein described and/or a desorber as herein described.

Brief Description of the Accompanying Figures

25

The present invention will become better understood from the following detailed description of various non-limiting embodiments thereof, described in connection with the accompanying figures, wherein:

- 9 -

Figure 1 is a schematic diagram of a solvent based post combustion carbon capture process;

Figure 2 illustrates a functional block diagram of an example processing system that can be utilised to embody or give effect to a particular embodiment;

5. Figure 3 illustrates an example network infrastructure that can be utilised to embody or give effect to a particular embodiment;

Figure 4 is a schematic diagram of a solvent based post combustion carbon capture process designed for a 300 MWe coal-fired power plant;

10 Figure 5 is a flow chart outlining a method of one embodiment of the present invention;

Figure 6 is a flowchart representing a simplified column design algorithm;

Figure 7 is a flowchart representing a design algorithm of an absorber column;

Figure 8 is a flowchart representing a design algorithm of a desorber column;

Figure 9 is a schematic of a packed column model structure;

15 Figure 10 is a schematic of a desorber;

Figure 11 is a schematic of solvent-based PCC process configuration for a 300 MWe coal fired power plant for example 1;

Figure 12 is a flowchart representing an overall PCC process synthesis and design methodology for example 2;

20 Figure 13 is a flowchart representing a simplified column design algorithm for example 2;

Figure 14 is a flowchart representing a design algorithm of an absorber column for example 2; and

25 Figure 15 is a flowchart representing a design algorithm of a desorber column for example 2.

Detailed Description of Embodiments and the Accompanying Figures

30 The foregoing describes only some embodiments of the present invention, and modifications and/or changes can be made thereto without departing from the scope and spirit of the invention, the embodiments being illustrative and not restrictive.

- 10 -

In the context of this specification, the word "comprising" means "including principally but not necessarily solely" or "having" or "including", and not "consisting only of". Variations of the word "comprising", such as "comprise" and "comprises" have
5 correspondingly varied meanings.

In accordance with certain embodiments, the present invention provides a method of determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas which includes carbon dioxide. The source of gas may be from
10 any source that provides a gas including carbon dioxide such as for example flue gas exiting from a combustion reaction, a natural gas source, and synthesis gas from a gasification reaction.

It was found that when the size and/or configuration of a solvent based carbon capture
15 process and/or plant was determined by means of the method described herein, the capital cost of the solvent based carbon capture plant and the ongoing operational cost of the plant were greatly reduced compared to when the size and configuration of a solvent based carbon capture process is designed using conventional methods which use simplistic calculations based upon the mere size of the absorber required to remove the carbon
20 dioxide from a flue gas stream of a particular volumetric flow rate of flue gas.

A typical example of a solvent based carbon capture process is depicted in the schematic of Figure 1. The source of gas including carbon dioxide (denoted as flue gas 10) is reduced in temperature by via indirect heat transfer 12 before entering the absorber column 15 (which
25 may be packed or tray). A lean solvent 20 enters the top of the absorber 15 providing a countercurrent arrangement with the flue gas 21 within the absorber 15. The solvent removes carbon dioxide from the flue gas through physico-chemical interaction and the rich solvent then exits from the bottom 22 of the absorber 15 while the cleaned flue gas leaves the absorber 15 overhead 25. The rich solvent then passes to the stripper column 35
30 where it is stripped of carbon dioxide by thermal treatment. The lean solvent 20 is recycled

- 11 -

to the absorber 15 while the carbon dioxide is sent from the overhead 38 of the stripper 35 column to a compression unit.

In certain embodiments the method of determining the size and/or configuration of a solvent based carbon capture process includes a two step method wherein the first step (step a) includes providing information or values on parameters of the solvent based carbon capture process which are constant and are not able to be varied. The next step (step b) involves determining the size and/or configuration of the solvent based carbon capture process using a mathematical model of the solvent based carbon capture process which is dependent upon a combination of various design parameters and various operational parameters.

Step b further includes determining one or more objective functions which are associated with the mathematical model of the solvent based carbon capture process. Once the one or more objective functions are determined, step b then further includes determining the one or more of the design parameters and operational parameters until a result is achieved which corresponds with or aligns with the objective function. Once correspondence has been achieved, the values provided for the various design parameters and the various operational parameters provide the basis for the size and configuration of the solvent based carbon capture process or plant that corresponds with the one or more objective functions.

The flow chart of Figure 5 outlines the method of the present invention in accordance with one embodiment.

As can be seen from the flow chart, in certain embodiments the first step (step a) of the method involves providing information or values on parameters of the solvent based carbon capture process which are constant and are not to be, or not able to be varied: Such values include:

- Planned rate of CO₂ capture (tonne/annum base case), or percentage of carbon dioxide capture,
- Flue gas flowrate

- 12 -

- Flue gas composition (for example CO₂, H₂O, N₂, etc.)
- Solvent type and its composition (for example monoethanolamine or MEA).
- List of candidate packings and their properties (arbitrary)
- Maximum acceptable diameter and height for and absorber and desorber column constrained for example by the equipment manufacturer.

The next step (step b) involves determining the size and/or configuration of the solvent based carbon capture process using a mathematical model of the solvent based carbon capture process which is dependent upon a combination of various design parameters and various operational parameters. In this respect there are two types of parameters. Some parameters of the mathematical model which may be directly linked with the optimization stage ($X_1 - X_9$). There are however, some other parameters which may be determined in the synthesis/design stage ($X_{10} - X_{24}$) and are indirectly linked with the optimization stage.

Once the values of the constant parameters are provided in step a, as an initial starting point, step b provides a set of values for the following parameters (item B in the flow chart):

- CO₂ capture rate (%) of the process (X_1)
- Inlet flue gas (to absorber) temperature (X_2)
- Inlet gas (to desorber) temperature (X_3)
- Inlet liquid (to absorber) temperature (X_4)
- Lean and rich loadings (X_5, X_6)
- Packing type and size (X_7)
- Absorber and desorber columns' pressure (X_8, X_9)

With this information it is now possible to begin the design and/or configuration of the solvent based carbon capture process.

In this embodiment the mathematical model of the solvent based carbon capture process begins with the absorber design (item C in Figure 5). With the given values of the parameters outlined above, the absorber model is calculated using the mathematical model and steps shown in the following absorber flow charts. The details on the parameters providing the design and configuration of the absorber model (item D in Figure 5) are given below. The main model outputs are the following parameters:

- 13 -

- Number of columns (X_{10})
- Column diameter and packing height (X_{11}, X_{12})
- Liquid flowrate (X_{13})
- pressure drop (X_{14})
- 5 • Gas-to-liquid ratio
- Profile of pressure
- Profiles of gas and liquid temperature
- Profiles of gas and liquid components

10 If the diameter and/or height of the absorber which is calculated exceed the maximum limits given by the designer at step a above, the number of columns is increased, the inlet gas flowrate for each column is updated and then the absorber model calculation is repeated. Once the satisfactory results are achieved, the synthesis/design task is continued with the desorber column(s) synthesis/design.

15 A methodology similar to the absorber is followed for the desorber (item E in Figure 5) using the model shown in the desorber flow chart appearing and discussed below. The main outputs of the desorber model are the following parameters:

- Number of columns (X_{15})
- Column diameter and packing height (X_{16}, X_{17})
- 20 • Gas flowrate (X_{18})
- Pressure drop (X_{19})
- Reboiler temperature (X_{20})
- Gas-to-liquid ratio
- Profile of pressure
- 25 • Profiles of gas and liquid temperature
- Profiles of gas and liquid components

30 Having defined the process values of absorber(s) and desorber(s), the size of other auxiliary processes i.e. condenser (X_{21}), reboiler (X_{22}), heat-exchanger (X_{23}), and pump (X_{24}) are sized (See item G in the flow chart above).

With this the first stage of determining the size and/or configuration of the solvent based carbon capture process and/or plant is completed.

- 14 -

Once all of the design and operation parameters ($X_{10} - X_{24}$) are obtained at the synthesis/design stage, the second stage of the methodology i.e. optimization is pursued.

- 5 One or more objective functions (e.g.: max NPV, min CAPEX, min OPEX, etc.) is determined. An example of objective function formulation as net-present-value of cash flow (CF-NPV) is given below.

The mathematical model then proceeds to determine the value of the variable parameters to
 10 select a new set of values (for parameters $X_1 - X_9$) and repeat the methodology until the value of the variable parameters corresponds with the one or more objective functions. Therefore the final optimized parameters would be:

- CO₂ capture rate (%) of the process (X_1)
- 15 • Inlet flue gas (to absorber) temperature (X_2)
- Inlet gas (to desorber) temperature (X_3)
- Inlet liquid (to absorber) temperature (X_4)
- Lean and rich loadings (X_5, X_6)
- Packing type and size (X_7)
- 20 • Absorber and desorber columns' pressure (X_8, X_9)
- Number of absorber columns (X_{10})
- Column diameter and packing height of absorber (X_{11}, X_{12})
- Liquid flowrate of absorber (X_{13})
- pressure drop of absorber column (X_{14})
- 25 • Number of desorber columns (X_{15})
- Column diameter and packing height of desorber (X_{16}, X_{17})
- Gas flowrate of desorber (X_{18})
- Pressure drop of desorber (X_{19})
- Reboiler temperature (X_{20})
- 30 • Condenser size (X_{21})
- Reboiler size (X_{22})
- Heat-exchanger size (X_{23})
- Pump size (X_{24})

- 15 -

The various design parameters and the various operational parameters then provide the basis for the size and configuration of the solvent based carbon capture process and/or plant that corresponds with the one or more objective functions

- 5 The absorber flow charts show an example embodiment of a part of the mathematical model of the solvent based carbon capture process and/or plant detailing an integrated modeling and design methodology of an absorber column. The compositions and temperatures of inlet gas (T_{int}^{TG}) and liquid T_{int}^L as well as gas flowrate are considered to be known (item B in Figure 5). The column pressure and the loading of liquid at column
- 10 inlet and outlet are also known (item B in Figure 5). The inlet liquid coming from the desorber has a loading of α_{int}^{Abs} and enters from top of the column at temperature of T_{int}^L . The design policy is to have certain loading at the column exit (α_{out}^{Abs}) from where the composition of the solvent ($c_{j,int}$) can be obtained. It is set to capture $\beta\%$ of the target component, $j = TG$, from the flue gas ($Y_{TG,out} = \left(1 - \frac{\beta}{100}\right) Y_{TG,int}$). Therefore the quantity
- 15 of component j at the column exit is also known.

The main model equations are:

1- Rate-based mass transfer equation of each component j in the gas phase

$$F_{mol,1} \frac{dY_j}{dz} = N_j = \left(\frac{1}{\frac{1}{k_j^G a_g} + \frac{m_j}{k_j^L a_{gE}}} \right) (p_j - m_j c_j) \quad j=2, 3, \dots, J \quad (1)$$

- in which $F_{mol,1}$ is the molar flux (molar flowrate divided by column cross section area) of non-reacting/non-diffusing component (e.g. nitrogen in PCC process), and Y_j is defined as
- 20 the mole fraction of component j (y_j) divided by that of the non-reacting/non-diffusing component 1 (y_j/y_1). The reverse relation would be:

$$Y_1 = 1 \text{ and } y_j = \frac{Y_j}{\sum_{j=1}^J Y_j} \quad j=2, 3, \dots, J$$

- 16 -

2- Rate-based mass transfer equation of each component j in the liquid phase

$$u^L \frac{dc_j}{dz} = N_j + \sum_{p=1}^P R_{jp} \quad j = 2, \dots, J; \quad p = 1, \dots, P \quad (2)$$

- 5 where N_j is the rate of diffusion of component j and R_{jp} is the rate of generation of component j by the reactions p of the system of P reaction ($p = 1, \dots, P$). u^L is the superficial velocity of the liquid (volumetric flowrate divided by the column cross section area).

3- Energy balance equation of gas phase

10
$$F_G(\text{mol}, 1) \left(\sum_{j=1}^J Y_j [C_p]_j \right) (dT^G)/dz = -h_G a_1 e (T^G - T^L) \quad (3)$$

where the heat transfer coefficient, h_G , is obtained from dimensionless analogies such as the well-known Chilton-Colburn j -factor, or using various empirical formulas. According to Chilton-Colburn:

15
$$J_D = St \cdot Sc^{\frac{2}{3}} \quad \text{and} \quad J_H = St_H \cdot Pr^{\frac{2}{3}}$$

Then

$$h_G = k_G (\rho_G C_{pG})^{\frac{2}{3}} \left(\frac{k_G^{Th}}{D_G} \right)^{\frac{2}{3}}$$

where k_G^{Th} is the thermal conductivity of the gas.

20 *4- Energy balance equation of liquid phase*

$$u^L \left(\sum_{j=1}^J Y_j [C_p]_j \right) (dT^L)/dz = +h_G a_1 e (T^G - T^L) + \sum_{j=1}^J N_j [C_{pj}(T^L - T_0) + \Delta H_j^v] + \sum_{p=1}^P R_p \Delta H_p \quad (4)$$

- 17 -

where ΔH_j^v is the heat of vaporization of component j . It is noteworthy that in the case of CO₂ capture with amines, the heat of vaporization will be applicable for one component only, i.e. water, in the desorber.

5-pressure profile of the column

$$5 \quad \left(\frac{dP}{dz}\right)_{\text{loaded}} = \left(\frac{dP}{dz}\right)_{\text{dry}} + \left(\frac{dP}{dz}\right)_{\text{liquid}} \quad (5)$$

where

$$\left(\frac{dP}{dz}\right)_{\text{dry}} = \text{cst}_1 G_f^{\text{cst}_1} \times 10^{\text{cst}_2 \times L_f}$$

$$\left(\frac{dP}{dz}\right)_{\text{liquid}} = \text{cst}_3 \left(\frac{L_f}{\text{cst}_4}\right)^{\text{cst}_5} \left[\left(\frac{dP}{dz}\right)_{\text{dry}}\right]^{\text{cst}_6}$$

$$L_f = \begin{cases} F_{\text{mass}}^L \left(\frac{\text{cst}_7}{\rho_L}\right) \left(\frac{F_p}{\text{cst}_8}\right)^{\text{cst}_9} (\text{cst}_{10} \mu_L)^{\text{cst}_{11}} & F_{pd} \leq \text{cst}_{12} \\ F_{\text{mass}}^L \left(\frac{\text{cst}_7}{\rho_L}\right) \left(\frac{F_p}{\text{cst}_8}\right)^{\text{cst}_9} (\text{cst}_{10} \mu_L)^{\text{cst}_{13}} & F_{pd} > \text{cst}_{12} \end{cases}$$

$$10 \quad G_f = \text{cst}_{13} u_D^G \sqrt{\rho_G} \left(\frac{F_{pd}}{\text{cst}_8}\right)^{\text{cst}_9}$$

L_f and G_f are liquid and gas loading factors, respectively. F_{pd} is the dry-bed packing factor (sometimes called "Robbins factor") which is usually provided by manufacturers. $\text{cst}_1 - \text{cst}_{13}$ are constants.

15 The problem of integrated modeling and design compared with modeling only is the unavailability of column diameter and height. To solve the ordinary differential equations (ODEs), i.e. Eqs. (1), (2), (3), (4) and (5), the initial parameters must be available. This is whilst the properties of liquid at the bottom of the column are not available (temperature, concentrations and flowrate). To initiation of the solution, smart initial values (IVs) are used and the average values of gas and liquid parameters (temperatures and concentration)

- 18 -

are computed at the top and bottom of the column from where the physico-chemical properties (viscosities, densities thermal conductivities, surface tensions, heat of reaction, diffusivities specific heat) are computed using the rigorous equations detailed in Table 1 (see appendix). Once these parameters are computed, Eq. (6) is used to obtain the flood
5 velocity.

$$\log Y_{fl} = -0.289(\log F_{LV})^2 - 1.081 \log F_{LV} - 1.682 \quad (6)$$

where

$$F_{LV} = \frac{L_{mass}}{G_{mass}} \sqrt{\frac{\rho_L}{\rho_G}}$$

Having obtained the value of Y_{fl} , the flux of gas at flooding ($F_{mass, fl}^G$) is calculated by a
10 relation given by

$$Y_{fl} = \frac{(F_{mass, fl}^G)^2}{g \rho_G \rho_L} \frac{\rho_w}{\rho_L} (1000 \mu_L)^{0.2} F_p$$

The packing factor, F_p , is usually given by manufacturers and when unavailable is approximated by $F_p = \frac{a_{\square}}{\varepsilon^3}$.

Then the design velocity (u_D^G) is calculated based on the method discussed earlier (70-85
15 % of u_{fl}^G). Having these, the column diameter is obtained.

The next step is to find the column height which will be obtained using the ODEs system. The finite difference (FD) method is used to solve the ODEs. The step size of Δz is defined for column height and the equations are solved for every stage s as illustrated in the schematic of a packed column model structure. There are also numerous physico-chemical
20 parameters (viscosities, densities, thermal conductivities, surface tensions, heat of reaction, diffusivities, specific heat, mass and heat transfer coefficients) required for solving the ODEs. Each and every of these physic-chemical properties are function of function of various parameters. The accurate values of these parameters are critical in the modeling.

- 19 -

The detailed equations of these properties are given in Table 1 (see appendix). These parameters are computed at every stage s of the column. As enhancement factor (E) and interfacial vapor pressure of the component ($p_{i,i}$) have implicit functionality, a loop is used to find the both parameters.

- 5 With all these the ODEs are solved for the first stage $s = 1$. The order of equations is: concentration equation of gas, Eq. (1), concentration equation of liquid, Eq. (2), temperature of gas, Eq. (3), temperature of liquid, Eq. (4), and finally pressure drop, Eq. (5). Once these equations are solved their output will be used as IVs for the next stage $s+1$. The ODEs for every stage are solved, continuing until condition is reached when the value
- 10 of the target component at stage s ($Y_{TG, s}$) becomes equal to the given value $Y_{TG, out}$. This stage is $s = S$ and the packing height of the column is $Z = S \times \Delta z$. At this point, if the output values match with the initial assumption, the design task is accomplished; otherwise, it is necessary to update the guess with the new values and proceed to the next iteration. Once the goal is achieved, the real packing height can be calculated from the
- 15 results of the final run.

The overall pressure drop of the column is calculated by summation of the pressure drops

in all stages ($\sum_{s=1}^S \Delta P_s$). If the overall pressure drop is higher than the desired value, then the packing type or size can be changed and another analysis carried out.

- 20 The desorber flow chart shows a further example embodiment of a part of the mathematical model of the solvent based carbon capture process detailing an integrated modelling and design methodology of a desorber column. The desorber model, though fundamentally similar to the absorber, has further complexity. Basically, the objective of the desorber is to heat the liquid in order to force the reverse kinetics to release the target component (CO_2). In the cases when the saturation temperature of solvent and solute are
- 25 very different then part of the solvent mixture can be evaporated to provide the stripping fluid. For instance, MEA has a saturation temperature of 170.4°C at atmospheric pressure, which is quite remote from that of water, being 100°C . Therefore, part of the liquid reaching the bottom of column can be heated (reboiled) to a certain temperature to generate steam. Further to this the recycled liquid from the condenser can also be used for

- 20 -

steam generation as shown below in the schematic of a desorber. Reboiler pressure and temperature are among the most critical (if not the main) parameters in desorber design. The reboiler pressure should be enough to compensate for the pressure drop of the packed column and allow the gas phase (steam) to move up. That is the minimum requirement for reboiler pressure. However, high pressures are usually preferred for the desorber because of the intensity of the process and the ability to work at higher temperatures without fear of more liquid vaporization. However, the pressure has an upper limit, enforced by (1) the economics of the system, either capital expenditures (CAPEX) or operational expenditures (OPEX), and (2) the solvent degradation temperature (T^{Deg}). Therefore, it is expected to have an optimum operating pressure based on any solvent used.

The reboiler temperature is more complex than the pressure. Theoretically, a higher desorber (reboiler) temperature will expedite the regeneration process. However, there is an upper bound for this temperature, enforced by process techno-economics and by two main issues: (1) a higher reboiler temperature means higher quality of utility steam, which increases the operating costs of the process; (2) high temperatures result in solvent degradation. Therefore, for any solvent there is a degradation temperature (T^{Deg}) that the reboiler temperature should not exceed. As such, the reboiler temperature constraint can be written as

$$T_{Resb} < \min(T^{sat}(P), T^{Deg})$$

where $T^{sat}(P)$ is obtained from flash vapor-liquid equilibrium calculations at pressure P . A simple $T^{sat}(P)$ can be obtained from Raoult's law, given by

$$T^{sat}(P) = x_{solvent} T_{solvent}^{sat}(P) + (1 - x_{MEA}) T_{H_2O}^{sat}(P)$$

in which $x_{solvent}$ is the molar fraction of the solvent in the solution. $T_{solvent}^{sat}(P)$ and $T_{H_2O}^{sat}(P)$ are the saturation temperatures of pure solvent and water at pressure P , respectively, which are usually presented by an Antoine equation.

The flowrate of the gas stream is another concern. Certainly, the objective is to use just enough steam to achieve the regeneration objective, as any extra steam will just add to

- 21 -

costs. The exit gas from the column top will contain mainly the target gas (CO_2) and steam. Obviously the partial pressure of the target gas (TG) should be less than the equilibrium partial pressure of the inlet liquid $\left[(p_{\text{CO}_2})_{s=S} \leq p_{\text{CO}_2} @ \alpha = \alpha_{\text{int}}^{\text{Des}} \right]$. Otherwise, reverse mass flow will occur, meaning that instead of desorbing the target component from the liquid, more TC will be absorbed. As such, the flow rate of steam at the bottom of the reboiler should be such that at the top of column the partial pressure of steam becomes greater than " $P - p_{\text{CO}_2}(\alpha_{\text{int}}^{\text{Des}})$ ". Moreover, the temperature of the gas at the top of the column should be higher than the inlet liquid temperature ($T_{\text{int}}^{\text{L}} < T_{\text{out}}^{\text{G}}$), to allow regeneration to initiate from the column top. It is therefore necessary that before desorber modeling the reboiler temperature is decided and fed into the program. The liquid temperature at the bottom of the column can also be set a few degrees below the reboiler temperature ($T_{\text{out}}^{\text{L}} = T_{\text{Reb}} - \Delta T$).

- Considering all these constraints, the gas flowrate is computed. Then the average physico-chemical properties are computed, from where the Y_{CO_2} (and thus $Y_{\text{H}_2\text{O}}$) computed.
- 15 If the assumptions have been selected inappropriately, it is possible that this value becomes meaningless or even negative. Therefore, it is necessary to check that the value is within the constraints; if it is not, the program should be updated with a new guess. The remainder of this algorithm is similar to that of the absorber, being calculation of the diameter, ODEs solution, and pressure drop computation.
- 20 The ODEs will solve until the loading of step $s = S$ becomes similar to loading of inlet liquid.

The development of objective function may be related to external factors that can influence the economic viability of the operation of the solvent based carbon capture process. For example, the objective function may be the minimization of OPEX, minimization of CAPEX, maximization of internal rate of return (IRR), maximization of NPV, etc. As an example a formulation for NPV of cash flow is outlined below:

The net present value (NPV) is given by,

- 22 -

$$NPV = -TCX + \sum_{t=M+1}^{M+N} \frac{CF_t}{(1+r)^t}$$

Where, TCX is the total capital expenditure of the PCC plant; CF_t is cash flow at year t ; r is the discount rate; M is number of plant construction years and N is the number of years for plant operation. The cash flow at year t (CF_t) is given by,

$$CF_t = EBT_t - \tau TI_t + \varepsilon_n TCX$$

Where EBT_t is earning before tax at year t , TI_t is the taxable income ($TI_t \geq EBT_t$), τ is fractional tax, ε_t represent depreciation fractional values for MACRS method (Modified Accelerated Cost Recovery System). The earnings before taxes at year t (EBT_t) is given by,

$$EBT_t = R_t - OX_t - \varepsilon_n TCX$$

where, R_t is the revenue (\$) from captured CO_2 during year t , and OX_t is the OPEX of PCC plant during year t .

The total CAPEX (TCX) and the annual OPEX (OX_t) are both function of the operation and design variables and are represented as follows:

$$OX_t = f_0(X_1, X_2, X_3, X_4, X_5, X_6, X_7, X_8, X_9, X_{10}, X_{11}, X_{12}, X_{13}, X_{14}, X_{15}, X_{16}, X_{17}, X_{18}, X_{19}, X_{20}, X_{21}, X_{22}, X_{23})$$

$$TCX = CX_{Absorber(s)} + CX_{Desorber(s)} + CX_{Reboiler(s)} + CX_{Condenser(s)} + CX_{Heat Ex.} + CX_{pump(s)} \quad X_{24}$$

Where φ is price escalation factor and,

$$CX_{Absorber(s)} = f_1(\varphi, X_1, X_2, X_4, X_5, X_6, X_7, X_8, X_{10}, X_{11}, X_{12}, X_{13}, X_{14})$$

$$CX_{Desorber(s)} = f_2(\varphi, X_3, X_5, X_6, X_7, X_9, X_{15}, X_{16}, X_{17}, X_{18}, X_{19}, X_{20})$$

$$CX_{Reboiler(s)} = f_3(\varphi, X_3, X_5, X_6, X_9, X_{15}, X_{18}, X_{19}, X_{20}, X_{23})$$

$$CX_{Condenser(s)} = f_4(\varphi, X_5, X_6, X_9, X_{18}, X_{19}, X_{21})$$

- 23 -

$$CX_{Heat\ Ex.} = f_5(X_2, X_3, X_4, X_5, X_6, X_{10}, X_{13}, X_{15}, X_{20}, X_{22})$$

$$CX_{pump(s)} = f_6(X_4, X_5, X_6, X_8, X_9, X_{10}, X_{12}, X_{24})$$

$f_0 - f_6$ are cost functions. With this the formulation for NPV of cash flow is complete.

5

The method described can be performed using a processing system, an example of which is shown in Figure 2. In particular, the processing system 100 generally includes at least one processor 102, or processing unit or plurality of processors, memory 104, at least one input device 106 and at least one output device 108, coupled together via a bus or group of buses 110. In certain embodiments, input device 106 and output device 108 could be the same device. An interface 112 can also be provided for coupling the processing system 100 to one or more peripheral devices, for example interface 112 could be a PCI card or PC card. At least one storage device 114 which houses at least one database 116 can also be provided. The memory 104 can be any form of memory device, for example, volatile or non-volatile memory, solid state storage devices, magnetic devices, etc. The processor 102 could include more than one distinct processing device, for example to handle different functions within the processing system 100.

Input device 106 receives input data 118 and can include, for example, a keyboard, a pointer device such as a pen-like device or a mouse, audio receiving device for voice controlled activation such as a microphone, data receiver or antenna such as a modem or wireless data adaptor, data acquisition card, etc. Input data 118 could come from different sources, for example keyboard instructions in conjunction with data received via a network. Output device 108 produces or generates output data 120 and can include, for example, a display device or monitor in which case output data 120 is visual, a printer in which case output data 120 is printed, a port for example a USB port, a peripheral component adaptor, a data transmitter or antenna such as a modem or wireless network adaptor, etc. Output data 120 could be distinct and derived from different output devices, for example a visual display on a monitor in conjunction with data transmitted to a

- 24 -

network. A user could view data output, or an interpretation of the data output, on, for example, a monitor or using a printer. The storage device 114 can be any form of data or information storage means, for example, volatile or non-volatile memory, solid state storage devices, magnetic devices, etc.

5

In use, the processing system 100 is adapted to allow data or information to be stored in and/or retrieved from, via wired or wireless communication means, the at least one database 116. The interface 112 may allow wired and/or wireless communication between the processing unit 102 and peripheral components that may serve a specialised purpose.

10 The processor 102 receives instructions as input data 118 via input device 106 and can display processed results or other output to a user by utilising output device 108. More than one input device 106 and/or output device 108 can be provided. It should be appreciated that the processing system 100 may be any form of terminal, server, specialised hardware, or the like.

15

The processing system 100 may be a part of a networked communications system 200, as shown in Figure 3. Processing system 100 could connect to network 202, for example the Internet or a WAN. Input data 118 and output data 120 could be communicated to other devices via network 202. Other terminals, for example, thin client 204, further processing
20 systems 206 and 208, notebook computer 210, mainframe computer 212, PDA 214, pen-based computer 216, server 218, etc., can be connected to network 202. A large variety of other types of terminals or configurations could be utilised. The transfer of information and/or data over network 202 can be achieved using wired communications means 220 or wireless communications means 222. Server 218 can facilitate the transfer of data between
25 network 202 and one or more databases 224. Server 218 and one or more databases 224 provide an example of an information source.

Other networks may communicate with network 202. For example, telecommunications network 230 could facilitate the transfer of data between network 202 and mobile or
30 cellular telephone 232 or a PDA-type device 234, by utilising wireless communication means 236 and receiving/transmitting station 238. Satellite communications network 240

- 25 -

could communicate with satellite signal receiver 242 which receives data signals from satellite 244 which in turn is in remote communication with satellite signal transmitter 246. Terminals, for example further processing system 248, notebook computer 250 or satellite telephone 252, can thereby communicate with network 202. A local network 260, which
5 for example may be a private network, LAN, etc., may also be connected to network 202. For example, network 202 could be connected with ethernet 262 which connects terminals 264, server 266 which controls the transfer of data to and/or from database 268, and printer 270. Various other types of networks could be utilised.

10 The processing system 100 is adapted to communicate with other terminals, for example further processing systems 206, 208, by sending and receiving data, 118, 120, to and from the network 202, thereby facilitating possible communication with other components of the networked communications system 200.

15 Thus, for example, the networks 202, 230, 240 may form part of, or be connected to, the Internet, in which case, the terminals 206, 212, 218, for example, may be web servers, Internet terminals or the like. The networks 202, 230, 240, 260 may be or form part of other communication networks, such as LAN, WAN, ethernet, token ring, FDDI ring, star, etc., networks, or mobile telephone networks, such as GSM, CDMA or 3G, etc., networks,
20 and may be wholly or partially wired, including for example optical fibre, or wireless networks, depending on a particular implementation.

The processing system 100 may be configured by a computer readable medium including executable instructions in the form of a computer program. Upon execution of the
25 computer program, the processing system 100 is configured to perform the method described above. The computer readable medium may be memory such as a hard drive of the processing system 100, or portable memory (CD, DVD, etc.) or the like.

In use, the user inputs via the input device 106 of the processing system 100 the
30 information including percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type. This information is stored in memory

- 26 -

104 of the processing system 100 together with the mathematical model. This information are known parameters for the specific solvent based carbon capture process. The mathematical model is dependent upon the one or more design parameters and the one or more operational parameters.

5

The processing system 100 then determines the size and/or configuration of the solvent based carbon capture process. In particular, the determination step includes the processor 102 iteratively manipulating the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process in accordance with one or more objective functions that have been selected by the user. More specifically, the user is presented via the output device 108 of the processing system a list indicative of the one or more objective functions that can be applied by the processor 102 of the processing system 100 to iteratively manipulate the value of the one or more design parameters and the one or more operational parameters within the mathematical model until the one or more objective functions are optimised. By iteratively optimising the objective functions, the processor 102 is able to determine the size and/or configuration of the solvent based carbon capture process which meets the one or more objective functions. The determine the size and/or configuration of the solvent based carbon capture process can then be output via the output device 108. It will be appreciated that due to the complex nature of the optimisation process, such a technique cannot be performed manually in any feasible timeframe and thus an automated process using the processing system is quite advantageous in achieving this outcome.

25 The processing system 100 preferably has stored in memory optimisation software for manipulating the one or more design parameters and the one or more operational parameters in order to optimise the objective functions. Such optimisation software may apply a global optimisation technique which can include techniques such as deterministic methods, stochastic methods, or heuristics and metaheuristics.

30 It will be appreciated that a non-transient computer readable medium can be provided for configuring the processing system to perform the described method.

- 27 -

The present invention will become better understood from the following example of a preferred but non-limiting embodiment thereof:

5 Example 1

The following example involves the design and/or configuration of a solvent based carbon capture process for use with a 300 MWe coal-fired power plant in Australia burning pulverized black coal.

- 10 The plant uses black coal with 25% ash, 8% moisture, and dry-ash-free (DAF) composition of 83.3% carbon, 5.4% hydrogen, 1.9% nitrogen, 0.6% sulfur, and 8.8% oxygen. The power plant emits 12260 mol/s (353.3 kg/s) of flue gas when operating at full capacity. The flue gas composition is 13.0 vol% (19.8 wt%) CO₂, 70.37% N₂, 13.52% H₂O, 3.11% O₂ and ppm level of SO_x and NO_x. The SO_x and NO_x are removed from the flue gas prior
- 15 to the emission entering into the proposed solvent based carbon capture process. As the temperature of flue gas is usually above 100 °C, and as the optimal operation of the absorber is far below 100 °C (40-60 °C for MEA), the flue gas will be required to be cooled to a temperature, T_{int}^G , prior to entering the absorber.
- 20 The example objective is to design and/or configure a solvent based carbon capture process with base-case capacity or the ability to capture 90% of the CO₂ in the flue gas.

Absorber

- By running the model as discussed above, the results of the first run of the program
- 25 showed that the size of an absorber operating under the above-mentioned base-case conditions would be very large, with a diameter in the order of 20 meters. For this reason it was decided to design two absorbers of similar sizes to work in parallel. Therefore, the

- 28 -

inlet flowrate of each absorber would be half of the total flowrate i.e. 6130 mol/s (176.65 kg/s).

The temperature of 50 °C (T_{inr}^G) was used for the inlet flue gas. Together with this inlet temperature, the remainder of the inlet gas information is known.

5

The quantity of CO₂ in the outlet gas stream is calculated as having the 90% capture objective. Estimate values are made for the composition of other gas components and also the outlet gas temperature. The inlet liquid temperature is taken as 53 °C (a few degrees higher than inlet gas temperature). The CO₂ loading of liquid is taken as 0.2 and 0.45 at
10 inlet and outlet of column, respectively. With these loadings, the concentrations of components are computed from where the outlet temperature of the liquid is calculated.

Two-inch (50 mm) ceramic berl saddle packing was chosen as a packing with specific area of 105 m², porosity of 0.72, packing factor of 150 (F_p) and dry bed packing factor (F_{pd})
15 of 102. The pressure at the column bottom was designed to be 1.1 bar. The gas should leave the column, on top, at a pressure greater than or equal to atmospheric pressure. Therefore, the maximum allowable pressure drop over the column is about 0.1 bar.

With the above data, the model outlined above was solved and converged in just the
20 second iteration. In this example, the column has a diameter of 9.80 m and length of 18.4 m. For this design, liquid with the flowrate of 0.573 m³/s or 25112 mol/s is required. This translates to L_{mol}/G_{mol} of 4.1. The diameter design was taken based on gas velocity being 80% of that in flooding (2.11 m/s vs. 2.64 m/s). That velocity guarantees that the inlet gas with pressure of 1.1 bar will leave the column at above atmospheric pressure (not below).
25 Figure 1-1 illustrates the profiles of pressure and pressure-drop over column height. The gas enters the column bottom at 1.1 bar and exits from the top at 1.0148 bar, being slightly above atmospheric pressure.

- 29 -

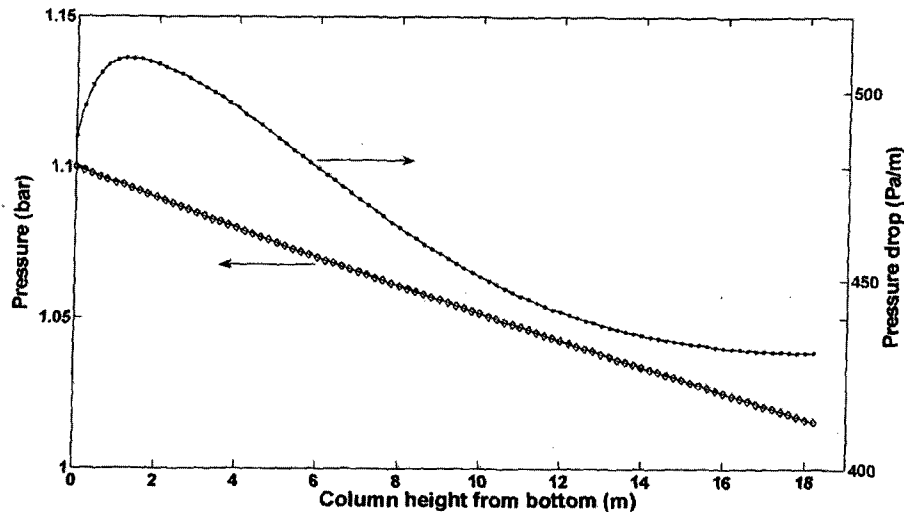


Figure 1-1: Profile of flue gas pressure and pressure-drop over absorber column height, using berl saddle 50 mm and at design velocity of 80% flooding. The liquid inlet and outlet loadings are 0.2 and 0.45 with the carbon capture target of 90%. The inlet liquid temperature is 53 °C and the gas temperature is 50 °C.

5

Figure 1-2 shows the temperature profile of gas and liquid over the column height. Once the lean liquid enters the column from top (at 326.15 K), it starts reacting with the flue gas as the gas progresses from the bottom upward. As the reaction is exothermic, the temperature increases down to the bottom of the column. The reduction in temperature in a small section at the bottom of column is due to the gas entering the column at lower temperature. The so-called “rich” liquid leaves the column from the bottom at the temperature of $T_{out}^L = 336.15$ K (63 °C) while the clean flue gas is emitted to the atmosphere from the top of the column at $T_{out}^G = 326.31$ K (63.16 °C).

15

The concentration profiles of the main components of the liquid are illustrated in Figure 1-3. The 30 wt% MEA solvent with a loading of 0.2 enters the column height at a concentration of 3.0 kmol/ m³. Moving down the height of the column it reacts with CO₂ and the concentration of pure MEA is decreased until the value of 0.5 kmol/m³ is reached at the bottom of the column. At the same time the composition of reaction products such as carbamate increases from an initial value of 1.0 kmol/m³ to 2.2 kmol/ m³ at the bottom of the column until the rich liquid leaves the column towards the regeneration unit.

20

- 30 -

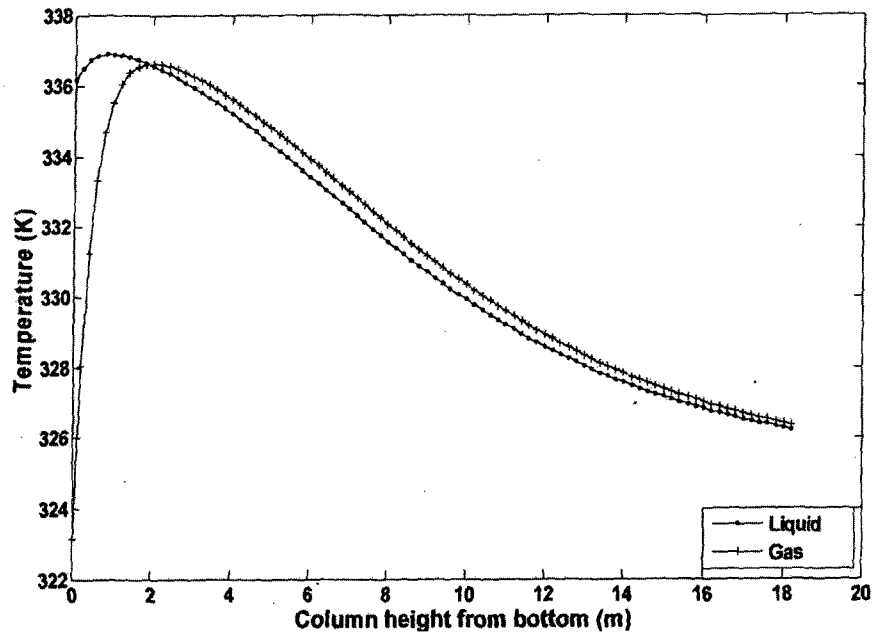


Figure 1-2: Temperature profile of the flue gas and liquid over absorber column height, using berl saddle 50 mm packing and at design velocity of 80% flooding. The liquid inlet and outlet loadings are 0.2 and 0.45 with the carbon capture target of 90%.

5

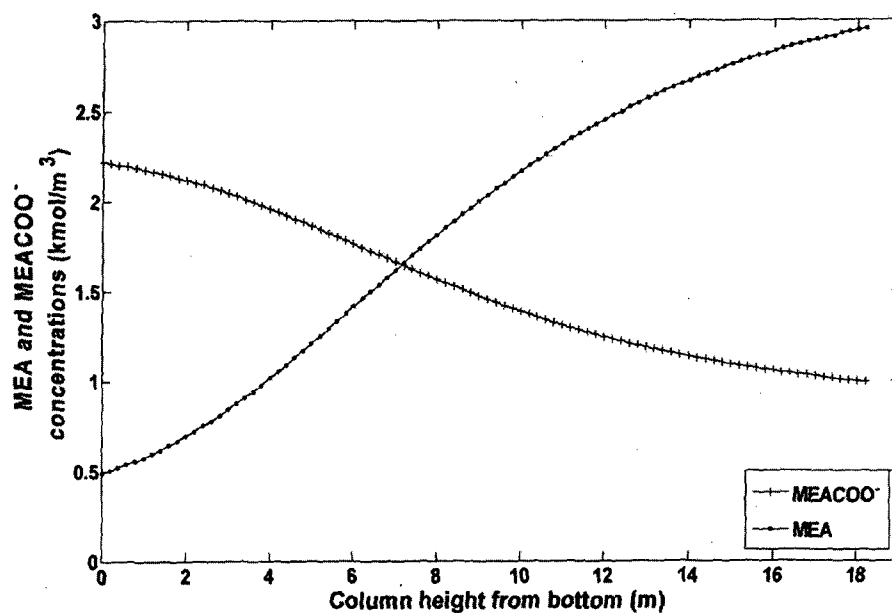


Figure 1-3: Concentration profiles of MEA and MEACOO⁻ of liquid temperatures over absorber column height when the liquid inlet and outlet loadings are 0.2 and 0.45 respectively.

10

Figure 1-4 illustrates the profile of loading of CO₂ in the liquid (hereinafter referred to as “loading”). The liquid enters the column with a loading of $\alpha_{\text{inat}}^{\text{Abs}} = 0.20$ (lean liquid) and

- 31 -

leaves with a loading of $\alpha_{out}^{Abs} = 0.45$ (rich liquid). The figure illustrates how the loading changes over the column and how it corresponds with the rate of CO₂ capture. The importance of the figure is that if the outlet loading is fixed at 0.45, and all the design and operating parameters are fixed at the values discussed earlier, then to capture 90% of the CO₂ from the flue gas (the design objective), the CO₂ loading of the inlet liquid must be 0.2. However, to capture 80% of the CO₂, the inlet loading of 0.23 and column height of 14.0 meter will achieve the goal. Similarly, to capturing only 50% of the CO₂, the corresponding inlet loading is 0.31 and a column height of 8.2 m will satisfy the goal.

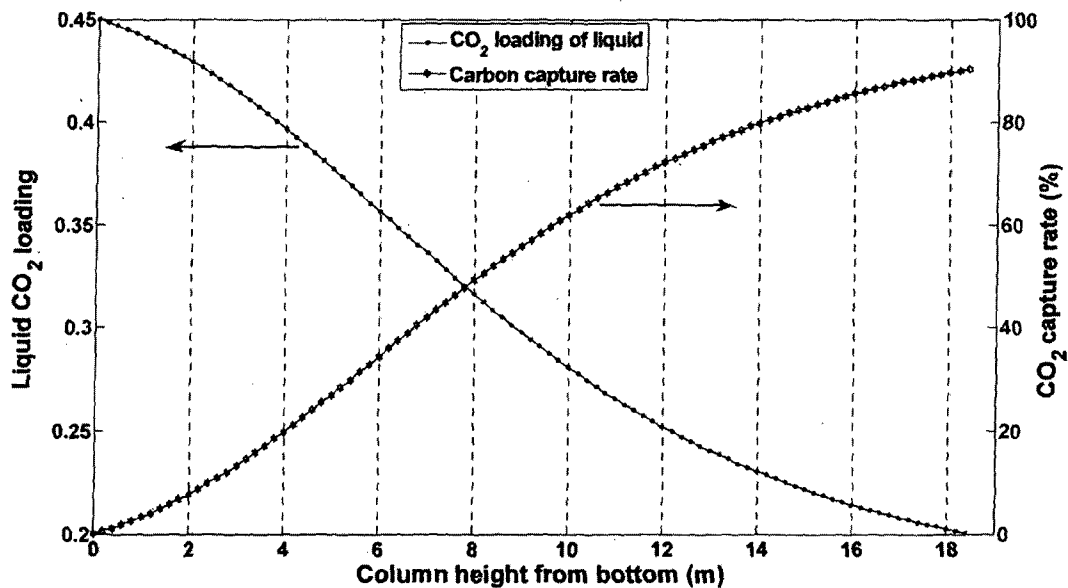
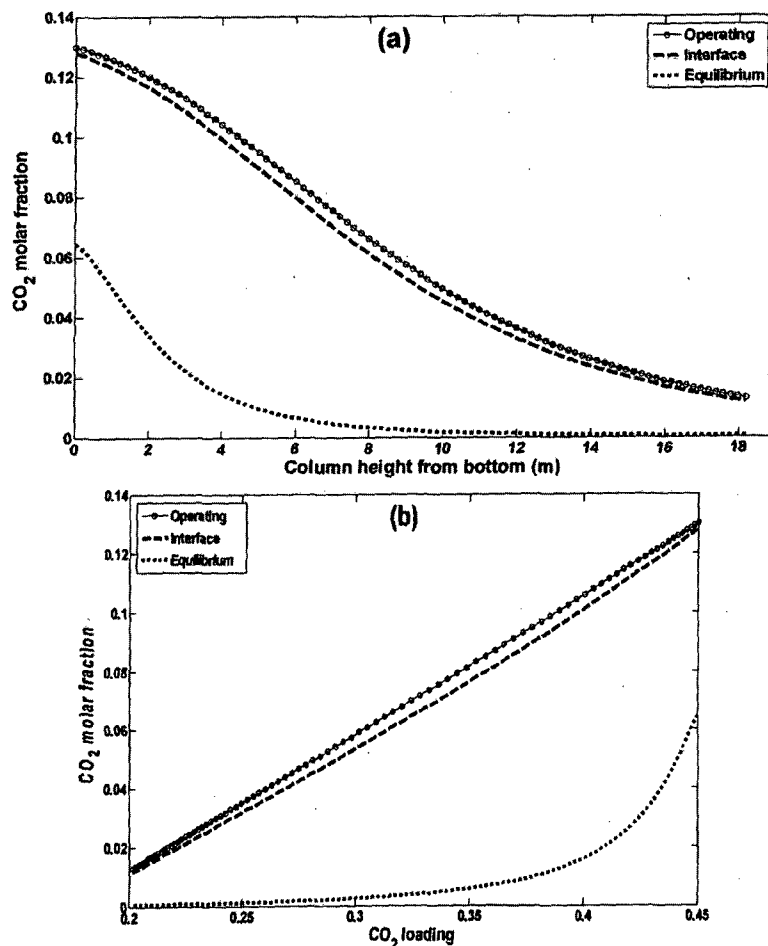


Figure 1-4: Relation between inlet liquid loading and capture rate using berl saddle 50 mm packing and at design velocity of 80% flooding for all examples; the base case design diameter of 9.8 m, flue gas with 13% CO₂.

The next profiles presented are the equilibrium, interface and operating curves. The assumption of equilibrium-based column modeling is that at every stage, the gas and liquid phases reach their equilibrium or a percentage of it. In the rate-based modeling, however, the gas-liquid mass transfer rate is function of numerous parameters which were earlier discussed. The gas-liquid equilibrium is therefore just one of the (limiting) conditions which might occur. Figure 1-5 illustrates the profiles of CO₂ at equilibrium, interface and operation over (a) column height, and (b) loading. The flue gas enters the column with the

- 32 -

CO₂ fraction of 13% and leaves it with $\beta = 90\%$ of the CO₂ having been captured (outlet fraction of 0.0131). From fundamental separation textbooks it is known that for absorption the operating line is above the equilibrium line and the difference between these two curves is an indication of the mass transfer driving force. It is of interest that the operating curve is remote from equilibrium. From these triple curves that the author believes have not been previously presented, reverse-engineering is possible to present the corresponding mass transfer efficiency over the column height or loading. Figure 1-5c illustrates the efficiency profile and implies that the mass transfer rate efficiency is maximum 10.3% at the top of column and reaches 3.14% at the bottom of the column when CO₂ loading is high.



- 33 -

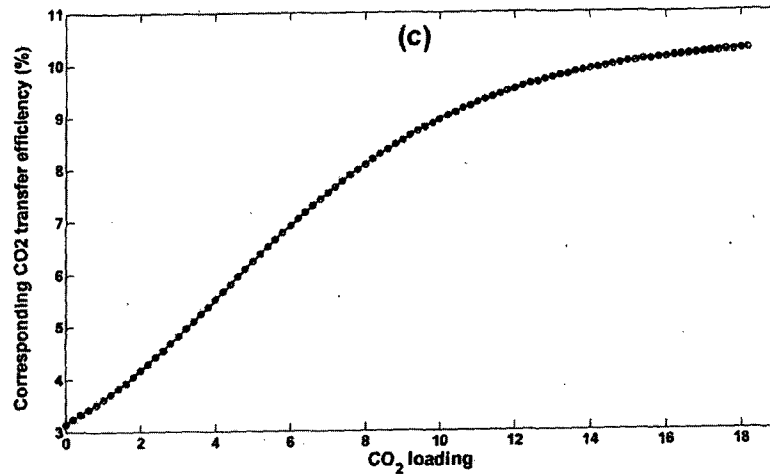


Figure 1-5: Profiles of CO₂ equilibrium, interface and operation over (a) the column height, and (b) loading. Figure (c) is the corresponding efficiency (%).

- 5 It would be of interest to study the mass transfer efficiency (Figure 4-11c) along with the enhancement factor. Figure 1-6 illustrates the enhancement factor over the column height, being at its lowest value of 25.9 at the bottom of the column and at its highest value of 85.3 at the top of the column. Therefore the mass transfer efficiency is identical to the enhancement factor.

10

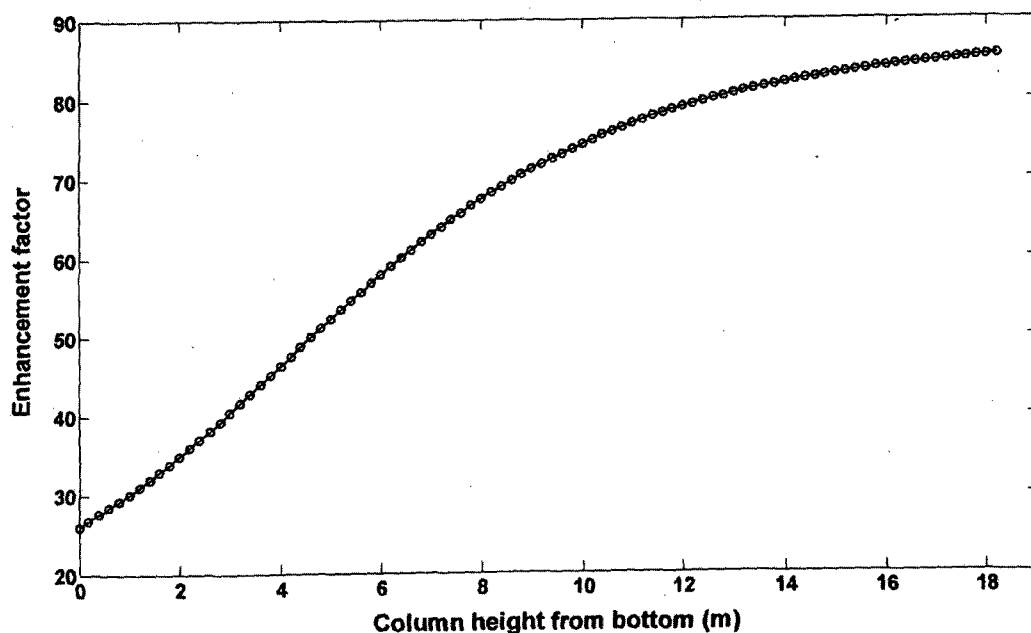


Figure 1-6: Profile of mass transfer enhancement factor (due to solvent and CO₂ reaction) over the absorber column height.

Parametric analysis of absorber

- 5 Parametric analysis is performed over some important parameters to quantitatively evaluate the impact of key parameters in the design and performance of the absorber column.

Impact of packing type and size

- 10 The impact of six different packing types on column sizing was investigated. The results are illustrated in Figure 1-7. It is clear from the figures that packing type has a notable impact on both operating and design parameters. Although all the given conditions except packing type are similar, the application of 50 mm berl saddle packing results in an absorber column with size of 9.8 m diameter and 18.4 m height along with a total pressure
- 15 drop of 8.5 kPa over the column. The second best option is a 50 mm Intalex saddle application, which results in a column with diameter of 9.5 m and height of 18.2 m. However, it has a greater pressure drop of 14.35 kPa. Therefore, if this packing is used the column pressure at the bottom should be above 1.157 bar. It is of interest that a 75 mm raschig ring results in a relatively tall column with diameter of 9.3 m and height of 33 m.
- 20 This is not the only difference; it further results in a notable pressure drop of 66.19 kPa over the column. Therefore, the significant impact of packing type on column design is demonstrated, and based on the few examples investigated a 50 mm berl saddle was selected as a preferred packing type.

- 35 -

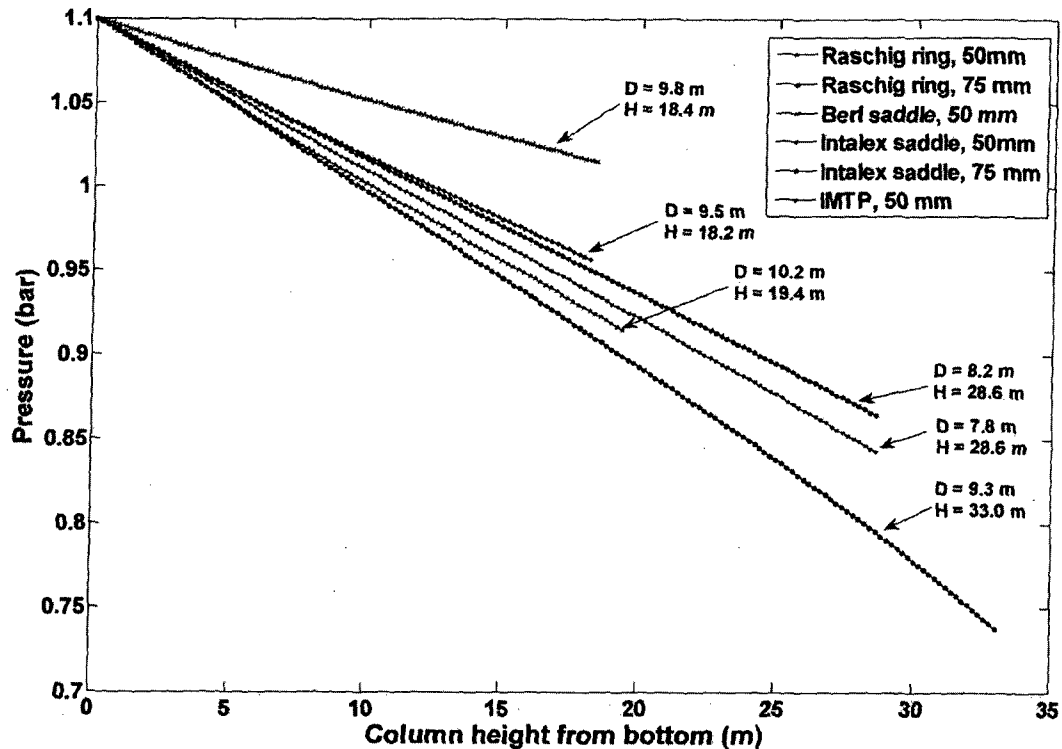


Figure 1-7: Impact of packing type/properties on column size and pressure drop at design velocity of 80% flooding for all examples where the flue gas inlet pressure is 1.1 bar.

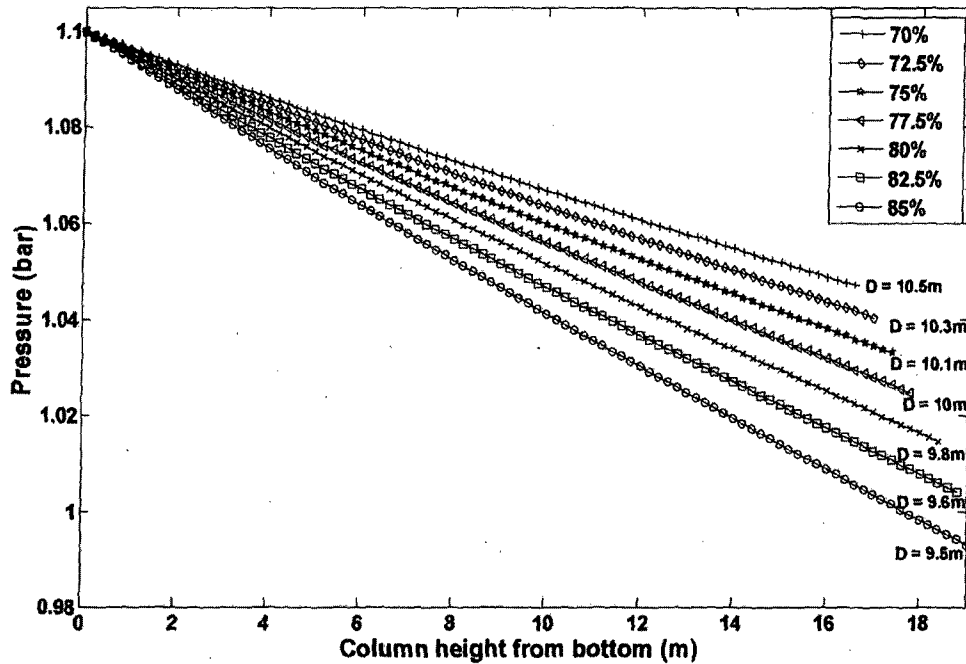
5

Impact of gas velocity/flux

As discussed earlier, in order to prevent flooding, the gas velocity is typically designed to be in the conservative range of about 70-85% of flooding velocity. Here the impact of the design velocity on both column design and performance is investigated. Seven different velocities are used, i.e. 70%, 72.5%, 75%, 77.5%, 80%, 82.5% and 85% of the flooding velocity. The results are given in Figure 1-8. It is evident that an increase of velocity has a slight impact on column size, in that it slightly increases height and slightly decreases column diameter. For instance, the design velocity of 70% results in a column with diameter of 10.5 m and height of 16.6 m, whereas at 85%, the size becomes 9.5 m diameter (9.5% decrease) and 19.2 m height (15.7% increase). The pressure drop for the 70% scenario will be 5.30 kPa through the packing height, while for the 85% scenario it will be 10.80 kPa, almost double of that for 70%.

- 36 -

If a higher velocity is selected, it will result in higher pressure drop and the outlet pressure will be less than atmospheric pressure. This means that a higher inlet pressure (higher than 1.1 bar) would be required, translating to higher operating costs.



5

Figure 1-8: Impact of design velocity on column size and pressure profile.

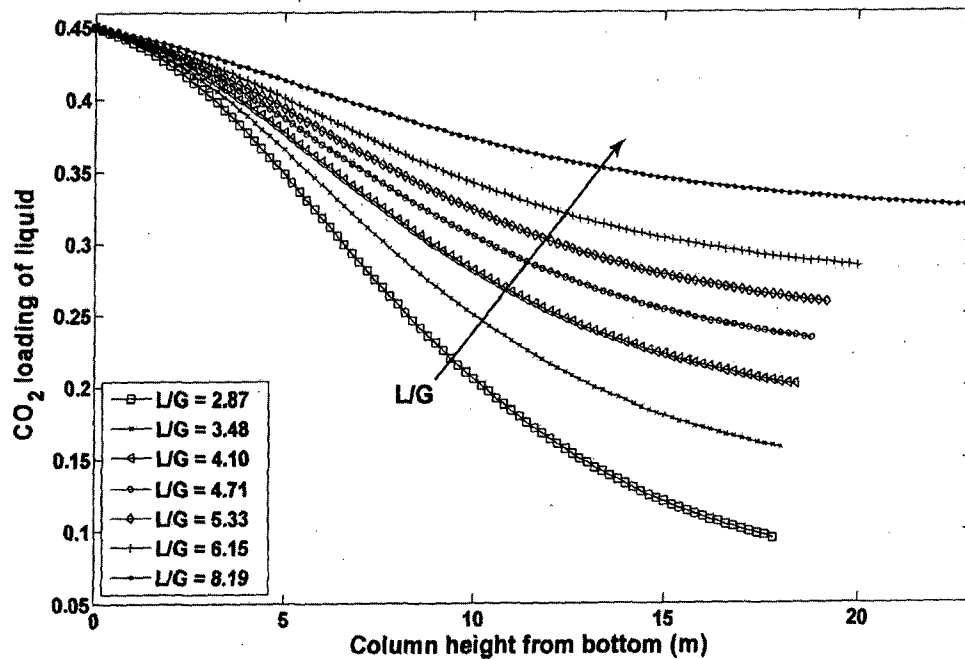
Impact of gas/liquid ratio

- 10 The gas/liquid flowrates are also among the key parameters of column design and operation. Their impact on at least the operating cost of the solvent based carbon capture process (OPEX) is non-trivial. For the base model, when the the liquid loadings at 0.2 and 0.45 are fixed, the corresponding L_{mol}/G_{mol} ratio becomes 4.10. The same example has been worked with the same given parameters, and studying the impact of liquid flow rate on
- 15 column design and operating parameters. The column diameter was fixed at 9.8 m for all parameters so that the impact on design was revealed via the column height. Figure 1-9 illustrates the impact of seven different liquid flowrates (flue gas flowrate, G_{mol} , being fixed for all cases) on process. The outlet liquid loading is fixed at 0.45 and the inlet loading is obtained by the model. It is evident from the figure that an increase in liquid
- 20 flowrate results in an increase in column height and inlet liquid loading. To facilitate

- 37 -

quantitative discussion, Table 1-1 lists some of the key parameters for each of the seven scenarios. For instance, whereas for flowrate of 17800.54 mol/s ($L_{mol}/G_{mol} = 2.87$) the column height is 17.8 m, it increases to 18.4 m at $L_{mol}/G_{mol} = 4.10$ and to 22.8 m at $L_{mol}/G_{mol} = 8.19$. The inlet liquid loading for $L_{mol}/G_{mol} = 2.87$ is 0.095 and it becomes

5 0.200 for $L_{mol}/G_{mol} = 4.10$ and 0.326 for $L_{mol}/G_{mol} = 8.19$. The implication of these values is that for a certain CO₂ capture objective (e.g. 90% of flue gas), when the liquid flowrate is increased, then lower inlet loading is required (considering a fixed outlet loading).



10 Figure 1-9: Profile of CO₂ loading along the column length and its impact on column height at fixed diameter of 9.8 m.

15

Table 1-1: Design output data for various L/G studies.

L_{mol} (mol/s)	L_{mol}/G_{mol}	$\alpha_{in}^{abs} \rightarrow \alpha_{out}^{abs}$	T_{out}^L (K)	T_{in}^L (K)	$y_{CO_2, out}$	D (K)	Z (K)	P_{out} (bar)	ΔP (kPa)	Q^L (lit/s)
$L_{mol}=0.70 L_{mol, base}$ =17800.54	2.87	0.095 $\rightarrow$ 0.45	336.15	321.26	0.0129	9.8	17.8	1.0301	6.99	401.60
$L_{mol}=0.85 L_{mol, base}$ =21567.47	3.48	0.157 $\rightarrow$ 0.45	336.15	324.22	0.0128	9.8	18	1.0231	7.69	487.66
$L_{mol}=1.15 L_{mol, base}$ =29101.47	4.10	0.200 $\rightarrow$ 0.45	336.15	326.19	0.0126	9.8	18.4	1.0143	8.57	573.72
$L_{mol}=1.30 L_{mol, base}$ =32867.96	4.71	0.2333 $\rightarrow$ 0.45	336.15	327.62	0.0126	9.8	18.8	1.0043	9.57	659.78
$L_{mol}=1.5 L_{mol, base}$ =37890.47	5.33	0.2586 $\rightarrow$ 0.45	336.15	328.71	0.0127	9.8	19.2	0.9928	10.72	745.84
$L_{mol}=1.5 L_{mol, base}$ =37890.47	6.15	0.2840 $\rightarrow$ 0.45	336.15	329.80	0.0127	9.8	20	0.9725	12.75	860.58
$L_{mol}=2.0 L_{mol, base}$ =50446.1	8.19	0.3257 $\rightarrow$ 0.45	336.15	331.53	0.0128	9.8	22.8	0.8814	21.86	1147.44

- 38 -

To capture 90% of CO₂ under the given condition $L_{mol}/G_{mol} = 4.10$ is required. With inlet and outlet loadings fixed, it is possible to use lower L_{mol}/G_{mol} and study the impact on the rate of carbon capture. Figure 1-10 illustrates the results of six different L_{mol}/G_{mol} scenarios within the range of 2.05 to 4.10. The details of the results are given in Table 1-2. It is evident that when L_{mol}/G_{mol} is 4.10 the rate of capture is 90%, but with reduction in L_{mol}/G_{mol} value the capture rate decreases until it reaches the lowest value of 46.55% at $L_{mol}/G_{mol} = 2.05$. However, reduction of L_{mol}/G_{mol} also reduces the column height requirement. While the capture of 90% CO₂ requires a column with height of 18.4 m, only 7.6 m column height is required to capture 46% of CO₂ when L_{mol}/G_{mol} is 2.05.

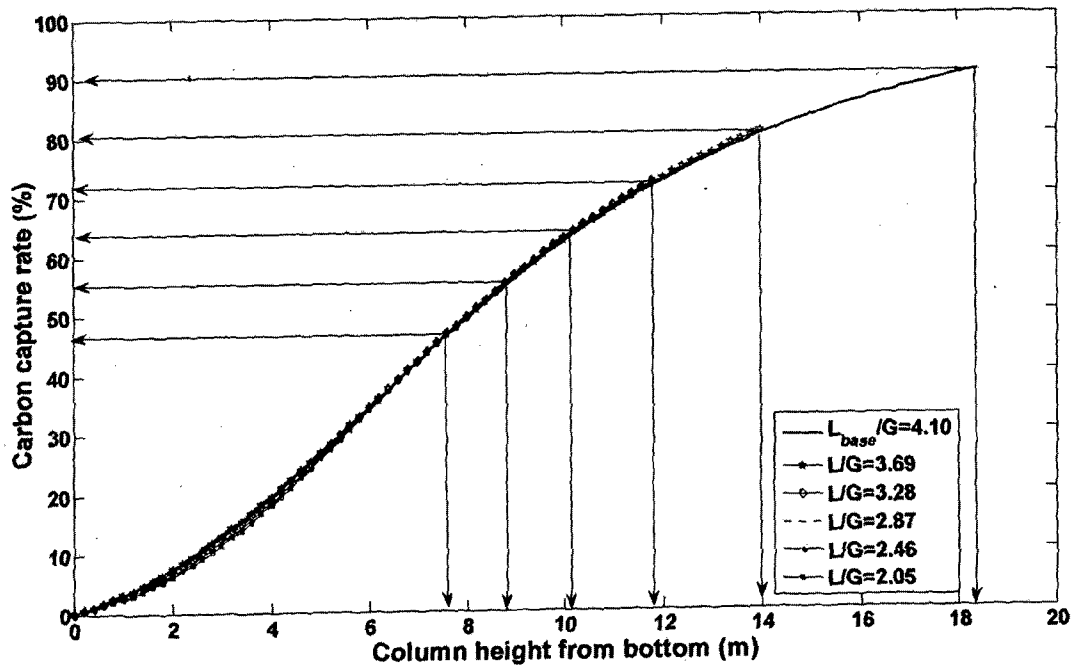


Figure 1-10: Impact of inlet liquid flowrate on column size and carbon capture rate using berl saddle 50 mm packing and at design velocity of 80% flooding for all examples; the base case design diameter of 9.8 m is fixed for all the scenarios. The liquid inlet and outlet alpha is fixed at 0.2 and 0.45.

Table 1-2: Design output data for various L/G studies.

L_{mol} (mol/s)	L_{mol}/G_{mol}	α_{int}	T_{out}^L (K)	T_{int}^L (K)	$y_{CO_2, out}$	Capture rate (%)	D (K)	Z (K)	P_{out} (bar)	ΔP (bar)	Q^L (lit/s)
$L_{mol, base} = 25334.68$	4.10	0.200 → 0.45	336.15	326.19	0.0126	90	9.8	18.4	1.0143	8.57	573.72
$L_{mol} = 0.9 L_{mol, base} = 22797.98$	3.69	0.200 → 0.45	336.15	326.43	0.0257	80.3	9.8	14	1.0372	6.28	516.35
$L_{mol} = 0.8 L_{mol, base} = 20265.88$	3.28	0.200 → 0.45	336.15	326.44	0.0364	72.0	9.8	11.8	1.0492	5.08	458.98
$L_{mol} = 0.70 L_{mol, base} = 17734.63$	2.87	0.200 → 0.45	336.15	326.40	0.0469	63.95	9.8	10.2	1.0580	4.20	401.60
$L_{mol} = 0.60 L_{mol, base} = 15201.88$	2.46	0.200 → 0.45	336.15	326.49	0.0581	55.28	9.8	8.8	1.0652	3.48	344.23
$L_{mol} = 0.50 L_{mol, base} = 12669.12$	2.05	0.200 → 0.45	336.15	326.65	0.0695	46.55	9.8	7.6	1.0712	2.88	286.86

5

Impact of gas/liquid temperature

The absorption process is highly exothermic. Therefore the application of correct liquid and gas temperatures was found to be critical in process design and operation. Here the individual impacts of inlet liquid temperature (T_{int}^L) and inlet flue gas temperature (T_{int}^{IG}) are studied. For analysis of the inlet liquid temperature, inlet flue gas temperature is maintained at 323.15 K (50 °C) and liquids with wide temperature ranges are used. Figure 1-11 illustrates the gas/liquid temperature profiles for four different scenarios. The other details of these scenarios are given in Table 1-3. It is evident that temperature has some small impact on column diameter, which may stem from physical expansion of fluids at high temperatures. However, the effect of liquid temperature on column height is interesting. At the low temperature of 313.15 K (40 °C), the column height is 21.2 m. When the temperature is increased the column height reduces to a minimum, from where it increases again at higher temperatures.

- 40 -

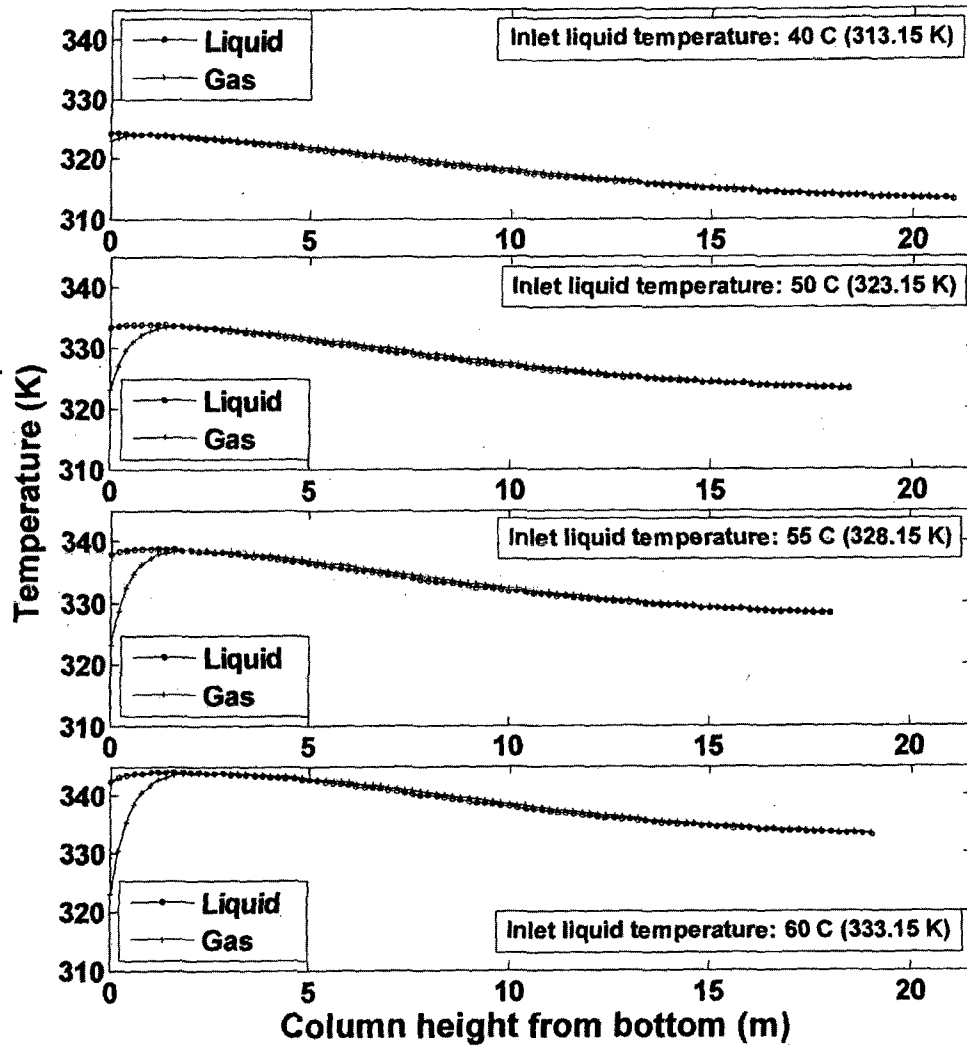


Figure 1-11: Impact of inlet liquid temperature on column size and pressure drop using berl saddle 50 mm packing and at design velocity of 80% flooding for all examples.

5

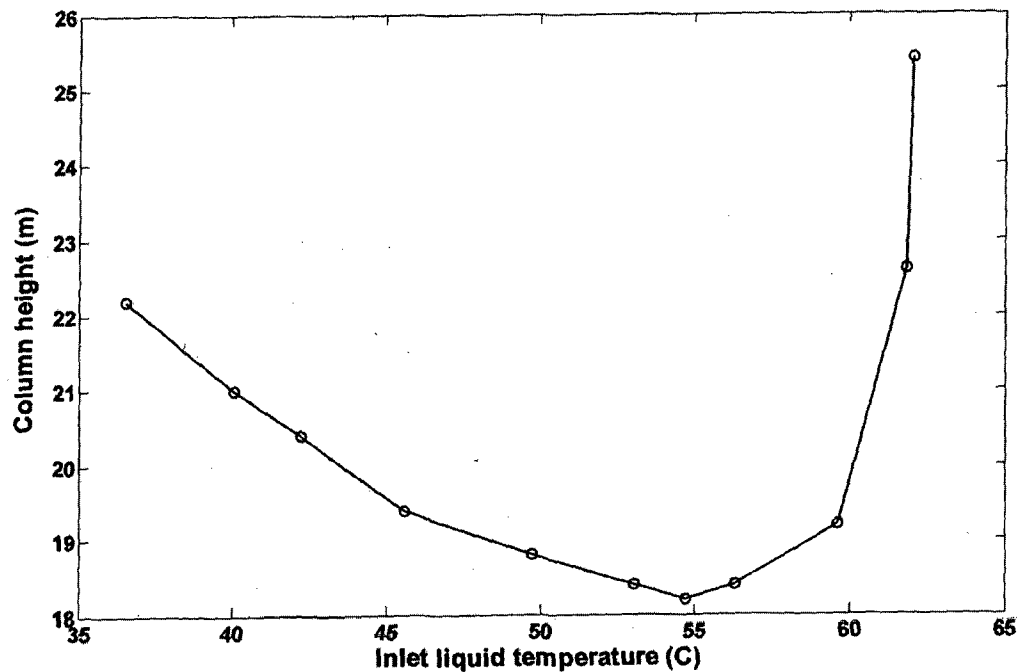
Table 1-3: Design output data for various inlet liquid temperatures studies.

T_{int}^L	T_{out}^L	T_{int}^G	$\alpha_{int}^{Abs} \rightarrow \alpha_{out}^{Abs}$	$y_{CO_2, out}$	D (K)	Z (K)	P_{out} (bar)	ΔP (bar)	Q^L (lit/s)	L_{mol} (mol/s)	L_{mol}/G_{mol}
313.17	324.35	323.15	0.200 $\rightarrow$ 0.45	0.0126	9.77	21.2	1.0029	9.71	573.72	25334.75	4.10
323.13	333.35	323.15	0.200 $\rightarrow$ 0.45	0.0129	9.80	18.6	1.0142	8.58	573.72	25334.18	4.10
328.14	337.9	323.15	0.200 $\rightarrow$ 0.45	0.0128	9.81	18.2	1.0153	8.47	573.72	25334.24	4.10
333.21	342.5	323.15	0.200 $\rightarrow$ 0.45	0.0129	9.83	19.2	1.0087	9.13	573.72	25334.08	4.10

To further analyse the impact of inlet liquid temperature on column height, the column diameter was fixed at 9.8 m and a number of runs were performed over numerous inlet

- 41 -

liquid temperatures. The profile is illustrated in Figure 1-12. It is evident that under the given conditions, the minimum column height is at an inlet liquid temperature of about 327.85 K (54.7 °C).



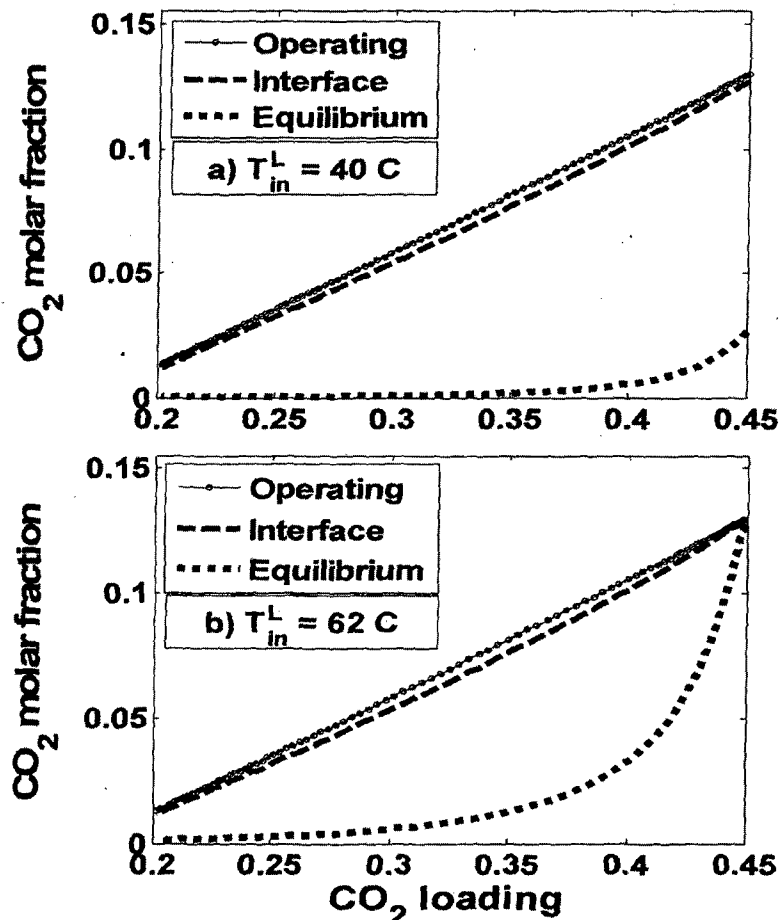
5

Figure 1-12: Impact of inlet liquid temperature on column size using berl saddle 50 mm packing and at design velocity of 80% flooding for all examples; diameter is fixed at 9.8 m.

- 10 There is also another important issue for selection of inlet liquid temperature; as the temperature increases, the equilibrium partial pressure of CO₂ increases. This reduces the mass transfer driving force. The limiting condition can occur when the equilibrium partial pressure of CO₂ at the bottom of the column becomes equal to or higher than the partial pressure of the inlet flue gas. To illustrate this, the equilibrium, interface and operating profiles of the column, over loading, are shown at two different temperatures (Figure 1-13). At the low liquid temperature of 40 °C it is evident that the interface and operating curves are well above the equilibrium curve, indicating that there is a good driving force for mass transfer (Figure 1-13a). However, as the temperature increases the equilibrium curve moves upwards and closer to the other two curves, resulting in the reduction of mass transfer driving force. For the scenario with inlet liquid temperature of 62 °C, when the
- 15
- 20

- 42 -

liquid leaves the bottom of the column at a loading of 0.45 and temperature of 71.2 °C, its equilibrium CO₂ fraction is equal to that of the inlet flue gas. This means that if, under the given conditions, an inlet liquid temperature higher than 62 °C is used, part of the bottom of the column will dysfunction. Therefore, consideration of the upper bound for inlet liquid temperature is necessary for robust/optimal column design and operation. One approach for dealing with high-temperature liquid is to select lower loadings at the column outlet, e.g. $\alpha_{out}^{Abs} < 0.45$ for the above scenario.



10 Figure 1-13: Impact of inlet liquid temperature on equilibrium, interface and operating curves at (a) 40 °C and (b) 62 °C using berl saddle 50 mm packing and at design velocity of 80% flooding for all examples; diameter is fixed at 9.8 m and $P_{int} \approx 1.1$ bar.

- 43 -

- Similarly, a parametric analysis was performed on the impact of flue gas temperature on absorber performance. It was noted that, compared with liquid temperature, the flue gas temperature has less impact on either operating or design variables. Table 1-3 lists the results for five different flue gas temperatures (40 °C, 50 °C, 60 °C, 70 °C, 80 °C) while the inlet liquid temperature is fixed at 53 °C. With increase in the temperature, the column diameter increases slightly (around 1% over a 40 degree increase in temperature). The pressure drop also decreases slightly. To maintain a similar performance when the inlet flue gas temperature increases, the column requires the inlet liquid temperature to be low or the outlet liquid temperature to be high. In this example, when for all scenarios the outlet liquid temperatures are fixed at 63 °C (336.15 K), by an increase in the inlet flue gas temperatures the inlet liquid temperature decreases. As is evident from the table, for inlet gas temperatures of 40 °C the inlet liquid temperature is 54 °C (327.13 K), whereas it decreases by around 7% to 50.2 °C (323.36 K) when the inlet gas temperature reaches 80 °C.
- It can be concluded, therefore, that absorber design and operation are much more sensitive to inlet liquid temperature than to inlet gas temperature, and this can be linked physico-chemically with the higher specific heat capacity of liquid compared with gas.

Table 1-3: Design output data for various inlet liquid temperature studies.

T_{int}^G (°C)	T_{int}^L (K)	T_{out}^L (K)	$\alpha_{int}^{Abs} \rightarrow \alpha_{out}^{Abs}$	$y_{CO_2,out}$	D (m)	Z (m)	P_{out} (bar)	ΔP (kPa)	Q^L (lit/s)	L_{mol} (mol/s)	L_{mol}/G_{mol}
40	327.13	336.15	0.20 $\rightarrow$ 0.45	0.0126	9.78	18.4	1.0131	8.69	573.72	25334.74	4.10
50	326.18	336.15	0.20 $\rightarrow$ 0.45	0.0126	9.81	18.4	1.0148	8.52	573.72	25334.78	4.10
60	325.24	336.15	0.20 $\rightarrow$ 0.45	0.0126	9.83	18.4	1.0163	8.37	573.72	25334.71	4.10
70	324.30	336.15	0.20 $\rightarrow$ 0.45	0.0127	9.86	18.4	1.0177	8.23	573.72	25334.55	4.10
80	323.36	336.15	0.20 $\rightarrow$ 0.45	0.0128	9.88	18.4	1.0189	8.11	573.72	25334.3	4.10

Desorber

- The rich liquid coming from the absorber is the feed to the desorber. Based on the base-case design of the absorber, the liquid enters the absorber with the flowrate of 25111.85 mol/s (0.573 m³/s) and after physico-chemical interaction with flue gas leaves the column

- 44 -

bottom rich with CO₂ loading of 0.45. The objective is to purify the liquid to the extent that it reaches the lean liquid loading predefined by the design as 0.20. As discussed earlier, these loadings may be any desired values selected by design.

As discussed in the absorber section, the design condition enforced selection of two
5 absorbers. Therefore, here it is desired to mix the rich liquids from both absorbers and send them to a single desorber unless the design suggests application of more than one desorber. As such the inlet liquid flowrate for desorber design is about 1.146 m³/s (1220.4 kg/s).

The rich liquid is assumed to be heated to 103 °C through heat exchange with the lean amine recycling back from desorber to absorber. It is expected that during this temperature
10 rise some desorption occurs in the heat exchanger and the loading decreases slightly. For ease of analysis it is assumed that the rich amine enters the desorber at the same loading as it leaves the absorber. The most critical techno-economical parameter in the desorber column design is the steam (referred to as gas) stream. The gas stream is assumed to enter the column at 126 °C, which is the temperature of the reboiler.

15 Unlike many other studies that reboil part of the effluent liquid from the desorber (lean liquid) to generate steam and use the recycled water (from the condenser) as wash-water at the top of the column, the preference here is not to use condenser water for the water-washing task. Water-washing is necessary for the absorber column as it reduces the amount of liquid droplets in the emitted gas and thus reduces the rate of MEA loss.
20 However, the effluent gas from the desorber column is mainly water and CO₂, and the gas stream moves towards the condenser where liquid components can be eliminated by cooling. Therefore, the application of a mechanical mesh at the top of the column may suffice. On this basis, the condensed recycled water from the condenser (being a few degrees below its saturation temperature) is used for reboiling and generating steam. Then
25 the remaining amount of steam required is supplied by reboiling a portion of lean liquid. The significant advantages of this configuration are (1) energy efficiency as a result of better application of recycled water, (2) prevention of salt formation by solvent reboiling, and (3) most importantly, application of steam with a lower CO₂ concentration (compared

- 45 -

with the case of generating 100% of the steam from solvent reboiling) results in a larger mass transfer driving force and thus higher system efficiency.

A desorber column should operate at pressure in the range of 1.7-2.5 bar. In this example pressure of 1.7 bar is used, but later a parametric study of pressure is presented. With these
5 given parameters, with the selected 50 mm berl saddle packing type and the design velocity of 70% the design model outlined above is solved for the desorber. The outcome is a column with diameter of 10.53 m and height of 15.8 m. This implies that one column is sufficient and there is no need for the use of two or more desorber columns. For this design, a gas stream with inlet flowrate of 8499.2 mol/s (160.0 kg/s) is required translating
10 to the L_{mol}/G_{mol} of 5.80. The CO₂ concentration (mole fraction) of the gas under the given condition is 0.031 at the bottom of the column and 0.203 at the outlet of the column (towards the condenser).

The condenser temperature depends on the required CO₂ purity. When a higher CO₂ purity is required, the separated CO₂ may undergo other dehydration processes (e.g., Glycol
15 system or molecular sieves). This is required when the captured CO₂ will be sent by pipeline to remote places for underground sequestration. In such cases the existence of water in the CO₂ stream can result in the formation of hydrates and thus erosion of the pipeline or compressors. Therefore, the condenser temperature selected depends on downstream application of the captured CO₂ and technically is defined by VLE
20 formulations. Reducing the gas temperature down to its bubble-point will separate almost all of the water, but at the cost of higher cooling utility expenses. In this analysis for the given example, a condenser temperature of 80 °C is used. The reboiler duty for the given scenario is 6.04 MJ/kg-CO₂.

Figure 1-14 illustrates the pressure and pressure-drop profiles over the desorber column
25 bed. The pressure drop over the column is 8.27 kPa, meaning that the inlet gas with pressure of 1.7 bar leaves the column to the desorber at 1.617 bar.

- 46 -

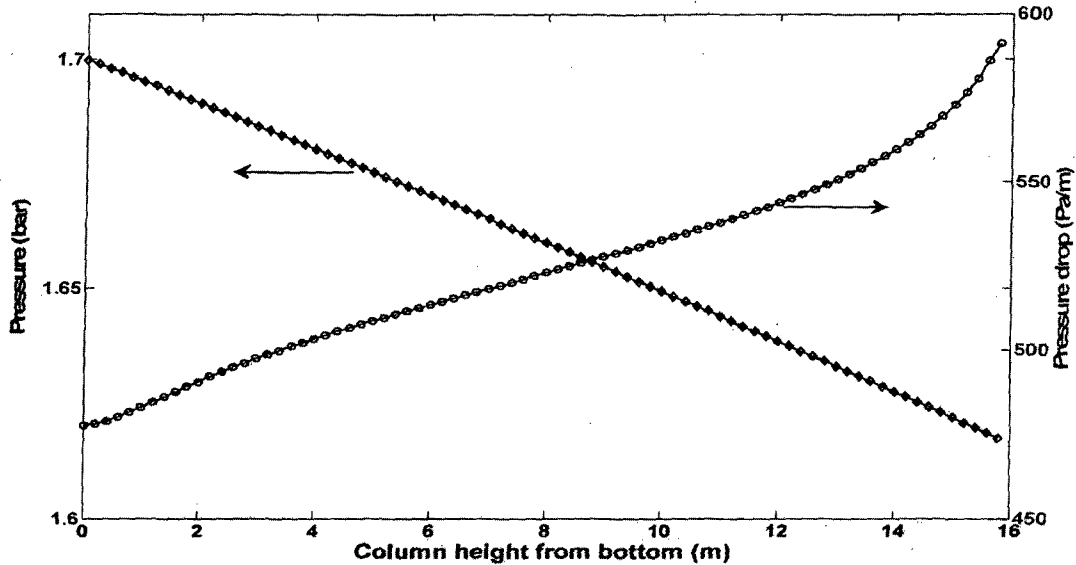


Figure 1-14: Profile of gas pressure and pressure drop over desorber column height, using berl saddle 50 mm packing and at design velocity of 70% flooding. The liquid inlet and outlet loadings are 0.45 and 0.20. The inlet liquid temperature is 103 °C and the steam temperature is 126 °C.

5

Figure 1-15 shows the temperature profiles of gas and liquid over the column. The gas enters the bottom of the column at 399.15 K (126 °C) and, while moving upward in the column towards the condenser, it passes its heat to the liquid stream and cools/condenses until leaving the column at temperature 380.22 K (107 °C). The inlet liquid enters the top of the column at 379.15 K (106 °C) and during movement towards the bottom, takes energy from gas for CO₂ regeneration. Its temperature is increased through the column and it reaches the bottom at the temperature of 391.6 K (118.5 °C).

15

- 47 -

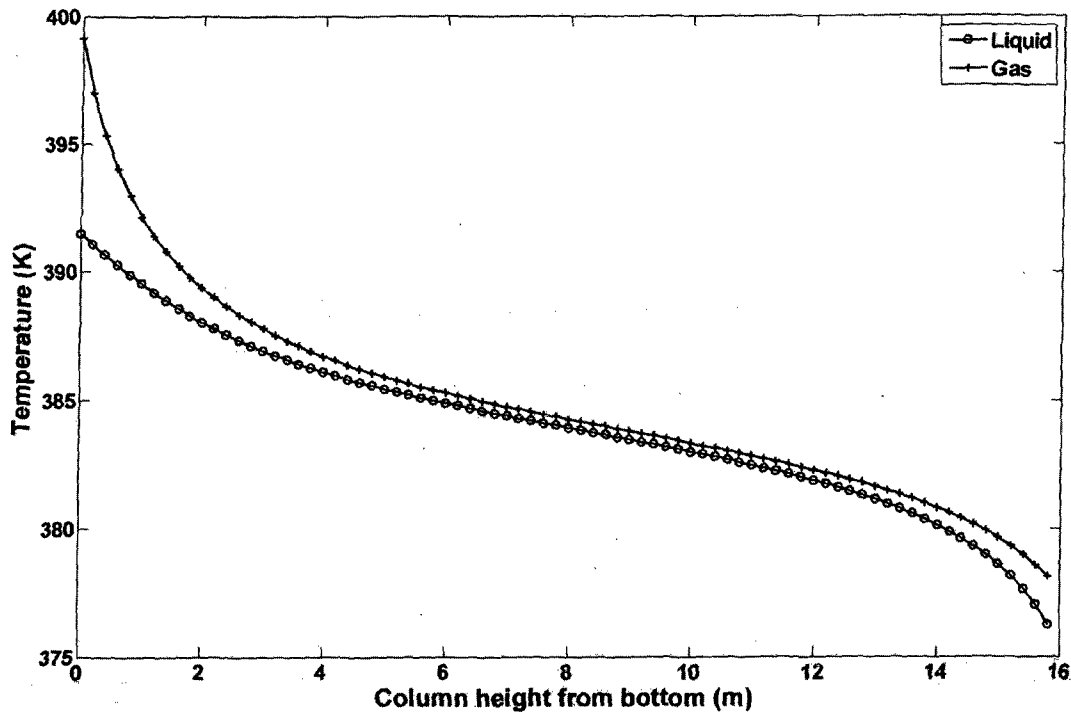


Figure 1-15: Gas and liquid temperature profile over column height using berl saddle 50 mm packing and at design velocity of 70% flooding for all examples; the liquid inlet and outlet loadings are fixed at 0.45 and 0.2, respectively. The inlet liquid temperature is 103 °C and the steam temperature is 126 °C.

5

Figure 1-16 shows the profiles of MEA and carbamate over the column. The rich liquid with low concentration of free MEA (0.5 kmol/m^3) and high concentration of bounded MEA such as carbamate (2.2 kmol/m^3) enters the top of the column and regenerates through downward movement in the column when the CO_2 bounds are broken, the concentration of bounded ions decreases, and free MEA increases. The lean liquid, with MEA of about 3 kmol/m^3 and carbamate of about 1 kmol/m^3 , leaves the column towards the absorber (after heat exchange with the rich liquid coming from the absorber).

10

- 48 -

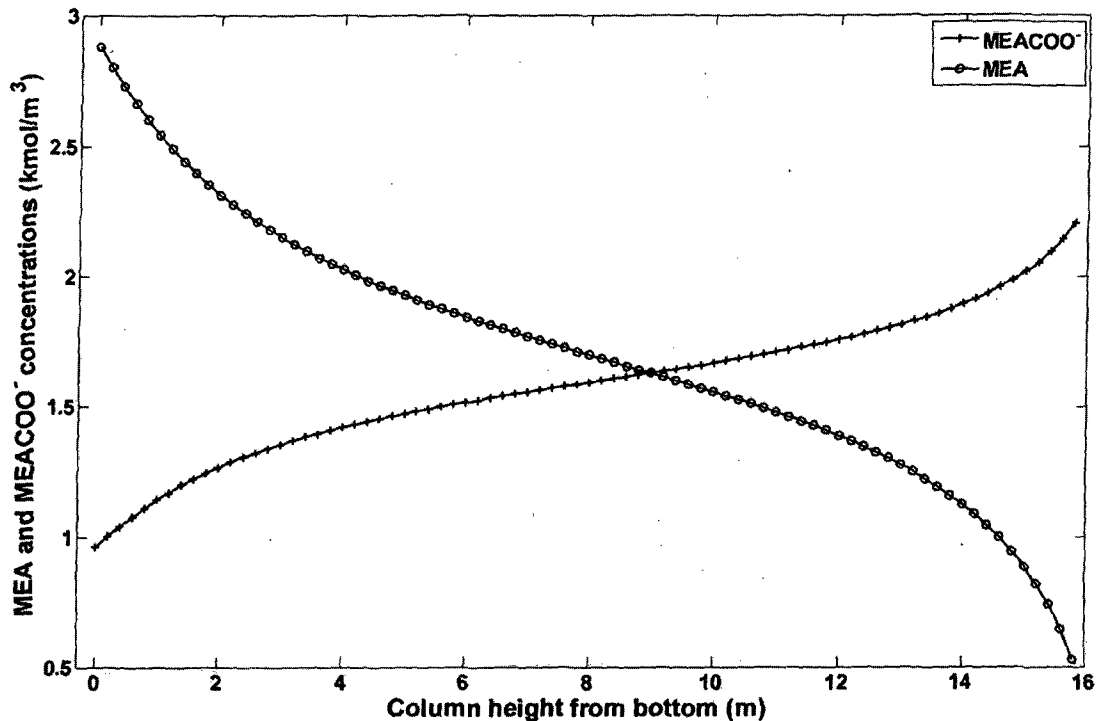


Figure 1-16: Profiles of liquid components over column height using berl saddle 50 mm packing and at design velocity of 70% flooding for all examples; the liquid inlet and outlet loadings are fixed at 0.45 and 0.2, respectively. The inlet liquid temperature is 103 °C and the steam temperature is 126 °C.

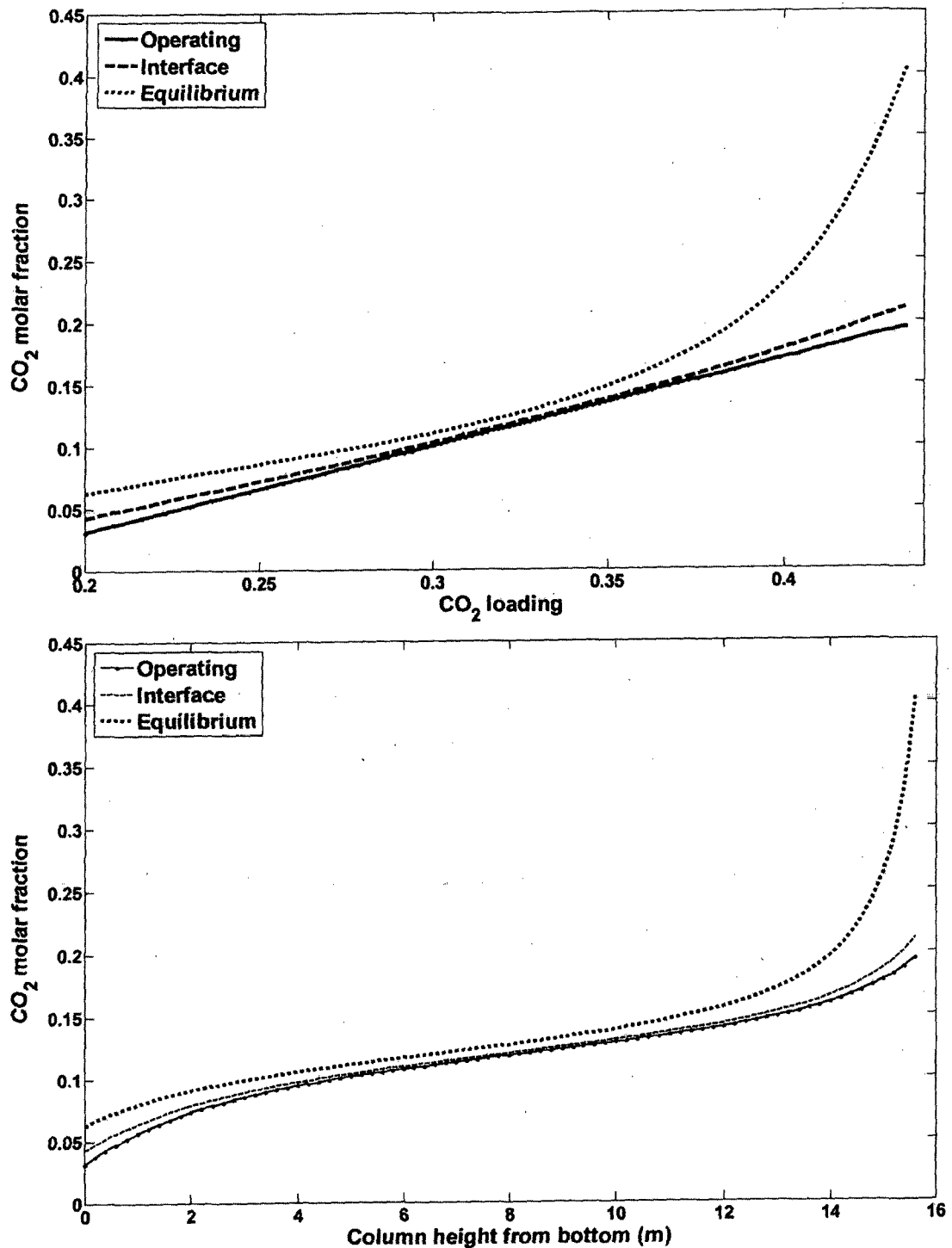
5

Figure 1-17 illustrates the three curves, i.e. operating, interface and equilibrium fraction of CO₂ over the column height and also with respect to loading of liquid. It is evident that the equilibrium and operating line are very close at loadings of about 0.30-0.35. From the figure it is clear why a higher CO₂ concentration in effluent gas is not possible under the condition that inlet liquid has a loading of 0.45 and an outlet loading of 0.2. From the figure, if outlet loadings higher than 0.2 are assumed then the operating line has the potential to move up from right-hand side, translating to outlet gas with a higher content of CO₂. In the parametric analysis section this issue is elaborated.

10

15

- 49 -



5 Figure 1-17: Profiles of operating, interface and equilibrium concentration of CO₂ over column height and loading using berl saddle 50 mm packing and at design velocity of 70% flooding for all examples; the liquid inlet and outlet loadings are fixed at 0.45 and 0.2, respectively. The inlet liquid temperature is 103 °C and the steam temperature is 126 °C.

- 50 -

Parametric analysis for desorber

The base-case scenario studied earlier resulted in a column with packing size of
5 10.53 m diameter and 15.8 m height. Under this scenario the L_{mol}/G_{mol} is about 5.80 and
the gas exiting from the desorber has CO₂ molar fraction of about 0.20. The overall
reboiler duty is 6.04 MJ/kg-CO₂. The reboiler duty is huge: a normal coal-fired power
plant produces high quantity of CO₂, in the range of a few hundred tonnes per hour, and
requires a huge amount of steam to satisfy this obligation. It is critical, therefore, to find
10 the optimum condition under which the reboiler duty is minimized.

The impact of packing type was investigated in the section on absorber column design, and
was also found to apply to the desorber.

Impact of gas velocity/flux

15 As with absorber, the impact of design velocity on column size and performance was
studied. Figure 1-18 illustrates the profiles of column diameter and pressure drop over the
design velocities in the range of 70% to 85% of flooding velocity. The detailed results for
each velocity scenario are given in Table 1-5. As expected, the increase of velocity
decreases the column diameter. However, the pressure drop is notably elevated with an
20 increase of design velocity. For instance, at the design velocity of 70%, the column
diameter is 10.53 m and 15.8 m, and the pressure drop is 8.27 kPa. When the velocity
reaches 85%, the diameter is 9.39 m, with a 10.8% decrease. Meanwhile the corresponding
pressure drop becomes 23.43 kPa, being 183.3% higher than that for the 70% scenario.
Velocity has affects the reboiler duty and the column height. Therefore, selection of the
25 optimal design velocity may be said to be a compromise between capital expenditure
CAPEX (size) and operating expenditure OPEX (pressure drop).

- 51 -

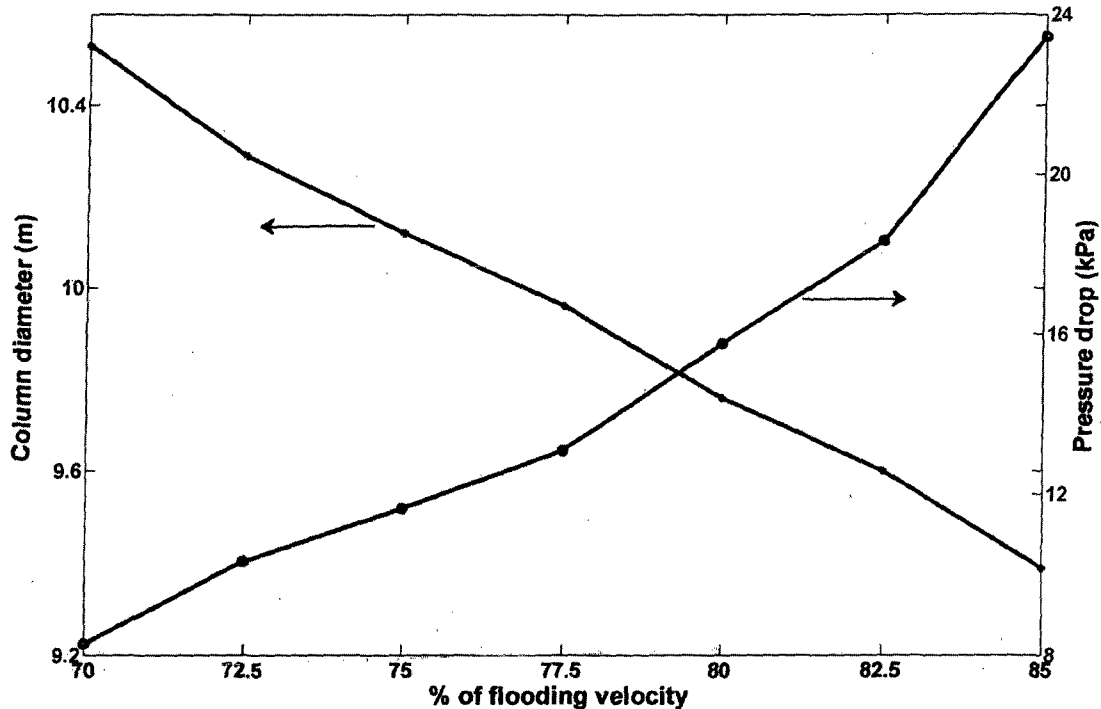


Figure 1-18: Impact of gas design velocity on column size and pressure drop, using berl saddle 50 mm packing for all examples; the liquid inlet and outlet loadings are fixed at 0.45 and 0.2, respectively.

5

Table 1-5: Impact of design velocity on desorber column design and performance.

%flood	T_{bot}^G (K)	T_{bot}^L (K)	T_{bot}^G (K)	T_{bot}^L (K)	$y_{\text{CO}_2}^{\text{gas}}$	$y_{\text{CO}_2}^{\text{liq}}$	$y_{\text{CO}_2}^{\text{out}}$	D (K)	Z (K)	P_{tot} (bar)	ΔP (kPa)	Q^i (lit/s)	L_{mol} (mol/s)	G_{mol} (mol/s)	$L_{\text{mol}}/G_{\text{mol}}$	Q (MJ/kg-CO ₂)
70.0	399.15	376.26	378.1134	391.5	0.45	0.2	0.20	10.53	15.80	1.617334	8.2666	1147.44	47703.20	8499.19	5.80	6.04
72.5	399.15	376.23	378.043	391.5	0.45	0.2	0.21	10.29	17.40	1.596769	10.3231	1147.44	47704.25	8350.76	5.91	5.93
75.0	399.15	375.80	377.8002	391.5	0.45	0.2	0.21	10.12	17.20	1.583499	11.6501	1147.44	47662.06	8350.76	5.91	5.77
77.5	399.15	376.03	377.9424	391.5	0.45	0.2	0.21	9.96	16.80	1.569293	13.0707	1147.44	47685.46	8350.76	5.91	5.86
80.0	399.15	375.88	377.8337	391.5	0.45	0.2	0.21	9.76	17.40	1.542459	15.7541	1147.44	47673.09	8254.64	5.97	5.74
82.5	399.15	376.15	377.9972	391.5	0.45	0.2	0.21	9.60	17.20	1.516767	18.3233	1147.44	47701.62	8207.40	6.01	5.82
85.0	399.15	375.87	377.8095	391.5	0.45	0.2	0.22	9.39	18.20	1.465738	23.4262	1147.44	47679.97	8023.75	6.15	5.60

Impact of gas/liquid ratio

- 10 The base-case scenario studied earlier resulted in $L_{\text{mol}}/G_{\text{mol}}$ of about 5.80. A scenario study was performed over various ranges of $L_{\text{mol}}/G_{\text{mol}}$. Figure 1-18 illustrates the results and shows how the changes of $L_{\text{mol}}/G_{\text{mol}}$ values impact on the reboiler duty, CO₂ fraction of the outlet gas and the column height. The other operating and design specifications of the scenarios are given in Table 1-6. As evident in Figure 1-18a, with the increase of the
- 15 $L_{\text{mol}}/G_{\text{mol}}$, or in other words with the decrease of gas flowrate, the reboiler duty expectedly

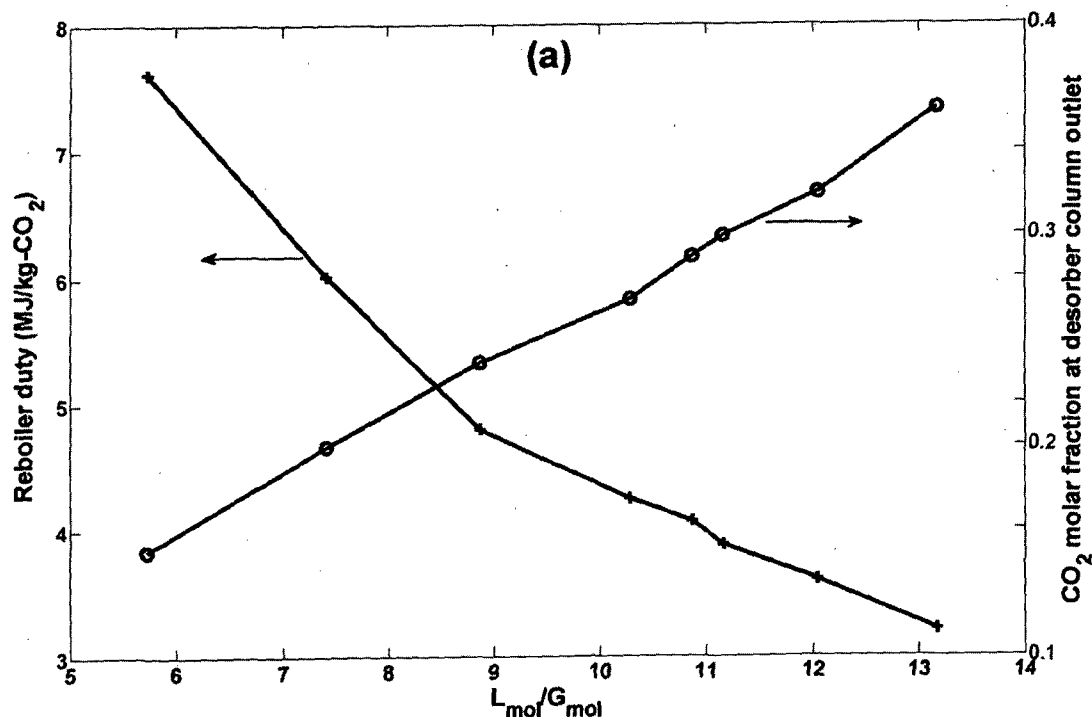
- 52 -

decreases. This results in an increase in the concentration of CO₂ in the outlet gas. For instance, for L_{mol}/G_{mol} of 5.73 the reboiler duty is 7.61 MJ/kg-CO₂ and the CO₂ fraction is 0.15, whereas at L_{mol}/G_{mol} of 13.17 the values become 3.21 MJ/kg-CO₂ and 0.36, respectively.

- 5 If just these three parameters were the basis for decision, it could be concluded that the higher the L_{mol}/G_{mol} (the lower the gas flowrate G_{mol}), the better. The gas flow rate could be designed as low as possible, with a lower bound such that the CO₂ fraction at the column top does not reach or exceed the equilibrium fraction of CO₂ for inlet rich liquid. However, from the column size graphing (Figure 4-26b), it is evident that although the reboiler duty
- 10 decreases with an increase in L_{mol}/G_{mol} , the column height meanwhile notably increases. The height for $L_{mol}/G_{mol} = 5.73$ is 2.20 m, but it skyrockets to 67.40 m at $L_{mol}/G_{mol} = 13.17$. The pressure drop also accordingly increases with height elevation.

The column diameter is a direct function of fluid flowrate. With decline of G_{mol} (increase of L_{mol}/G_{mol}), the column diameter also decreases. The diameter at $L_{mol}/G_{mol} = 5.73$ is

15 10.67 m whilst it is 8.29 m at $L_{mol}/G_{mol} = 13.17$ m.



- 53 -

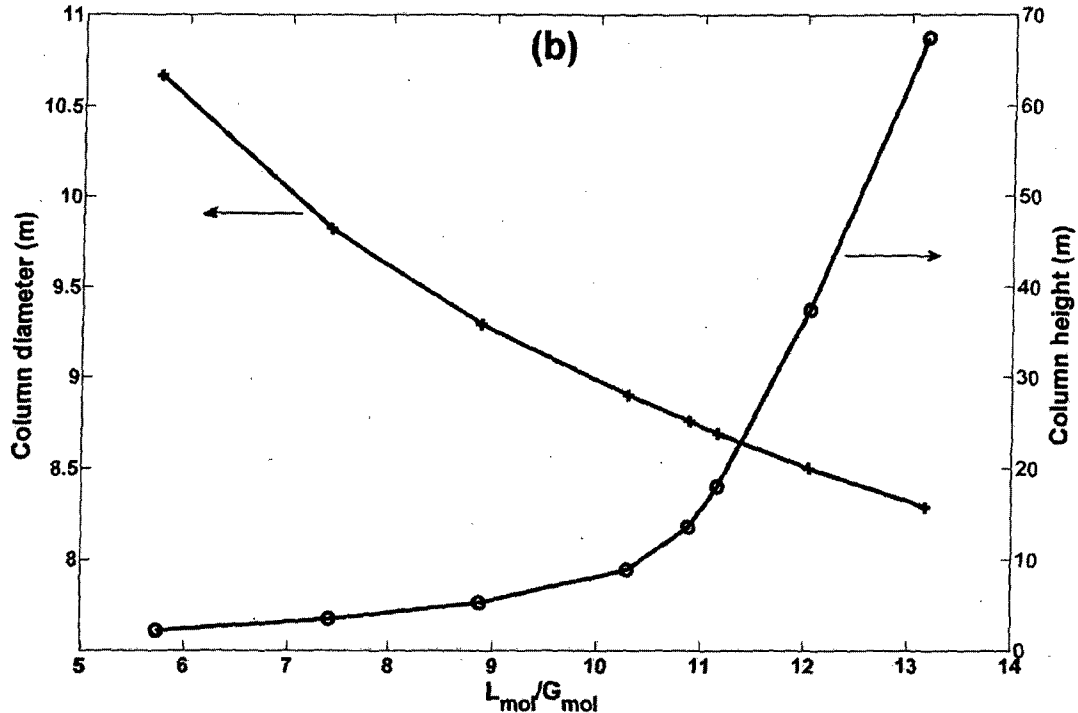


Figure 1-18: Impact of liquid-gas ratio (L_{mol}/G_{mol}) on (a) reboiler duty and CO_2 concentration, and (b) column size, using berl saddle 50 mm packing and at design velocity of 70% flooding for all examples; the liquid inlet and outlet loadings are fixed at 0.45 and 0.2, respectively. The inlet steam temperature is 126 °C.

Table 4-6: Impact of inlet gas (steam) flowrate on column design and performance.

G_{mol} (mol/s)	L_{mol} (mol/s)	L_{mol}/G_{mol}	T_{in}^s (K)	T_{in}^l (K)	T_{out}^s (K)	T_{out}^l (K)	$x_{in}^{CO_2}$	$x_{out}^{CO_2}$	τ_{CO_2}	D (K)	Z (K)	P_{in} (bar)	ΔP (kPa)	Q^s (W/s)	Q (MJ/kg- CO_2)
3745.22	48037.63	13.17	399.15	378.21	379.4663	391.5	0.45	0.25	0.36	8.29	67.40	1.379307	32.0693	1147.44	3.21
4095.08	48054.82	12.04	399.15	378.60	379.7798	391.5	0.45	0.25	0.32	8.50	37.40	1.527343	17.2657	1147.44	3.60
4420.03	48048.80	11.16	399.15	378.66	379.9098	391.5	0.45	0.25	0.30	8.69	18.00	1.618579	8.1421	1147.44	3.88
4533.85	48070.45	10.88	399.15	378.94	380.1494	391.5	0.45	0.25	0.29	8.76	13.60	1.639273	6.0727	1147.44	4.07
4794.13	48044.89	10.29	399.15	378.83	380.2579	391.5	0.45	0.25	0.27	8.90	8.80	1.660806	3.9194	1147.44	4.25
5559.77	47985.16	8.87	399.15	378.63	380.6433	391.5	0.45	0.25	0.24	9.29	5.20	1.676627	2.3373	1147.44	4.80
6658.68	48018.90	7.41	399.15	379.39	381.9072	391.5	0.45	0.25	0.20	9.82	3.40	1.684438	1.5562	1147.44	6.01
8604.87	47944.97	5.73	399.15	379.41	383.5008	391.5	0.45	0.25	0.15	10.67	2.20	1.68972	1.0280	1147.44	7.61

To further analyze the reality behind the competitive relation of height and L_{mol}/G_{mol} the operating, interface and equilibrium profiles of six different L_{mol}/G_{mol} scenarios are illustrated. At high L_{mol}/G_{mol} (or low G_{mol} values), the concentration of CO_2 in the gas phase increases and as a result the driving force between gas and liquid decreases. As evident in Figure 1-19a, in the middle part of the figure all three curves almost touch each other, indicating that the column is operating at its lowest efficiency. This results in a

- 54 -

notable increase in column height. With a decrease of L_{mol}/G_{mol} (or increase of G_{mol}), the concentration of CO_2 in the gas stream reduces and thus the driving force between gas and liquid increases. This is easily observed through Figure 1-19a-f.

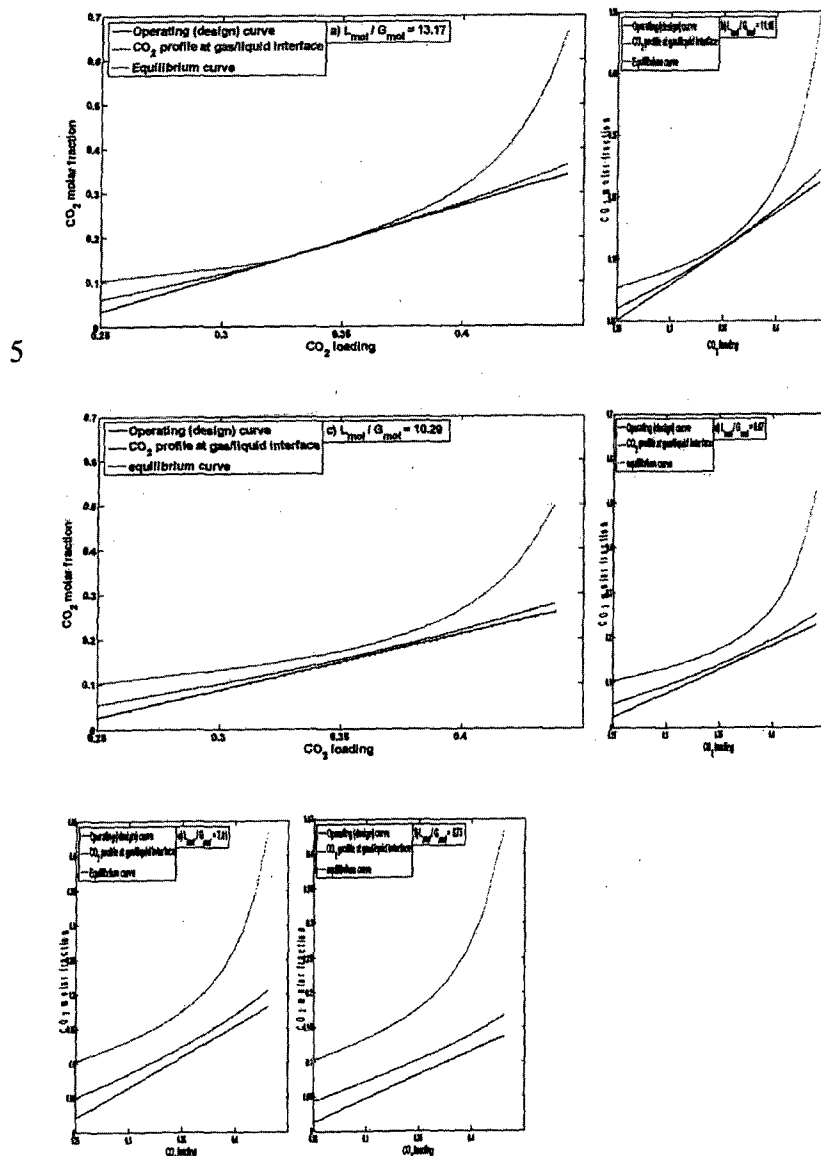


Figure 1-19: Impact of L_{mol}/G_{mol} on operating, interface and equilibrium curves, using berl saddle 50 mm packing and at design velocity of 70% flooding for all examples; the liquid inlet and outlet loadings are fixed at 0.45 and 0.25, respectively. The inlet steam temperature is 126 °C.

- 55 -

Therefore, a high G_{mol} or low L_{mol}/G_{mol} is required to enhance the mass transfer driving force and improve the performance of the column, with concurrent costs in higher reboiler duty and higher column diameter. The minimization of reboiler duty is linked with column size and the selection of correct values requires a techno-economic study that includes both

5 CAPEX and OPEX parameters.

Impact of gas/liquid temperature

The desorber performance is based on heat transfer from gas phase to liquid phase and breaking the reaction of the solvent with the target component and stripping it out of the

10 liquid. Therefore, selection of the right temperatures for inlet gas and liquids are important. Numerous scenarios were performed with inlet gas temperatures and liquids, which are presented in Table 1-7 and Table 1-8. The analyses were carried out in two different ways. In the first approach, the outlet CO₂ molar fraction was fixed to be almost the same for all scenarios (in the range of 0.295-0.305). Then the impact of temperatures on other

15 parameters was investigated.

Table 1-7 lists the results of this first approach. It is evident that the main impact of the fluid temperature is on column height. The temperature of the liquid has a more significant impact on column size than does the temperature of the gas. For instance, at inlet gas temperature of 397.15 K (124 °C), when the liquid enters the column at $T_{int}^L = 374.69$ K

20 (101.44 °C), the regeneration of the liquid (from a loading of 0.45 down to 0.25) requires a column 86.6 m in height, whereas if the inlet liquid temperature rises about two degrees to 376.56 K (103.41 °C) the column height will be 51.40 m. Further increase of the inlet liquid temperature to 379.52 K (106.37 °C) will require a column 17.4 m in height whereas at T_{int}^L of 381.36 (108.21 °C) a column only 6.2 m in height is required.

25 The impact of gas temperature on column height, despite being positive, is not as severe as that of liquid temperature. For instance, at an outlet liquid temperature of $T_{out}^L = 391.5$ K, when the inlet gas temperature is 394.15 K, a column 18.2 m in height is required, whereas with an increase of the gas temperature to 397.15 K, 400.15 K and 403.15 K, the column height decreases slightly to 17.2 m, 16.2 m and 15.2 m, respectively.

- 56 -

The very slight increase of column diameter with temperature rise (Table 4-7) can be attributed to the physical expansion of fluids at high temperatures.

5 **Table 1-7: Impact of inlet gas and liquid temperatures on column design and performance while the outlet CO₂ concentration is constant (at about 0.295-0.305).**

T_{in}^G (K)	T_{in}^L (K)	T_{out}^G (K)	T_{out}^L (K)	$\alpha_{in}^{CO_2} \rightarrow \alpha_{out}^{CO_2}$	$y_{CO_2, out}$	D (K)	Z (K)	P_{out} (bar)	ΔP (kPa)	Q^L (lit/s)	L_{mol} (mol/s)	G_{mol} (mol/s)	L_{mol}/G_{mol}	Q (MJ/kg- CO ₂)
387.50	376.08	374.99	0.25 → 0.45	0.29	8.62	88.20	1.270427	42.9573	1147.4424	48119.15	4302.14	11.48	3.83	
388.50	376.95	375.82	0.25 → 0.45	0.30	8.63	71.80	1.357544	34.2456	1147.4424	48095.77	4321.31	11.42	3.82	
389.50	377.95	376.85	0.25 → 0.45	0.29	8.64	54.00	1.446952	25.3048	1147.4424	48095.81	4349.24	11.35	3.86	
394.15	390.50	378.82	0.25 → 0.45	0.30	8.66	34.40	1.541923	15.8077	1147.4424	48070.23	4367.23	11.30	3.84	
391.50	379.56	378.23	0.25 → 0.45	0.30	8.66	18.20	1.617753	8.2247	1147.4424	48020.09	4381.71	11.26	3.75	
392.50	380.54	379.08	0.25 → 0.45	0.30	8.68	11.40	1.648706	5.1294	1147.4424	48002.05	4409.08	11.18	3.74	
387.50	375.92	374.69	0.25 → 0.45	0.30	8.62	86.60	1.281502	41.8498	1147.4424	48074.78	4294.11	11.50	3.73	
388.50	377.14	376.06	0.25 → 0.45	0.29	8.63	69.80	1.369368	33.0632	1147.4424	48105.86	4317.74	11.43	3.88	
389.50	377.80	376.56	0.25 → 0.45	0.30	8.64	51.40	1.461491	23.8509	1147.4424	48048.93	4331.97	11.39	3.76	
390.50	378.86	377.67	0.25 → 0.45	0.30	8.65	31.80	1.554932	14.5068	1147.4424	48055.95	4358.85	11.32	3.83	
397.15	391.50	379.84	0.25 → 0.45	0.30	8.67	17.20	1.622419	7.7581	1147.4424	48048.00	4393.46	11.23	3.85	
392.50	380.98	379.69	0.25 → 0.45	0.29	8.68	10.80	1.651603	4.8397	1147.4424	48049.62	4413.15	11.17	3.92	
393.50	381.68	380.06	0.25 → 0.45	0.31	8.70	8.00	1.664083	3.5917	1147.4424	47984.77	4437.44	11.11	3.77	
394.50	383.00	381.36	0.25 → 0.45	0.30	8.71	6.20	1.672154	2.7846	1147.4424	48009.63	4467.39	11.03	3.90	
387.50	376.17	375.04	0.25 → 0.45	0.30	8.62	84.80	1.293245	40.6755	1147.4424	48094.18	4289.97	11.51	3.82	
388.50	377.10	375.94	0.25 → 0.45	0.30	8.64	68.00	1.379434	32.0566	1147.4424	48083.14	4318.25	11.43	3.83	
389.50	378.16	377.05	0.25 → 0.45	0.29	8.65	49.40	1.471559	22.8441	1147.4424	48087.85	4339.89	11.37	3.90	
400.15	390.50	379.06	0.25 → 0.45	0.30	8.66	29.40	1.566709	13.3291	1147.4424	48066.33	4355.85	11.33	3.89	
391.50	380.01	378.79	0.25 → 0.45	0.30	8.68	16.20	1.627133	7.2867	1147.4424	48057.06	4398.93	11.21	3.91	
392.50	381.12	379.84	0.25 → 0.45	0.29	8.69	10.40	1.653461	4.6539	1147.4424	48054.98	4421.24	11.15	3.96	
395.50	384.06	382.15	0.25 → 0.45	0.30	8.73	5.00	1.677513	2.2487	1147.4424	47973.53	4492.71	10.97	3.87	
387.50	376.08	374.83	0.25 → 0.45	0.30	8.62	83.20	1.303008	39.6992	1147.4424	48062.70	4289.12	11.51	3.75	
388.50	377.20	376.04	0.25 → 0.45	0.30	8.64	66.20	1.389078	31.0922	1147.4424	48083.46	4322.95	11.42	3.86	
389.50	378.12	376.94	0.25 → 0.45	0.30	8.65	47.00	1.484176	21.5824	1147.4424	48062.09	4332.34	11.39	3.85	
403.15	390.50	379.02	0.25 → 0.45	0.30	8.66	27.40	1.576263	12.3737	1147.4424	48043.68	4357.60	11.32	3.85	
391.50	379.88	378.50	0.25 → 0.45	0.30	8.68	15.20	1.631928	6.8072	1147.4424	48015.29	4390.55	11.23	3.81	
392.50	380.97	379.52	0.25 → 0.45	0.30	8.69	10.00	1.655419	4.4581	1147.4424	48008.25	4407.90	11.19	3.84	
394.50	382.97	381.14	0.25 → 0.45	0.31	8.72	6.00	1.673096	2.6904	1147.4424	47967.34	4470.03	11.03	3.82	

In the second approach, the outlet CO₂ concentration was allowed to reach its highest possible value under the given conditions of each scenario. The main understanding from this approach, further to the previous scenario, is that at higher temperatures the CO₂ fraction of the equilibrium condition increases and thus the equilibrium curve moves upward in the CO₂ fraction vs. loading graph. This allows the CO₂ fraction of the gas phase to also reach higher values. With this fact the rate of steam as the “filler of the partial pressure gap” reduces, resulting in a decline in reboiler duty. For instance, at T_{int}^G of

- 57 -

397.15 K (124 °C) when the T_{int}^L is 375.48 K (102.33 °C), the L_{mol}/G_{mol} is 9.16, resulting in reboiler duty of 4.88 MJ/kg-CO₂, whereas with an increase of the inlet liquid temperature to T_{int}^L of 381.09, the L_{mol}/G_{mol} increases (G_{mol} decreases) to 12.74 and the duty decreases to 3.31 MJ/kg-CO₂. An increase of the inlet gas temperature has a similar but milder impact than that of the inlet liquid temperature.

Table 1-8: Impact of inlet gas and liquid temperature on column design and performance while the outlet CO₂ concentration is allowed to reach the maximum possible.

T_{int}^G (K)	T_{int}^L (K)	T_{int}^H (K)	T_{int}^L (K)	$\alpha_{int}^{gas} \rightarrow \alpha_{int}^{liq}$	$\nu_{CO_2, max}$	D (K)	Z (K)	P _{int} (bar)	ΔP (kPa)	$\dot{Q}^L$ (lit/s)	L_{mol} (mol/s)	G_{mol} (mol/s)	L_{mol}/G_{mol}	Q (MJ/kg-CO ₂)
394.15	387.50	376.50	375.28	0.25 $\rightarrow$ 0.45	0.23	9.26	16.00	1.628729	7.1271	1147.44	48083.28	5472.96	9.02	4.89
	388.50	377.05	375.68	0.25 $\rightarrow$ 0.45	0.25	9.11	16.80	1.625114	7.4886	1147.44	48040.69	5173.93	9.54	4.46
	389.50	378.13	376.90	0.25 $\rightarrow$ 0.45	0.26	8.98	16.40	1.626729	7.3271	1147.44	48070.43	4937.41	10.00	4.37
	390.50	378.84	377.55	0.25 $\rightarrow$ 0.45	0.28	8.82	16.80	1.624425	7.5575	1147.44	48048.64	4650.45	10.61	4.04
	391.50	379.80	378.58	0.25 $\rightarrow$ 0.45	0.29	8.72	17.20	1.622392	7.7608	1147.44	48055.90	4470.83	11.03	3.92
	392.50	380.41	379.05	0.25 $\rightarrow$ 0.45	0.32	8.59	18.20	1.61699	8.3010	1147.44	48014.92	4256.91	11.58	3.60
397.15	387.50	376.65	375.48	0.25 $\rightarrow$ 0.45	0.23	9.22	17.00	1.624469	7.5531	1147.44	48095.05	5390.57	9.16	4.88
	388.50	377.16	375.84	0.25 $\rightarrow$ 0.45	0.26	9.04	17.60	1.621487	7.8513	1147.44	48044.11	5061.10	9.75	4.41
	389.50	378.06	376.74	0.25 $\rightarrow$ 0.45	0.27	8.96	16.80	1.624718	7.5282	1147.44	48040.12	4909.26	10.05	4.27
	390.50	378.88	377.57	0.25 $\rightarrow$ 0.45	0.29	8.79	17.20	1.622644	7.7356	1147.44	48035.19	4609.66	10.70	4.00
	391.50	379.61	378.21	0.25 $\rightarrow$ 0.45	0.30	8.71	16.80	1.624274	7.5726	1147.44	48007.66	4453.73	11.07	3.78
	392.50	380.48	379.12	0.25 $\rightarrow$ 0.45	0.32	8.57	18.40	1.616532	8.3468	1147.44	48011.44	4211.89	11.71	3.58
400.15	393.50	381.36	380.00	0.25 $\rightarrow$ 0.45	0.34	8.47	18.00	1.617995	8.2005	1147.44	48009.03	4037.19	12.21	3.42
	394.50	382.37	381.09	0.25 $\rightarrow$ 0.45	0.35	8.37	17.20	1.621251	7.8749	1147.44	48019.20	3868.71	12.74	3.31
	387.50	376.46	375.13	0.25 $\rightarrow$ 0.45	0.25	9.19	17.60	1.621846	7.8154	1147.44	48048.58	5333.18	9.26	4.68
	388.50	377.21	375.81	0.25 $\rightarrow$ 0.45	0.26	9.07	16.80	1.625469	7.4531	1147.44	48029.20	5096.63	9.68	4.41
	389.50	378.29	377.07	0.25 $\rightarrow$ 0.45	0.27	8.93	17.80	1.620721	7.9279	1147.44	48066.28	4830.52	10.22	4.31
	390.50	379.02	377.74	0.25 $\rightarrow$ 0.45	0.29	8.80	17.60	1.621232	7.8768	1147.44	48041.88	4608.89	10.70	4.04
403.15	391.50	379.90	378.64	0.25 $\rightarrow$ 0.45	0.30	8.67	17.00	1.623484	7.6516	1147.44	48044.19	4375.53	11.27	3.84
	392.50	380.80	379.58	0.25 $\rightarrow$ 0.45	0.32	8.54	18.00	1.618421	8.1579	1147.44	48046.32	4157.83	11.86	3.66
	394.50	382.50	381.23	0.25 $\rightarrow$ 0.45	0.35	8.37	17.20	1.621125	7.8875	1147.44	48023.55	3866.14	12.75	3.34
	387.50	376.68	375.43	0.25 $\rightarrow$ 0.45	0.24	9.14	17.60	1.622191	7.7809	1147.44	48067.88	5232.36	9.44	4.70
	388.50	377.60	376.36	0.25 $\rightarrow$ 0.45	0.25	9.06	16.20	1.628374	7.1626	1147.44	48067.29	5077.28	9.72	4.57
	389.50	378.33	377.04	0.25 $\rightarrow$ 0.45	0.27	8.93	16.40	1.627192	7.2808	1147.44	48048.30	4832.13	10.21	4.29
403.15	390.50	379.31	378.12	0.25 $\rightarrow$ 0.45	0.28	8.80	16.80	1.625168	7.4832	1147.44	48069.46	4598.48	10.73	4.14
	391.50	380.13	378.93	0.25 $\rightarrow$ 0.45	0.30	8.67	17.40	1.622032	7.7968	1147.44	48059.51	4373.12	11.28	3.92
	392.50	380.76	379.44	0.25 $\rightarrow$ 0.45	0.32	8.56	17.60	1.62064	7.9360	1147.44	48021.57	4175.48	11.81	3.62
	394.50	382.51	381.24	0.25 $\rightarrow$ 0.45	0.36	8.31	18.60	1.614747	8.5253	1147.44	48018.87	3775.50	13.05	3.27

10

Impact of gas/liquid pressure

Pressure has a notable impact on column operation. Fundamentally, a pressurized process design is not attractive unless the feed streams are naturally under high pressure, e.g. crude

- 58 -

oil/gas streams. This is mainly because pressurization is energy-intensive further to increased construction costs. However, reactions are a function of temperature, and generally high temperature improves the rate of any physico-chemical interaction.

For a solvent based carbon capture processes, it is desired to operate the desorber at near-atmospheric pressures to prevent pressurization costs. Meanwhile, operation of the column at high temperatures will be advantageous due to expediting the regeneration pressure. The problem is, however, the fact that at atmospheric pressure, the liquid will vaporize at a few degrees higher (depending on solvent concentration) than 100 °C. Therefore, the only option to operate the process at higher temperatures is to increase the column pressure. It was found that the desire to high-temperature process is what justifies having high-pressure process. To illustrate this, a scenario-based study was carried out over column pressure in the range of 1.4-2.4 bar. Table 1-9 lists results of a few scenarios at various pressure and temperatures.

In one scenario, the inlet gas temperature (T_{int}^G) and outlet liquid temperature (T_{out}^L) are fixed at 390.15 K (117 °C) and 385.5 K (112.35 °C), respectively, and the column pressure is varied from 1.4 bar to 2.4 bar. As it was expected the column height decreases with increase of the pressure. The height is 18.0 m at $P = 1.4$ bar whereas it is 13.2 m at $P = 2.4$ bar. The reboiler duty, however, increases notably from 4.47 MJ/kg-CO₂ at $P = 1.4$ bar to 7.54 bar at $P = 2.4$ bar. The reason is that by increase of the column pressure, the partial pressure of target component decreases (and thus the equilibrium curve) declines resulting in the reduction of molar fraction of target component at the outlet of the column. To compensate this, the gas flowrate should be higher (lower L_{mol}/G_{mol}) which increases the reboiler duty notably.

With the increase of process pressure, the desorber can operate in higher temperatures without fear of vaporization. In another scenario, along with the pressure rise the fluids' temperatures are also elevated to values close to the saturation temperature of the given pressure. It is evident from the results that under this scenario both column height and reboiler duty decrease. For instance, at $P = 1.4$, when the inlet gas temperature is 390.15 K (117 °C), the column height is 18.0 m and duty is 4.47 MJ/kg-CO₂. With increase of

- 59 -

pressure to 2.0 bar ($T_{int}^G = 399.15 \text{ K}$), the height and duty are 16.4 m and 3.82 MJ/kg-CO₂. Further increase of the pressure to 2.4 bar ($T_{int}^G = 405.15 \text{ K}$) is attributed to height of 16.0 m and duty of 3.38 MJ/kg-CO₂. Though the results suggest operating the desorber column at relatively higher temperature and pressure, there are few concerns about column pressure selection:

- 1- It is expected that this pressurization has an upper limit, meaning that there could be a pressure above which the OPEX due to pressurization becomes greater than the benefit from efficient operation of the column at high temperatures. This has already been discussed by others.
- 10 2- Again it should be remembered that although high pressure seems to have a positive impact on column size and reboiler duty, it is limited by the solvent operation requirement to prevent solvent degeneration.
- 3- Moreover, a column with high reboiler pressure requires steam of higher quality, which is more costly for power plants.
- 15 Therefore, as with other parameters, the optimal pressure value of a column is a function of the overall techno-economics of the process.

Table 1-9: Impact of pressure on desorber column design and performance.

P (bar)	T_{int}^G (K)	T_{int}^L (K)	T_{int}^G (K)	T_{int}^L (K)	$\alpha_{int}^{G,G}$	$\alpha_{int}^{L,L}$	$y_{CO_2, \text{reboiler}}$	D (K)	Z (K)	P_{int} (bar)	ΔP (kPa)	Q^L (lit/s)	L_{int} (mol/s)	G_{int} (mol/s)	L_{int}/G_{int}	Q (MJ/kg-CO ₂)
1.4	388.15	383.50	372.45	371.18	0.45	0.25	0.23	9.62	19.40	1.31041	8.9590	1147.44	48099.72	5656.30	8.74	4.89
	390.15	385.50	374.30	373.06	0.45	0.25	0.25	9.33	18.00	1.316722	8.3278	1147.44	48099.79	5135.46	9.62	4.47
1.6	388.15	383.50	372.87	371.67	0.45	0.25	0.19	9.84	16.40	1.52637	7.3630	1147.44	48120.07	6466.44	7.64	5.77
	390.15	385.50	374.29	372.92	0.45	0.25	0.23	9.49	17.00	1.523659	7.6341	1147.44	48059.21	5766.78	8.57	4.95
	393.15	388.50	377.06	375.78	0.45	0.25	0.26	9.05	16.80	1.523853	7.6147	1147.44	48061.88	4944.08	9.98	4.30
	396.15	391.50	379.74	378.51	0.45	0.25	0.31	8.65	17.80	1.518592	8.1408	1147.44	48052.51	4218.54	11.69	3.67
1.8	388.15	383.50	372.76	371.35	0.45	0.25	0.18	9.99	16.00	1.729329	7.0671	1147.44	48061.31	7178.47	6.88	6.20
	390.15	385.50	374.67	373.39	0.45	0.25	0.20	9.64	16.20	1.728529	7.1471	1147.44	48080.23	6414.27	7.70	5.68
	393.15	388.50	377.25	375.98	0.45	0.25	0.24	9.17	16.60	1.726541	7.3459	1147.44	48061.09	5458.16	9.04	4.81
	396.15	391.50	379.88	378.61	0.45	0.25	0.28	8.77	16.20	1.727982	7.2018	1147.44	48040.17	4686.13	10.53	4.11
2.0	388.15	383.50	373.23	371.99	0.45	0.25	0.15	10.15	15.00	1.934709	6.5291	1147.44	48104.74	7909.11	6.25	7.15
	390.15	385.50	374.78	373.41	0.45	0.25	0.18	9.81	14.40	1.937632	6.2368	1147.44	48058.60	7105.47	6.95	6.26
	393.15	388.50	377.40	376.07	0.45	0.25	0.21	9.31	14.80	1.935938	6.4062	1147.44	48045.75	6020.36	8.20	5.33
	396.15	391.50	380.04	378.73	0.45	0.25	0.25	8.89	15.00	1.934754	6.5246	1147.44	48030.17	5147.28	9.58	4.55

- 60 -

	399.15	394.50	382.61	381.31	0.45	0.25	0.30	8.47	16.40	1.927759	7.2241	1147.44	48011.27	4363.25	11.30	3.82
	388.15	383.50	373.25	371.82	0.45	0.25	0.14	10.30	13.00	2.144006	5.5994	1147.44	48059.82	8680.02	5.69	7.67
	390.15	385.50	374.86	373.42	0.45	0.25	0.16	9.94	14.00	2.139891	6.0109	1147.44	48036.83	7750.47	6.37	6.79
	393.15	388.50	377.52	376.16	0.45	0.25	0.19	9.43	14.20	2.139433	6.0567	1147.44	48036.22	6553.72	7.53	5.81
2.2	396.15	391.50	380.20	378.89	0.45	0.25	0.23	8.99	14.40	2.138466	6.1534	1147.44	48027.76	5591.31	8.82	5.00
	399.15	394.50	382.62	381.22	0.45	0.25	0.28	8.58	15.80	2.131891	6.8109	1147.44	47981.35	4773.53	10.32	4.15
	402.15	397.50	385.32	383.97	0.45	0.25	0.33	8.24	16.60	2.127588	7.2412	1147.44	47971.03	4124.86	11.94	3.58
	388.15	383.50	373.34	371.83	0.45	0.25	0.13	10.46	12.80	2.345136	5.4864	1147.44	48037.57	9472.75	5.22	8.30
	390.15	385.50	375.17	373.80	0.45	0.25	0.15	10.06	13.20	2.34403	5.5970	1147.44	48059.44	8393.28	5.89	7.54
	393.15	388.50	377.81	376.52	0.45	0.25	0.17	9.54	13.60	2.342672	5.7328	1147.44	48052.25	7083.62	6.97	6.44
2.4	396.15	391.50	380.01	378.49	0.45	0.25	0.22	9.07	14.80	2.337547	6.2453	1147.44	47972.71	5998.89	8.22	5.18
	399.15	394.50	382.63	381.13	0.45	0.25	0.26	8.67	15.20	2.33534	6.4660	1147.44	47954.93	5158.82	9.55	4.44
	402.15	397.50	385.55	384.25	0.45	0.25	0.30	8.32	15.40	2.333896	6.6104	1147.44	47977.48	4457.13	11.05	3.95
	405.15	400.50	388.18	386.86	0.45	0.25	0.35	8.01	16.00	2.330735	6.9265	1147.44	47958.45	3867.68	12.72	3.38

Integrated analysis of absorber and desorber

- 5 In this section the impact of loadings on the overall performance of the two columns is studied. The given parameters are the same flue gas flowrate and compositions, with the objective of 90% carbon capture. As discussed earlier, the overall configuration has two absorber columns of equal size and one desorber column, as shown in Figure 4-28.
- 10 In the following integrated analysis, 24 scenario analyses are performed for absorbers with lean loading (α_{int}^{Abs}) in the range of 0.15-30 and rich loading (α_{int}^{Abs}) of 0.4-0.48. For each scenario, the design model requires a specific liquid flowrate, given the rich and lean liquid quality and quantity. The desorber analysis is then carried out for each scenario (24 scenarios) using the rich liquid information from the absorber and knowing the lean liquid
- 15 quality requirements. It is then possible to study the overall PCC process performance based on each scenario and to investigate the optimality conditions.

The results of the absorber analyses are given in Table 1-10 and the desorber results are presented in Table 1-11. It is evident that at any fixed lean liquid loading (α_{int}^{Abs}), with an

20 increase in loading of rich liquid (α_{int}^{Abs}), the column diameter decreases slightly while the

- 61 -

column height notably increases. For instance, for a lean loading of 0.2, when the rich loading is 0.4, the column diameter becomes 10.11 m and the height is 14 m. When the rich loading increases to 0.48, the column diameter decreases slightly to 9.67 m (4.4% decrease) and the height becomes 24.4 m (74.3% increase). This notable increase in column height is, however, to the benefit of application of a lower flowrate of liquid for higher rich loading (α_{out}^{Abs}). In the two scenarios discussed, when the rich loading is 0.4, the L_{mol}/G_{mol} becomes 5.12, whereas it is 3.66 (28.0% reduction) for a loading of 0.48. Now consider the impact of lean loading (α_{inlet}^{Abs}) for a fixed rich loading (α_{inlet}^{Abs}). Interestingly, it is evident that a lean loading does not have a notable impact on column height, unlike the rich loading. For instance, at a rich loading of 0.48, doubling the lean loading from 0.15 to 0.30 reduces the column height by only 2.2% from 18.6 m to 18.2m. The diameter also increases by only 7.5% from 9.51 m to 10.22 m. The increase in the liquid flowrate is, however, notable. A column with an inlet loading of 0.3 requires 83.5% more liquid flowrate compared than with a lean loading of 0.15 (L_{mol}/G_{mol} 5.69 vs. 3.10).

15

It may be concluded, therefore, that from a design point of view and in order to have a shorter column height (less pressure drop), the application of lower rich loading is advantageous. Lean loading has the opposite effect on column diameter and height, and the decision as to the correct lean loading requires techno-economic analysis.

20

In this analysis the rich liquid from the outlet of the two absorbers, for each scenario, is used for the analysis of desorber design and performance, with the objective of producing lean liquid of the same quality as that entering into the absorber. The results of 24 desorber scenarios identical to the 24 scenarios of Table 1-10 are given in Table 1-11. It is evident that for any given lean loading, the desorber column height increases noticeably with an increase of rich loading and the diameter decreases. For instance, for a lean loading of 0.2, at rich loading of 0.4 the diameter and height are 11.47m and 15.4 m, respectively, whereas at the rich loading of 0.48 the values become 9.72m (18.0% less) and 17.3m (12.3% more). The gas (steam) flow rate, however, decreases with an increase of rich

- 62 -

- loading, resulting in lower reboiler duty. For instance, for a lean loading of 0.2 when the rich loading is 0.4, the reboiler duty is 6.57 MJ/kg-CO₂, whereas at the rich loading of 0.48 it becomes 4.73 MJ/kg-CO₂ (28% reduction), translating to a noticeable reduction in operating costs. A similar trend is noticed when the rich loading is fixed and the lean loading is varied. Therefore, it can be concluded that for the desorber, an increase of both lean and rich loadings notably reduces the reboiler duty, and whereas low lean/rich combinations result in a thicker and shorter desorber column, higher lean/rich combinations require a relatively thin and taller column.
- 10 It is again pertinent to remember that the inlet liquid flowrates for each of the scenarios of Table 1-11 were different and obtained from output results of absorbers given in Table 1-10. Combining the results of both tables and the analysis discussed, it seems that a high rich loading results in a high absorber column height with lower reboiler duty. Therefore, finding the optimal lean and rich loading requires a techno-economic analysis containing
- 15 both CAPEX and OPEX in the objective.

Table 1-10: Impact of lean and rich amine loading on absorber column design and performance.

α_{out}	α_{in}	T_{in}^c (K)	T_{in}^l (K)	T_{out}^c (K)	T_{out}^l (K)	$y_{CO_2,out}$	$y_{CO_2,out}$	D (K)	Z (K)	P _{out} (bar)	ΔP (bar)	Q^L (lit/s)	L _{mol} (mol/s)	G _{mol} (mol/s)	L _{mol} /G _{mol}
0.4	0.15	323.15	336.15	324.97	324.80	0.02	0.01	9.83	14.6	1.0351	6.49	573.72	25334.85	6129.50	4.10
	0.20	323.15	336.15	327.48	327.35	0.02	0.01	10.11	14	1.0398	6.02	717.15	31612.94	6129.50	5.12
	0.23	323.15	336.15	328.95	328.85	0.02	0.01	10.34	13.4	1.0439	5.61	843.71	37151.50	6129.50	6.02
	0.25	323.15	336.15	329.86	329.77	0.02	0.01	10.54	13.2	1.0460	5.40	956.20	42075.95	6129.50	6.83
	0.28	323.15	336.15	331.22	331.15	0.02	0.01	10.93	12.6	1.0507	4.93	1195.25	52538.95	6129.50	8.54
	0.30	323.15	336.15	332.09	332.04	0.02	0.01	11.29	12.2	1.0541	4.59	1434.30	63002.35	6129.50	10.24
0.43	0.15	323.15	336.15	323.68	323.50	0.02	0.01	9.69	16.6	1.0237	7.63	512.25	22644.20	6129.50	3.66
	0.20	323.15	336.15	326.27	326.13	0.02	0.01	9.91	16	1.0282	7.18	623.61	27517.74	6129.50	4.45
	0.23	323.15	336.15	327.71	327.59	0.02	0.01	10.09	15.8	1.0306	6.94	717.15	31612.40	6129.50	5.12
	0.25	323.15	336.15	328.65	328.55	0.02	0.01	10.24	15.6	1.0327	6.73	796.84	35100.11	6129.50	5.69
	0.28	323.15	336.15	330.00	329.93	0.02	0.01	10.53	15.4	1.0358	6.42	956.20	42075.98	6129.50	6.83
	0.30	323.15	336.15	330.89	330.83	0.02	0.01	10.77	15.2	1.0384	6.16	1103.31	48514.82	6129.50	7.88
0.45	0.15	323.15	336.15	322.88	322.68	0.02	0.01	9.61	18.6	1.0120	8.80	478.10	21148.75	6129.50	3.41
	0.20	323.15	336.15	325.42	325.27	0.02	0.01	9.81	18.4	1.0147	8.53	573.72	25334.42	6129.50	4.10
	0.23	323.15	336.15	326.90	326.78	0.02	0.01	9.96	18.2	1.0171	8.29	651.96	28758.55	6129.50	4.66
	0.25	323.15	336.15	327.83	327.73	0.02	0.01	10.09	18.2	1.0184	8.16	717.15	31612.49	6129.50	5.12
	0.28	323.15	336.15	329.20	329.12	0.02	0.01	10.32	18.2	1.0207	7.93	843.71	37152.14	6129.50	6.02
	0.30	323.15	336.15	330.10	330.04	0.02	0.01	10.52	18.2	1.0225	7.75	956.20	42075.96	6129.50	6.83
0.48	0.15	323.15	336.15	321.58	321.38	0.02	0.01	9.51	24.4	0.9777	12.23	434.64	19246.64	6129.50	3.10
	0.20	323.15	336.15	324.20	324.04	0.02	0.01	9.67	24.4	0.9796	12.04	512.25	22643.70	6129.50	3.66
	0.23	323.15	336.15	325.67	325.53	0.02	0.01	9.79	24.6	0.9803	11.97	573.72	25334.55	6129.50	4.10
	0.25	323.15	336.15	326.66	326.54	0.02	0.01	9.89	24.6	0.9816	11.84	623.61	27517.69	6129.50	4.45
	0.28	323.15	336.15	328.05	327.96	0.02	0.01	10.07	25	0.9823	11.77	717.15	31612.16	6129.50	5.12
	0.30	323.15	336.15	328.95	328.87	0.02	0.01	10.22	25.4	0.9826	11.74	796.84	35099.96	6129.50	5.69

Table 1-11: Impact of lean and rich amine loading on desorber column design and performance.

α_{out}	α_{in}	T_{in}^G (K)	T_{in}^L (K)	T_{out}^G (K)	T_{out}^L (K)	$x_{CO_2, out}$	D (K)	Z (K)	P_{out} (bar)	ΔP (bar)	Q^L (lit/s)	L_{mol} (mol/s)	G_{mol} (mol/s)	L_{mol}/G_{mol}	Q (MJ/ kg-CO ₂)
0.4	0.15	397.15	391.15	375.58	374.60	0.12	12.09	14.7	1.6233	7.67	1147.44	49377.53	12423.63	3.97	8.81
	0.2	397.15	391.15	378.16	377.59	0.16	11.47	15.4	1.6254	7.46	1434.30	61672.06	9459.46	6.52	6.57
	0.23	397.15	391.15	379.89	379.50	0.19	11.28	15.5	1.6287	7.13	1687.42	72515.66	8076.67	8.98	5.48
	0.25	397.15	391.15	381.09	380.79	0.21	11.27	14.6	1.6354	6.46	1912.40	82153.39	7322.34	11.22	4.87
	0.28	397.15	391.15	383.03	382.87	0.25	11.40	14.7	1.6393	6.07	2390.50	102634.14	6288.51	16.32	4.05
	0.3	397.15	391.15	384.38	384.28	0.28	11.70	12.6	1.6512	4.88	2868.60	123113.66	5772.48	21.33	3.63
0.43	0.15	397.15	391.15	374.22	372.64	0.13	11.42	15.4	1.6179	8.21	1024.50	44111.45	11155.45	3.95	7.88
	0.2	397.15	391.15	376.58	375.43	0.18	10.73	15.7	1.6219	7.81	1247.22	53659.75	8371.63	6.41	5.72
	0.23	397.15	391.15	378.26	377.38	0.22	10.42	17.3	1.6175	8.25	1434.30	61673.78	6969.01	8.85	4.66
	0.25	397.15	391.15	379.40	378.66	0.25	10.31	17.5	1.6192	8.08	1593.68	68502.60	6202.61	11.04	4.05
	0.28	397.15	391.15	381.29	380.76	0.30	10.25	18	1.6215	7.85	1912.40	82156.61	5197.32	15.81	3.28
	0.3	397.15	391.15	382.52	382.10	0.35	10.34	17.7	1.6264	7.36	2206.62	94758.27	4637.88	20.43	2.82
0.45	0.15	397.15	391.15	373.48	371.41	0.14	11.04	15.6	1.6158	8.42	956.20	41185.48	10454.38	3.94	7.39
	0.2	397.15	391.15	375.74	374.13	0.20	10.28	16.7	1.6154	8.46	1147.44	49385.85	7715.81	6.40	5.26
	0.23	397.15	391.15	377.23	375.81	0.24	9.96	17.3	1.6159	8.41	1303.92	56089.99	6418.60	8.74	4.21
	0.25	397.15	391.15	378.58	377.43	0.27	9.81	17.2	1.6189	8.11	1434.30	61674.18	5685.76	10.85	3.73
	0.28	397.15	391.15	380.24	379.26	0.34	9.67	18.8	1.6157	8.43	1687.42	72518.04	4683.73	15.48	2.90
	0.3	397.15	391.15	381.60	380.83	0.39	9.68	18.9	1.6191	8.09	1912.40	82158.24	4139.19	19.85	2.50
0.48	0.15	397.15	391.15	371.87	369.58	0.15	10.50	16.6	1.6086	9.14	869.28	37462.57	9498.79	3.94	6.70
	0.2	397.15	391.15	373.94	372.23	0.22	9.72	17.3	1.6101	8.99	1024.50	44116.95	6959.95	6.34	4.73
	0.23	397.15	391.15	376.07	374.36	0.25	9.87	17.4	1.6078	9.22	1147.44	49260.76	6882.54	7.16	4.04
	0.25	397.15	391.15	376.87	375.43	0.29	9.66	17.6	1.6121	8.79	1247.22	53544.42	5983.14	8.95	3.42
	0.28	397.15	391.15	378.24	377.34	0.39	8.97	18.8	1.6122	8.78	1434.30	61674.01	4122.01	14.96	2.54
	0.3	397.15	391.15	379.51	378.78	0.45	8.90	20.5	1.6079	9.21	1593.68	68500.96	3590.53	19.08	2.13

- 5 Solvent-based carbon capture technology is considered by industries to be the best available technology for implementation in large-scale carbon capture and storage projects, including power plants. This technology has the drawback of high energy intensity. Many technological parameters have been identified with good potential for improving the efficiency of this technology. The complexity of solvent-based PCC technology stems
- 10 from the physico-chemical operation of absorber and desorber columns, addressed as “reactive separation”.

Example 2

- 65 -

A 300 MWe coal-fired power plant in Australia burning pulverized black coal. The plant uses black coal with 25% ash, 8% moisture, and dry-ash-free (DAF) composition of 83.3% carbon, 5.4% hydrogen, 1.9% nitrogen, 0.6% sulfur, and 8.8% oxygen. The power plant emits 12260 mol/s (353.3 kg/s) of flue gas when operating at full capacity. The flue gas composition is 13.0 vol% (19.8 wt%) CO₂, 70.37% N₂, 13.52% H₂O, 3.11% O₂ and ppm levels of SO_x and NO_x. The SO_x and NO_x are removed from the flue gas prior to the emission entering into the PCC process. As the temperature of flue gas is usually above 100 °C, and as the optimal operation of the absorber is far below 100 °C (40-60 °C for MEA), the flue gas needs to be cooled to the desired temperature, prior to entering the absorber.

The aim is to design a solvent based post carbon capture plant with base-case capacity or the ability to capture 90% of the CO₂ in the flue gas. The maximum allowable column diameter and packing height are 12 m and 25 meters respectively. The captured CO₂ is aimed to be compressed to 100 bar and sent for sequestration.

It is desired to assess two different scenarios. In Scenario 1, the objective is to design the plant with minimum possible total capital expenditure (CAPEX). In scenario 2, the objective is to design the plant to minimize the levelized cost of captured CO₂ per unit weight.

The overall schematic of the synthesis/design algorithm is given in Figure 12. The steps are as follows:

Firstly the following parameters are provided with values which are constant and do not change. (item A in Figure 12):

- What percentage of CO₂ they aim to capture with base-case design?
- Their flue gas flowrate and detailed composition (CO₂, H₂O, N₂, O₂, NO_x and SO_x)
- The solvent to capture the CO₂ and its composition (our base case solvent is the most well-known amine called monoethanolamine or MEA).

- 66 -

Once the program receives the constant parameters the next step begins with the determination of the size and configuration of the absorber (item B in Figure 12). If is not clear how many absorber columns are required to process the given inlet flowrate to satisfy the given CO₂ capture requirements, there will be a calculation loop to find firstly the
5 number of columns and secondly the diameter of the columns. The model of each absorber column (item C) is given in simplified form in Figure 12 and in detailed form in Figure 14.

The next stage after determining the size and configuration of the absorber column(s) is to determining the size and configuration of the desorber column(s).

10

A similar methodology to the absorber case will be followed for the desorber column(s) (item D in Figure 12). The model of the desorber (item E) is however more complex than that of the absorber as it is complicated by the presence of a condenser and a reboiler. The model in simplified form could still be explained by Figure 13, but the detailed model is
15 given by Figure 15.

The next stage (item F) will be to determine the size and configuration of the auxiliaries in the plant, i.e. the heat exchanger between absorber and desorber, pumps, compressors, knock-out drums and inter-coolers.

20

Up till now the design methodology has achieved the goal of determining the size and configuration of a solvent based carbon capture plant which both satisfies the operation and design constraints and requirements. The next stage is to determin one or more objective functions. An objective function may be related to a company's policies and
25 planning. For example it could be to minimise levelized costs, minimise CAPEX, maximise the internal rate of return (IRR), maximise net present value (NPV).

Results

30 Table 1 shows the results of the determination of the size and/or configuration of a solvent based carbon capture process which corresponds for two different objective functions. The

- 67 -

method is capable of finding the optimal values of both technical (design and operational) and economic variables. As a benchmark, Scenario 0 has been provided in Table 1 for designing with some typical operational parameters from open literature. As evident from Table 1, the given "typically good" operational parameters have resulted in a process with
5 CAPEX of \$ 366.99 million and levelized cost of 76.53\$/tonne-CO₂. Scenarios 1 and 2 which design the plant with concurrent consideration of both design and operational parameters as optimization variables will be discussed below.

The first objective function is to minimise CAPEX (Scenario 1). In scenario 1, the method
10 has determined the size and configuration of a solvent based carbon capture plant with two absorber columns (Diameter: 10.54 m, Packing height: 13.2 m) and one desorber column (Diameter: 11.27 m, Packing height: 14.6 m). The reboiler duty of the desorber is 4.87 GJ/tonne-CO₂. With the given design and operational values in Table 1, the minimum CAPEX is found to be \$338.27 million. The levelized cost of CO₂ capture and
15 compression, under this scenario, is found to be 73.56\$/tonne-CO₂ being 2.97\$/tonne-CO₂ less than Scenario 0.

In Scenario 2, the objective function is to minimise the levelised cost of CO₂ capture and compression which is combination of both CAPEX and OPEX. The method has
20 determined the size and configuration of the solvent based carbon capture plant to be similar to scenario 1, i.e. two absorber columns and one desorber column. However, the optimal values of techno-economic variables are different. Absorber columns were found to have a diameter of 10.52 m and a packing height of 18.2 m. The diameter of desorber column is 9.68 with packing height being 18.98. Therefore, the absorber and desorber
25 columns are larger than those for Scenario 1. This, however, results in lower reboiler duty of 2.5 GJ/tonne-CO₂. The CAPEX, under this scenario, is around \$13.86 million higher than previous scenario (\$352.13 million vs \$338.27 million). However, the minimum levelized cost is found to be 67.44 \$/tonne-CO₂ which is 6.12\$/tonne-CO₂ lower than scenario 1.

30

- 68 -

This example clearly demonstrates the interactivity between design and operational parameters and shows how the plant design can be affected with different objectives of power generating companies. It also highlights that the optimal design can vary for different geographic locations with the change of their local economical input parameters.

- 5 The method is therefore capable of determining the size and configuration of a solvent based carbon capture plant according to various objectives.

Table 1: The optimization results

Objective	Scenario 0: Design with typical operation data (non- optimized)	Optimized Scenario 1: Minimum CAPEX	Optimized Scenario 2: Minimum levelized capture cost
Technical variables (optimal values)			
Number of absorber column(s)	2	2	2
Number of Desorber column(s)	1	1	1
Absorber column(s) diameter (m)	9.67	10.54	10.52
Packing height of absorber column(s) (m)	24.4	13.2	18.2
Desorber column(s) diameter (m)	9.72	11.27	9.68
Packing height of desorber column(s) (m)	17.3	14.6	18.9
Packing type	berl saddle 50 mm	berl saddle 50 mm	berl saddle 50 mm
Absorber column(s) pressure drop (kPa)	12.04	5.4	7.75
Rich solvent loading	0.48	0.40	0.45
Lean solvent loading	0.2	0.25	0.30
Absorber L/G (mol/mol)	3.66	6.83	6.83
Reboiler duty (GJ/tonne-CO ₂)	4.73	4.87	2.5
Economic variables (optimal values)			
FOM (\$million/y)	27.17	25.27	25.62
VOM (\$million/y)	42.23	43.80	30.87
Total OPEX (\$million/y)	69.41	69.07	56.49
Total capital cost, TCI (\$million)	366.99	<u>338.27</u> (objective)	352.13
Levelized capture cost (\$/tonne-	76.53	73.56	<u>67.44</u>

- 69 -

CO2)			(objective)
------	--	--	-------------

Many modifications will be apparent to those skilled in the art without departing from the
5 scope of the present invention.

Notation

Symbols

	a	actual interfacial areas of packing per volume of column, m^2/m^3
	a_e	effective interfacial areas of packing per volume of column, m^2/m^3
5	c_j	concentration of component j , mol/m^3
	c_{ji}	concentration of component j at interface, mol/m^3
	c_p^L	specific heat capacity of liquid, $\text{kJ}/(\text{kg.K})$
	$c_{p,j}$	specific heat capacity of gas component j , $\text{kJ}/(\text{kg.K})$
	D_j^L	diffusivity of component j through liquid
10	D_j^G	diffusivity of component j through gas
	d_p	packing diameter, m
	F_p	packing factor, $1/\text{m}$
	F_{mass}^G	mass flux of gas, $\text{kg}/\text{m}^2.\text{s}$
	$F_{mass,D}^G$	design flux of gas, $\text{kg}/\text{m}^2.\text{s}$
15	$F_{mass,fl}^G$	mass flux of gas at flooding condition, $\text{kg}/\text{m}^2.\text{s}$
	F_{mass}^L	mass flux of liquid, $\text{kg}/\text{m}^2.\text{s}$
	$F_{mol,1}$	molar flux of non-reacting/non-diffusing component, $\text{mol}/\text{m}^2.\text{s}$
	G	gas flow rate; molar (G_{mol}) or mass (G_{mass}), mol/s (kg/s)
	h_G	heat transfer coefficient
20	Ha	Hatta number
	He_j	Henry constant for component j , $\text{Pa.m}^3/\text{mol}$
	$k_j^{Th.}$	thermal conductivity of gas component j , $\text{W}/\text{m.K}$
	k_j^G	mass transfer coefficients of component j through gas
	k_j^L	mass transfer coefficients of component j through liquid
25	K_j^G	overall mass transfer coefficients of component j

- 71 -

	K_j^L	overall mass transfer coefficients of component j
	L	liquid flow rate; molar (L_{mol}) or mass (L_{mass}), mol/s (kg/s)
	m_j	vapor-liquid equilibrium constant for component j
	MW_G	molecular weight of gas
5	MW_L	molecular weight of liquid
	N_j	mass transfer flux of component j , mol/m ² .s
	p_j	partial pressure of component j , bar
	p_{ji}	partial pressure of component j at interface, bar
	P	gas pressure, bar
10	Pr	Prandtl number
	R	ideal gas constant
	R_{jp}	the rate of generation of component j by the reactions p , kmol/m ² .s
	Re_G	gas Reynolds number
	St	Stanton number
15	T	temperature, K (or °C)
	T_0	273.15K °C
	T_s^G	temperature of gas at stage s , K (or °C)
	T_s^L	temperature of liquid at stage s , K (or °C)
	T^{sat}	saturation temperature of liquid, K (or °C)
20	T_j^{sat}	saturation temperature of component j , K (or °C)
	u^L	superficial velocity of the liquid, m/s
	u_D^G	design velocity of gas, m/s
	u_{fl}^G	flooding velocity of gas, m/s
	u_{MOC}^G	velocity of gas at maximum operational capacity, m/s
25	x_j	mole fraction of component j in liquid
	y_j	mole fraction of component j in gas
	Y_j	mole fraction of component j in gas (excluding non-reacting component)
	Z	packing height of column

30 Greek letters

- 72 -

	ε	void fraction (porosity) of packed bed
	σ_j	surface tension of component j , N/m
	σ_L	surface tension of liquid, N/m
	σ_c	surface tension of packing, N/m
5	α	CO ₂ loading of liquid
	μ_G	gas viscosity, Pa.s
	μ_j	viscosity of component j , Pa.s
	μ_L	liquid viscosity, Pa.s
	ρ_G	gas density, kg/m ³
10	ρ_j	density of gas components j , kg/m ³
	ρ_L	liquid density
	ΔH_p	heat of reaction p , kJ/mol
	ΔH_j^v	heat of vaporization of component j , kJ/mol

15 **Subscripts**

	0	initial condition (entrance)
	D	design
	G	gas
	i	interface
20	int	inlet
	L	liquid
	out	out
	p	reaction number
	Reb	reboiler
25	s	stage

Superscripts

	Abs	absorber
	Deg	degeneration
30	Des	desorber
	G	gas

- 73 -

L	liquid
sat	saturation

Appendix

Table 1: Physico-chemical properties of gas, liquid and the components involved in MEA-CO₂ interaction.

Property	Formula	Dimension	Reference and remark
Solubility	$p_{CO_2, l} = H_{CO_2-Sol} \cdot C_{CO_2, l}$ $H_{CO_2-Sol} = \frac{H_{CO_2-H_2O}}{H_{N_2O-H_2O}} H_{N_2O-Sol}$ where $H_{CO_2-H_2O} \{Pa \cdot m^3 \cdot mol^{-1}\} = 2.82 \times 10^5 \exp\left(\frac{-2044}{T}\right)$ $H_{N_2O-H_2O} \{Pa \cdot m^3 \cdot mol^{-1}\} = 8.55 \times 10^5 \exp\left(\frac{-2284}{T}\right)$ $H_{N_2O-Sol} \{Pa \cdot m^3 \cdot mol^{-1}\} = \exp\left(\phi_2 \phi_3 \beta_{23} + \sum_{j=2}^3 \phi_j \ln(H_{N_2O-j})\right)$ where 1: N₂O, 2: MEA, 3: H₂O $\phi_j = \frac{x_j V_j}{\sum_{j=2}^3 x_j V_j}, V_j: \text{molar volume of pure solvent } j$ $\beta_{23} = 4.793 - 7.446 \times 10^{-3} T - 2.201 \phi_3$ $H_{N_2O-MEA} \{Pa \cdot m^3 \cdot mol^{-1}\} = 1.207 \times 10^5 \exp\left(\frac{-1136.5}{T}\right)$	Pa.m ³ .mol ⁻¹	$H_{CO_2-H_2O}$ and $H_{N_2O-H_2O}$ from (Versteeg and Vanswaaij, 1988, Versteeg et al., 1996) H_{N_2O-Sol} from (Wang et al., 1992) and (Tsai et al., 2000)
Enhancement factor	$E = \frac{H_a \frac{E_{\infty}-E}{E_{\infty}-1}}{\tanh\left(H_a \sqrt{\frac{E_{\infty}-E}{E_{\infty}-1}}\right)}, E_{\infty} = \left(\frac{D_A^*}{D_B^*}\right)^{1-m} + \left(\frac{D_A^*}{D_B^*}\right)^{-m} \frac{C_B}{C_{Al}}, Ha = \frac{\sqrt{K D_A^* C_B}}{k_L}$ Yermanian simplification for penetration (using Danckwerts PSFO E_{ps}): $E = \frac{E_{ps}^2}{2(E_{\infty}-1)} \left[\left(1 + 4 \frac{E_{\infty}(E_{\infty}-1)}{E_{ps}^2}\right)^{0.5} - 1 \right]$	dimensionless	(Vanswaaij and Versteeg, 1992); (Yermanian et al., 1970); (Decourse, 1974); (Last and Stichlmair, 2002); (Wellek et al., 1978)

- 75 -

	Wellek modified Yermanian method: $\frac{1}{(E-1)^{1.25}} = \frac{1}{(E_{\infty}-1)^{1.25}} + \left(\frac{Ha}{\tanh(Ha)} - 1\right)^{1.25}$ $E = \frac{1}{\left(\frac{1}{Ha^2} + \frac{1}{E_{\infty}^2}\right)^{\frac{2}{2.25}}}$ Last and Stichlmair method: $E = \frac{-Ha^2}{2(E_{\infty}-1)} + \sqrt{\frac{Ha^4}{4(E_{\infty}-1)^2} + \frac{E_{\infty}Ha^2}{(E_{\infty}-1)} + 1}$ Another method by Decourse: Binary diffusivity (component A through component B) using Fuller method: $D_{AB} = \frac{P M_{AB}^{0.5} \left[(\sum A DV)^{\frac{1}{3}} + (\sum B DV)^{\frac{1}{3}} \right]}{N_A - Y_A \sum_j \frac{N_j}{D_{Aj}}}$ The effective diffusivity of component A in gas mixture from Bird et al.: $D_{Ae} = \frac{N_A - Y_A \sum_j \frac{N_j}{D_{Aj}}}{\sum_j Y_j \frac{N_A - Y_A \sum_j \frac{N_j}{D_{Aj}}}{D_{Aj}}}$			(Fuller et al., 1966) (Poling et al., 2001a) (Bird et al., 2002)
Diffusivity in gas		$m^2.s^{-1}$		
Diffusivity of CO ₂ in the solution	1) N₂O analogy $D_{CO_2-sol.} = \frac{D_{CO_2-H_2O}}{D_{N_2O-H_2O}} D_{N_2O-sol.}$ $D_{CO_2-H_2O} \{m^2.s^{-1}\} = 2.35 \times 10^{-6} \exp\left(\frac{-2119}{T}\right)$ $D_{N_2O-H_2O} \{m^2.s^{-1}\} = 5.07 \times 10^{-2} \exp\left(\frac{-2371}{T}\right)$ $D_{N_2O-sol.} \{m^2.s^{-1}\} = 10^{-4} \times (5.07 \times 10^{-2} + 8.65 \times 10^{-3} c_{MEA} + 2.78 \times 10^{-3} c_{MEA}^2) \exp\left(\frac{-2371}{T}\right)$ 2) Stokes-Einstein relationship $D_{CO_2-sol.} = D_{CO_2-H_2O} \left(\frac{\mu_{H_2O}}{\mu_{sol.}}\right)^n$ $n = 0.545$ (Freguia and Rochelle, 2003).	Diffusivities: $m^2.s^{-1}$ MEA molarity (c_1 (MEA) $mol.m^{-3}$):	$D_{CO_2-H_2O}$ and $D_{N_2O-H_2O}$ from (Versteeg and Vanswaaij, 1988, Versteeg et al., 1996, Versteeg et al., 1987) $D_{N_2O-sol.}$ from (Ko et al., 2001)	

Diffusivity of MEA in the solution	1) Simple prediction ($c_{MEA_0} : 0.04 - 5.02 \text{ M, } T: 298 - 333 \text{ K}$): $D_{MEA-sol.} = \exp\left(-13.275 - \frac{2198.3}{T} - 7.8142 \times 10^{-3} c_{MEA_0}\right)$ 2) Stokes-Einstein relationship $D_1(MEA - sol.) = D_1(MEA - H_2O) (\mu_1(H_2O) / \mu_1(MEA))^{\eta} \quad \eta = 0.6 \text{ @ ADDIN EN.CITE <}$ $D_{MEA-H_2O} = \frac{1.4 \times 10^{-5}}{\mu_{H_2O}^{0.599}}$	Diffusivities: $m^2.s^{-1}$ MEA molarity ($c_1([MEA], 0)$): $kmol.m^{-3}$	(Snijder et al., 1993) (Versteeg et al., 1996)
Density and molar volume of fluid	$\rho_L = \rho_{sol} = \frac{x_{MEA} MW_{MEA} + x_{H_2O} MW_{H_2O} + x_{CO_2} MW_{CO_2}}{V_{sol}}$ $V_{sol} [m^3.kmol^{-1}] = x_{MEA} V_{MEA} + x_{H_2O} V_{H_2O} + x_{CO_2} V_{CO_2} + x_{MEA} x_{H_2O} V^*$ $V_j = \frac{MW_j}{\rho_j}, \quad V^* = -1.8218 \times 10^{-3}, \quad V_{CO_2} = 0.04747 \times 10^{-3}$ $\rho_{H_2O} = 1000 \times (0.863559 + 1.21494 \times 10^{-3} T - 2.57080 \times 10^{-6} T^2)$ $\rho_{MEA} = 1000 \times (1.19093 - 4.29990 \times 10^{-4} T - 5.66040 \times 10^{-7} T^2)$	density: kg/m^3 molar volume (V): $m^3/kmol$	(Amundsen et al., 2009, Weiland et al., 1998) densities of pure components from (Hsu and Li, 1997a)
Density of gas components	$\rho_j = \frac{M_j P}{RT}, \quad \rho_G = \sum y_j \rho_j$	$kg.m^{-3}$	
Viscosity	$\mu_L = \mu_{sol} = \mu_{H_2O} \times \exp\left(\frac{(21.186.wt\% + 2373)[\alpha(0.01015.wt\% + 0.0093T - 2.2589) + 1]}{T^2}\right)$ $\mu_{MEA} = 10^{-6} \times \rho_{MEA} \times \exp\left(-3.51312 + \frac{8.93173 \times 10^2}{T - 1.59612 \times 10^2}\right)$ $\mu_{H_2O} = 10^{-6} \times \rho_{H_2O} \times \exp\left(-3.28285 + \frac{4.56029 \times 10^2}{T - 1.54576 \times 10^2}\right)$	$Pa.s (10^3 \text{ cP})$	Viscosity of solution (wt% <40) from (Weiland et al., 1998, Amundsen et al., 2009) For pure components (Hsu and Li, 1997b) (Hilliard, 2008)
Specific heat capacity of	$c_p^L = 4.008 + 2.474 \times 10^{-3} T + 5.907 \times 10^{-6} T^2 - 5.234 \times 10^{-3} m_{MEA} - 0.5575 \alpha + 0.767$	$J/(kg.K)$	

- 77 -

solution		w_{MEA} : molality (mol/kg water) α : CO ₂ loading (mol CO ₂ /mol MEA) c_p^L : kJ/(kmol.K)	
Specific heat capacity of gas components	$c_{p,CO_2} = 19.1 + 7.342 \times 10^{-3} T - 5.602 \times 10^{-5} T^2 + 1.715 \times 10^{-8} T^3$ $c_{p,MEA} = 9.311 + 0.30097 - 1.818 \times 10^{-4} T^2 + 4.656 \times 10^{-8} T^3$ $c_{p,H_2O} = 32.2 + 1.924 \times 10^{-3} T + 1.055 \times 10^{-5} T^2 - 3.596 \times 10^{-9} T^3$ $c_{p,N_2} = 31.15 - 1.357 \times 10^{-2} T + 2.68 \times 10^{-5} T^2 - 1.168 \times 10^{-8} T^3$	(Poling et al., 2001b)	
Surface tension of solvent (MEA+H ₂ O)	$\frac{\sigma_{MEA} - \sigma_{H_2O}}{\sigma_{H_2O}} = \left(1 + \frac{a \cdot x_{H_2O}}{1 - b \cdot x_{H_2O}} \right)$ Where, $\sigma_{H_2O} = 10^{-3} \times [76.0852 - 0.1609 \times (T - 273.15)]$ $\sigma_{MEA} = 10^{-3} \times [53.0800 - 0.1648 \times (T - 273.15)]$ $a = 0.63036 - 0.00013 \times (T - 273.15)$ $b = 0.94700 - 0.00002 \times (T - 273.15)$	N/m	(Vazquez et al., 1997) (Aboudheir, 2002)
Thermal conductivity of gas components	$k_{H_2O}^{T_A} = 10^{-3} \times (-3.9 + 0.075 \times T)$ $k_{CO_2}^{T_A} = 10^{-3} \times (-6.337 + 0.076 \times T)$ $k_{N_2}^{T_A} = 10^{-3} \times (4.84 + 0.069 \times T)$ $k_G^{T_A} = \sum y_j k_j^{T_A}$	W/m.K	(Greer, 2008)

- 78 -

The Claims:

1. A method of determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, the method including the following steps:
- 5 a) providing information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,
- 10 b) determining the size and/or configuration of the solvent based carbon capture process,
- wherein, step b includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.
- 15 2. A method according to claim 1, wherein the method includes implementing a solvent based carbon capture plant according to the determined size and configuration of the solvent based carbon capture process and/or plant.
- 20 3. A method according to claim 1 wherein the one or more design parameters of the solvent based carbon capture process and/or plant are selected from the following: the number of absorber columns, the number of desorber columns, the diameter of the absorber column(s), the diameter of the desorber column(s), the height of the absorber column(s), the height of the desorber column(s), the type of packing of the absorber column(s), and the type of packing of the desorber column(s).
- 25 4. A method according to any one of the preceding claims wherein the one or more operational parameters of the solvent based carbon capture process and/or plant are selected from the following: the ratio of gas to liquid flow rates entering the absorber and/or the desorber, the ratio of the source of gas including carbon dioxide to solvent, the carbon dioxide capture rate, the temperature of the gas including
- 30

- 79 -

carbon dioxide, the temperature of the inlet gas to the desorber, the solvent temperature, the lean and rich loadings of the solvent, the operating pressure of the absorber column(s), the operating pressure of the desorber column(s), the duty of the boiler associated with the desorber column(s).

5

5. A method according to any one of the preceding claims wherein step b includes the following steps:

- 10
- i. providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters;
 - ii. determining one or more objective functions associated with the mathematical model;
 - 15 iii. determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.

20 6. A method according to claim 5 wherein the mathematical model of the solvent based carbon capture process and/or plant includes the following:

- 25
- calculation of the number of absorber columns;
 - calculation of the size of the absorber column(s);
 - calculation of the number of desorber columns required;
 - calculation of the size of the desorber column(s); and
 - optionally calculation of the size and/or configuration of the auxiliaries required.

30 7. A method according to claim 5 or 6 wherein the one or more objective functions associated with the mathematical model includes the minimisation of the capital installation cost and/or the operating cost of the solvent based carbon capture process and/or plant.

- 80 -

- 5 8. A method according to claims 5 or 6 wherein the objective function associated with the mathematical model includes one or more of the following: economic objective functions, techno-economic objective functions, and/or technical objective functions.
- 10 9. A method according to claim 8 wherein the one or more objective functions is selected from: minimising the capital cost of the solvent based carbon capture process, minimising the operating cost of the solvent based carbon capture process, the net present value and/or combinations thereof.
10. A method according to any one of the preceding claims wherein the source of the gas is derived from a fossil fuel combustion process.
- 15 11. A solvent based carbon capture process and/or plant that has been sized and/or configured by the method according to any one of the preceding claims.
- 20 12. A method of determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, wherein the method includes:
- 25 a. a processing system obtaining information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,
- b. the processing system determining the size and/or configuration of the solvent based carbon capture process,
- wherein, step b includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.
- 30

- 81 -

13. The method according to claim 12, wherein the method includes the processing system:
- i. providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters;
 - ii. determining one or more objective functions associated with the mathematical model; and
 - iii. determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which corresponds to the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.
14. The method according to claim 13, wherein the method includes the processing system iteratively manipulating the one or more design parameters and the one or more operational parameters within the mathematical model to determine the value of the one or more design parameters and the one or more operational parameters which corresponds to the one or more objective functions.
15. A processing system for determining the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, wherein the processing system is configured to:
- a. obtain information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,
 - b. determine the size and/or configuration of the solvent based carbon capture process;
- wherein the determination by the processing system includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

- 82 -

16. The processing system according to claim 15, wherein the method includes the processing system:
- 5 i. providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters;
 - ii. determining one or more objective functions associated with the mathematical; and
 - 10 iii. determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.
- 15 17. The processing system according to claim 16, wherein the processing system is configured to iteratively manipulate the one or more design parameters and the one or more operational parameters within the mathematical model to determine the value of the one or more design parameters and the one or more operational parameters which corresponds to the one or more objective functions.
- 20 18. A computer readable medium for configuring a processing system to determine the size and/or configuration of a solvent based carbon capture process and/or plant for a source of gas including carbon dioxide wherein the configuration includes one or more absorber columns and one or more desorber columns, wherein the computer
- 25 readable medium includes executable instructions for configuring the processing system to:
- a. obtain information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and,
 - 30 b. determine the size and/or configuration of the solvent based carbon capture process;

- 83 -

wherein the determination by the processing system includes determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

- 5 19. The computer readable medium according to claim 18, wherein the executable instructions configure the processing system to:
- i. provide a mathematical model of the solvent based carbon capture process dependent upon the one or more design parameters and the one or more operational parameters;
 - 10 ii. determine one or more objective functions associated with the mathematical model; and
 - 15 iii. determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond with the one or more objective functions to thereby determine the size and/or configuration of the solvent based carbon capture process and/or plant.
- 20 20. The computer readable medium according to claim 19, wherein the executable instructions configure the processing system to iteratively manipulate the one or more design parameters and the one or more operational parameters within the mathematical model until the one or more objective functions are optimised.
- 25 21. An absorber associated with a solvent based carbon capture process and/or plant wherein the absorber has been sized by providing information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.
- 30 22. An absorber according to claim 21 wherein determining the value of one or more design parameters and one or more operational parameters of the solvent based

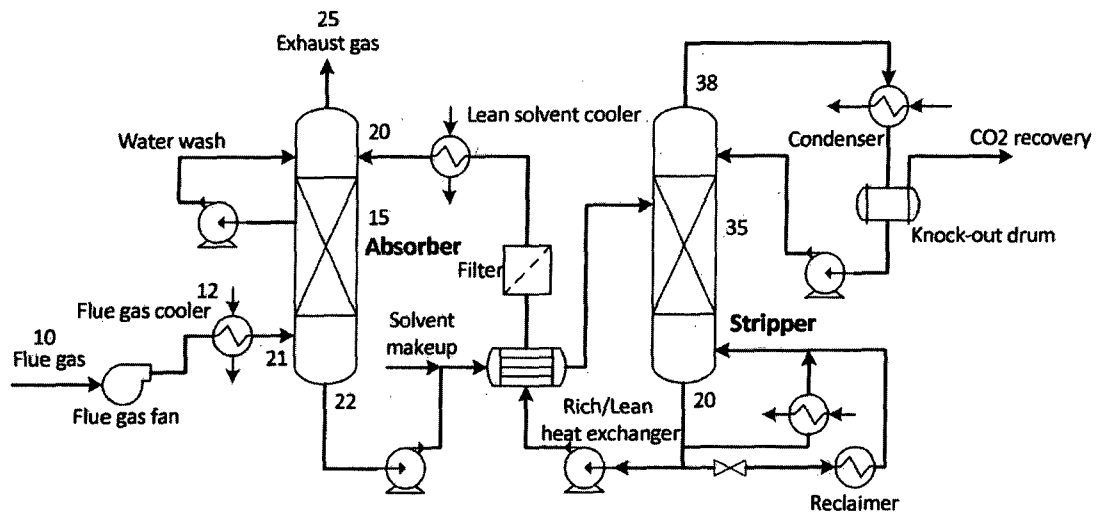
- 84 -

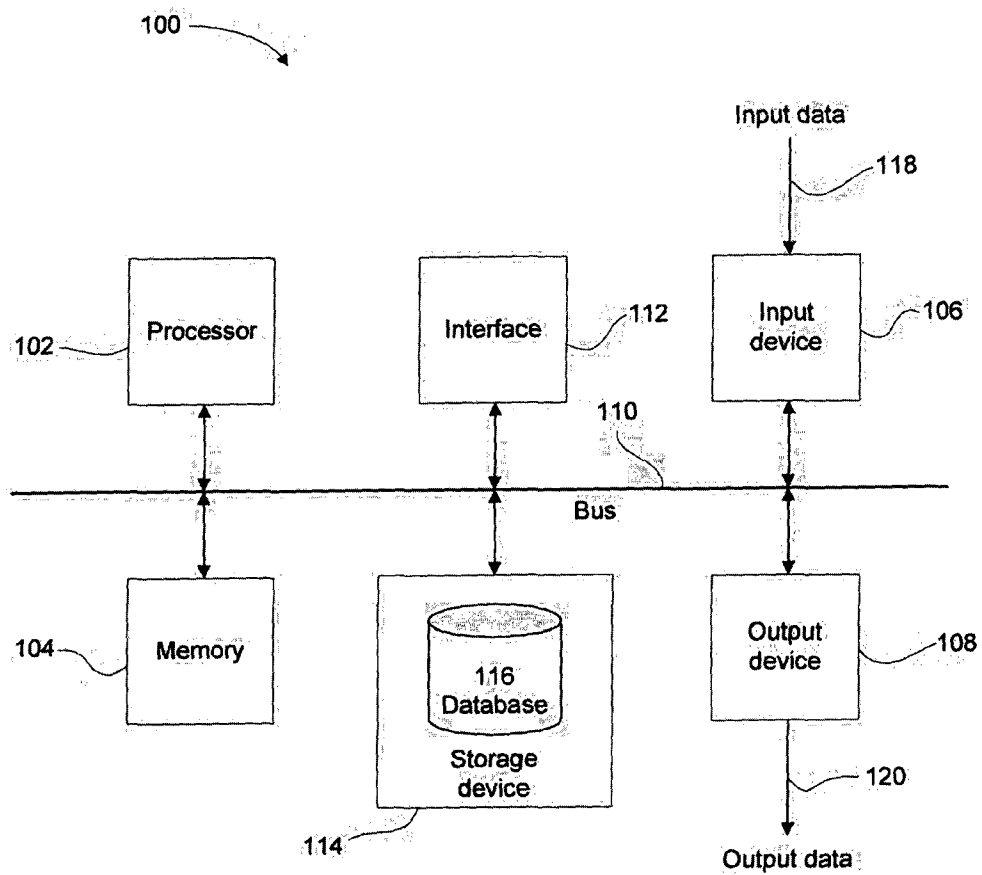
5 carbon capture process and/or plant includes providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters, determining one or more objective functions associated with the mathematical model; and determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size of the absorber.

10 23. A desorber associated with a solvent based carbon capture process and/or plant wherein the desorber has been sized by providing information including: percentage of carbon dioxide capture, composition of the source of gas, flow rate of the source of gas, and solvent type; and determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant.

15 24. A desorber according to claim 23 wherein determining the value of one or more design parameters and one or more operational parameters of the solvent based carbon capture process and/or plant includes providing a mathematical model of the solvent based carbon capture process and/or plant dependent upon the one or more design parameters and the one or more operational parameters, determining one or more objective functions associated with the mathematical model; and determining the value of the one or more design parameters and the one or more operational parameters within the mathematical model which correspond to the one or more objective functions to thereby determine the size of the absorber.

20 25. A solvent based carbon capture plant including an absorber according to claim 21 or claim 22 and/or a desorber according to claim 23 or claim 24.

**Figure 1**

**Figure 2**

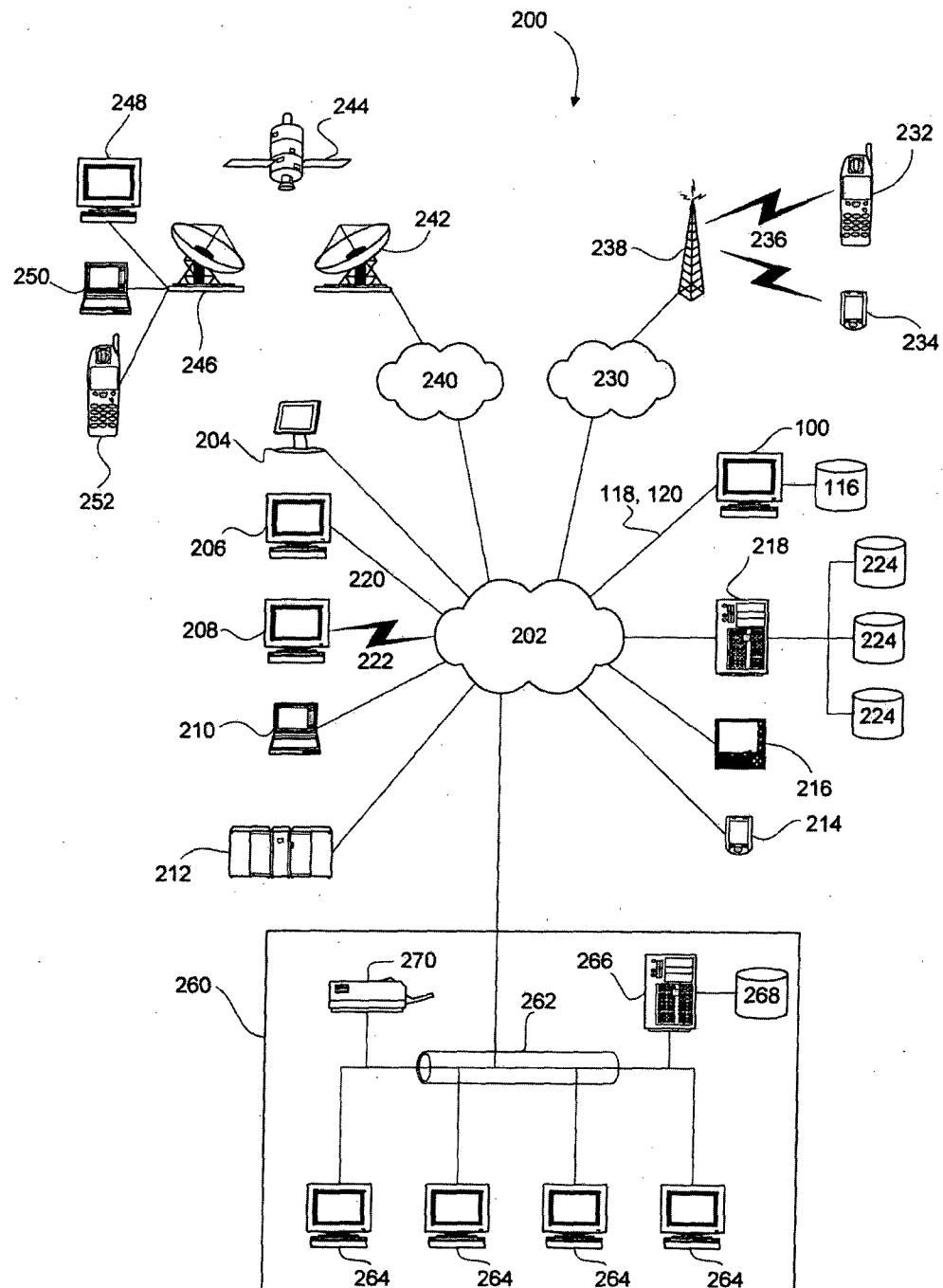
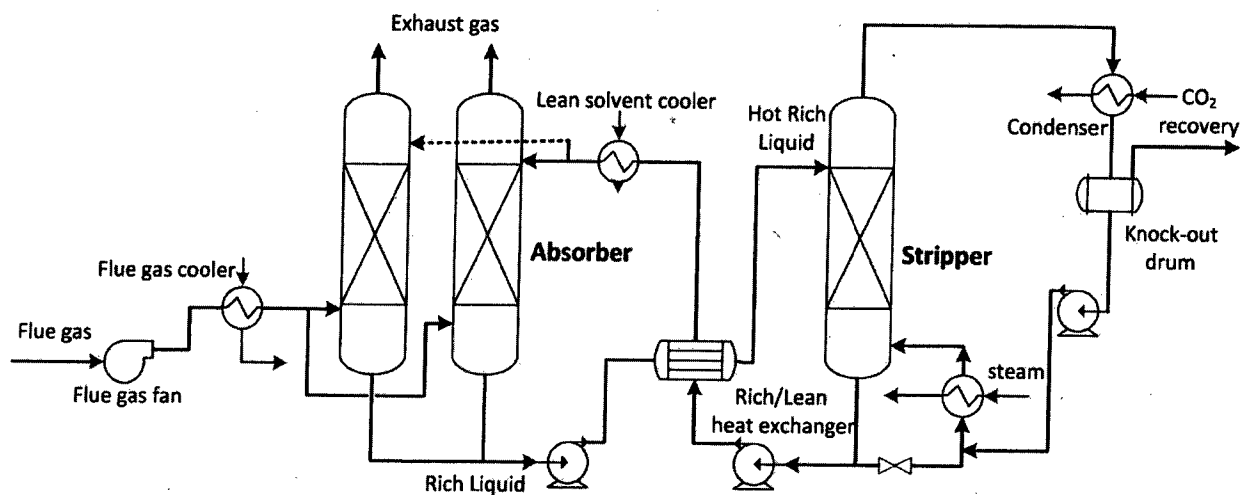


Figure 3

**Figure 4**

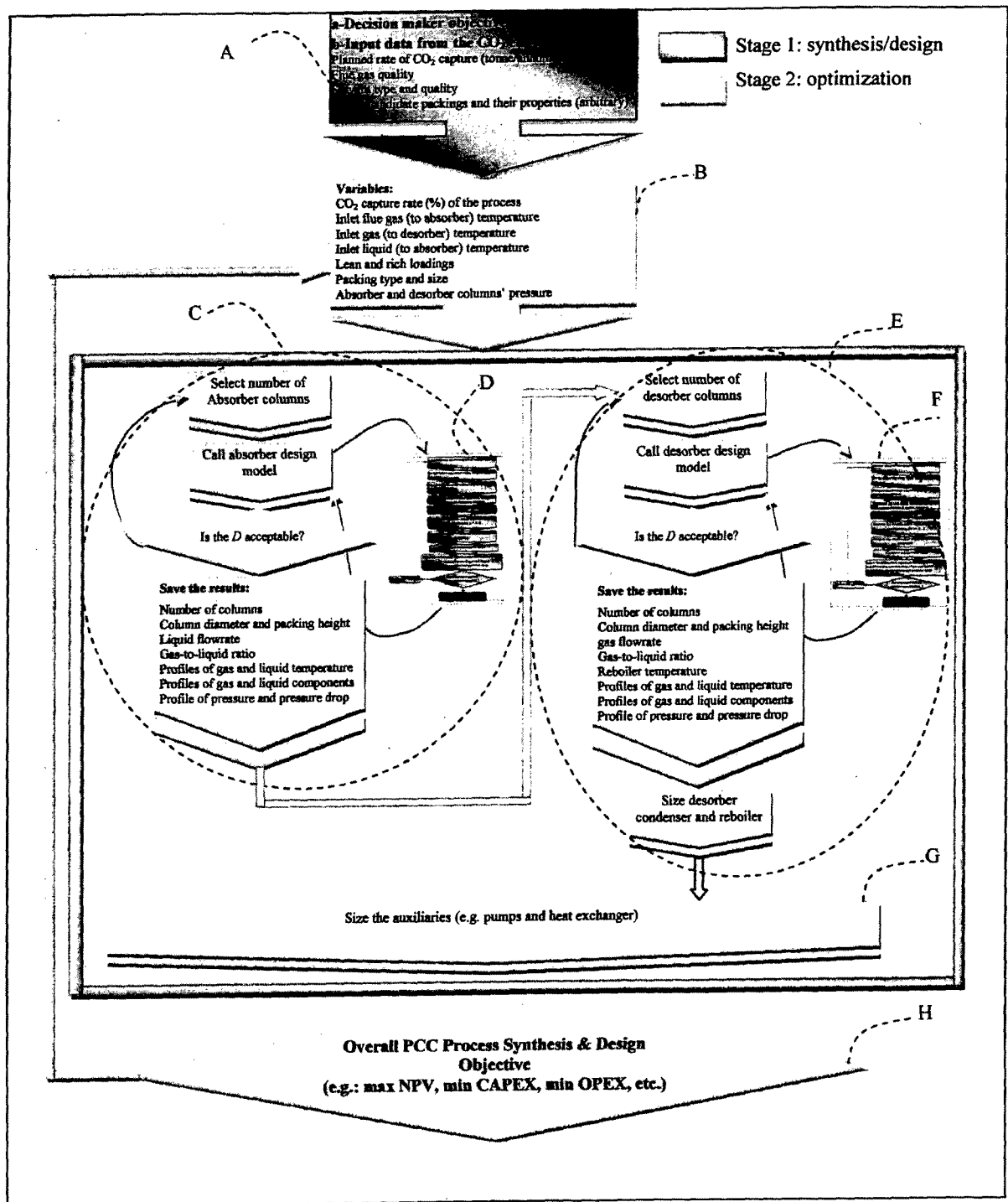


Figure 5

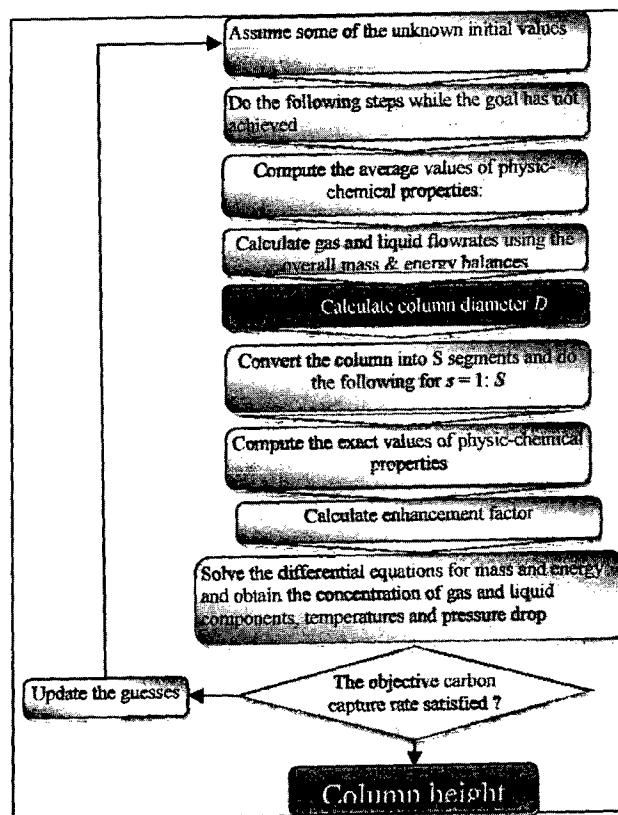


Figure 6

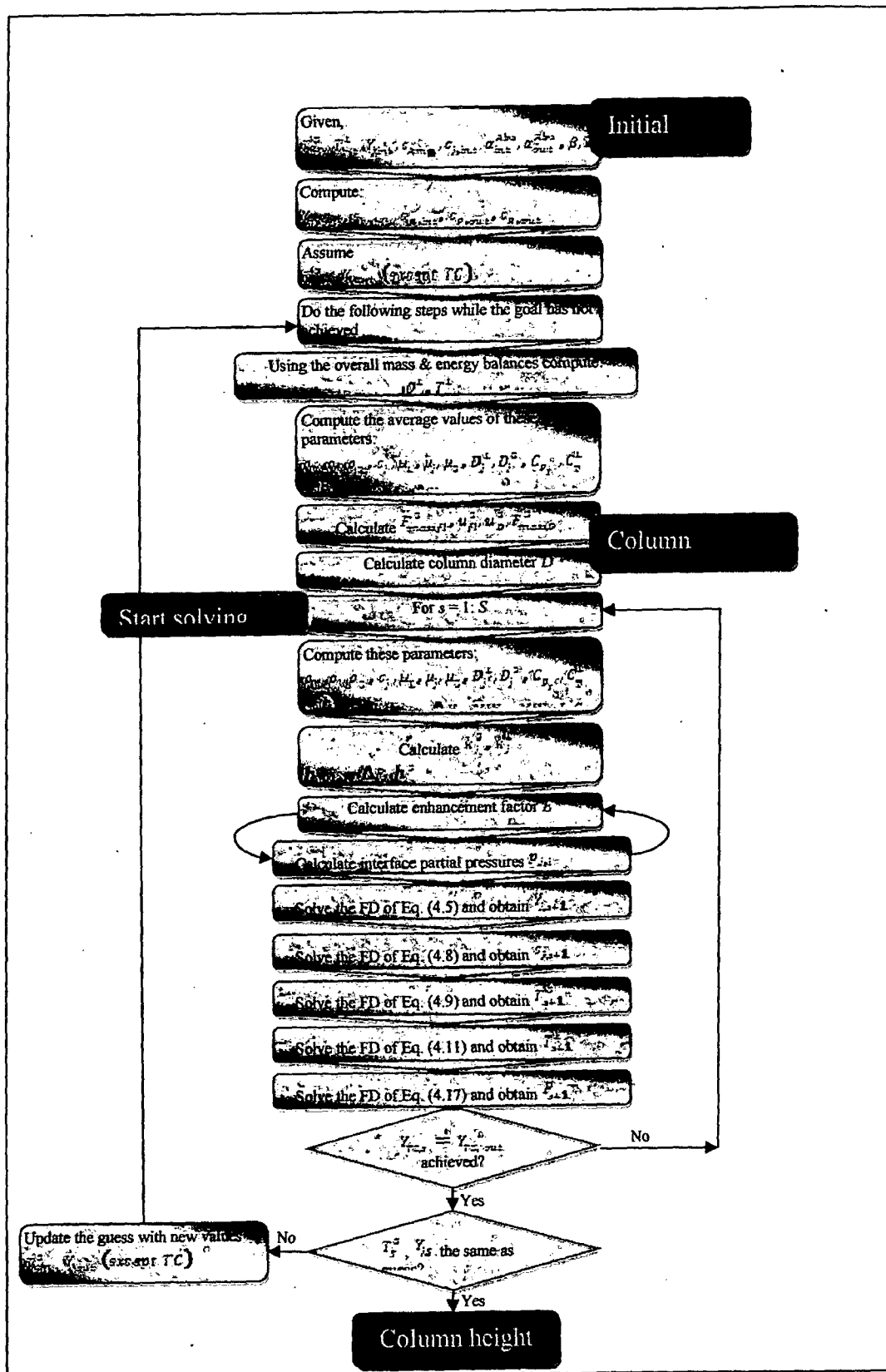


Figure 7

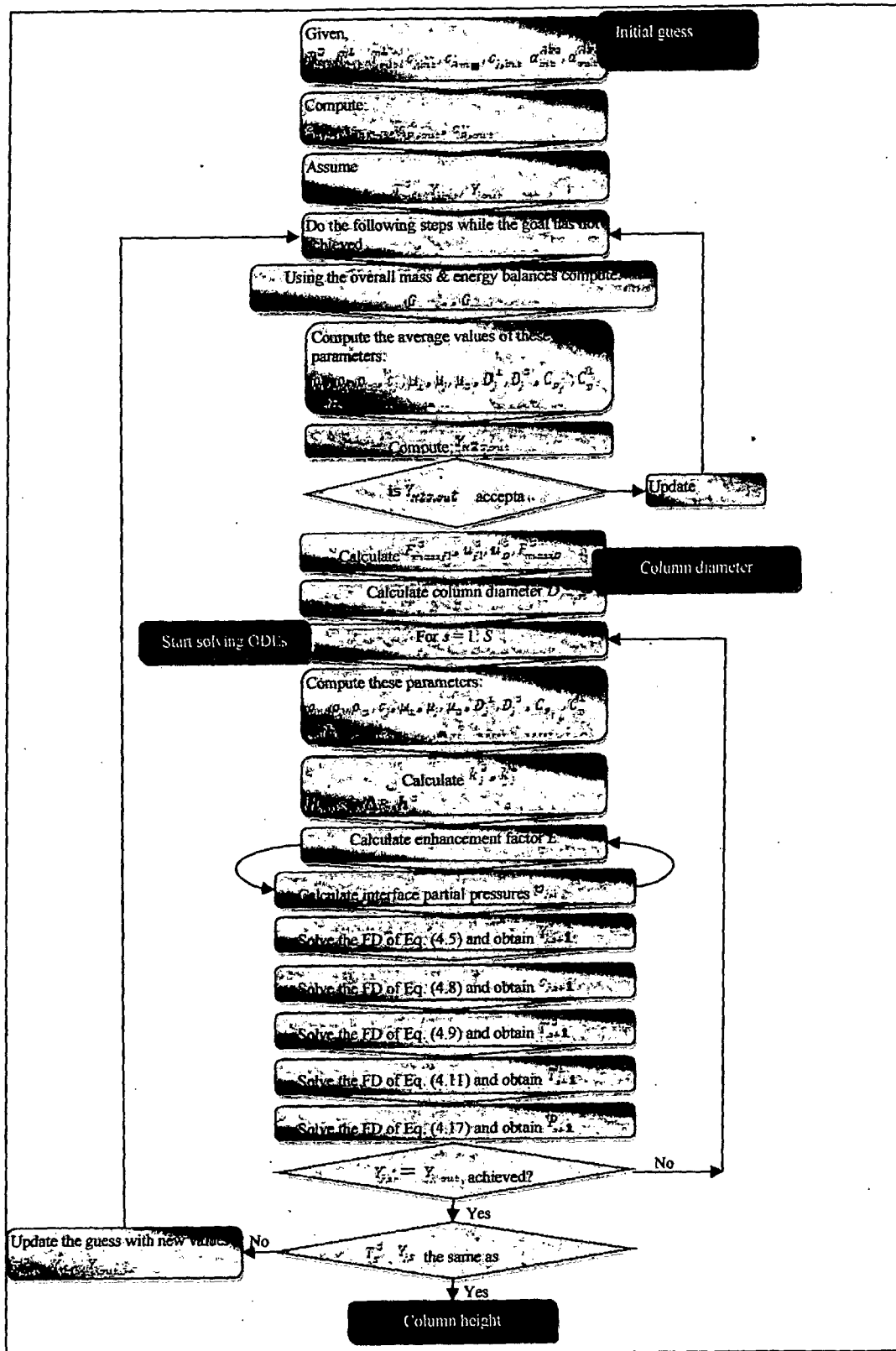
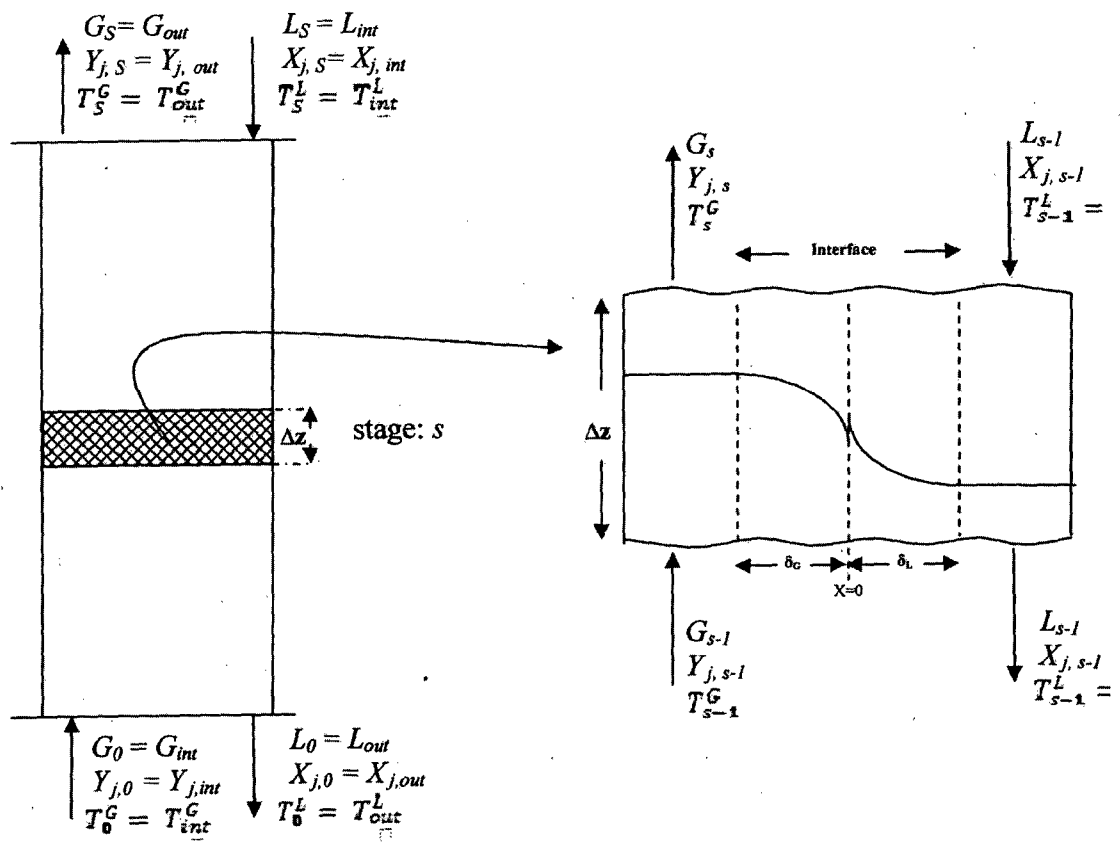
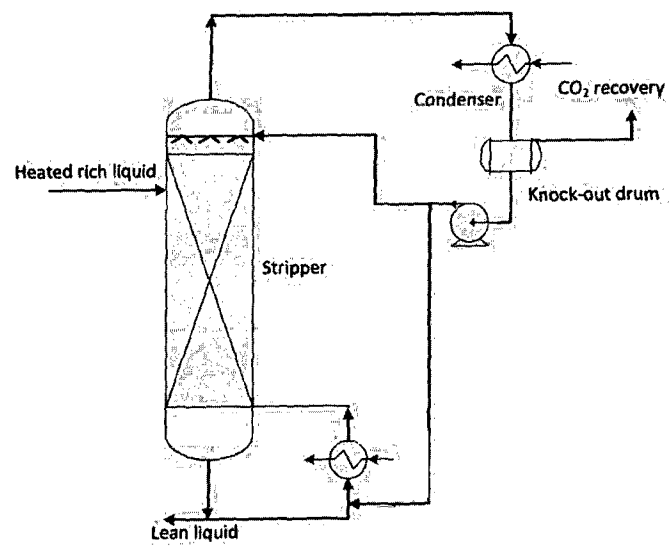
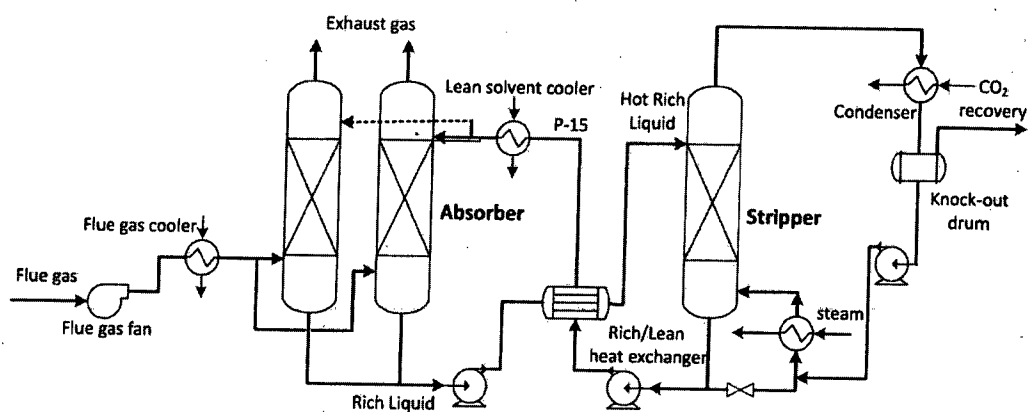


Figure 8

**Figure 9**

**Figure 10**

**Figure 11**

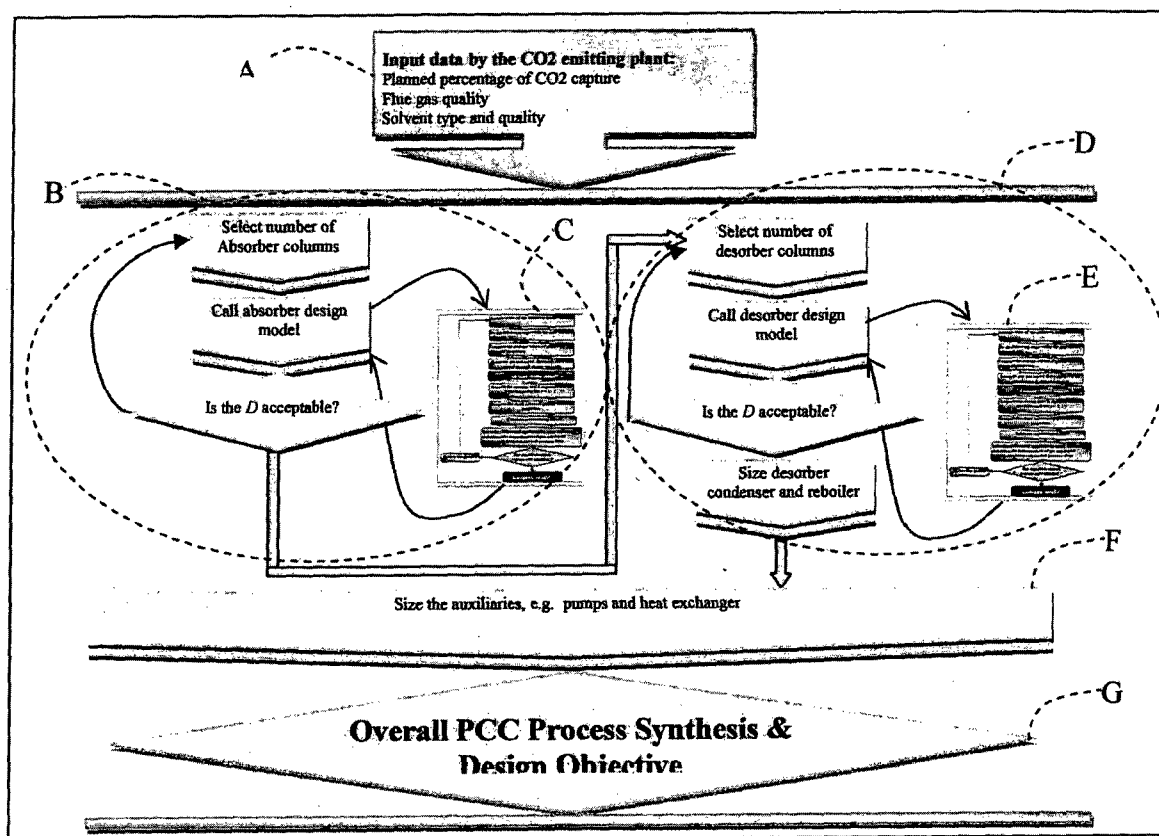


Figure 12

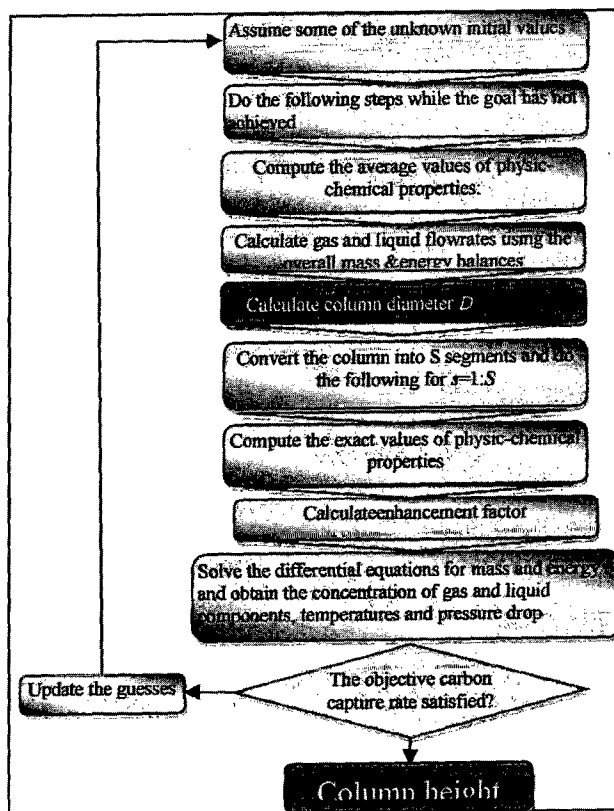


Figure 13

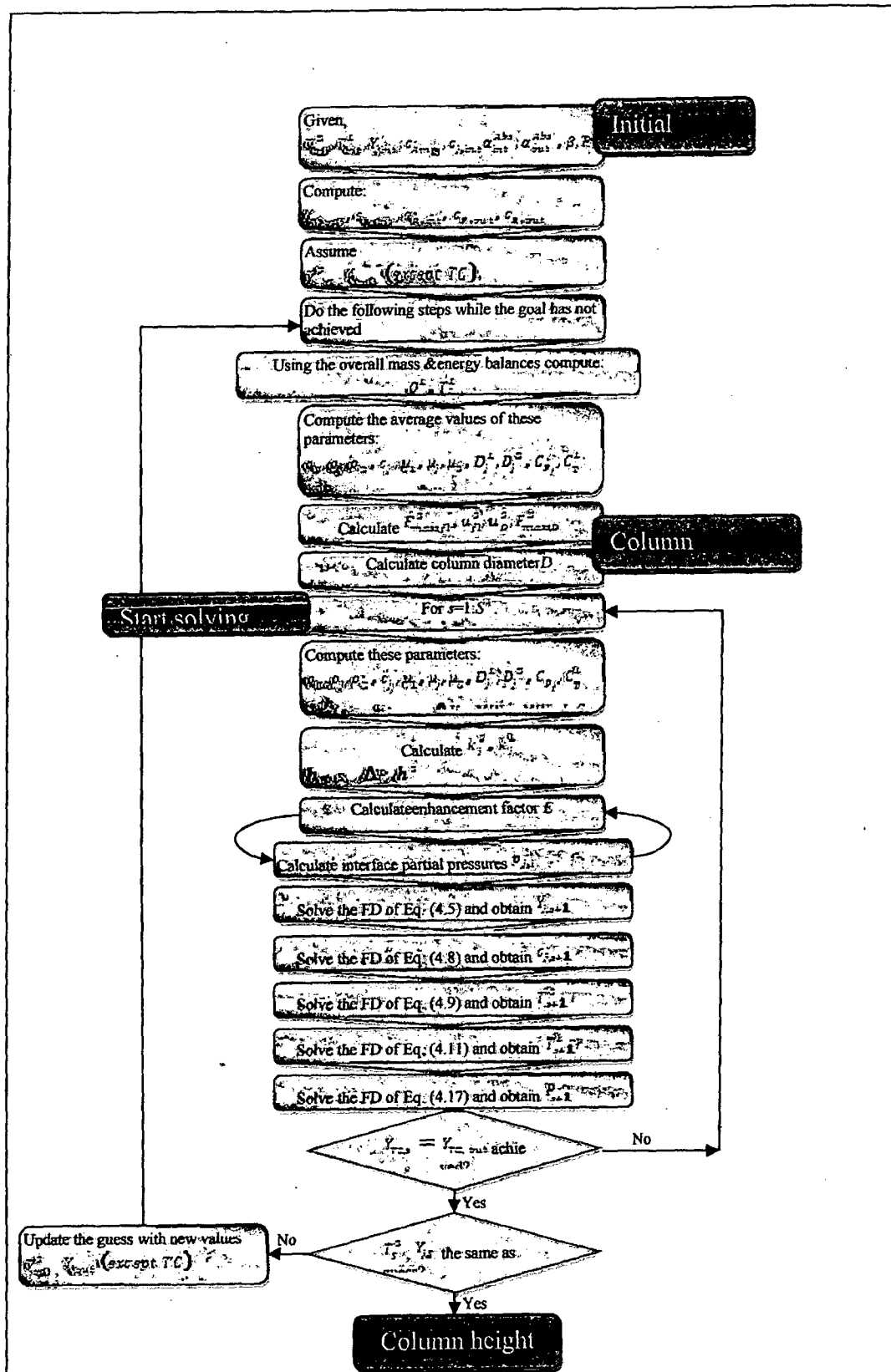


Figure 14

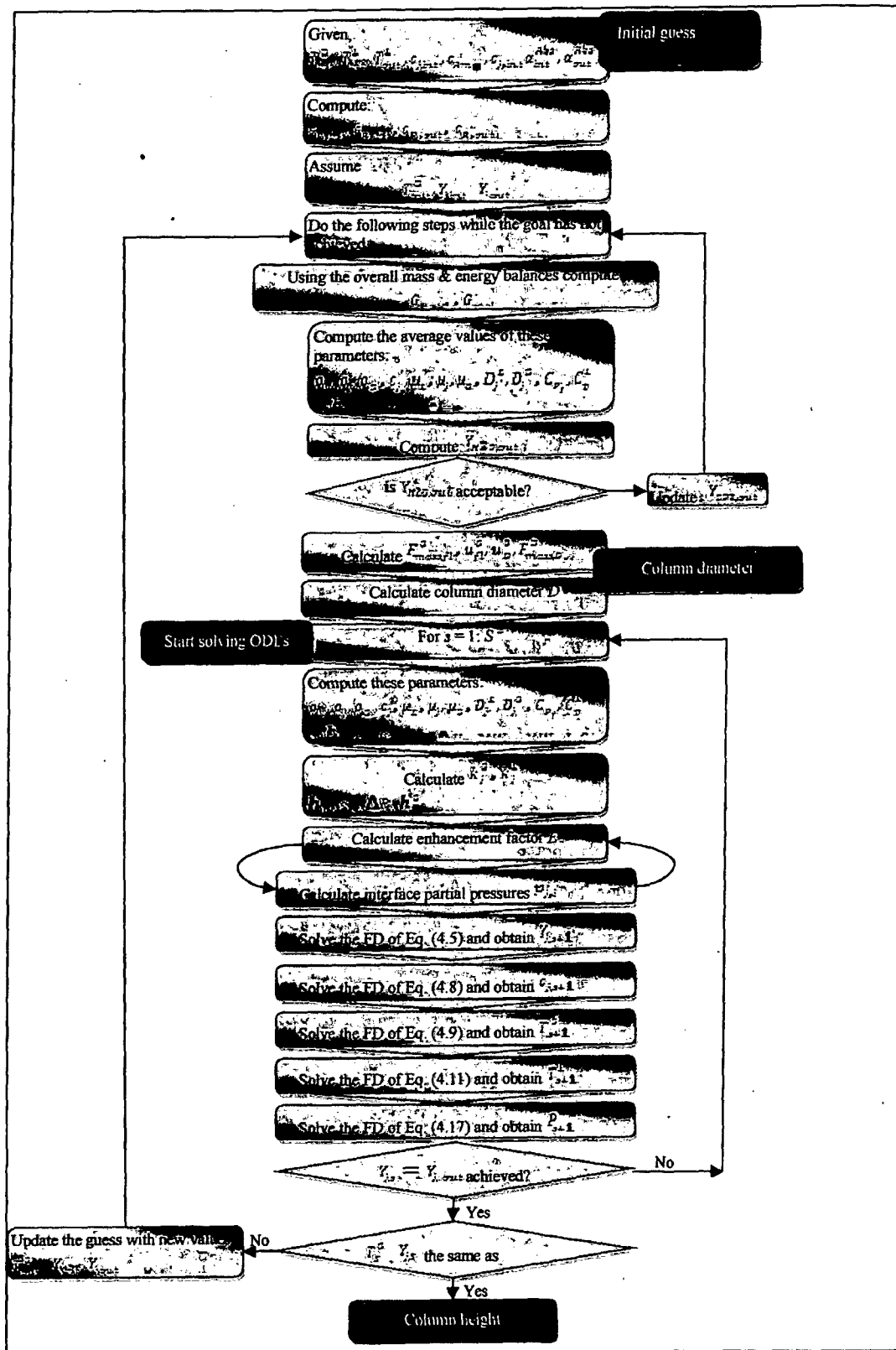


Figure 15

INTERNATIONAL SEARCH REPORT

International application No.
PCT/AU2013/000978

A. CLASSIFICATION OF SUBJECT MATTER

B01D 53/14 (2006.01) B01D 53/62 (2006.01) B01D 53/78 (2006.01) C01B 31/20 (2006.01)

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practicable, search terms used)

WPI, EPODOC:IC/CC B01D53/14, 53/62, 53/78, C01B31/20, Y02C10/06 and keywords (carbon w dioxide, CO2, captur+, stor+, sequest+, remov+, measur+, determin+, monitor+, comput+, calculat+, process+, formulat+, flow+, composit+, solvent, height, diamet+, absorb+, desorb+, design, config, mode:, optimiz+, algorithm), google patent Search, Esp@cenet, with above key words.

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category*	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
	Documents are listed in the continuation of Box C	

☒ Further documents are listed in the continuation of Box C ☒ See patent family annex

* Special categories of cited documents:	
"A" document defining the general state of the art which is not considered to be of particular relevance	"T" later document published after the international filing date or priority date and not in conflict with the application but cited to understand the principle or theory underlying the invention
"E" earlier application or patent but published on or after the international filing date	"X" document of particular relevance; the claimed invention cannot be considered novel or cannot be considered to involve an inventive step when the document is taken alone
"L" document which may throw doubts on priority claim(s) or which is cited to establish the publication date of another citation or other special reason (as specified)	"Y" document of particular relevance; the claimed invention cannot be considered to involve an inventive step when the document is combined with one or more other such documents, such combination being obvious to a person skilled in the art
"O" document referring to an oral disclosure, use, exhibition or other means	"&" document member of the same patent family
"P" document published prior to the international filing date but later than the priority date claimed	

Date of the actual completion of the international search 13 November 2013	Date of mailing of the international search report 15 November 2013
Name and mailing address of the ISA/AU AUSTRALIAN PATENT OFFICE PO BOX 200, WODEN ACT 2606, AUSTRALIA Email address: pct@ipaustalia.gov.au Facsimile No.: +61 2 6283 7999	Authorised officer Asoka Dias-Abey AUSTRALIAN PATENT OFFICE (ISO 9001 Quality Certified Service) Telephone No. 0262832141

INTERNATIONAL SEARCH REPORT		International application No.
C (Continuation). DOCUMENTS CONSIDERED TO BE RELEVANT		PCT/AU2013/000978
Category*	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
X	ANUSHA KOTHANDARAMAN, Carbon Dioxide Capture by Chemical Absorption: A Solvent Comparison Study, published June 2010, [retrieved on 12 November 2013]. Retrieved from the Internet, <URL: http://www.sequestration.mit.edu/pdf/Anusha_Kothandaraman_thesis_June_2010.pdf > Abstract pages 3-4, 28-29, 43-44, 89-104, fig. 3-1, page 69	1-25
X	ZEINAB AMROLLAHI et al, Thermodynamic analysis on post combustion CO2 capture of natural gas fired power plant, [retrieved on 12 November 2013]. Retrieved from the Internet, <URL: http://www.folk.ntnu.no/ivarse/art/Qtar_jan2010.pdf > Abstract, paragraphs 3-4, 5.2, Fig. 1, 3a, 3b	1-25
X	US 2011/0245937 A1 (RAWSON et al) 06 October 2011 Abstract, paragraphs 0052-0054, 0058, fig. 9, claim 1, 4, 17, 20	1-25
A	WO 2011/062710 A2 (ALSTOM TECHNOLOGY LTD) 26 May 2011 Abstract	1-25

INTERNATIONAL SEARCH REPORT		International application No.	
Information on patent family members		PCT/AU2013/000978	
This Annex lists known patent family members relating to the patent documents cited in the above-mentioned international search report. The Australian Patent Office is in no way liable for these particulars which are merely given for the purpose of information.			
Patent Document/s Cited in Search Report		Patent Family Member/s	
Publication Number	Publication Date	Publication Number	Publication Date
US 2011/0245937 A1	06 Oct 2011	CA 2793719 A1	06 Oct 2011
		CN 102811795 A	05 Dec 2012
		EP 2552572 A1	06 Feb 2013
		US 2011245937 A1	06 Oct 2011
		WO 2011123193 A1	06 Oct 2011
WO 2011/062710 A2	26 May 2011	CA 2781266 A1	26 May 2011
		CN 102713166 A	03 Oct 2012
		EP 2501903 A2	26 Sep 2012
		JP 2013511387 A	04 Apr 2013
		KR 20120093383 A	22 Aug 2012
		MA 33887 B1	02 Jan 2013
		MX 2012005843 A	03 Aug 2012
		US 2011120128 A1	26 May 2011
		WO 2011062710 A2	26 May 2011
End of Annex			
Due to data integration issues this family listing may not include 10 digit Australian applications filed since May 2001.			