Dynamic analysis and control of sieve tray gas absorption column using MATLAB and SIMULINK

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The present work highlights the powerful combination of SIMULINK/MATLAB software as an effective flowsheeting tool which was used to simulate steady state, open and closed loop dynamics of a sieve tray gas absorption column. A complete mathematical model, which consists of a system of differential and algebraic equations was developed. The S-Functions were used to build user defined blocks for steady state and dynamic column models which were programmed using MATLAB and SIMULINK flowsheeting environment. As a case study, the dynamic behaviour and control of a sieve tray column to absorb ethanol from CO\textsubscript{2} stream in a fermentation process were analysed. The linear difference equation relating the actual and equilibrium gas phase compositions was solved analytically to relate the actual gas phase composition to the liquid phase with Murphree tray efficiency as a parameter. The steady state mathematical model was found to be nonlinear (w.r.t. number of stages) due to the introduction of the Murphree tray efficiency. To avoid the solution of large linear algebraic system, a sequential steady state solution algorithm was developed and tested through the idea of tearing the recycle stream in the closed loop configuration. The number of iterations needed to achieve a given tolerance was found to be function of the Murphree tray efficiency. The open-loop dynamic analysis showed that the gas phase composition response was nonlinear with respect to the inlet gas flow rate, while it was linear with respect to inlet gas composition. The nonlinearity increased along the column height and was maximum at the top tray. On the other hand, the Murphree tray efficiency had little effect on the dynamic behaviour of the column. The controlled variable was found to exhibit fairly large overshoots due to step change in the inlet gas flow rate, while the PID controller performance was satisfactory for step change in the inlet gas composition. The closed-loop dynamic analysis showed that the controlled variable (outlet gas phase composition) had a fairly linear dynamics due to step changes in the set point.

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1. Introduction

Steady state design of chemical equipment is confronted by dynamic and controllability issues. In this regard, it is often easy to design a chemical process based on steady state conditions, which is practically uncontrollable and unrealistic. In order to avoid any wrong assumption during process synthesis and design, and to ensure safe start-up, shutdown and stable plant operation, the dynamic behaviour of the relevant units should be known. Dynamic simulation is known to be a slow process, in particular when used on the flowsheet level, where the challenge is dealing with processing units having different time constants. However, dynamic simulation makes use of recent advances in computers power in terms of memory and computational speed, and the advances in information technology where user friendly GUI and flowsheeting packages are used intensively \cite{1,2,21,28,29}. Kvamsdal et al. \cite{2} presented three important factors, which contribute to the importance of the dynamic simulation to improve the overall design and optimize the operation of chemical processing units. These factors are: (1) the coupled absorber/stripper system is complex, with higher degree of two-way interaction between these two units. (2) The upstream processing units might operate under a varying load operation. (3) New process approaches, with energy integration mean more complex operations. Therefore, a dynamic process simulator enables the study of most of these isolated or coupled effects.

The dynamic simulation and control of gas absorption process attracted many researchers’ attention on both individual and flowsheet (planwide) levels \cite{2,3,28,29}. Kvamsdal et al. \cite{2} presented a dynamic model of a CO\textsubscript{2} absorption column that is intended to
be coupled with models of other individual processes to form a complete model of a power generation plant with CO₂ removal. In their model, the operational challenges, such as load variation and high degree of heat integration between the power plant and the absorber/stripper process were studied. Lin et al. [3] presented planwide control of a reactive CO₂ absorption/stripping process with monoethanol-amine as a solvent using dynamic simulation. These authors proposed a new control structure, where the liquid absorbent flow rate (at the top of the column), the liquid level and temperature at the bottom of the stripping column were controlled. Through the help of dynamic simulations, the developed model and control structure was found to achieve removal targets and stabilize quickly under the influence of external disturbances.

Robinson and Luyben [28,29] presented a hybrid power/chemical plant model for the purpose of dynamic simulation and planwide control structure design. This hybrid power/chemical plant is the gasification process producing synthesis gas, which under standard operation conditions feeds a combustion turbine to generate electricity or feed a chemical plant during periods of lower power demand. These authors used process simulator ASPEN to perform dynamic simulations of the H₂S and CO₂ absorption/stripping processes and water–gas shift reactors, which are essential for the development of stable and robust plantwide control structures of this hybrid plant.

From the above review, it is obvious that dynamic simulation and control of gas absorption process is essential. Moreover, it is clear that gas absorption/stripping is one of the main and important processing blocks in many chemical and power generation plants. It is used in pollution control devices such as wet scrubbers and spray-dryer-type, dry scrubbers for the removal of acid gas compounds and water-soluble organic compounds [4,5]. This process is usually carried out through tray columns that contain multiple numbers of trays, which bring gas and liquid into intimate contact. If the gas leaving the tray is in thermodynamic equilibrium (which is a rather rare situation) with the liquid leaving the tray, then theoretical stage is provided. To account for the failure to achieve equilibrium, Murphree tray efficiency is used. The computational approach is to determine the theoretical stages and then correct to actual stages by means of tray efficiency [6]. The use of the component Murphree tray efficiency for separation of binary mixtures has been described by several authors: Hinze and Maddox [7], Van Winkle [8], Edmister [9] and Holland and McMahon [10]. However, a simple way to apply Murphree tray efficiency to liquid–vapour separation processes has not been shown. Edmister [9] presented a description of different types and uses of tray efficiencies. A tray efficiency was defined as a multiplier of the absorption or stripping factor on each stage. Takamatsu and Kinoshita [11] showed a new solution process for multi-component distillation columns. They took the liquid mole fractions to be the independent variables and their difference between consecutive iterations to be the functions set to zero. Their procedure seems to be stable and fast, and the application of the Murphree efficiency includes a new step in solving the liquid–vapor composition non-equilibrium equations with a successive substitution method. In general, the values of the Murphree tray efficiencies are not equal throughout the column. Further, they usually are not equal for the different components in a mixture, even on the same stage.

The gas absorption process is modelled through a system of mathematical equations to enable prediction of the process behaviour [12]. These equations do not have a general analytical solution and some approximations and numerical methods can be used for solving them [13]. The implementation of a control scheme for such a process is vital to achieve optimal operation despite the presence of significant uncertainty about the plant behaviour and disturbances. The purpose of any control system is to suppress the influence of external disturbances, ensure the stability of process and optimize process performance. The feedback system is a common control configuration where it uses direct measurements of the controlled variables to adjust the values of the manipulated variables. The objective is to keep the controlled variables at desired levels (set points) [14,15]. There are various controllers that can be considered for implementation such as: proportional–integral–derivative (PID) controller and Artificial Neural Networks (ANN) controller [16]. As the capability of a certain controller is not the main issue of the present work, the traditional PID controller was selected for its simplicity when compared with ANN and being able to achieve the required targets. Tuning the PID feedback controllers is the adjustment of the controller parameters to match the characteristics of the rest of the components of the loop. One of the popular methods is the on-line or closed-loop tuning method. For the desired response of the closed loop, Ziegler and Nichols specified a decay ratio of one-fourth. The decay ratio is the ratio of the amplitudes of two successive oscillations [17,18].

MATLAB is a software for mathematical computation, whereas SIMULINK is a powerful software for modelling, simulation, and analysis of dynamical systems in a flowsheeting environment. It supports linear and nonlinear systems, modelled in continuous time, sampled time, or a hybrid of the two. Bequette [19], illustrated that the interactive MATLAB/SIMULINK tool enhances the ability to learn new model-based techniques and provide an inside depth of the dynamic nature and control of chemical processes [20]. For modelling, SIMULINK provides a graphical user interface (GUI) for building models as block diagrams, using click-and-drag mouse
operations. With this interface, you can draw the models just as you would with pencil and paper. SIMULINK includes a comprehensive block library of sinks, sources, linear and nonlinear components, and connectors. You can also customize and create your own block models using the S-Function format. This approach provides insight into how a model is organized and how its parts interact. After you define a model, you can simulate it, using a choice of integration methods, either from the SIMULINK menus or by using MATLAB’s m-files. The menus are particularly convenient for interactive work, while the m-file approach is very useful for running a batch of simulations. Using scopes and other display blocks, you can see the simulation results while the simulation is running. In addition, you can change parameters and immediately see what happens, for “what if” exploration. The simulation results can be put in the MATLAB m-files for post processing and visualization. And because MATLAB and SIMULINK are integrated, you can simulate, analyse, and revise your models in either environment at any point [19]. Therefore, modeling and control of stagewise processes is now common using SIMULINK with its versatile environment to get rapid and accurate simulation for models of varying degree of complexity. For example, Mjalli [21] conducted neural network-based control algorithms to control the product compositions of a Scheibel agitated extractor through modelling and simulation. While, population balance SIMULINK model for a crystallization process was developed by Ward and Yu [22].

The objective of this work is to develop a steady state and dynamic models for the coupled hydrodynamics and mass transfer of a sieve tray gas absorption column, where an existing bioethanol process [16] is simulated as a real-life example. The developed models are programmed in MATLAB and SIMULINK flow sheeting software using the standards of S-Functions. The combination of SIMULINK and MATLAB is utilized to develop an industrial feedback control system for a general tray gas absorption column with only one solute transfer. This environment allows for a practical way to develop a control block diagram including measuring device, feedback controllers and final control element.

2. Dynamic and steady state models

In this section we present a coupled tray hydrodynamics and mass transfer model for a general binary gas absorption column. The dynamic model is derived first and then followed by the steady state model derivation.

2.1. Dynamic model derivation

To simplify the derivation of the tray gas absorption dynamic model, the following set of assumptions is used:

1. Liquid on the tray is perfectly mixed and incompressible.
2. Tray vapor holdup is negligible.
3. Vapor and liquid are in thermal equilibrium (same temperature) but not in phase equilibrium. A Murphree tray efficiency is used to describe the departure from equilibrium.
4. Total gas flow rate \( G \) is constant.
5. The equilibrium relationship is linear.

Using the above assumptions, the dynamic mathematical model for the tray gas absorption column is derived using unsteady state total material and component balances on the jth tray inside the column as shown in Fig. 1. This figure shows the schematic diagram of the tray gas absorption column under consideration with a focus on single cross-flow tray. The liquid flow from each tray is predicted using Francis equation [18].

Feed gas

![Schematic diagram of a tray gas absorption column with a focus on single cross-flow tray.](image)

To take into account the deviation of tray performance from ideal behaviour (equilibrium), the Murphree tray efficiency is used:

\[
\eta = \frac{y_j - y_{j+1}}{y_j^* - y_{j+1}} \tag{1}
\]

where \( y_j^* \) is the equilibrium solute concentration in the gas phase leaving stage \( j \) while \( y_j \) is the actual (non-equilibrium) solute concentration in the gas phase leaving stage \( j \). The equilibrium solute concentration in the gas phase is assumed linear with constant distribution coefficient. This is a reasonable assumption for dilute solutions. Usually, the equilibrium data are represented in the form of Eq. (2):

\[
y_j^* = K x_j \tag{2}
\]

where \( x_j \) is the mole fraction of solute in the liquid phase. Although, ideally, the \( K \) values can be derived from pure-component vapor pressure using Raoult’s law, in fact, the \( K \)-values vary with total system pressure, temperature, and composition. Fortunately, extensive charts and correlations have been developed for predicting \( K \) values for many components, particularly those associated with the natural gas and oil refining industries [23].

2.1.1. Total mass and component material balances on tray \( j \)

By referring to the jth tray schematic diagram, which is shown in Fig. 1, the unsteady state material balance reads:

\[
\frac{dM_j}{dt} = L_{j-1} + G - L_j - G
\tag{3}
\]

In the above equation, \( M_j \) is the total liquid holdup on the jth tray, while \( L_{j-1}, L_j \) (mol/s) are the inlet and outlet liquid flow rates respectively, while the total gas flow rate, which is assumed constant, is \( G \) (mol/s). Similarly, a solute balance on the jth tray can be written as:

\[
\frac{d}{dt}(M_j x_j) = x_{j-1} L_{j-1} - x_j L_j + G(y_{j+1} - y_j) \tag{4}
\]
where \( x_i \) and \( y_i \) are the solute mole fractions in the liquid and gas phases respectively. The tray index \( j \) runs from 1 (top tray) to \( n \) (the bottom) tray.

2.1.2. Tray fluid dynamics

From dynamic point of view, the liquid flow rates throughout the column will not be the same. They will depend on the fluid mechanics of the tray. Often a simple Francis weir formula relationship is used to relate the liquid holdup on the \( j \)th tray \( (M_j) \) to the liquid flow rate leaving the tray \( (L) \) [24]:

\[
L_j = \frac{\rho_L L_w}{c} \left[ \frac{M_j}{\rho_L A_p} - h_w \right]^{3/2}
\]

where \( h_w \) is weir height \((m)\), \( L_w \) is weir length \((m)\), \( A_p \) is active plate area \((m^2)\), \( \rho_L \) is the molar density of the liquid mixture \((mol/m^3)\), \( c \) is a weir empirical constant \((m^{-1/3} s^{2/3})\).

The Murphree tray efficiency can be written explicitly in terms of the equilibrium solute mass fraction \( y_j^* \) as follows:

\[
y_j = (1 - \eta)^{N+1} y_j^* + \eta \sum_{m=0}^{N-j-1} (1 - \eta)^{N-j-m} y_N^* \]

where \( j = 1, 2, \ldots, N \). Note that when \( \eta \to 1 \), \( y_j \to y_j^* \) and hence the ideal tray model is recovered. In the above equation the first term is the solution of the homogeneous finite difference equation (Eq. (1)), while the second two terms are the particular solution. It is clear that the first term accounts for the liquid boundary condition, which is required to solve the Murphree finite difference equation. Note that all the terms in Eq. (6) are nonlinear with respect to tray efficiency \((\eta)\), except the second term. The second two terms are nonlinear with respect departure of the tray from equilibrium conditions. The derivation of Eq. (6) as a solution to the finite difference Eq. (1) is shown in Appendix A.

It is obvious that the above system of equations is a differential algebraic equation (DAE) system, which can be solved sequentially in time thanks to the explicit form of Eq. (6). Note that the solute mole fraction \( y_j \) in Eq. (4) can be eliminated using Eqs. (6) and (2), which results in a system of ODE in terms of solute mole fraction \( (x) \) in the liquid phase.

2.1.3. Initial conditions

The initial conditions for the system of Eqs. (3) and (4) are given by the solution of the steady state versions of these equations. The initial conditions can be stated mathematically as:

\[
M_j(0) = M_{0,j} \quad L_j(0) = L_{0,j} \quad x_j(0) = x_{0,j} \quad y_j(0) = y_{0,j} \quad \forall j = 1, 2, \ldots, N
\]

To examine the system degrees of freedom, the feed gas flow rate \((G)\) and its inlet composition \((y_{in})\) are assumed to be given from the upstream unit. The Murphree tray efficiency is assumed to be known empirically. The total number of variables is equal to \( 5N + 2 \) and the total number of equations is \( 5N \). From control engineering point of view, there are only two variables that can be controlled. These two variables are the inlet absorbent flow rate and composition. The inlet absorbent composition \((x_{in})\) is specified from the upstream unit (stripper or distillation column) and hence it is a disturbance imposed on the process. Accordingly, we are left only with the absorbent flow rate \((L_0)\) as a manipulated variable. The absorbent flow rate is manipulated using a feedback controller to control the composition of the gas stream leaving the top of the tray absorption column as shown in Fig. 2. This adds an extra equation to close the dynamic model. The equation should relate the absorbent flow rate to the gas stream composition leaving the column using the feedback (PID) controller equation:

\[
L_0(t) = f(y_N)
\]

Fig. 2 shows the block diagram of the controlled gas absorption column with total inlet gas flow rate \((G)\), composition \((y_{in})\) and the inlet absorbent composition \((x_{in})\) as the main three disturbances to the closed loop control system. The controlled (measured) variable
is the gas outlet concentration leaving the top tray (yN), which is controlled by manipulating the absorber flow rate (L0).

2.2. Steady state model derivation

The steady state model of the tray gas absorption column is derived using the steady state versions of Eqs. (3) and (4) coupled with the algebraic system given by Eq. (6), where Lj−1 = Lj = L at steady state. For ease of presentation, the steady state model is put in compact matrix form as follows:

\[ A_x x = Bu - \Delta y \]  

(9)

The above equation is a result of combining Eqs. (3) and (4), while the Murphree tray efficiency (Eq. (6)) is expressed in compact matrix form as follows:

\[ A_y y = -\eta K x - (1 - \eta)y_m v \]  

(10)

In the above equations, the elements of \(A_x\) and \(A_y\) are given by:

\[
(a_x)_{i,j} = \begin{cases} 
\left( \frac{L_0}{G} \right) & \text{if } i = j \\
\left( \frac{L_0}{G} \right) & \text{if } j = i + 1 \\
0 & \text{else}
\end{cases}, \quad \forall i, j = 1, 2, \ldots, N
\]

\[
(a_y)_{i,j} = \begin{cases} 
1 & \text{if } i = j \\
1 - \eta & \text{if } j = i + 1 \\
0 & \text{else}
\end{cases}, \quad \forall i, j = 1, 2, \ldots, N
\]

While the elements of the vectors \(B\), \(u\), \(\Delta y\), \(v\) and \(y\) are given by:

\[
B = \begin{bmatrix} 
-1 & 0 \\
0 & 0 \\
\vdots & \vdots \\
0 & -1 
\end{bmatrix}, \quad u = \begin{bmatrix} 
L_0 \\
y_m 
\end{bmatrix}, \quad \Delta y = \begin{bmatrix} 
\Delta y_1 \\
\Delta y_2 \\
\vdots \\
\Delta y_N-1 \\
-y_N 
\end{bmatrix}, \quad \Delta y_j = y_{j+1} - y_j, \quad j = 1, 2, \ldots, N \]

\[
v = \begin{bmatrix} 
0 \\
0 \\
\vdots \\
0 \\
1 
\end{bmatrix}, \quad x = \begin{bmatrix} 
x_1 \\
x_2 \\
\vdots \\
x_{N-1} \\
x_N 
\end{bmatrix}, \quad y = \begin{bmatrix} 
y_1 \\
y_2 \\
\vdots \\
y_{N-1} \\
y_N 
\end{bmatrix}
\]

Note that the system input variables (the manipulated variable: \(L_0\) and the disturbances: \(G\), \(x_m\) and \(y_m\)) are all contained in the input vector \(u\), while the system outputs (the liquid and gas phase compositions) are contained in the vectors: \(x\) and \(y\) respectively.

This decomposed system consists of two parts: the x-phase and the y-phase systems, where each of them has an N × N dimension. On the other hand, the coupled system of linear equations has a dimension of 2N × 2N. This results in the solution of large and dense linear system, which is undesirable from numerical point of view. Fig. 3 presents the conceptual flow diagram of the coupled system in both closed loop and cycle tearing configurations. This is to highlight the sequential solution algorithm of this system. By doing this, an efficient iterative solution algorithm is developed. The algorithm starts by neglecting the correction terms (the two nonlinear terms) in Eq. (6) to get the following approximation:

\[ y = \eta v \]  

(11)

Moreover, by dropping the \((1 - \eta)\) term from Eq. (10) one gets \(A_y y = \eta K x\), where the matrix \(A_y = I (I \text{ is the identity matrix})\). Accordingly, using the last result, the vector \(\Delta y\) in Eq. (9) becomes \(\Delta y = \eta K \Delta x\). Substituting this in Eq. (9) to get \(A_x x + \eta K \Delta x = Bu\), which can be written in a more compact way to solve for the initial guess vector \(x_0\):

\[ A x_0 = Bu \]  

(12)

Due to the appearance of the vector \(\eta K \Delta x\), the elements of the matrix \(A\) are given by:

\[
(a)_{i,j} = \begin{cases} 
-\left( \frac{\eta K + L_0}{G} \right) & \text{if } i = j \\
L_0 & \text{if } i = j + 1 \\
0 & \text{else}
\end{cases}
\]

And the elements of \(B\) and \(u\) are the same as those given by Eq. (9). The steady state solution algorithm can be summarized as follows:

1. Solve the system: \(A x_0 = Bu\) to get the initial guess \(x_0\).
2. Calculate the vector \(y\) by solving the system: \(A_y y = -\eta K x - (1 - \eta)y_m v\).
3. Calculate the new vector \(x\) by solving the system: \(A x = Bu - \Delta y\).
4. Check for convergence: error = |\(x - x_0\)|.
5. if (error < Tolerance) then
   \text{STOP}
else
   \(x = x_0\)
   GO TO STEP 2
End if

2.3. Dynamic model derivation and solution algorithm

After expanding the derivative vector on the left hand side of Eq. (4), the dynamic model given by Eqs. (3)–(5) is cast into the following state-space form, which is suitable for MATLAB/SIMULINK implementation:

\[ \frac{dx}{dt} = A_x x + A_y y + Bu \]  

(13)

where the elements of \(A_x\) and \(A_y\) are given by:

\[
(a_x)_{i,j} = \begin{cases} 
-\left( \frac{L_i}{M_i} \right) & \text{if } i = j \\
L_i & \text{if } i = j + 1 \\
0 & \text{else}
\end{cases}, \quad j = 1, 2, \ldots, N
\]

\[
(a_y)_{i,j} = \begin{cases} 
\frac{1}{m} \frac{dM_j}{dt} & \text{if } i = j \\
L_i & \text{if } i = j + 1 \\
0 & \text{else}
\end{cases}, \quad j = 1, 2, \ldots, N
\]

\[
u = \begin{bmatrix} 
x_m \\
y_m 
\end{bmatrix}
\]
and \( y \) is given by the solution of Eq. (6) or (10). Note that the DAE system above, which results from the slow dynamics of the tray liquid composition \((dx/dt)\) and the negligible dynamics of the gas phase composition \((dy/dt \approx 0)\) is represented by the two state variables \( x \) and \( y \) which are transformed by the matrices \( A'_x \) and \( A'_y \), while the system disturbances and manipulated variables are augmented in the vector \( B'u \).

The dynamic solution algorithm starts by filling the trays with \( M_j \) kmol liquid (absorbent) followed by solving the steady state system given by Eqs. (9) and (10) using the algorithm of Section 2.2.1 to provide an initial condition to the dynamic model. Then, the system of ODEs given by Eq. (13) is integrated using the standard MATLAB ordinary differential equation solvers.

It is interesting to explore again the nonlinear structure of the dynamic model by examining the elements of the governing matrices in the system of Eq. (13). The matrices \( A'_x, A'_y \) and \( B' \) are functions of time and space (stage number) due to the coupled mass transfer process and tray hydrodynamics. This is exacerbated by the nonlinear behaviour of the equilibrium gas phase mole fraction as function of liquid tray composition, which is reflected by Eq. (6). This nonlinearity is expected to propagate along the column height even when the system attains its steady state value due the appearance of the fractional efficiency \( \eta \) in Eq. (6). For ideal tray \((\eta=1)\) the gas phase composition is given in Eq. (2) with constant equilibrium constant \((K)\). This reduces the right hand side of the dynamic model (13) to \( [A'_x + KA'_y]x + B'u \). Since the two matrices \( A'_x \) and \( A'_y \) are both bidiagonal, then the eigenvalues of \([A'_x + KA'_y]\) are simply the diagonal elements, which are given by \( \lambda_i = (a'_{x,i}) + (a'_{y,i}) < 0 \). This insures the stability of the dynamic model (13). Moreover, the speed of response is controlled by the system time constants, which are given by \( \tau_i = [(a'_{x,i}) + (a'_{y,i})]^{-1} \). This shows that the dependence of the time constants on the gas and liquid flow rates, the dynamic tray holdup and the equilibrium distribution coefficient \((K)\). Therefore, the ideal sieve tray absorption column is expected to show a nonlinear dynamic behaviour characterized by variable
time constants. This variability depends on the absorbent flow rate ($L$) and the gas phase flow rate ($G$). For the general case (nonideal tray), the matrix $A^0$ in Eq. (13) is modified by the elements of the $A^{-1}$ as a solution to the system of Eq. (10). This results in new elements which are functions of $\eta$ and $K$. Since these two parameters are considered constant, one expect no effect on the system nonlinearity, which is inherited to the try hydrodynamic model as shown above.

3. MATLAB/SIMULINK software: control of gas absorption column

To highlight the applicability of the developed steady state and dynamic solution algorithms, a real industrial bioethanol gas absorption process is chosen as a case study [16]. By referring to Fig. 4, this process is described as follows: The collected gas mixture ($CO_2 + ethanol$) from the sugarcane fermentation tanks is fed at the bottom and the absorbent liquid at the top of the ethanol recovery column, where the two phases flow counter currently. A distillation column supplies the absorbent liquid, consisting of water containing some alcohol (around 100 ppm). This absorption column is composed of nine stages and operates at 40 °C and 1 atm. This equipment must process 3 mol/s of gas mixture in order to reduce the alcohol concentration from approximately 20,000 to 300 ppm.

The initial concentration depends on the fermentation broth conditions and also on specific features of the fermentation tanks. Therefore, this is considered as a disturbance to the absorption process and the absorbent flow rate must be manipulated to maintain a high ethanol recovery. The effluent gas from the first column, mainly composed of carbon dioxide, a small amount of ethanol, and around 2000 ppm of water, is fed into a depurator to oxidize the organic impurities. A 10-tray absorption column is then used to reduce the water concentration in the gas mixture to 300 ppm. The control purpose is: to keep low ethanol concentration in the effluent gas phase from the first absorption column (ethanol recovery column) despite the changes in the two main disturbances: the

Fig. 4. (a) The complete bioethanol process [16] and (b) the input–output structure of the absorption column.
main gas flow rate and the composition of the inlet gas stream (ethanol concentration). Therefore, the measured variable is the ethanol concentration ($y_{out}$) in the effluent gas stream and the manipulated variable is the inlet absorber flow rate ($L_0$).

To get initial conditions for the solution of the dynamic model, the developed steady state solution algorithm (steps 1–5) is implemented using MATLAB programming language. The input data are written in an input file and loaded at the beginning of the main programme during the execution phase.

Table 1 shows the definition and typical values of the input variables to simulate the bioethanol process. The input data is used by both steady state solver and the dynamic simulation programme that calls the SIMULINK flowsheet, whereas, the main output data resulting from calling the steady state solution algorithm is the steady state liquid and gas phases mole fractions ($x$ and $y$ respectively) and the steady state liquid holdup ($M_0$) on each tray. These are used as initial condition vectors for the dynamic SIMULINK flowsheet as given by Eq. (7).

One of the attractive features of SIMULINK is its flowsheeting capability to construct block flow diagrams, which provide an intuitive and hierarchical approach to simulating feedback control systems. The "sfun" SIMULINK block can be used for building user defined blocks which are nonlinear and complicated in nature by invoking the "nlin" library in SIMULINK. This block can then be used to pass the input data from other SIMULINK blocks to the S-Function. The S-Function can be written in M (level-1) a programming language and must conform to S-Function standards. The variables: time ($t$), the state variables ($x$), the input system parameters ($u$), and flags are automatically passed to the S-Function by SIMULINK. The nonlinear model can be defined in a MATLAB m-function, while the SIMULINK block itself has a library extension "mxd1". Accordingly, for control purposes the tray gas absorption column dynamic model, which is given by Eq. (13) is implemented as an imbedded S-Function in SIMULINK. This is because the coupled system of equations comprises a nonlinear dynamic model and hence the linear blocks in SIMULINK are not able to model it directly. However, with the help of the powerful S-Function blocks from the nonlinear library in SIMULINK one can implement nonlinear and new models and embed them as new SIMULINK blocks [25]. For the synthesis of the tray gas absorption column using SIMULINK environment, a control system synthesis analogy similar to that of process synthesis is followed [26]. The modified process synthesis steps are restated here to highlight its applicability to control system design:

1. Identify the input–output process structure (manipulated variables, disturbances and the output state variables).
2. Write the gas absorption dynamic model in an S-Function and prepare the process SIMULINK block.
3. Feed preparation step: choose the manipulated variable and the final control element and put it in SIMULINK blocks.
4. Design the recycle by choosing suitable measuring device for the controlled variable.
5. Choose the feedback controller: P (proportional), PI (proportional plus integral) or PID (proportional plus integral plus derivative) and prepare its SIMULINK block.
6. Close the loop (task integration).

The input–output structure of the bioethanol process given in Fig. 4b is constructed by drawing a circle around the absorption column (Fig. 4a) where all the system internal details are hidden. The input–output diagram provides essential information about the process under consideration such as the input and (disturbances: $y_{in}$, $x_{in}$ and manipulated variable: $L_0$) and output variables (controlled variable: $y_{out}$ and uncontrolled variables: $L$ and $x_{out}$). Moreover, at this early stage of control system synthesis one can identify the process controllability; in other words, whether the output variable ($y_{out}$) is adequately sensitive to changes in the manipulated variable ($L_0$). This can be measured (or calculated) from the steady state analysis of the process gain, which is defined as:

$$K_p = \frac{\Delta y_{out}}{\Delta L_0}$$

(14)

Therefore, the input–output structure can bridge the gap between steady state process design and control system synthesis. Fig. 5 shows the variation of the controlled variable ($y_{out}$) as function of step changes in the manipulated variable ($L_0$), and the slope of the curve is actually the process gain. The sensitivity of the controlled variable to changes in the manipulated variable is evident and becomes less sensitive at high absorber flow rates where the thermodynamic equilibrium limits the solute removal. It is clear that the process is only linear up to absorber (liquid) flow rate of approximately 1.5 mol/s. Mathematically, the matrix

<table>
<thead>
<tr>
<th>Variable</th>
<th>Description</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>$L_0$</td>
<td>Initial absorber flow rate (mol/s)</td>
<td>2.25</td>
</tr>
<tr>
<td>$G$</td>
<td>Inlet gas flow rate (mol/s)</td>
<td>1.2</td>
</tr>
<tr>
<td>$x_{in}$</td>
<td>Inlet solute concentration in the absorber (mol/mol)</td>
<td>0.0</td>
</tr>
<tr>
<td>$y_{in}$</td>
<td>Inlet solute concentration in the feed (mol/mol)</td>
<td>0.02</td>
</tr>
<tr>
<td>$l_{min}$</td>
<td>Minimum liquid flow rate entering the column (mol/s)</td>
<td>0.5</td>
</tr>
<tr>
<td>$l_{max}$</td>
<td>Maximum liquid flow rate entering the column (mol/s)</td>
<td>8.0</td>
</tr>
<tr>
<td>SlopeOfEquilibriumCurve</td>
<td>Slope of equilibrium line</td>
<td>1.0682</td>
</tr>
<tr>
<td>LiquidMolarDensity</td>
<td>Absorbent molar density (kmol/m³)</td>
<td>55.11</td>
</tr>
<tr>
<td>$T$</td>
<td>Simulation time (s)</td>
<td>2000</td>
</tr>
<tr>
<td>$P$</td>
<td>Temperature (°C)</td>
<td>40.0</td>
</tr>
<tr>
<td>$A_p$</td>
<td>Pressure (atm)</td>
<td>1.0</td>
</tr>
<tr>
<td>$H_s$</td>
<td>Effective sieve plate area (m²)</td>
<td>1.21</td>
</tr>
<tr>
<td>$H_0$</td>
<td>weir height (m)</td>
<td>0.10</td>
</tr>
<tr>
<td>$L_0$</td>
<td>weir length (m)</td>
<td>1.0</td>
</tr>
<tr>
<td>$C_p$</td>
<td>Weir constant (m²² s⁻¹)</td>
<td>1.848</td>
</tr>
<tr>
<td>$M$</td>
<td>Liquid tray holdup (kmol)</td>
<td>Calculated</td>
</tr>
<tr>
<td>$N$</td>
<td>Number of stages</td>
<td>10</td>
</tr>
<tr>
<td>Ethanol</td>
<td>Murphree plate efficiency</td>
<td>0.6</td>
</tr>
<tr>
<td>$Y$</td>
<td>Gas phase composition vector of length (N)</td>
<td>Calculated</td>
</tr>
<tr>
<td>$X$</td>
<td>Gas phase composition vector of length (N)</td>
<td>Calculated</td>
</tr>
</tbody>
</table>

Global variables:
- $L_0$, $T$, $P$, $C_p$, $M$, $N$, $Y$, $X$.
Table 2
Basic description of the MATLAB/SIMULINK software blocks given in Fig. 6.

<table>
<thead>
<tr>
<th>Block name</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>Input data</td>
<td>SimulationDataSimulink: this is a MATLAB file to input the data shown in Table 1</td>
</tr>
<tr>
<td>Steady state solution</td>
<td>SteadyStatesolution.m: MATLAB m-function that implements the steady state solution algorithm (Section 2.2.1)</td>
</tr>
<tr>
<td>Numerical parameters</td>
<td>Built in SIMULINK flowsheet numerical parameters, which can be accessed from the “Simulation” menu</td>
</tr>
<tr>
<td>PID controller parameters</td>
<td>Industrial PID controller tuning parameters: $K_c$, $t_1$, $t_0$, and the realization parameter $0 &lt; \alpha &lt; 1$</td>
</tr>
<tr>
<td>ODE: m-function</td>
<td>GasAbsorptionODE.m: this a MATLAB file where the ODE system (Eq. (13))</td>
</tr>
<tr>
<td>ODE: S-function</td>
<td>GasAbsorp_sfcn.m: this is an s_func.m, which serves as the user defined nonlinear SIMULINK block as can be seen in Fig. 12</td>
</tr>
<tr>
<td>SIMULINK flowsheet</td>
<td>Gas_Absorption_sfcn.m: this is the complete SIMULINK flowsheet and is run by the MATLAB main driver: FDynamicSolutionSimulinkControl.m</td>
</tr>
<tr>
<td>SIMULINK output</td>
<td>Output from SIMULINK flowsheet using floating scopes and “mat” MATLAB files</td>
</tr>
<tr>
<td>MATLAB output</td>
<td>Output using standard MATLAB graphical output by loading the exported data from the SIMULINK flowsheet</td>
</tr>
</tbody>
</table>

$A_s$ in Eq. (12) is function of the absorbent flow rate $(L_0)$ which is not constant.

Now, the main MATLAB programme acts as an interface between the steady state module and the dynamic SIMULINK flowsheet. The general structure of the SIMULINK/MATLAB software is developed to simulate the closed loop response of the gas absorption column. This is shown in Fig. 6. The basic blocks of this software are briefly described in Table 2.

3.1. SIMULINK gas absorption software

The heart of the MATLAB/SIMULINK is the user defined nonlinear SIMULINK block which is used to implement the dynamic solution algorithm given in Section 2.3.

To implement this dynamic model in an S-Function, the dynamic model is written first in an ordinary differential equation (ODE) m-file. The S-Function (GasAbsorp_sfcn.m), which is implemented in the gas absorption dynamic model shown in Fig. 7 is used to call the dynamic model of the gas absorption column (GasAbsorptionODE.m). The SIMULINK block shown in Fig. 7 receives the
input, which consists of disturbances (inlet gas flow rate and composition of both liquid and gas phases) and manipulated (absorbent liquid flow rate) variables. This block sends its output to the column output S-Function as shown by the second block in Fig. 7. The output consists of the liquid phase composition \( x \), the gas phase composition \( y \), sieve tray liquid holdup \( M \) and the top tray gas composition \( y_{out} \). This nonlinear SIMULINK block makes use of the global variables shown in Table 1.

The details of the column output S-Function block are shown in Fig. 8, where the S-Function SIMULINK user defined block is used for algebraic calculation of the equilibrium gas phase composition. This is based on the use of Eq. (6) using the liquid phase composition \( x \). The input to this block (green node 1) consists of the output from the ODE MATLAB solver, which is basically: liquid and gas phase compositions and sieve tray liquid holdup at each instant of time. These are the state variables appearing in the system of Eq. (13).

The measuring device (gas composition analyser [16]) is installed in the recycle loop to measure the gas composition leaving the top tray and is approximated by a first-order linear dynamic model with standard input of 0–1 mole fraction and standard output of 4–20 mA. This SIMULINK block is shown in Fig. 9 with a limiter to keep the input variables within the acceptable physical range.

The expanded SIMULINK block flow diagram of the final control element, which is a pneumatic control valve is shown in Fig. 10.

This is approximated by a first-order linear transfer function having a time constant of 2 s with a limiter to keep the input signal in the standard range input range. The negative and positive biases are used to convert the control valve opening into absorbent flow (mol/s) and to keep the flow rate within the desired physical limits. The expanded SIMULINK feedback controller block diagram, which is composed of an industrial PID controller is shown in Fig. 11. 
An industrial version of the PID controller is used here to realize the controller transfer function that is given by:

$$G_c(s) = K_c \frac{\tau_1 s + 1}{\tau_1 s} \frac{\tau_2 s + 1}{\alpha \tau_2 s + 1}$$  (15)

where $K_c$ is the controller gain, $\tau_1$ and $\tau_2$ are the integral and derivative actions respectively. The parameter $\alpha$ is introduced to realize the controller transfer function and has a value that ranges from 0.05 to 0.2 [18]. These parameters can be adjusted by the user from this SIMULINK subsystem. Note that the PID controller built block in SIMULINK does not support the industrial PID controller given by Eq. (15).

Now, all the major elements in the closed control loop of Fig. 2 are modelled using SIMULINK. The last step in the control system synthesis is the task integration, where all the elements are wired together using the SIMULINK flowsheeting environment. Fig. 12 shows the complete SIMULINK flowsheet of the gas absorption process, with source and sink blocks to facilitate data input and output. The main inputs to the control system are the disturbances in inlet gas flow rate, composition and the set point. These are accomplished by SIMULINK step change blocks, where the step change magnitude, direction (positive or negative) and the time at which the step change begins can be easily set by the user in the SIMULINK flowsheet. These inputs are recorded as a function of time by saving their data to a MATLAB formatted data files. These data are then loaded and plotted using MATLAB. On the other hand, the floating scope SIMULINK block can be used as recorders for the data output at any point along the directed edges as shown in Fig. 12.
4. Numerical results and discussion

The numerical results from the simulation of the bioethanol process are divided into two parts: the steady state analysis and the dynamic analysis for open and closed loops.

4.1. Steady state analysis

All the simulation results from the steady state solution of the tray gas absorption model show that the gas phase composition profiles is a function of tray number when using different Murphree tray efficiency as a parameter. The proposed sequential solution algorithm converges smoothly to the steady state solution as shown in Table 3. By referring to Eq. (10), it is clear that the degree of coupling between the x- and y-phase compositions decreases as $\eta \to 0$ and as $\eta \to 1$. Accordingly, the number of iterations to get the desired accuracy (tolerance = $10^{-12}$) decreases as the two above limits are approached.

Fig. 13 shows the effect of Murphree tray efficiency on the steady state gas composition leaving trays along the gas

![Fig. 13. Steady state gas phase composition profiles as function of tray number with Murphree tray efficiency as a parameter.](image)

![Fig. 14. Dynamic response of the gas phase composition at top, middle and bottom tray along the column height due to positive and negative step changes in the inlet gas flow rate of magnitude $\pm 0.5$ mol/s with tray efficiencies $\eta = 1$ and 0.65.](image)
absorption column. As expected, the solute removal from the gas phase is the highest when the tray efficiency \( \eta = 1 \). It is evident that even though the Murphree tray efficiency is the same for all trays, the shift in the gas composition profiles is nonlinear along the column. The maximum deviation occurs between trays 6 and 8 as shown in Fig. 13. This means that the introduction of the Murphree tray efficiency introduced a clear nonlinearity in the steady state and dynamic model as well. This nonlinearity is evident by referring to Eq. (6) where the term \((1 - \eta)\) is raised to different powers as function of tray number, thanks to the explicit form of Eq. (6).

### Table 3

Number of iterations required to converge to the steady state solution as function of tray efficiency.

<table>
<thead>
<tr>
<th>( \eta )</th>
<th>Number of iterations</th>
<th>( \eta )</th>
<th>Number of iterations</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.0</td>
<td>1</td>
<td>0.5</td>
<td>32</td>
</tr>
<tr>
<td>0.9</td>
<td>37</td>
<td>0.4</td>
<td>28</td>
</tr>
<tr>
<td>0.8</td>
<td>37</td>
<td>0.3</td>
<td>24</td>
</tr>
<tr>
<td>0.7</td>
<td>36</td>
<td>0.2</td>
<td>19</td>
</tr>
<tr>
<td>0.6</td>
<td>36</td>
<td>0.1</td>
<td>13</td>
</tr>
</tbody>
</table>

#### 4.2. Dynamic analysis

It is a well-known fact that the hydrodynamics (flow by convection) is much faster than mass transfer processes taking place by the very slow mixing and diffusion. This fact is usually represented quantitatively by the Peclet number which is in the order of \(10^3 - 10^6\) [27]. The aim of this dynamic analysis is to discover the extent to which the controlled variable (the gas phase composition) is nonlinear. Moreover, the dynamic analysis sheds some light on the dominant time constant for the gas absorption process. This will be very useful in process shutdown, start-up and operation of the gas absorption column or the operation of the column under varying load as in the case of hybrid power plant/gas absorption process [28,29].

#### 4.2.1. Open loop dynamic analysis

The open loop dynamic behaviour of the tray gas absorption column is investigated by studying the open loop response due to step changes in the major disturbances to the column. These disturbances are the total gas flow rate entering the column and its composition. Positive and negative step changes are introduced to test the process linearity. Fig. 14 shows the dynamic response of the
gas phase composition at the bottom, middle and top trays with the Murphree tray efficiency as a parameter due to step changes in the inlet gas flow rate. It is obvious that the tray’s location have a pronounced effect on the column open loop dynamics. The nonlinearity increases from the bottom to the top trays as reflected by the non-symmetric profiles due to the positive and negative step changes in the inlet gas flow rate. This nonlinearity is expected due to the nonlinear dependence of the gas composition on the gas flow rate (Eq. (4)) and the time dependence of the system time constants as discussed in Section 2.3. It seems that there is little dependence of the response nonlinearity on the tray efficiency, which is obvious in the steady state case. On the other hand, the open loop dynamic response due to positive and negative step changes in the inlet gas composition ($y_{in}$) is linear as can be seen from the symmetric gas composition profiles shown in Fig. 15. This is again due to the linear dependence of the gas composition on the inlet gas composition. Again, the tray efficiency has no obvious effect on the dynamic response of the gas phase composition due to step changes in $y_{in}$. In terms of sensitivity, it is clear that the gas phase composition is more sensitive to disturbances in the inlet gas flow rate than that in the inlet gas composition. This is clear by comparing Figs. 14 and 15. The significant liquid holdup on each tray presents large time constants, which results in slow composition response. To compare the dynamic behaviour of the liquid holdup on the different trays and the gas phase composition response, a step change is introduced in the manipulated variable (absorbent flow rate) and the outputs are recorded. Fig. 16 compares the responses of the gas phase composition and the liquid holdup due to positive step change in the absorbent flow rate. As expected, the liquid holdup responds very fast when compared to the gas phase composition due to the fast (convection) and slow changes in the tray solute concentrations because of the significant liquid holdup on each tray. Therefore, the liquid holdup responds faster than the liquid composition on each tray. This means that the largest time constant lies in the liquid phase composition. This results in two sets of time constants: large time constants corresponding to the liquid composition and small ones corresponding to the liquid holdup itself. These properties, which are essential for process start-up, shutdown and control, could not be explored without this open loop dynamic analysis. However, this difference in time scales presents numerical difficulties in solving the differential algebraic equations (DAE) comprising this model. Actually, a special stiff MATLAB solver (ode15s) is chosen to overcome this numerical difficulty. From numerical point of view, this presents a drawback of dynamic simulation where very small time steps are usually required with significant increase in the CPU time.

Moreover, Fig. 16 shows that the delay time increases as the output variable gets further from the disturbance point. This is clear by examining the liquid holdup response on each tray, where the top tray responds almost immediately because it is very close to the
absorbent inlet. This is one of the main characteristics of stagewise (or distributed) systems [1,18].

4.2.2. Closed loop dynamic analysis

The closed loop dynamic behaviour is investigated in the presence of a PID controller. The major difficulty lies in selecting (adjusting) the best controller parameters to produce a specified closed loop response. This process is known in the control literature as tuning feedback controllers. The quarter–decay ratio or the continuous cycling method is used to estimate the best controller parameters: namely, the proportional gain, the integral and the derivative time constants. The continuous cycling method is used here by switching off the integral and derivative control actions and adjusting the value of the proportional gain until the controlled variable cycles continuously [18]. The value of the proportional controller gain at which this continuous cycling occurs is called the ultimate controller gain: $K_u$. By measuring the ultimate oscillation ($P_u$) the best controller parameters are found.

Accordingly, Fig. 17 depicts the continuous oscillation of the controlled variable (top tray gas phase composition), where $K_u = 9000$ and $P_u = 17$ s. So, the best controller parameters using the continuous cycling method are calculated using the values of $K_u$ and $P_u$, where these values are shown in Table 4.

First, the dynamic behaviour of the liquid holdup on each tray under the optimum PID control action is shown in Fig. 18. Since positive step change is introduced in the inlet gas composition as seen in Fig. 19, the controller increases the flow of the absorbent to absorb the extra quantity of solute entering with the gas feed. As a result, the liquid holdup increases in an oscillatory behaviour characterizing the overdamped pre–specified response. The settling time of the liquid holdup is in the same order of magnitude for the controlled variable (top tray gas composition). This fact is clear by comparing Figs. 18 and 19, where the liquid holdup and the controlled variable settling times are around 80 s. The overshoot in the top tray gas composition is relatively small and the damping is very fast as shown in Fig. 18. So, the PID controller performance is satisfactory in cancelling the inlet gas composition disturbance.

On the other hand, the closed loop response due to positive step change in the inlet gas flow rate is characterized by large overshoots around the set point as can be seen in Fig. 20. This is due to the high sensitivity of the gas phase composition to the inlet gas flow rate.

<table>
<thead>
<tr>
<th>Feedback controller</th>
<th>$K_u$</th>
<th>$r_1$ (s)</th>
<th>$r_2$ (s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>P</td>
<td>4500</td>
<td></td>
<td></td>
</tr>
<tr>
<td>PI</td>
<td>4090</td>
<td>14.2</td>
<td></td>
</tr>
<tr>
<td>PID</td>
<td>5294</td>
<td>8.5</td>
<td>2.13</td>
</tr>
</tbody>
</table>

Fig. 17. The ultimate controller gain ($K_u = 9000$) and period of oscillation ($P_u = 17$ s) using the continuous cycling method applied to the closed-loop tray gas absorption column.

Fig. 18. Closed-loop response of the liquid holdup using the optimum PID controller settings given in Table 4.

Fig. 19. Controlled system performance due step change in the inlet gas composition using PID controller with optimum settings estimated using the continuous cycling method (Table 4).
Fig. 20. Controlled system performance due step change in the inlet gas flow rate using PID controller with optimum settings estimated using the continuous cycling method (Table 4).

Fig. 21. Controlled system performance due to positive and negative step changes in the set point using PID controller with optimum settings estimated using the continuous cycling method (Table 4).

Fig. 22. Controlled system performance due simultaneous positive step changes in the inlet gas flow rate and composition using PID controller with optimum settings estimated using the continuous cycling method (Table 4).

However, the settling time is the same as in the case of step change in the inlet gas composition, thanks to the fast response of the trays' liquid holdup. Fig. 21 shows the PID controller performance due to positive and negative step changes in the set point (servo-mechanism). In both cases the controlled variable follows the set point with 60 s as a settling time. The controlled process seems to be approximately linear due to step changes in the set point. This is clear from the image of the positive step response as shown in Fig. 21.

Finally, the optimum PID controller is tested by introducing simultaneous step changes in the two major disturbances (inlet gas composition and flow rate) as in Fig. 22. Again, the dominant effect of the inlet gas flow rate is clear due to the relatively large overshoot in the controlled variable. On the other hand, the controller performs well in terms of settling time under the influence of the two disturbances.

5. Summary and conclusions

In this work, mathematical modelling and design of a MATLAB/SIMULINK flowsheeting software were developed and implemented to simulate and control a sieve tray gas absorption column. The developed MATLAB/SIMULINK software is found to be an effective tool to simulate and control nonlinear dynamic models not offered by the built-in SIMULINK library. Further work is under progress to build a graphical user interface (GUI) to enter the basic simulations inputs for the developed software which is a combination of the MATLAB code and SIMULINK flowsheet programme.

From the detailed steady state and dynamic analysis the following conclusions can be drawn:

1) The steady state mathematical model was found to be nonlinear due to the introduction of the Murphree tray efficiency. Also, due to the linear difference equation relating the actual and equilibrium gas phase compositions, a closed analytical form is derived to relate the actual gas phase composition to the liquid phase one with Murphree tray efficiency as a parameter. Furthermore, to avoid the solution of large linear system, a sequential steady state solution algorithm is developed and tested.

2) By using steady state sensitivity analysis and the number of degrees of freedom, the outlet gas phase composition was found to be a suitable candidate for controlling the outlet gas composition. This is supported by the degrees of freedom analysis, which showed that only two variables can be manipulated: absorbent flow rate and its composition.
3) The dynamic model was characterized by two sets of time constants: the smallest time was associated with tray hydodynamics, while the larger one was related to the change in the gas phase composition. Accordingly, special MATLAB ODE solver was used (ode15s) to integrate the open and closed loop models.

4) The open-loop dynamic analysis showed that the gas phase composition response is nonlinear with respect to the inlet gas flow rate, while it was linear with respect to inlet gas composition. The nonlinearity increases along the column height and it was found to be maximum at the top tray. On the other hand, the Murphree tray efficiency had little effect on the dynamic behaviour of the column.

5) The closed-loop dynamic analysis showed that the controlled variable (outlet gas phase composition) had a fairly linear dynamics due to step changes in the set point.

6) The continuous cycling method was used to estimate the PID controller parameters for the control of the bioethanol gas absorption column. The controlled variable was found to exhibit fairly large overshoots due to step change in the inlet gas flow rate, while the PID controller performance was satisfactory for step change in the inlet gas composition.

References


