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Advance Process Control of Binary Distillation Column

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Abstract: *Distillation columns are important unit operations in chemical process plants. This paper reviews advance control techniques of modeling and simulation of distillation columns. The aim of this paper is the advance control design of the distillation column for separation of binary mixture. Advance control strategy is effective for systems that have large time constants and disturbances and advance control strategy is fit for a system that lacks an accurate model. It aims at providing simple recommendations to assist the engineer in designing control systems for distillation columns. The standard LV-configuration for level control combined with a fast temperature loop is recommended for most columns. The response to change in feed composition has larger gain than the response to change in feed flow rate. The paper also compares the simulation results with some of other works.*

Keywords: *Binary Distillation Column, Fundamental Modeling, Simulation.*

1. INTRODUCTION

A Distributed Control System (DCS) is an enabling technology that provides a platform for further improvement to realize greater benefits from the Unit. It also serves to facilitate further process and operational improvement that is otherwise limited by the nonavailability of technology that is mature and easy to use.

With Advanced Process Control, the DCS is able to push the process to a higher level of productivity, and is able to deliver bottom line improvement. Currently gaining popularity in the process industry is Multi-Variable Model Predictive Control Technology. Experts unanimously agree that APC is able to deliver sustainable measurable benefits simply by stabilizing the plant to yield consistent quality products. The logic for this is that control strategies should be developed with an understanding of the process and its nuances; the control system on which the APC would sit; the need for integration with wider plant objectives, as well as a knowledge of base layer control loops. It is also important to consistently translate economic objectives into operating objectives. APC allows for companies to operate its facilities with greater safety, cost effectiveness, reliability, and compliance with environmental factors. When jointly used with other unit-operation optimization technologies, APC can prove to be extremely beneficial. Optimization is not a one-time event, it has to be a continuous effort to enhance operating performance in ever-changing conditions. The attraction of adopting a model based approach to controller development is illustrated in the block diagram shown in Figure 1.

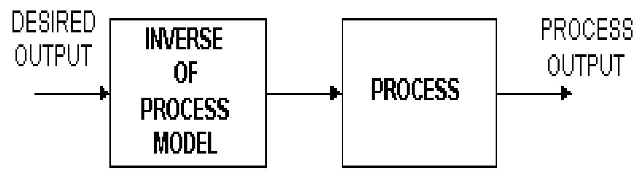

Fig. 1: Ideal Model Based Control

By implementing advanced control, benefits ranging from 2% to 6% of operating costs have been quoted [Anderson, 1992].These benefits are clearly enormous and are achieved by reducing process variability, hence allowing plants to be operated to their designed capacity. Advanced control comes into play from the level of basic control through that of process optimization. Instead of having the operators manually adjust control units for specific variables, advanced systems provide generalized models that automate regulatory and constraint control as well as process optimization. In regulatory control single loop feedback improvements such as feed forward, cascade control can be used to supplement PID algorithms. Time delay compensation techniques can also be applied to compensate for long delays permitting tighter control. At the level of constraint control, multivariable techniques can be used.

Advanced control technology is therefore a combination of:

- Advanced hardware (on line sensors, pneumatic or electronic analog with digital systems, computer hardware and digital control units)
- Advanced control algorithms at the regulatory, constrained and optimization levels.

In order to control a distillation column, the first step is to develop a model of the column. By modeling we can understand the behaviour of the column, predict future reaction and therefore devise a control structure for the column. This paper reviews some important techniques in distillation column modeling and then describes a model for a lab-scale binary continuous distillation column.

The paper is organized as follows. A review of advance process control in part I, modeling of continuous distillations is presented in part II, the modeling of our distillation column is detailed in part III, the simulation results are shown in part IV, and the conclusion and future work are presented in part V of this paper.

2. THE DISTILLATION COLUMN

Distillation column is considered one of the most important unit operations in chemical engineering, and also the most studied unit operations in terms of control. A distillation column is used to separate a mixture into its components by the application and removal of heat. It consumes a huge amount of energy in both heating and cooling operations. There are many types of distillation columns based on different classifications such as: batch, continuous, binary, multiproduct, tray, packed. In this paper we focus on continuous binary distillation columns since continuous columns are dominant in industry and binary columns are usually referred to as a foundation by the researchers when they examine other types of distillation columns.

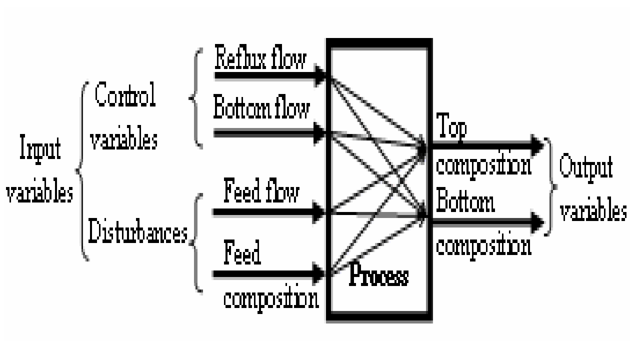

Fig. 2. Distillation process input and output variables.

A simple two-product continuous distillation column is shown in Fig .3 .The column has N stages on which the vapour-liquid equilibriums occur. The feed enters the column on the stage N_F. This stage divides the column into a rectifying section and a stripping section. Near the bottom of the column is a reboiler which provides energy to the column. The mixture is heated to form a flow of vapour rising up inside the column. In the stripping section, the less volatile component is enriched while in the rectifying section the more volatile component is enriched. The top product is condensed by the condenser from which there is a reflux flow back to the top of the column to enhance the purity of the product.

The difficulties in controlling distillation columns lie in their highly nonlinear characteristics; their multiple inputs multiple outputs (MIMO) structure and the presence of severe disturbances during operation. The nonlinearity of distillation columns is well known. It has been shown that the purer the products get, the more nonlinear the system becomes. A distillation column is also a typical example for an MIMO system in which there are strong interactions between the variables. The interactions occurring between the inputs and the outputs are difficult to identify. The disturbances to a distillation column can come from many sources. They can come from the feed (feed flow rate, feed composition), from the pressure inside the column, from the cooling water etc. These difficulties pose numerous challenging control problems and also attract a large number of researchers from different disciplines.

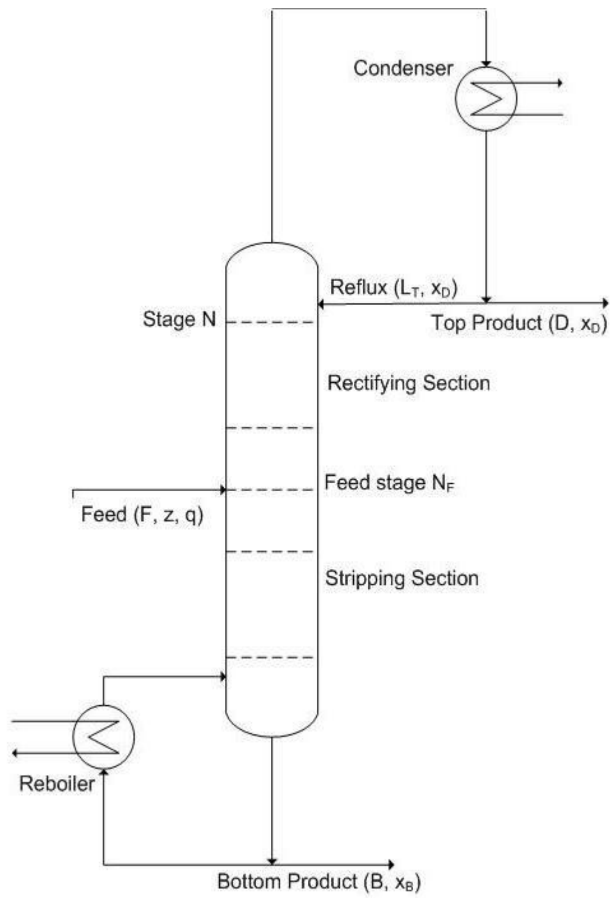

Fig. 3: A Two-product Distillation Column

3. MODELING OF CONTINUOUS DISTILLATION COLUMNS

Modeling of distillation columns is often classified into three groups: fundamental modeling, empirical modeling and hybrid modeling. In fundamental modeling, the model is constructed based on the physical properties of the system, such as the preservation of mass, energy and momentum. And depending on the levels of accuracy of the assumptions we will have different models ranging from simple to rigorous models. This method of modeling has the advantage of global validity, accuracy and it gives more complete process understanding. However, this method is quite complex for controller design with huge amount of computation and simplifications are often needed. The empirical modeling (sometimes called the black-box modeling) utilizes the input and output data from the operation of the column to build the relationship between the input and the output. With this method we do not need to understand the inner dynamics of the column, and the computation can be reduced. But in using this method we have to carry out experiments on the real column, and the results may not be applied for other column, even the results from one column can be different if the column's conditions are different between the experiment and the actual operation of the column. The hybrid modeling (or the 'grey-box' model) combines the fundamental modeling and the empirical modeling. This method utilizes the advantages of the other two, but in order to do that we need a well-structured model in which we have to decide which part of the model to use fundamental technique and which part to use empirical data. In our project we will focus mainly on the fundamental modeling, even though the empirical model is dominantly used in the industry. The reason is that we want to understand the dynamics of the distillation columns, and since the black-box model "may not be used to predict the behaviour of the system at other operating conditions".

The paper summarized the simplifications of the rigorous model since no references had been found on solving all the equations of the rigorous model. The simplifications are aimed to the vapour dynamics, to the energy balance and to the liquid flow dynamics. The paper recommends not neglecting liquid dynamics (i.e. not assuming constant liquid holdups) due to the fact that the initial response, an important factor in feedback control, is largely affected by the liquid holdups.

Abdulla *et al*. have done a quite complete review on the recent nonlinear modeling applications in continuous distillation column. The summary states that the empirical modeling has been preferred in industry because of its simplicity compared to the fundamental model; and the current development focuses on hybrid models, which can exploit the advantages of both fundamental model and empirical model; and that the neural network method is used the most to combine with the fundamental model in empirical modeling.In the case of fundamental modeling, the model is often simulated to understand the column's dynamic behaviour. The development of distillation column's simulation has been going along with the growth of computing capacity. As of 1930s and 1940s only graphical methods and simple short-cut models were used to get insights of the steady-state behaviour of the distillation columns. The fast-growing of computing power has allowed the use of more complex and rigorous models. Computer programming and the numerical methods to solve the differential equations play an important role.

4. MODELING OF THE APC

The APC (Advanced Process Control) column is a pilot distillation column that has 15 trays and equipped with a DCS control system. The feed is positioned at tray 7. The model is developed based on a model. In the model the following assumptions are made:

1. Binary mixture, the feed contains only two components
2. The pressure inside the column is fixed by controlling the cooling water
3. Constant relative volatility, $\alpha = 1.5$
4. Constant molar flows
5. No vapour holdup, the vapour holdup on each tray is negligible
6. Linear liquid dynamics
7. Equilibrium on all stages
8. Total condenser, there is no vapour holdup in the condenser

The total material balance equation on stage i is:

$$\frac{d}{dt} M_i = L_{i+1} - L_i + V_{i-1} - V_i \qquad (1)$$

where M_i is the liquid holdup on tray i, L_i and V_i are the liquid flow rate and vapor flow rate that come towards tray i.

The material balance for the light component on tray i is:

$$\frac{d}{dt}(M_i x_i) = L_{i+1}x_{i+1} - L_i x_i + V_{i-1} y_{i-1} - V_i y_i \qquad (2)$$

where x_i and y_i is the composition of the light component and heavy component on tray i respectively.

At the feed stage (NF=7) we have:

$$\frac{d}{dt} M_{NF} = L_{NF+1} - L_{NF} + V_{NF-1} - V_{NF} + F \qquad (3)$$

and

$$\frac{d}{dt}(M_{NF} x_{NF}) = L_{NF+1}x_{NF+1} - L_{NF}x_{NF} + V_{NF-1}y_{NF-1} - \qquad (4)$$

$- V_{NF}\, y_{NF-1} + F z_F$

in which F is the feed flow rate and z_F is the concentration of the light component in the feed.

The reboiler is also an equilibrium stage with i=1:

$$\frac{d}{dx} M_1 = L_2 - V_1 - B \quad (5)$$

$$\frac{d}{dx} M_1 x_1 = L_2 x_2 - V_1 y_1 - B x_1 \quad (6)$$

At the condenser we have $i = NT = 16$ and

$$\frac{d}{dx} M_{NT} = V_{NT-1} - L_{NT} - D \quad (7)$$

$\frac{d}{dx} M_{NT}\, x_{NT} = V_{NT-1}\, y_{NT-1} - L_{NT}\, x_{NT} - D x_{NT}$ (8) Where B is the bottom product flow rate and D is the distillate product flow rate. The composition of the heavy component is related to the composition of the light component via the relative volatility formula:

$$y_i = \frac{\alpha x_i}{1+(\alpha-1)x_i} \quad (9)$$

The liquid flow dynamics is considered due to its important effect on the initial response of the column. The formulas for the liquid holdup are:

$$L_i = L_{ob} + \frac{M_i - M_{oi}}{\tau} + (V_{i-1} - V_o)\,\lambda \quad (10)$$

$$L_i = L_o + \frac{M_i - M_{oi}}{\tau} + (V_{i-1} - V_{ot})\,\lambda \quad (11)$$

For i from NF+1 to NT-1.
where L_0 is the nominal reflux flow and M_{0i} is the nominal reboiler holdup (kmol) on stage i. These values are achieved after we do steady state simulation (see Table 1). τ is the time constant for liquid dynamics, in this model it is chosen to be 0.063 (min), and λ represents the effect of vapor flow on liquid flow. In the simulation we ignore this effect by setting $\lambda = 0$. L_{0b} is the nominal liquid flow below feed, given by the formula:

$$L_{ob} = L_o + q_{FO} F_o \quad (12)$$

in which F_0 = 1(kmol/min) is the nominal feed rate, $q_{F0} = 1$ is the nominal fraction of liquid in the feed.

5. SIMULATION RESULTS AND DISCUSSION

Because the distillation process is a nonlinear one it is wise to use a model based control structure, which can take account of the process nonlinearities by changing the process model according to the operating point. The process is a nonlinear one, but represented as a reunion of linear models one for each channel and operating point. At each simulation initialization step the bottom and top composition, the feed flow and composition are sent from the process to the control structure, in MATLAB. Here, using these data the model parameters (time constants and gain) are determined and loaded in controller. Using these model parameters, the two controlled variables (reflux flow and bottom product flow) are computed and sent at each sampling instant to the process.

The dynamic system behaviour analysis consisted of modifying the compositions set point, the disturbances and the controllers tuning parameters. (figures 4 to 7)

For top composition advance controller has the following default simulation parameters:

- prediction horizon – is variable, and it is calculated using the process model time constant T1.
- control variable time horizon – 30 sampling times;
- output weight (minimum value: 0, maximum value: 1);
- control variable weight – 0.2 (minimum value: 0, maximum value: 1).

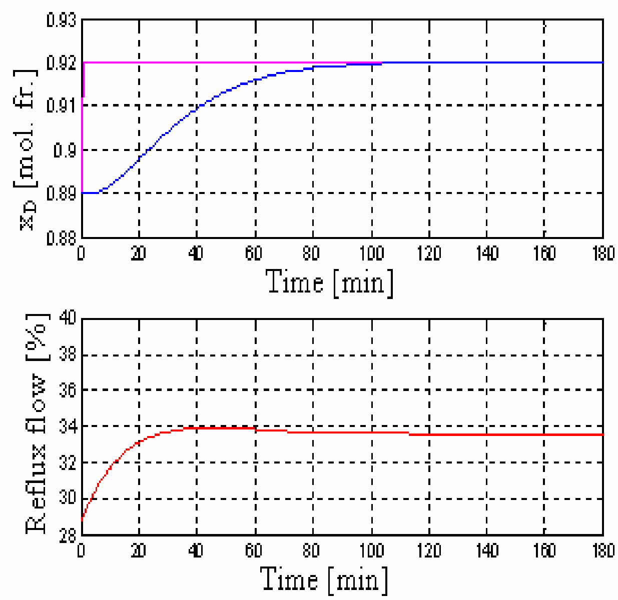

Fig. 4: Top composition trend when the controller setpoint increases from 0.89 mol. fr. to 0.92 mol. fr.

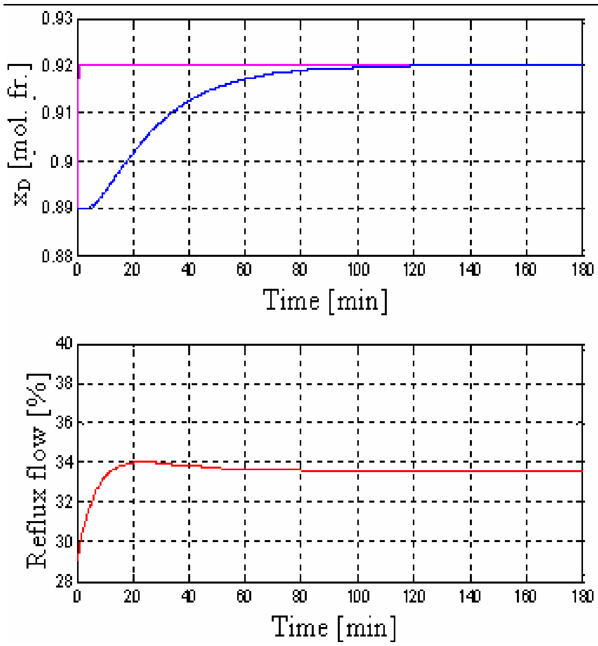

Fig.5: Top composition trend when the controller setpoint increases from 0.89 mol. fr. to 0.92, control variable time horizon is 25.

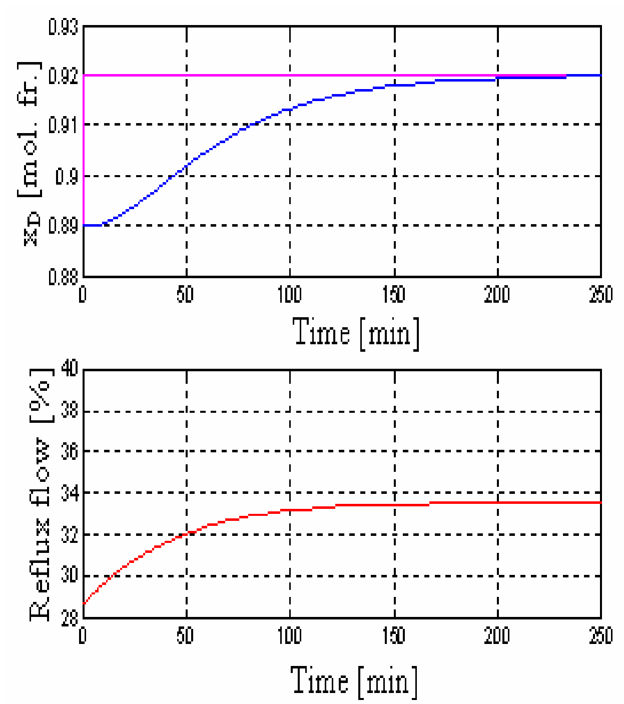

Fig.6: Top composition trend when the controller setpoint increases from 0.89 mol. fr. to 0.92, control variable weight is 0.7.

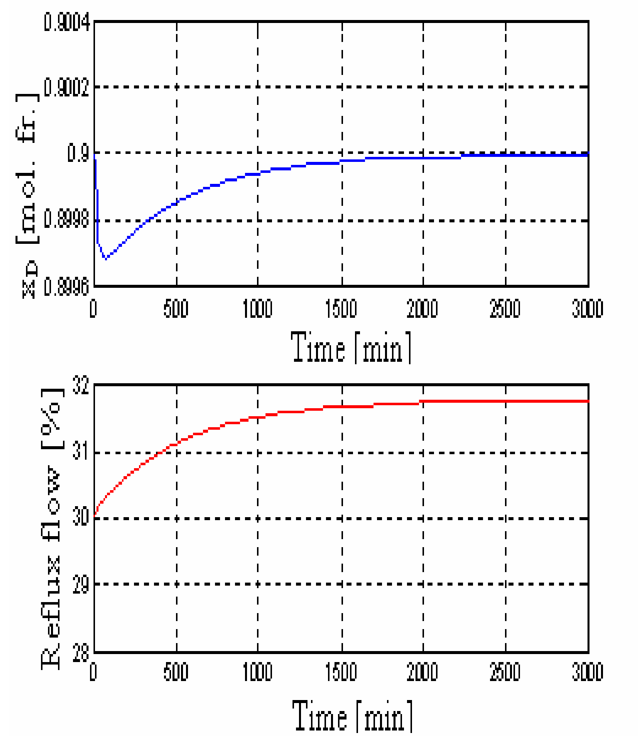

Fig.7: Top composition trend when the feed flow increases from 241.5 kmol/h to 246.5 kmol/h.

6. CONCLUSION AND FUTURE DIRECTIONS

This paper presented the advance process control design of a binary distillation column under disturbance. Control of the top and bottom compositions of the column is a difficult task due to presence of the control loop interactions and nonlinearities. The structure allows taking into account dynamic variations of the process and adapting the controller parameters to this various conditions. Advance controllers achieved a accurate performance in controlling the top and bottom compositions and also in controlling the feed rate, top and bottom rates. As can be seen from the presented trends, the behaviour of the process with the control system was studied for different values of the tuning parameters, observing that a decreasing of the control variable horizon or a decreasing of the control variable weight, from the default values, can lead to an increasing of the transient time. Also we can observe that the process output value reaches the set point value with the best dynamic performances for the default values of the tuning parameters. The control system has a robust behaviour when a disturbance appears in the process.

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