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Control of Complex Distillation Configurations

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Abstract

The dynamics and control of two complex column configurations (sidestream column with stripper; prefractionater/sidestream column configuration), which are multivariable, interacting and nonlinear, have been studied. A new control scheme developed by Han and Park (1993) to deal with the nonlinear and multivariable nature of distillation processes has been applied to these complex distillation configurations. The control scheme incorporates a nonlinear wave model into a generic model control framework. An observer based on the nonlinear wave model is used to determine the profile positions of distillation column sections. The control scheme enables tight control of the profile position of each column section that leads to fast stabilization of product compositions.

Introduction

Complex distillation configurations have been actively studied recently because they require much less energy than conventional designs. Despite their advantages, they have not often been used in real plants. It can be mainly attributed to control problems. The systems are characterized by multiple feeds and multiple products and are highly interactive and multivariable so that they are very difficult to operate and control. Therefore, the control scheme to overcome the difficulty has to be made for the applications of the energy conserving technology.

Steady-state design of complex configurations for ternary separation has been studied by Doukas and Luyben (1978), Tedder and Rudd (1978), Alatiqi and Luyben (1985) and Glinos and Malone (1985). Alatiqi and Luyben (1985) showed that for separating ternary mixtures that contain small amounts (less than 20 %) of the intermediate component in the feed, the complex sidestream column/stripper configuration was more energy efficient than other configurations. Glinos and Malone (1985) also proved that the sidestream columns require less total vapor than the conventional configuration does. In particular, they reported that the vapor savings are always large when the intermediate concentration is small. The maximum savings possible is 50 %, independent of the volatilities. Doukas and Luyben (1981) also reported that when there is a large amount of the intermediate component in the feed, the prefractionater/sidestream column configuration is attractive from an economic standpoint.

Both the complex sidestream/stripper configuration and the prefractionater/sidestream column present a challenging 4x4 multivariable control problem. Until now, very few studies have been done on the control of complex distillation configurations. The dynamics and control of the complex configurations have been studied by Doukas and Luyben (1981), Alatiqi and Luyben (1986) and Ding and Luyben (1990). Alatiqi and Luyben (1986) controlled the complex sidestream column/stripper configuration with SISO PI controllers and gave dynamic comparison with other conventional distillation configurations. Doukas and Luyben (1981) studied the dynamics and control of prefractionater and sidestream column. Ding and Luyben (1990) presented control of a complex heat integrated distillation system by using conventional diagonal SISO controllers. But, there has been no study applying a multivariable, nonlinear controller to complex distillation configurations.

Han and Park (1993) have developed a new control scheme using a nonlinear wave theory and applied it to simple high-purity binary distillation columns successfully. The purpose of this paper is to apply the control system developed by Han and Park (1993) to the complex sidestream/stripper configuration and prefractionater/sidestream column configuration. Significant improvement in the control performance over the conventional decentralized control has been achieved by applying the proposed control scheme to the configurations.

Profile Position Control of Distillation Column section

The complex distillation configurations are very nonlinear, interacting and multivariable. The nonlinear behavior is mainly due to the tendency of the column composition profile to move up or down the column ends as a result of feed disturbances and of control actions. The dynamic behavior of a distillation column is characterized by the propagation of concentration or temperature profile in the column sections. Figures 1 and 2 show the propagation of column composition profile after a step change of heat input to the reboiler of the main column of sidestream/stripper configuration and to the reboiler of the sidestream column of prefractionater/sidestream configuration, respectively. During propagation of the profile, the profile takes a nearly constant shape along a spatial coordinate, so the profile can be viewed as a "wave". If we control the profile of each column section of the distillation configuration preventing it from moving up or down the column ends, we can obtain fast stabilization of product purity. The profile position can be regarded as the location of the standing wave. It is more convenient to represent the location of a constant-pattern wave with a single point corresponding to a representative concentration. The control of the profile position can regulate the propagation of column profile to prevent the column from dropping into a nonlinear region and giving much errors in product purity.

Profile Position Controller Design

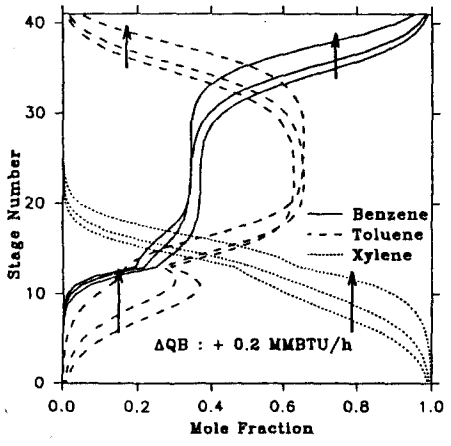

Fig. 1. Dynamic profiles of the main column after a step change in the heat input (sidestream/sidestripper configuration).

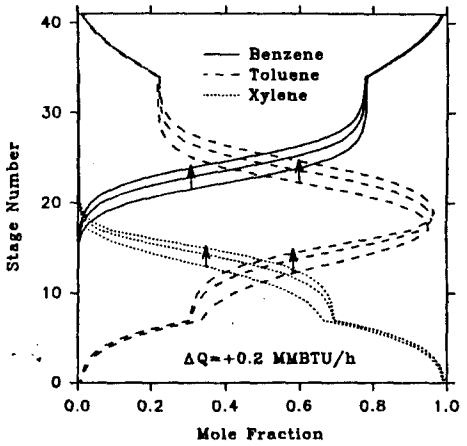

Fig. 2. Dynamic profiles of the sidestream column after a step change in the heat input (prefractionater/sidestream configuration).

Generic Model Control (GMC) introduced by Lee and Sullivan (1988) has several advantages to make it a good framework for the profile position control. The overall structure of GMC allows the incorporation of nonlinear process models directly into the control algorithm. Also, GMC makes feedback control of the rate of change of the controlled variable possible so that the rate of the profile propagation can be regulated.

The formulation of a GMC controller for profile position control of distillation column sections incorporates the nonlinear wave model equation proposed by Hwang (1991). Choosing the profile position as a state vector, the control equation can be written as follows:

$$dS/dt = K_1(S^* - S) + K_2\int_0^t (S^* - S)dt \qquad (1)$$

where S and S* are the profile position and its setpoint, respectively; dS/dt is a propagation rate of the profile; K_1 and K_2 are tuning constants. S is expressed in terms of the normalized distance from the bottom of the column (S=0 at bottom; S=1 at the top). The propagation rate can be expressed from the nonlinear wave model.

$$dS/dt = u_n = \frac{V}{F}\frac{\Delta y/\Delta x - L/V}{1 + r(\Delta y/\Delta x)} \qquad (2)$$

Combining these two equations gives the following equation for control of the profile position:

$$\frac{V}{F}\frac{\Delta y/\Delta x - L/V}{1 + r(\Delta y/\Delta x)} - K_1(S^* - S) - K_2\int_0^t (S^* - S)dt = 0 \qquad (3)$$

On-line Estimation of the Profile Position

For the profile position control, the estimation of the profile position is necessary since the profile position cannot be measured directly. The profile position of each column section can be estimated by the nonlinear profile position observer which is described by Han and Park (1993). They used the nonlinear wave model by Hwang (1991) with the feedback of temperature or concentration measurements for the profile position observer. Concentrations are measured at different points of the column and the error weighted by spatially distributed function between the measured concentration and the estimated concentration is used for the correction of the estimated profile position. The concentrations can be replaced by the temperatures since there exists one-to-one correspondence between temperatures and compositions, given by the thermodynamic equilibrium relationship in the binary distillation region.

In these complex distillation configurations, the column sections directly related to product purities are nearly binary so that the temperature of the tray can be used to estimate the composition of the tray.

The full description of the estimator is

$$\dot{S} = dS/dt = \frac{V}{F}\frac{\Delta y/\Delta x - L/V}{1 + r(\Delta y/\Delta x)} + \sum K_i(x_i - \hat{x}_i) \qquad (4)$$

$$\Delta y/\Delta x = \frac{\dot{S} + \frac{L}{F}}{\frac{V}{F} - r\dot{S}} \qquad (5)$$

$$\hat{x}_i = K_2(S_i - S) + x_s \qquad (6)$$

$$K_i = K_o \exp[-b(x_i - x_s)^2] \qquad (7)$$

where i is the measurement tray number and ^ indicates the estimated value. The details about the estimation of the profile position are given by Han and Park (1993).

Composition/Profile Position Cascade Control

The profile position controller (GMC) is used for instantaneous control to prevent the propagation of disturbances to column ends. However, the offset from the setpoint in the product composition may occur when the profile position of the column section is fixed at the specific point. So, the composition controller which is a simple PI controller is employed and cascaded as a primary controller to the profile position controller to remove the offset.

Nonlinear Dynamic Simulation

The dynamic simulation of the complex distilatillation configurations is based on rigorous tray-to-tray calculations by making the assumption of negligible pressure drop in the column and ideal vapor liquid equilibrium, but by taking into account the effects of nonequal molar overflow. The tray efficiency is assumed to be 100 % and the liquid tray hydraulics is taken from linearized Francis weir formula. Composition analyzer dead time is assumed to be 3 minutes for the composition control loops.

Control of Sidestream/Sidestripper configuration

Description of the System

The ternary system benzene/toluene/o-xylene has been chosen for testing the performance of the proposed controller. The complex sidestream stripper configuration (SSS) which Alatiqi and Luyben (1986) used has been chosen. They reported that the configuration consumes 30 % less energy than the conventional two-column system. Figure 3 shows the schematic diagram of the complex distillation configuration. The sidestream stripper configuration uses a sidestream column with the liquid sidedraw fed into a small stripping column. The stripper has a small reboiler that removes some of the lightest component from the sidestream product. Vapor from the stripper is fed back into the main column. The product purities of all components have been set to 99 mol %. The column configuration has high purity specification of product compared to the column of Alatiqi and Luyben (1986). Hence, the nonlinearity and interaction of the column configuration are severer than that of Alatiqi and Luyben (1986) since nonlinearity and internction increase dramatically as product purity increases. The details of the column are listed in Table 1A.

The column configuration consists of four column sections so that any specified recoveries and purities can be obtained. The process configuration has three compositions to be controlled and four manipulated variables. The manipulated variables are heat input to the main column (QB), heat input to the side column (QBS), reflux rate of the main column (R) and liquid sidedraw rate (LS). The amount of liquid sidedraw rate fed to the stripper is closely related to energy consumption. The higher this rate, the less the total energy consumption. However, there is a limiting value of LS beyond which the purity of toluene product from the stripper base can no longer be attained (Alatiqi and Luyben, 1986). The increase of LS makes the column profile move up, resulting the increase of the heavy component (xylene) around the sidedraw tray. Any xylene that enters the stripper leaves in the toluene product.

The sidedraw rate LS can be held constant and the other manipulated variables can be used to control the three product purities. However LS manipulation is necessary to maintain toluene product purity and to minimize energy consumption. It is because the sidedraw rate can be used to control the profile position in the middle section of the main column, preventing the column profile from moving up or down to column ends. The propagation of the column profile will produce a large deviation from its setpoint in all product purities. So, the tight control of the profile position in the middle section of the main column guarantees the stabilization of product purities and dynamic controllability. Also, the control produces a similar effect to dual composition control, leading to minimizing energy consumption.

Profile Position Controller Design

For sections I, III and IV, the derived equation for the profile position control can be written as follows:

For column section I,

$$\frac{V_1}{F}\frac{\Delta y/\Delta x - L_1/V_1}{1+r(\Delta y/\Delta x)} - K_{11}(S_1^* - S_1) - K_{12}$$

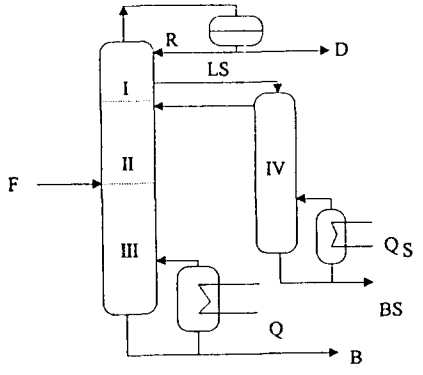

Fig. 3. Sidestream/sidestripper configuration

$$\int_0^t (S_1^* - S_1)dt' = 0 \qquad (8)$$

For column section III,

$$\frac{V_3}{F}\frac{\Delta y/\Delta x - L_3/V_3}{1+r(\Delta y/\Delta x)} - K_{31}(S_3^* - S_3) - K_{32}$$

$$\int_0^t (S_3^* - S_3)dt' = 0 \qquad (9)$$

For column section IV,

$$\frac{V_S}{F}\frac{\Delta y/\Delta x - L_S/V_S}{1+r(\Delta y/\Delta x)} - K_{41}(S_4^* - S_4) - K_{42}$$

$$\int_0^t (S_4^* - S_4)dt' = 0 \qquad (10)$$

$$V_3 = V_1 - V_S - (1-q)F \qquad (11)$$

$$L_3 = L_1 - L_S - qF \qquad (12)$$

where L and V represent the liquid and vapor flow rates respectively in that section. The profile position and the slope of the equilibrium curve at the representative concentration can be estimated from the profile position, respectively. Liquid and vapor flow rates can be calculated. For column section II, application of the nonlinear wave model is difficult since in the section the system is not binary anymore. Hence, simple PI controller is employed for the control of the profile position in the section. Also, the estimation of the profile position in the section is made without using the nonlinear wave model.

On-line estimation of the profile position

For the SSS configuration, almost no heavy component will appear at the top of the main column and at the bottom of the side column and the same thing is true at the bottom where the light component composition is nearly zero. From these, the proposed estimator can be used to these sections. The column section in the

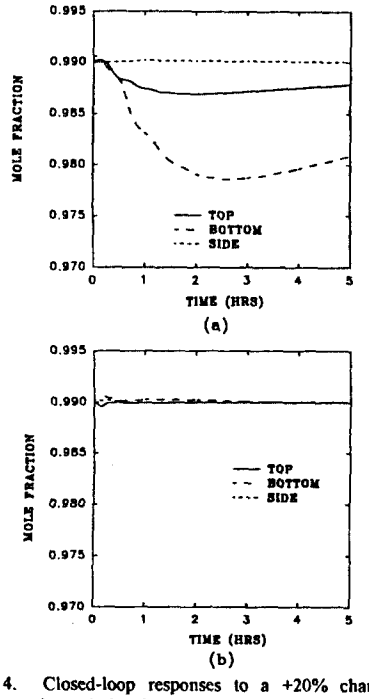

Fig. 4. Closed-loop responses to a +20% change in toluene concentration of the feed (SSS configuration): (a) conventional control scheme with SISO controllers
(b) proposed control scheme.

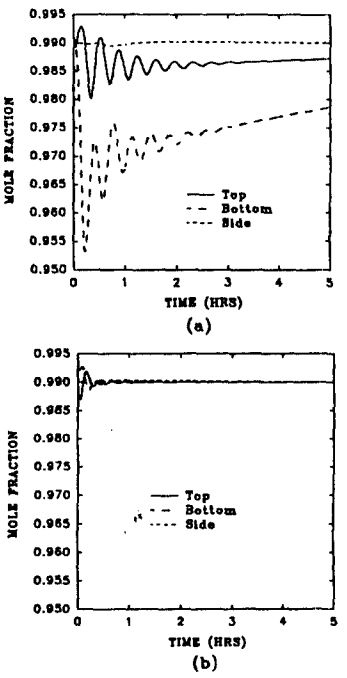

Fig. 5. Closed-loop responses to a +20 % change in feed flow rate (SSS configuration): (a) conventional control scheme with SISO controllers (b) proposed control scheme.

middle part of the main column for the SSS configuration is a multicomponent system which prevents the application of the nonlinear wave model for the profile position observer. In this section, we have chosen the front of temperature profile as the profile position. The temperature front is the position at which the temperature gradient is the largest in a column section. The position can be determined by the following method. By measuring each stage temperature, we can calculate the temperature difference between the neighboring measurement trays. The position can be determined by dividing the sum of the temperature differences multiplied by its position by the sum of the temperature differences:

$$S = \frac{\sum_{i=l}^{m-1} S_i \Delta T_i}{\sum_{i=l}^{m-1} \Delta T_i} \qquad (13)$$

where i is the measurement tray number: l is the first measurement tray number and m is the last measurement tray number in a column section. ΔT_i is the temperature difference between the i-th measurement tray and i+1-th measurement tray and S_i is the midpoint between i-th and i+1-th measurement tray.

Composition/profile position cascade control

For column sections I, III, and IV of sidestream/sidestripper configuration, the composition/profile position cascade control is used. However, for the control of the profile position in the column section II, a simple PI controller is used. The profile position control in column section I and one in column section III are paired to the distillate composition controller and the bottom composition controller, respectively. The profile position control in column section IV is cascaded to bottom composition controller of the sidestripper.

Conventional Control with SISO Controllers

For the control of the top composition of the main column, reflux flow rate is chosen as the manipulated variable. For the control of bottom composition of both columns, heat input to each column is employed. A constant temperature difference between trays above and below the sidedraw tray is maintained by manipulating LS. This control scheme was introduced by Alatiqi and Luyben (1986). However, in this scheme, the temperature difference does not match with any one profile position in most cases so that the gains of the temperature difference changes signs if the column profile moves too large. Also, the control of the temperature difference could be difficult since the temperature profile in the middle section is sharp and produces nonlinear output of the temperature difference.

Step testing has been used to get transfer function models for controller tuning. The composition responses have been assumed to be first order plus dead time. Multiloop PI controllers are used for the control of these configurations and the method of biggest log modulus tuning (BLT) (Yu and Luyben, 1986) is used for the tuning of control parameters.

Comparison of the Proposed Control Scheme with the Conventional Decentralized Control Scheme

Figures 4 through 7 show the results of the proposed control scheme and conventional decentralized control scheme for the SSS configuration. The responses of both the conventional control scheme and our control scheme to a +20% step change in the feed concentration are shown in Figure 4. The performance of our control scheme is much better than that of the conventional scheme with PI controllers. In Figure 5, the responses of both controllers to a +20 % feed flow disturbance are given. It indicates that the performance of our controller is far superior to that of the conventional control scheme although our controller controls the feed flow disturbance in a feedforward manner. Figure 6 shows the dynamic response of the profile positions tracking their setpoint

Table 1. Steady-state Designs

A) Sidestream/sidestripper configuration

feed flow rate	600 lb mol/h	
feed temperature	278 °F	
composition		B/T/X mol %
	feed	40/10/40
	distillate	99/1/0
	bottom	0/1/99
	side	1/99/0
feed tray (from base)	13	
side draw tray	27	
total no. of trays in the main column		40
total no. of trays in the side column		13
column pressure		29.4 psia

B) Prefractionater/sidestream column configuration

	Column 1	Column 2
feed flow rate	600 lb mol/h	326.7/273.3
feed temperature	255 °F	231.8/299.9 °F
composition	B/T/X mol%	
feed	40/40/10	73.05/26.45/0.05
		0.05/56.19/43.36
distillate	73.05/26.45/0.05	99/1/0
bottom	0.05/56.19/43.36	0/1/99
side	2.5/95/2.5	
feed tray	12	7/34
side draw tray		17
total trays	20	40
column pressure	29.4 psia	29.4 psia

given by the product composition controllers. Figure 7 compares the responses of the conventional decentralized control scheme and the responses of the proposed control scheme to a +100 % change in feed concentration. The result shows that the performance of the conventional control scheme is very poor and control of product purity becomes impossible eventually but our controller maintains stable operation, keeping the purities of all products to their setpoints. It proves that our control scheme is capable of handling larger disturbance than the conventional control scheme.

Control of complex prefractionater/sidestream column

Description of the System

A two column prefractionater/sidestream column (PF) configuration consists of two columns: one is a prefractionater that makes a preliminary split and the other column is a side stream column to which overhead and bottoms product of the prefractionter are fed at two different feed locations and at which final separation into three products is acquired. Both feed to the sidestream column are essentially binary mixtures since the prefractionater removes the lightest and heaviest components from distillate and bottom products, respectively. It makes the lightest and the heaviest component do not appear at the center section of the sidestream column as a liquid sidestream from a tray located between the two feed trays. It follows that the lightest component is the overhead product of this sidestream column, the heaviest component is the bottom product and the intermediate component is the side product. The schematic diagram of this configuration is given in Figure 8. The product purities in the product streams of the sidestream column are 1 % toluene purity in the distillate, 1 % toluene impurity in the bottom, 2.5 % benzene and 2.5 % xylene impurity in the sidedraw.

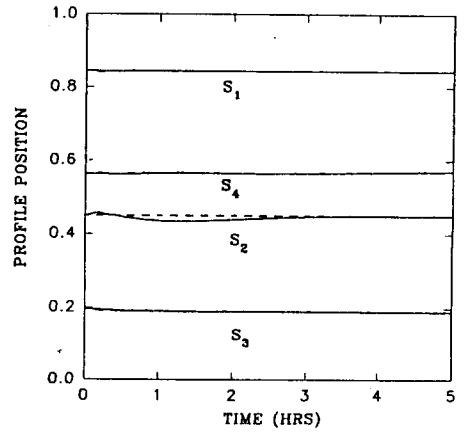

Fig. 6. Closed-loop responses of profile position to a +100 % change in toluene concentration of the feed (— , output; ----, setpoint; SSS configuration)

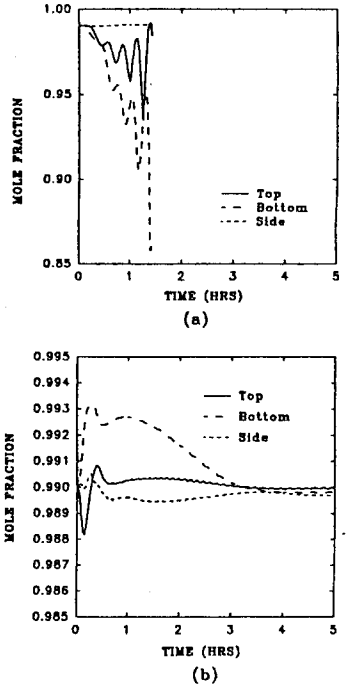

Fig. 7. Closed-loop responses to a +100 % change in toluene concentration of the feed (SSS configuration) :
(a) conventional control scheme with SISO controllers
(b) proposed control scheme.

There are five manipulated variables in this prefractionater/sidestream column configuration: reflux flow rate to each column, heat input to each column and sidestream draw rate. There could be five variables to be controlled: the benzene impurity of the first column, the xylene impurity of the overhead of the first column, the o-xylene impurity of the overhead of the second column, the benzene impurity of the sidedraw, the o-xylene impurity of the sidedraw of the second column, the toluene impurity of the bottom of

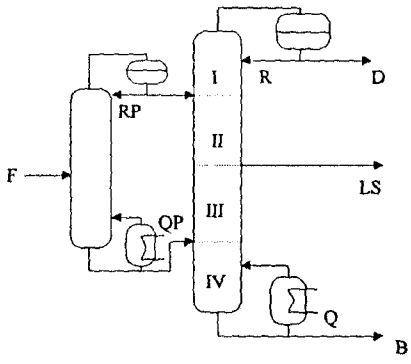

Fig. 8. Prefractionater/sidestream configuration

the second column. There must be at least one composition controller on prefractionater to prevent large amounts of benzene from dropping down to the bottom and large amounts of xylene from going overhead (Ding and Luyben, 1990). These events make the separation in the second column very difficult because the lightest or heaviest component appears at the center section of the sidestream column and give adverse effects on sidestream purity. The details of the column are listed in Table 1B.

Profile Position Controller Design

The prefractionater/sidestream column configuration has also four column sections. For each column section, the following equations can be written:

For column section I,

$$\frac{V_1}{F_1}\frac{\Delta y/\Delta x - L_1/V_1}{1 + r(\Delta y/\Delta x)} - K_{11}(S_1^* - S_1) - K_{12}$$

$$\int_0^t (S_1^* - S_1)dt' = 0 \qquad (14)$$

For column section II,

$$\frac{V_2}{F_1}\frac{\Delta y/\Delta x - L_2/V_2}{1 + r(\Delta y/\Delta x)} - K_{21}(S_2^* - S_2) - K_{22}$$

$$\int_0^t (S_2^* - S_2)dt' = 0 \qquad (15)$$

For column section III,

$$\frac{V_3}{F_2}\frac{\Delta y/\Delta x - L_3/V_3}{1 + r(\Delta y/\Delta x)} - K_{31}(S_3^* - S_3) - K_{32}$$

$$\int_0^t (S_3^* - S_3)dt' = 0 \qquad (16)$$

For column section IV,

$$\frac{V_4}{F_2}\frac{\Delta y/\Delta x - L_4/V_4}{1 + r(\Delta y/\Delta x)} - K_{41}(S_4^* - S_4) - K_{42}$$

$$\int_0^t (S_4^* - S_4)dt' = 0 \qquad (17)$$

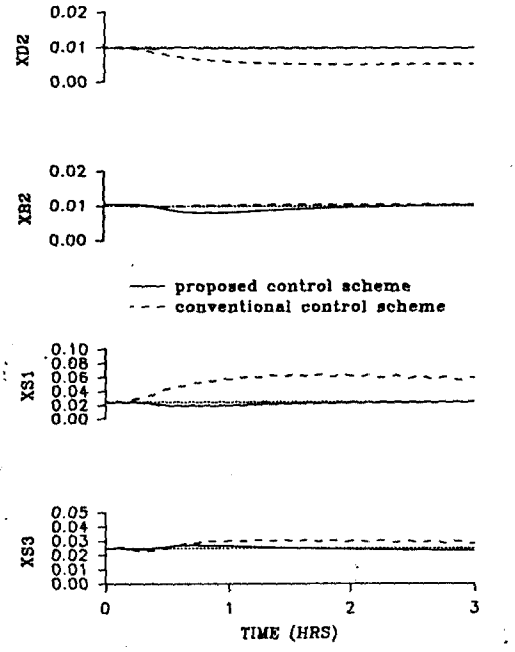

Fig. 9. Closed-loop responses to a +12.5 % change in benzene concentration of the feed (PF configuration): XF (0.4/0.4/0.2 → 0.5/0.3/0.2)

$$V_2 = V_1 - (1 - q_1)F_1 \qquad (18)$$

$$V_3 = V_2 \qquad (19)$$

$$V_4 = V_3 - (1 - q_2)F_2 \qquad (20)$$

$$L_2 = L_1 + q_1F_1 \qquad (21)$$

$$L_3 = L_2 - LS \qquad (22)$$

$$L_4 = L_3 + q_2F_2 \qquad (23)$$

$$F_1 = VP - LP \qquad (24)$$

$$F_2 = LP + F - VP \qquad (25)$$

From the above equations, liquid and vapor flow rate of each column, section can be determined to control the profile position of the section.

On-line estimation of the profile position

Two feeds from the prefractionater are nearly binary so the sidestream column can be regarded as a pseudobinary system. So, the proposed estimator has been used for all sections of the column.

Composition/Profile position cascade control

In the case of prefrationater/sidestream column configuration, the composition/profile position cascade system is used for all column sections. The profile position control in column section I is cascaded to the composition controller of XD2. The control of profiles in column section II and in section III are cascaded to the controller of XS1 and XS3, respectively. The profile position control in column section IV is cascaded to the controller of XB2. Benzene concentration in the bottom of prefractionater is regulated with simple PI controller.

Conventional Control with SISO Controllers

The four compositions to be controlled in the three product streams of the sidestream column are toluene impurity in the distillate (XD2), benzene impurity in the liquid side stream (XS1), o-xylene impurity (XS3) in the liquid sidestream and toluene impurity in the bottoms (XB2). Toluene impurity in the distillate is controlled by manipulating heat input to the sidestream reboiler (QB). Toluene impurity in sidestream column distillate is controlled by manipulating reflux flow rate to the sidestream column (R). The benzene impurity in the liquid sidestream is controlled by manipulating distillate rate from prefractionater (D1) and the xylene impurity in the liquid side stream is controlled by manipulating sidedraw rate LS. Benzene impurity in the bottom of the prefractionater is regulated by the heat input to the reboiler of the prefractionater.

Comparison of the Proposed Control Scheme with the Conventional Decentralized Control Scheme

The responses of both the conventional control scheme and the proposed control scheme to a step change in the feed concentration are given in Figure 9. The proposed control scheme is shown to improve the performance of the system greatly. In particular, the control performance of our control scheme is far surperior in our control scheme to the conventional decentralized control scheme in regulating impurity compositions in the sidedraw of side stream column. It is because nonlinearity is very severe for the sidestream toluene purity as Ding and Luyben (1990) indicated.

Conclusion

A profile position controller based on a nonlinear wave theory and generic model control has been developed and applied to the control of complex distillation configurations. A nonlinear profile position observer has been used to estimate the profile position of the column sections.

The nonlinear, interacting, multivariable complex distillation configurations have been successfully controlled by the proposed control scheme. Dynamic simulation experiments for the control of these two example systems show that the proposed control scheme can improve the control of complex distillation configurations, substantially and can handle larger disturbances than the conventional control scheme with PI controllers.

Nomenclature

B = bottoms flow in main column (SSS, lb mol/h)
BS = bottoms flow in side column (SSS, lb mol/h)
B1, B2 = bottoms flow from first and second column (PF, lb mol/h)
D = distillate flow, lb mol/h
D1, D2 = distillate flow from first and second column, respectively (PF, lb mol/h)
K = GMC controller constant
F = feed flow rate, lb mol/h
F1, F2 = feeds to sidestream column (PF, lb mol/h)
LP = liquid rate in prefractionater (PF, lb mol/h)
LS = liquid draw rate from main column to stripper in SSS configuration; liquid side stream flow rate in PF configuration, lb mol/h.
QB = main column heat-transfer rate MMBTU/h
QBS = side column reboiler heat-transfer rate (SSS, MMBTU/h)
t = time, h
q, q1, q2 = mole fraction of liquid in the feed, feed 1 and feed 2, respectively
R = reflux flow rate, lb mol/h
r = molar holdup ratio of vapor to liquid
S = profile position of column section
ΔT = temperature difference, F
u_Δ = normalized shock wave velocity
V = vapor boilup in the main column, lb mol/h
VP = vapor boil-up in prefractionater of PF configuration (PF, lb mol/h)
VS = vapor boilup in the side column (SSS, lb mol/h)
x_B = bottoms composition in the main column (SSS, mole fraction)
x_{BS} = bottoms composition in the side column (SSS, mole fraction)
XB1 = benzene concentration in bottoms of prefractionater (PF, mole fraction)
XB2 = toluene concentration in bottoms of second column (PF, mole fraction)
XD2 = toluene concentration in distillate of second column (PF, mole fraction)
XS1 = benzene concentration in side draw of second column (PF, mole fraction)
XS3 = xylene concentration in side draw of second column (PF, mole fraction)
x_D = distillate composition in the main column (SSS, mole fraction)
x_s = representative liquid mole fraction corresponding to profile position of self-sharpening standing wave

Greek Letters

Δ : prefix for difference between the two sides of self-sharpening wave
τ : reset time, h

Superscripts

* : setpoint
^ : observer prediction
· : derivative

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