Control of High-Purity Distillation Column Using a Nonlinear Wave Theory

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A new model-based controller for profile position control of high-purity distillation columns is presented. The controller has been developed by incorporating nonlinear wave model into the generic model control framework. An observer based on the nonlinear wave theory is also proposed to determine the profile position. The performance of the nonlinear control scheme has been tested in simulation experiments of high-purity binary distillation columns. Tight control of profile position, which leads to fast stabilization of product compositions, has been achieved. Since the profile position control alone produces a slight offset from the desired product concentration, the composition/profile position cascade system has been used to remove the offset. The control scheme can handle significant disturbances and model-plant mismatch.

Introduction

High-purity distillation columns have been well known to pose a formidable controller design problem, resulting from unique characteristics, such as complex dynamics, high nonlinearity and interaction between the control loops. Recent demands for a better product quality, waste reduction, and low energy consumption have made tight control of high-purity distillation columns an important issue.

The highly nonlinear and distributed character of the high-purity column is represented by its sharp composition and temperature profiles. It is because only a small portion of the column is used for the major separation while other parts are only responsible for refining the products (Hwang, 1991). Using the location of a sharp temperature profile as a controlled variable, Luyben (1972) pioneered a profile position control strategy. He insisted that the profile position control has several advantages over the temperature control on a tray. The basic problem in trying to control a temperature on a tray is the high process gain that requires a very small feedback controller gain to maintain closed-loop stability but the profile position control greatly reduces the process gain that permits higher controller gains and better response to load disturbance. Marquardt (1986, 1988) proposed a nonlinear wave concept to develop a model for binary distillation using wave position, wave shape parameters and product compositions as state variables. He showed that the model can be applied to stage columns with sufficient accuracy by adjusting the model parameters. Hwang (1991) established a nonlinear wave theory for the nonlinear and distributed dynamics of binary distillation columns that may be useful for model development and control system design.

The simulation model used in this work is a high-purity ($X_D = 0.99$, $X_B = 0.01$) methanol-ethanol separation column model, referred to as column C by McAvoy (1983). This process is shown by Alsop and Edgar (1990) to be sufficiently nonlinear, interactive and ill-conditioned to serve as a challenging test case for the control.

This article presents a new model-based controller design that deals with both nonlinear and multivariable characteristics of high-purity distillation columns. The controller has been developed by incorporating Hwang's (1991) nonlinear wave model of high-purity distillation column into the generic model control (GMC) framework by Lee and Sullivan (1988). To estimate the profile position, an observer of the profile position has also been developed. The controller performance becomes worse as the composition analyzer dead times increase. On the other hand, temperature measurement on a tray has faster dynamics so that a temperature/composition cascade control system can be used to improve the performance of the controller (Fuentes and Luyben, 1983). Similarly, the profile position controller may be cascaded to the composition controller to guarantee the tight product composition control. The per-

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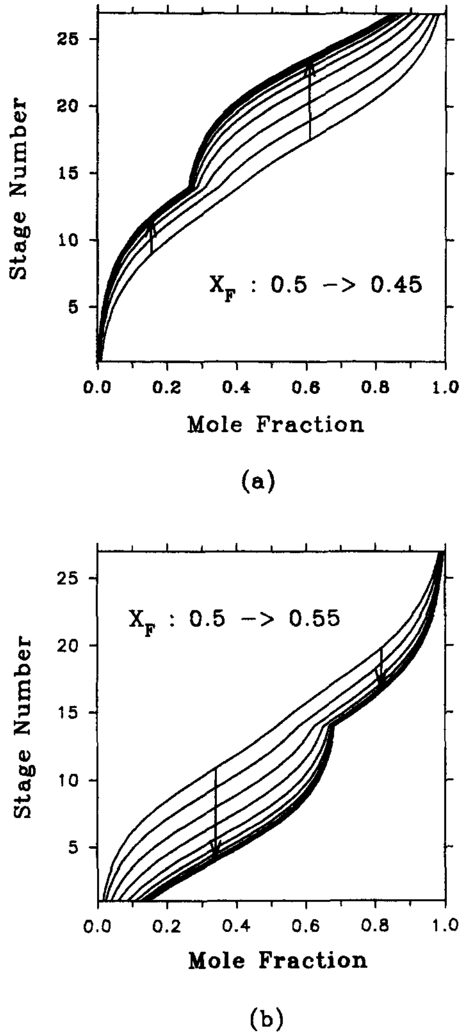


Figure 1. Dynamic profiles of the distillation column to a step disturbance of feed composition.

(Each curve is separated by 10 time units.)

formance of this control system has been tested in dynamic simulation experiments.

Nonlinear Wave Theory

Model based approaches have been often taken to get high quality control of distillation columns. Often these models are linear and thus inadequate for highly nonlinear processes such as high-purity distillation columns. This 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 distillation columns is characterized by the propagation of concentration or temperature profile in the column sections. During transients, the profile is preserved to be constant. This typical dynamic behavior is summarized by Marquardt (1986) to nonlinear wave propagation where nonlinear wave is a spatial structure moving with a constant propagation velocity and constant shape along a spatial coordinate. Figure 1 presents the numerical simulation results of the propagating wave for the column given in Table 1. To model the state of the column in terms of a continuous composition profile and its dynamics in terms of propagating composition waves can be a key for the control of a high-purity distillation column. Gilles and Retzbach (1980, 1983) and Marquardt (1986, 1988) successfully approximated the behavior of distillation columns with this approach. Recently, Hwang (1991) proposed a nonlinear wave theory to explain the nonlinear behavior of distillation columns.

Table 1. Specification for Methanol-Ethanol Separation Column C of McAvoy (1983)

Column Data:	
Relative volatility	$\alpha = 1.65$
Trays	27
Feed tray	14
Accumulator molar holdup	5.5
Reboiler molar holdup	5.5
Tray molar holdup	0.55
Tray efficiency	100%
Operating Variables:	
$X_D = 0.99$	
$X_B = 0.01$	
$X_F = 0.5$	
$F = 1.65$	
$D = 0.825$	
q feed	1
$V_B = 5.17575$	
R/D nominal	5.27
Measurement tray: 8, 11, 14, 17	

The nonlinear wave theory starts from the differential material balance for a section of a distillation column as follows:

$$\frac{\partial x}{\partial \tau} + r\frac{\partial y}{\partial \tau} - \left(\frac{L}{F}\right)\frac{\partial x}{\partial \sigma} + \left(\frac{V}{F}\right)\frac{\partial y}{\partial \sigma} = 0 \tag{1}$$

where r is vapor to liquid holdup ratio. Assuming the liquid flow is so slow that local equilibrium is attained, y in Eq. 1 can be substituted with the vapor-liquid equilibrium relation. Rearranging Eq. 1, an equation for the "wave velocity" tracking the propagation of a specific value of concentration x or y can be obtained as the following:

$$u \equiv \left(\frac{\partial \sigma}{\partial \tau}\right)_x = \frac{V}{F}\frac{dy/dx - L/V}{1 + r(dy/dx)} \tag{2}$$

where dy/dx is the slope of the equilibrium curve at the specific value of concentration. The wave velocity also can be viewed to be the relative convective transport at equilibrium. This velocity reflects the intrinsic behavior pertinent to the equilibrium between the two phases while disregarding the dissipative effects of nonequilibrium and dispersion. The wave velocity varies with concentration for a system with nonlinear equilibrium and therefore varies with location within a wave. So, a wave tends to sharpen if the velocity decreases with the location and this wave has been called as a "self-sharpening" wave or shock wave. The velocity of this wave is derived from the material balance across the wave:

$$u_\Delta = \left(\frac{\partial \sigma}{\partial \tau}\right)_\Delta = \frac{V}{F}\frac{\Delta y/\Delta x - L/V}{1 + r(\Delta y/\Delta x)} \tag{3}$$

where Δ denotes the difference between the two sides of the shock wave.

For most binary distillation, the wave is always self-sharpening since the corresponding portion of the equilibrium curve is as a whole concave toward the operating line. In a low-purity column, the column is too short to contain an entire self-sharpening wave. However, in a high-purity column, the wave is always self-sharpening and becomes a constant pattern wave with fixed shape as soon as its intrinsic sharpening tendency is balanced by the dissipative effects of nonequilibrium and dispersion (Hwang, 1991). It occupies only a small portion of the column with sharp gradient where major separation occurs and other parts of the column only take the role of refining the products.

The travel of such a constant-pattern wave can be characterized by shock wave velocity. The wave tends to travel to either one of column ends unless the balance of convective transports is carefully maintained to have a zero shock wave velocity with the compositions and flow rates of all streams entering a column section. So, the behavior of a high-purity column is severely nonlinear and sensitive since even a small upset of the balanced condition will lead to a large shift of the profile, giving dramatic changes in product purity (Hwang, 1991). During propagation of a profile, the profile takes a constant shape along a spatial coordinate so that movement of some specific concentration can well represent the dynamic of the profile. With simple mathematics, the nonlinear wave theory can predict the nonlinear dynamics of distillation columns. It makes possible to develop dynamic models and to design control systems for high purity distillation columns.

The attempts of using such a partial differential equation to describe the dynamic behavior of distillation columns have been tried by some other researchers (Wong and Luus, 1980; Cho and Joseph, 1983; Stewart et al., 1985; Marquardt, 1986, 1988). Most of them have taken an orthogonal collocation technique to approximate the composition profiles in the column as polynomials. The approaches do not consider the spatial structure evolving from the distillation column explained above. They cannot explain a highly nonlinear, distributed character of distillation columns which is resulted from its sharp composition and temperature profiles, only providing a polynomial approximation of the column profile by mathematical formalism. The resulting parameters do not have a physical meaning and cannot be easily adjusted (Skogestad, 1992). Gilles and Retzbach (1980) viewed the essential dynamic quality of their extractive column as the movement of the regions of high mass transfer that is reflected by temperature fronts. Therefore, they used the loci of temperature fronts as state variables. Marquardt (1986, 1988) generalized the approach starting from partial differential material balance of column section to develop a novel model for distillation. The approach is satisfactory but its disadvantage is the high complexity of the proposed model. Hwang's model explains the dynamics of high-purity distillation column within a concise mathematical framework. In this work, the model forms a basis of the observer for the profile position and the generic model controller in our control system.

Generic Model Controller Design

The control of profile position regulates the propagation of column profile to prevent the column from dropping into a nonlinear region and giving large errors in product purity. Generic model control (GMC) introduced by Lee and Sullivan (1988) has several advantages that 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 and thus the nonlinear wave model can be used for profile position control. 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. Any inaccuracies introduced by approximate modeling can be compensated by the integral term in the control algorithm. This integral term ensures that the controller is robust despite modeling errors.

The formulation of a GMC controller for profile position control of distillation columns incorporates the nonlinear wave model equation proposed by Hwang (1991). Consider a process described by the following equation:

$$\frac{dx}{dt}=f(x,\,u,\,d) \tag{4}$$

where x is the state vector, u is the vector of manipulated variables and d is the vector of disturbance variables. When the process is away from its desired steady state x^* we would like the rate of change of x to be such that the process is returning towards steady state and has zero offset, that is

$$dx/dt=K_1(x^*-x)+K_2\int_0^t(x^*-x)dt' \tag{5}$$

where K_1 and K_2 are tuning constants. Choosing the profile position of a column section as the state vector in Eq. 5, the equation can be rewritten by replacing x with S as follows:

$$dS/dt=K_1(S^*-S)+K_2\int_0^t(S^*-S)dt' \tag{6}$$

where S and S^* are the profile position and its setpoint respectively; dS/dt is the propagation rate of the profile. 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 as follows:

$$dS/dt=u_\Delta=\frac{V}{F}\frac{\Delta y/\Delta x-L/V}{1+r(\Delta y/\Delta x)} \tag{7}$$

Simple distillation columns, in general, have two sections: one is the rectifying section and the other is the stripping section. Combining Eqs. 6 and 7 gives one equation for each section as follows:

$$\frac{V}{F}\frac{\Delta y/\Delta x-L/V}{1+r(\Delta y/\Delta x)}-K_{11}(S_1^*-S_1)-K_{12}\int_0^t(S_1^*-S_1)dt'=0 \tag{8}$$

$$\frac{\overline{V}}{F}\frac{\Delta\bar{y}/\Delta\bar{x}-\overline{L}/\overline{V}}{1+r(\Delta\bar{y}/\Delta\bar{x})}-K_{21}(S_2^*-S_2)-K_{22}\int_0^t(S_2^*-S_2)dt'=0 \tag{9}$$

where subscripts 1 and 2 represent rectifying section and stripping section respectively, and L and V are the liquid and vapor flow rates respectively in the rectifying section and $\overline{L}$ and $\overline{V}$ in the stripping section. The profile position and the slope of the equilibrium curve at the representative concentration can be estimated by the profile position observer accurately. Mass balance around the feed tray gives

$$\overline{L} = L + qF \qquad (10)$$

$$V = \overline{V} + (1-q)F \qquad (11)$$

where q is the liquid mole fraction of the feed. Therefore, if we know the feed condition, L, V, $\overline{L}$ and $\overline{V}$ can be determined from Eqs. 8-11.

On-Line Estimation of the Profile Position

The success of the GMC profile position controller is mainly dependent on the ability to estimate the profile positions that cannot be measured directly. In this work, we propose a nonlinear profile position observer to estimate the profile positions where an observer is a dynamic model that estimates the state variables of a process.

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 profile position of the constant pattern wave can be determined easily by tracking the representative concentration instead of the entire wave.

For the determination of the profile position, several approaches have been taken. The measurement of composition or temperature on a chosen tray for tracking the representative concentration has a problem: the effective range of measurement is limited due to the saturation. The temperature of the measurement tray drops to the boiling point of the low boiling component as the profile drops below the measurement point but shows no change thereafter as the light component moves down the column. To reduce this saturation problem, Luyben (1972) located several temperature sensors up and down the column, above and below the desired location of the profile. He determined the position of the profile in a straightforward manner by having the computer scan several thermocouples installed up and down the column to find out among which trays a temperature in the middle of the break lies. Gilles (1986) developed a simplified dynamic model to estimate the profile position and other state variables. Wozny et al. (1989) estimate the temperature front location, which has the same meaning as the profile position, using the global material balance of the column section with the feedback of the tray temperature measurement. These methods are based on temperature measurements and material balances of column sections.

We have used Hwang's nonlinear wave model with the feedback of temperature or concentration measurements. Concentrations are measured at different points of the column and compared with the observer's output concentration. The concentration can be determined by the temperature in each stage through the equilibrium relationship in the binary distillation column. The error is weighted by a spatially distributed function. The resulting term is used to correct the mathematical model of the observer. A full-order state observer according to Luenberger (1964) consists of a model of the plant with additional feedback of the weighted output error as follows:

$$\dot{S} = \frac{dS}{dt} = \frac{V}{F}\,\frac{\Delta y/\Delta x - L/V}{1 + r(\Delta y/\Delta x)} + \sum_{i=l}^{m} K_1(x_i)(x_i - \hat{x}_i) \qquad (12)$$

$$\Delta y/\Delta x = \frac{\dot{S} + \dfrac{L}{F}}{\dfrac{V}{F} - r\dot{S}} \qquad (13)$$

where i is the measurement tray number and l and m are the number of first measurement tray and the number of last measurement tray in a column section, respectively; ^ indicates the estimated value. The representative slope, which is the slope of the equilibrium curve at the representative concentration, can be estimated from Eq. 13. The representative concentration can be determined as the concentration at which the equilibrium curve has the same slope as $\Delta y/\Delta x$. The observer computes the concentration of each measurement tray from the estimated profile position with the assumption that the profile has a linear form as follows:

$$\hat{x}_i = K_2(S_i - S) + x_s \qquad (14)$$

where S_i is the normalized distance at which ith tray is located and x_i is the concentration at the ith tray and x_s is the representative concentration. Waves in distillation columns tend to sharpen and eventually become constant-pattern waves when their intrinsic sharpening tendency is balanced by the dissipative effects of nonequilibrium and dispersion. It follows that the parameter K_2 can be obtained as an intrinsic system property from operation data or simulation.

Within the limits of the effective measurement range, the measured variables change nearly linearly with the deviation from the representative concentration. If the measurement tray goes out of the limits as the profile propagates, the weighting of the tray becomes smaller while other measurement trays within the limits become to have greater weighting. So, regardless of the profile position, the trays near the profile position, forming a sharp profile region, have strong weighting. The difference between the process and model has to be corrected by the measurement so that the sharp profile regions of both coincide. Thus the profile position can be estimated with sufficient accuracy and the saturation problem can be greatly reduced. The weighting factor $K_1(x_i)$ has the following form:

$$K_1 = K_0 \exp[-b(x_i - x_s)^2] \qquad (15)$$

The value of the weighting factor K_1 gets greater as the concentration of the measurement tray is closer to the representative concentration with an exponential form. Here, K_o denotes the gain of the weighting function and b adjusts the range of the weighting. The weighting gain K_o of the observer is chosen so that the observer is considerably faster than the process. The parameter b can be determined by considering the profile shape of the column section. The observer response time should be fast enough to provide convergence of the estimates within the time interval of the interest. Therefore, the sampling time of the observer is chosen as 0.1 time unit.

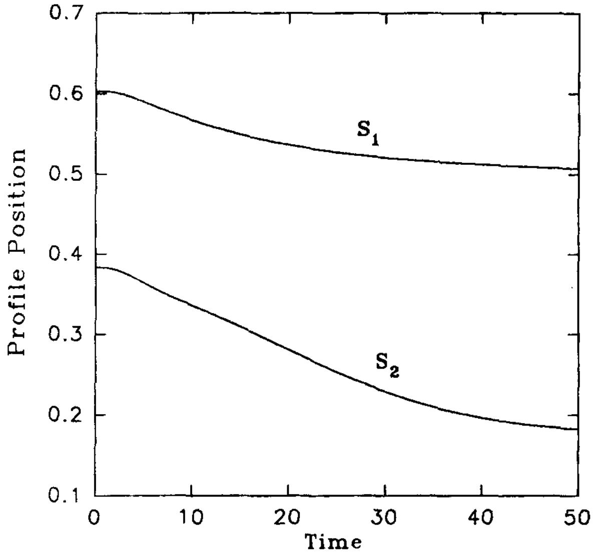


Figure 2. Open-loop step response of profile position to a step disturbance of X_F.

The algorithm of the observer is summarized in the following steps:

Step 0. Select initial estimates of profile position S and representative slope $\Delta y/\Delta x$.

Step 1. Measure temperatures or concentrations at measurement trays and calculate the errors between the measured and the estimated tray compositions.

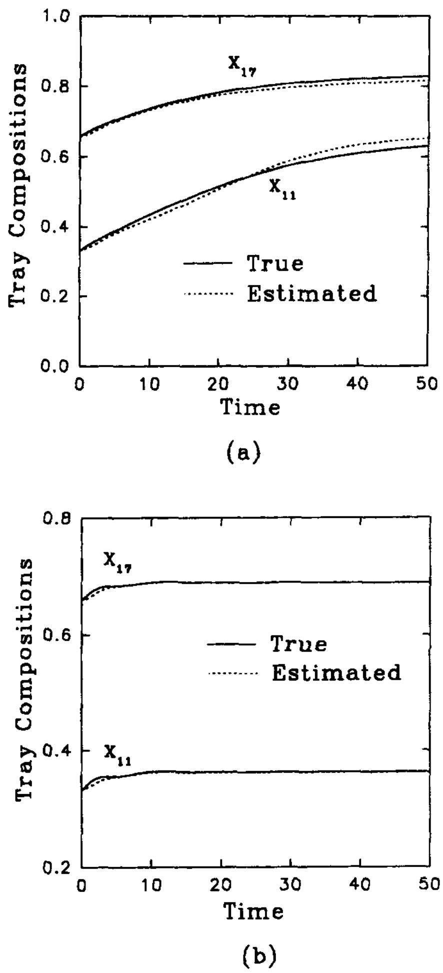
Step 2. Estimate the propagation velocity and the representative slope from the equation with the feedback of the weighted errors.

Step 3. From the equilibrium relation and the representative slope, find the representative concentration.

Step 4. Estimate the profile position by integrating the wave propagation velocity.

Step 5. Estimate tray compositions from the profile position and the representative concentration, return to step 1.

For testing the performance of the observer, a step change from 0.5 to 0.55 in the feed concentration is made. The dynamic responses of the column are shown in Figures 2–3. Figure 2 shows the transient behavior of the profile position estimated by the observer. The comparisons between the measured and the estimated concentration by the observer at the measurement trays are shown in Figure 3. Both comparisons are in good agreement and more accurate estimation is obtained in the closed-loop as expected. The accuracy of the estimated profile position may be supported by the agreement.



(a)

(b)

Figure 3. Step response of tray compositions to feed composition change, measured vs. estimated: (a) open-loop; (b) closed-loop.

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 product composition may occur when profile position of the column section is fixed at the specific point. Therefore, composition controller (PI) is employed and cascaded as the primary controller to the profile position controller to remove the offset. The comparison of GMC only and the cascade controller is shown in Figure 4 which shows the slight offset in the response under GMC only. It shows that tight control of profile position leads to fast stabilization of the product concentrations and that the composition controller can remove the slight offset from the setpoint in product composition. The sampling time of the composition controller was chosen as one time unit while that of the profile position controller was 0.1 time unit.

Marquardt (1988) showed that the shape of the concentration profile is altered significantly if the internal flow rates are changed, for example, as a result of the control action. However, in the case of keeping the ratio L/V (L and V are the internal liquid and vapor flow rates) in a column section, the variations in the concentration profile can be reduced (Roffel and Rijnsdorp, 1982). Since the profile position controller is a ratio controller of internal flow rates, the composition profile in the transient period can be preserved with little change. The smooth variations of the column profile in the case of the setpoint change and disturbances are shown in Figures 5–6.

Figure 5 shows the closed-loop response to a desired setpoint change in X_B from 0.010 to 0.001. This is a very large relative change in the setpoint for the system. With no offset, the product concentration can be moved to the setpoint. Feed composition and flow upsets are major disturbances affecting distillation columns. Compensation for feed flow disturbances can be achieved easily since, in our control scheme, ratios of the manipulated variables to the feed are manipulated in a simple feedforward manner. The ability of the control system to reject the effect of the feed composition change is shown in Figure 6. The proposed control scheme is shown to provide a good performance for the large upset such as a change in the feed composition from 0.5 to 0.1.

Robustness Evaluation

It is important to examine the robustness aspects of the controller about model-plant mismatch and input uncertainty for applications to real distillation plants. Since parameters for the proposed controller are vapor-to-liquid holdup ratio and relative volatility, we have tested the performance of the controller for the errors in the parameters in the case of feed composition change of −0.3.

The test for the errors in vapor-to-liquid holdup ratio has shown that the response of the product composition is essentially the same and the performance of the controller has also no degradation despite the change in the parameter value (from 0.01 to 0.05). The main source of nonlinearity in our model is the equilibrium relation, which is represented by the relative volatility. Thus, the great error of relative volatility between

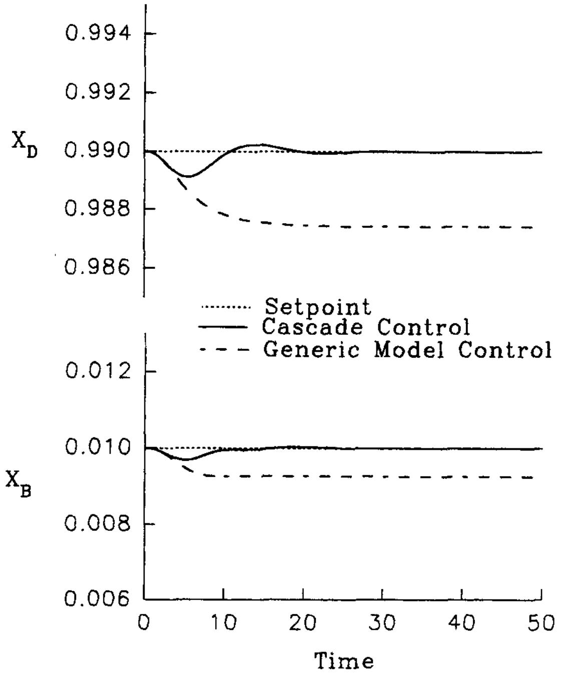
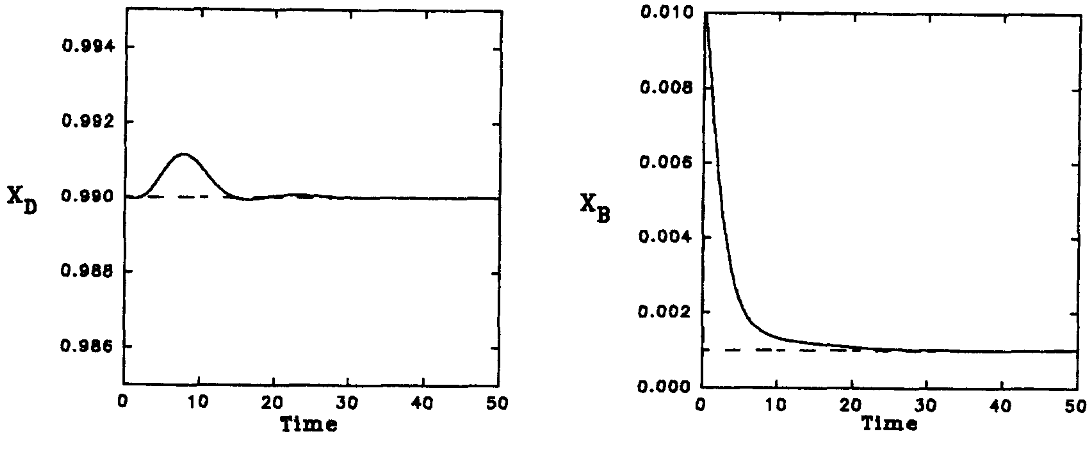


Figure 4. Comparison of cascade control and GMC only.



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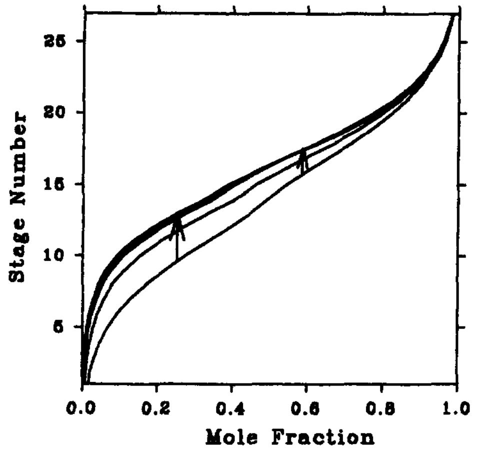


(c)

Figure 5. Closed-loop transient response for setpoint change in X_B of −0.009: (a) distillate; (b) bottom product; (c) column profile.

(Each curve is separated by 2 time units).

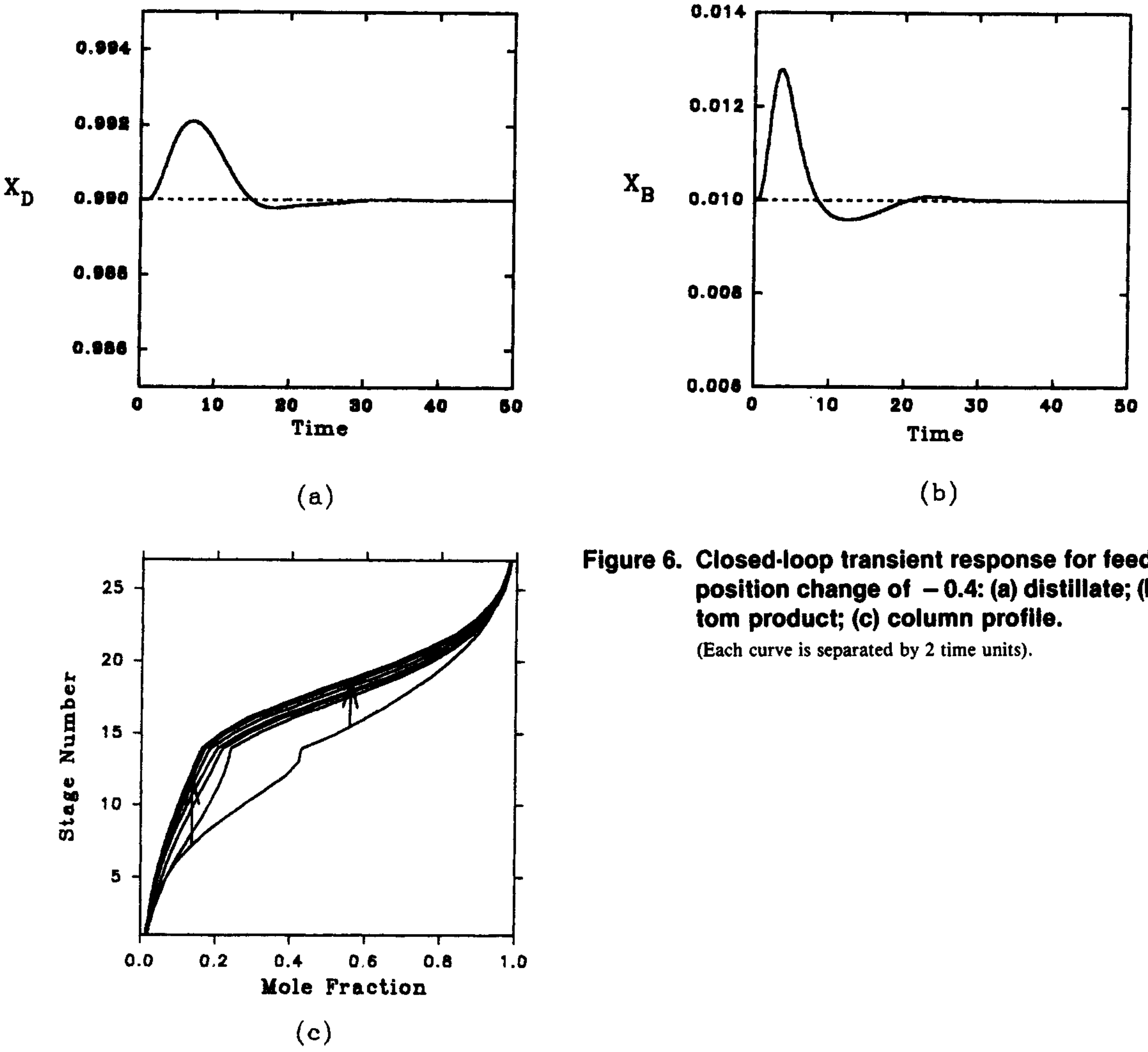


Figure 6. Closed-loop transient response for feed composition change of − 0.4: (a) distillate; (b) bottom product; (c) column profile.

(Each curve is separated by 2 time units).

the process and the model may lead to instability of the control system. Figure 7 shows the responses of the controlled product composition when there exists an error in the value of relative volatility. The controller is stable and shows slight degradation of the performance up to 15% error in the relative volatility. Therefore, this case shows that the control scheme is robust to model-plant mismatch.

Skogestad and Morari (1986) showed that the closed-loop system may be extremely sensitive to input uncertainty when the LV configuration is used. Therefore, it is necessary to consider input uncertainty for getting a realistic evaluation of the controllers (Skogestad and Morari, 1986, 1987). For testing the performance of the proposed control scheme in the presence of input uncertainties, the following uncertainties used by Skogestad and Morari (1988) have been chosen.

$$\Delta L = (1+\Delta_1)\Delta L_c \qquad (16)$$

$$\Delta V = (1+\Delta_2)\Delta V_c \qquad (17)$$

Here, ΔL and ΔV are the real changes in manipulated flow rates, while ΔL_c and ΔV_c are the desired values as computed

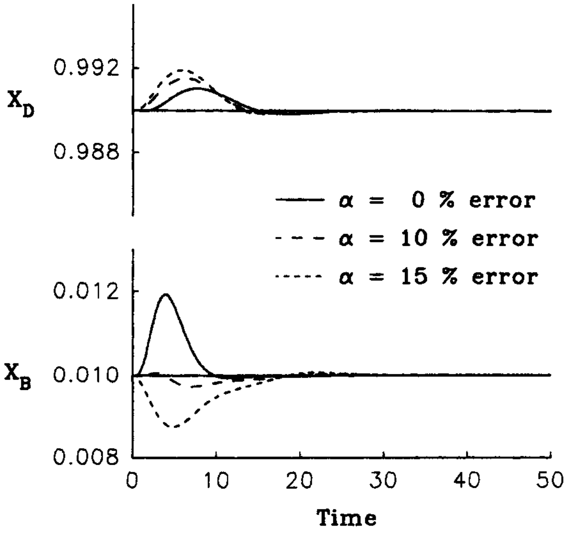


Figure 7. Closed-loop transient response for different errors of relative volatility.

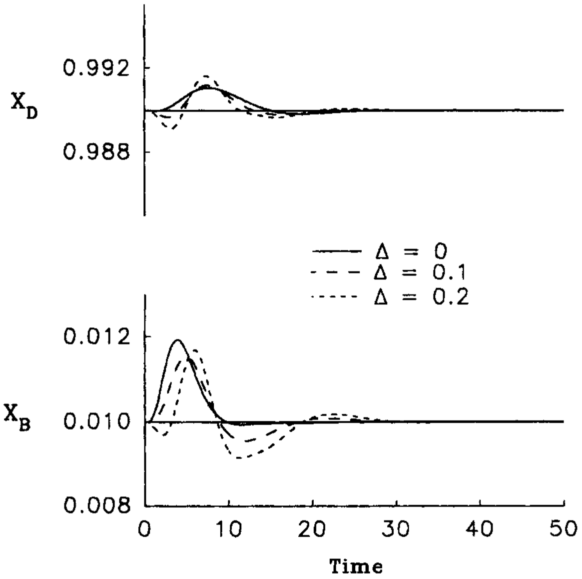


Figure 8. Closed-loop transient response for different errors of input uncertainty.

by the controller. $\Delta_1 = -\Delta_2$ has been selected to yield the worst combination of the uncertainties (Skogestad and Morari, 1986, 1987). Figure 8 shows the closed-loop transient response of different values of input uncertainty. The figure shows that in the face of input uncertainties the controller has remained stable up to 20% error. This example shows that the performance is satisfactory in spite of unmeasured feed composition disturbance (from 0.5 to 0.2) and model-plant mismatch (uncertainty).

Comparison of the Proposed Controller with Other Control Strategies

To get a strict evaluation of the performance of the proposed controller, three columns have been chosen and several control strategies have been applied to the columns. The performance of the control strategies has been compared with that of our controller.

Alsop and Edgar (1990) developed a new modeling/control approach which uses an approximate linearization and a nonlinear transformation using singular value decomposition. The specification of their column is listed in Table 1. Figure 9 shows the ability of several control systems to reject a feed composition change of -0.3 (-60%). Our controller is shown to provide superior performance to those of Alsop and Edgar.

Skogestad and Morari (1988) studied a high-purity distillation column whose details are shown in Table 2. Their approach is to base the controller design on a linear model. They used several controllers and analyzed their performances. However two controllers, simple diagonal PI-controller and μ optimal controller, have been chosen for comparison with our controller since the controllers show good performance. Figure 10a gives the closed-loop response to small setpoint change in X_B at operating point C. Our controller shows much better performance than other controllers. Figure 10b shows a transition from operating point A to operating point C using logarithmic compositions as controlled outputs. The logarithmic compositions are defined as follows:

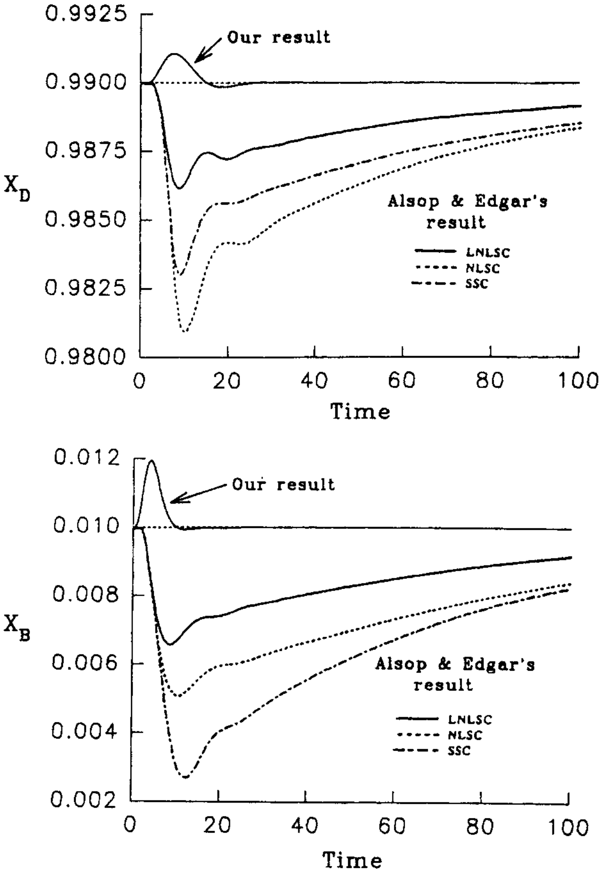


Figure 9. Comparison of control performances: (X_D, X_B for a feed composition change of -0.3 (see Figure 13, Alsop and Edgar, 1990).

$$Y_D = \ln(1 - y_D) \tag{18}$$

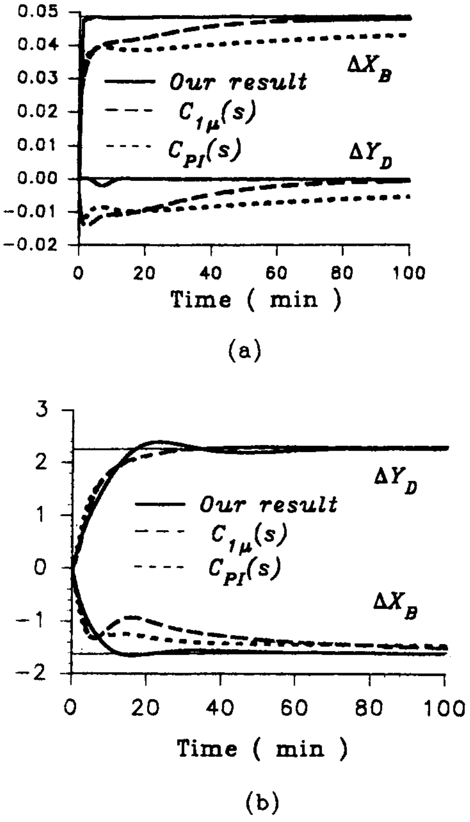
$$X_B = \ln x_B \tag{19}$$

The closed-loop responses of all controllers seem to be very good. However our controller brings the logarithmic compositions to the setpoints sooner than the other two controllers.

Fuentes and Luyben (1983) reported that a binary system with high relative volatility ($\alpha = 4$) is more difficult to control

Table 2. Steady-State Data for Distillation Column at Operating Points A and C (Skogestad and Morari, 1988)

Column Data:		
Relative Volatility		$\alpha = 1.5$
Number of theoretical trays		$N = 40$
Feed tray (1 = reboiler)		$N_F = 21$
Feed compositions		$X_F = 0.5$
Operating Variables:		
A	C	
$y_D = 0.99$	0.90	
$x_B = 0.01$	0.002	
$D/F = 0.5$	0.555	
$L/F = 2.706$	2.737	
Measurement tray: 16, 19, 22, 25, 28		



(a)

(b)

Figure 10. Comparison of control performances (logarithmic compositions as controlled outputs): (a) closed-loop response to small setpoint change in X_B of column C from 0.002 to 0.0021 (see Figure 11A, Skogestad and Morari, 1988); (b) transition from operating point A to C (see Figure 12, Skogestad and Morari, 1988).

than systems with low relative volatility ($\alpha = 2$). To test the applicability of the proposed controller to high relative volatility mixtures, we have selected the column *C* of Georgiou, Georgakis and Luyben (1988) which is very high purity column with high relative volatility ($\alpha = 4$) and tested by Fuentes and Luyben (1983). The column details are listed in Table 3. The column is known to be so nonlinear that modeling is very difficult. The regulatory responses of our controller and those of Georgiou et al. (1988) for a 5% step change in the feed composition are shown in Figure 11. The performance of our controller is quite good and better than the performance of the nonlinear dynamic matrix control (NLDMC) and the control configuration using *L* and *V* (LV) of Georgiou et al.'s. The three examples support that the proposed control scheme can improve the control of the high purity distillation column quite substantially.

Conclusion

A model-based control strategy based on a nonlinear wave theory and the GMC has been developed and applied to the profile position control of high-purity distillation columns. A nonlinear profile position observer has also been developed and proven to estimate the profile position of the column section with sufficient accuracy.

Table 3. Steady-State Design for Column C (Georgiou et al., 1988)

Composition (mol frac.)	Distillate	0.99999
	Bottom	0.00001
	Feed	0.5
Flow rates (kmol/min)	Feed	61.7
	Distillate	30.85
	Bottoms	30.85
	Reflux	21.579248
	Vapor	52.429248
Relative volatility		4
Trays (from bottom)	Total	40
	Feed	16
Holdups (kmol)		
Base		466 (5 min)
Trays in stripping section		28.04
Trays in rectifying section		19.75
Reflux drum		466
Tray hydraulic time constant		0.144 (min)
Measurement tray: 9, 10, 11, 19, 20, 21, 22		

The GMC controller works directly on the rate of the profile propagation. The tight control of the profile position helps overcome the nonlinearity of high-purity distillation columns. Before the sharp-profile region approaches the end of the column and it affects the concentration of products, the controller changes the ratios of the flow rates to prevent the column from dropping into the nonlinear region. To overcome the slight offset caused by GMC alone and the significant dead time in the composition analyzer, which degrades the performance of direct control schemes, the composition/profile position cascade configuration has been introduced.

The performance of the derived controller has been evaluated through nonlinear dynamic simulation of high-purity binary distillation columns. It seems that the controller is robust over a wide range of process nonlinearity and model-plant mismatch and has relatively wide tuning bands.

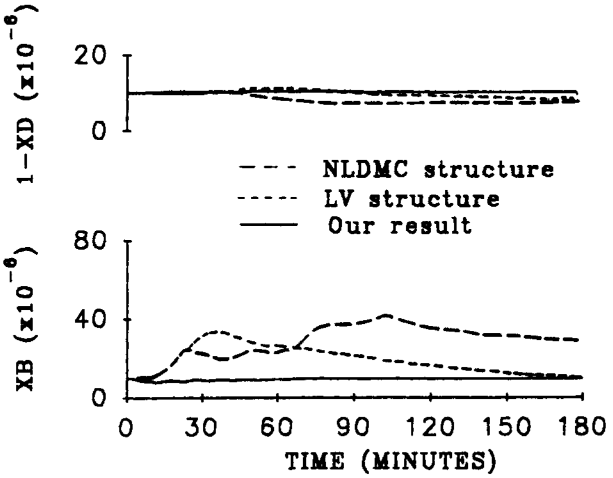


Figure 11. Closed-loop rigorous nonlinear dynamic simulation for very high-purity column C and +5% step change in feed concentration (0.5→0.525) (see Figure 15, Georgiou et al., 1988).

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Notation

B = molar flow rate of bottom product
$C_{I\mu}(s)$ = μ optimal controller
$C_{PI}(s)$ = simple diagonal PI controller
D = molar flow rate of distillate product
F = molar flow rate of feed
G = molar vapor holdup of plate
H = molar liquid holdup of plate
K = GMC controller constant
k = index of tray counted from bottom
L = molar flow rate of liquid stream
N = number of trays of entire column
q = liquid mole fraction of the feed
r = molar holdup ratio of vapor to liquid, $r \equiv G/H$
S = profile position of column section
u = normalized wave velocity for specific composition
u_Δ = normalized shock wave velocity of self-sharpening wave
V = molar flow rate of vapor stream
x = liquid mole fraction of light component
x_s = representative liquid mole fraction corresponding to profile position of self-sharpening standing wave
y = vapor mole fraction of light component or measured variable

Greek letters

α = relative volatility
Δ = prefix for difference between the two sides of self-sharpening wave
σ = normalized distance from bottom of the column, $\sigma = k/N$
τ = normalized time, $\tau = tF/NH$

Superscripts

$*$ = setpoint
$\hat{}$ = observer prediction
$\cdot$ = derivative

Subscripts

F = feed
i = tray i
R = rectifying section
S = stripping section
1 = rectifying section
2 = stripping section

Acronyms

GMC = generic model control
LNLSC = linearized nonlinear structural compensator
LV = energy balance control by manipulating L and V
NLDMC = nonlinear dynamic matrix control
NLSC = nonlinear structural compensator
SSC = static structural compensator

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