Rigorous Dynamics and Feedforward Control Design for Distillation Processes

Feedforward control has the theoretical potential for perfect control; therefore, research to simplify and improve the use of such a control strategy for possible industrial applications has potential value. A general model for simulation of controlled distillation columns is used to reproduce operating trends of some monitored variables for an industrial column. Comparisons between simulated and operating data show a general good agreement.

When a rigorous model is available and can be proved reliable for practical purposes, the design of feedforward control schemes becomes viable and can be incorporated in the more conventional approach. A specific industrial example shows the practical implementation problems and the real economic value of a feedforward control strategy to optimize the process behavior. In particular, the feedforward control action can reduce the inherent error from changes in feed composition when a basic feedback control structure is used to infer composition.

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Introduction

In distillation processes using feedback control loops, load variables enter the process and controlled variables are measured at different points. The longer the distance or the process residence time separating these points, the more difficult it will be to maintain the controlled variables at the desired set points. Therefore a more effective control approach becomes necessary when process residence dead times are large, when load upsets are frequent, and when very high bottom or top purities are required. In these cases, a relationship among load variables, manipulated variables, and desired set points of the controlled variables must be found in such a way that load variables directly affect the manipulated ones.

This approach, usually termed feedforward control system, evaluates the proper value for the manipulated variables in order to cancel the effects of input variations on the process. Consequently, a feedforward control system requires a proper knowledge of the process or a sound mathematical model for steady state and dynamic conditions.

Despite its potential for perfect control, feedforward control suffers from several inherent weaknesses, in particular:

- It requires identification of all the possible disturbances acting on the system, but these measures are not always practically available
- any change of internal parameters of the process cannot be compensated by a feedforward controller if their impact cannot be correctly modeled (e.g., reduction of the heat transfer coefficient due to fouling)

On the other hand, feedback control is rather insensitive to both of these drawbacks but has poor performance for some control objectives. Therefore a combined feedforward-feedback control systm can retain the superior performance of the former and the relative insensitivity of the latter to uncertainties and inaccuracies. As a matter of fact, any deviations caused by the possible weakness of the feedforward control can be corrected by the feedback controller.

Figure 1 shows two possible configurations of a combined feedforward-feedback control. Both include three main elements: model control computation, feedback correction, and dynamic compensation. In Figure 1a the feedback action is added to the feedforward computed value of the corresponding manipulated variable. Depending on the nature of the model computation, the feedback correction can also provide adjustment to the model coefficients in order to improve the control performance (Badawas, 1986; Congalidis et al., 1989). Alternatively, for frequent and large changes of the major load variables, (MacMullan and Shinskey, 1984), the feedforward control strategy is used to provide new process set points that satisfy required

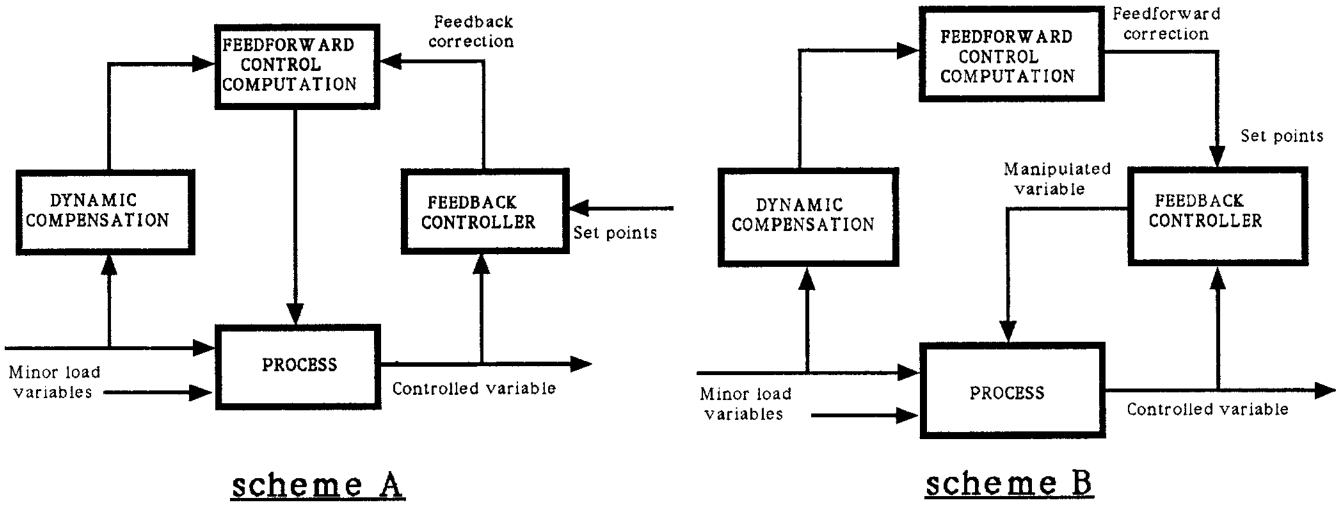

Figure 1. Components of feedforward control schemes.

control objectives, Figure 1b. In both cases a proportional-integral is usually sufficient for the feedforward correction.

As far as dynamic compensation is concerned, it can be observed that variables enter the process at different locations. This causes a dynamic imbalance between the two different actions of the incoming disturbances and the corresponding manipulated variable variation. Therefore, dynamic compensation in the form of lead/lags or dead time is required to improve the dynamic performance of a steady-state feedforward model and also to reduce computational errors due to modeling simplifications. As a matter of fact, even small time-constant errors between a predictive model and a real process can introduce a dynamic imbalance and a corresponding less-accurate control action. This has been in practice one of the main deterrents to the use of advanced feedforward control schemes. Over the last 20 years, all the different industrial applications have been developed on the basis of simplified or linear dynamic model and usually verified by a more or less complex nonlinear state-space model (Congalidis et al., 1989). However, no significant efforts have been addressed in the literature toward a design procedure based on a rigorous dynamic analysis. Conversely, in the same period the use of advanced control techniques and various aspects of digital control has grown tremendously. Moreover, during the last

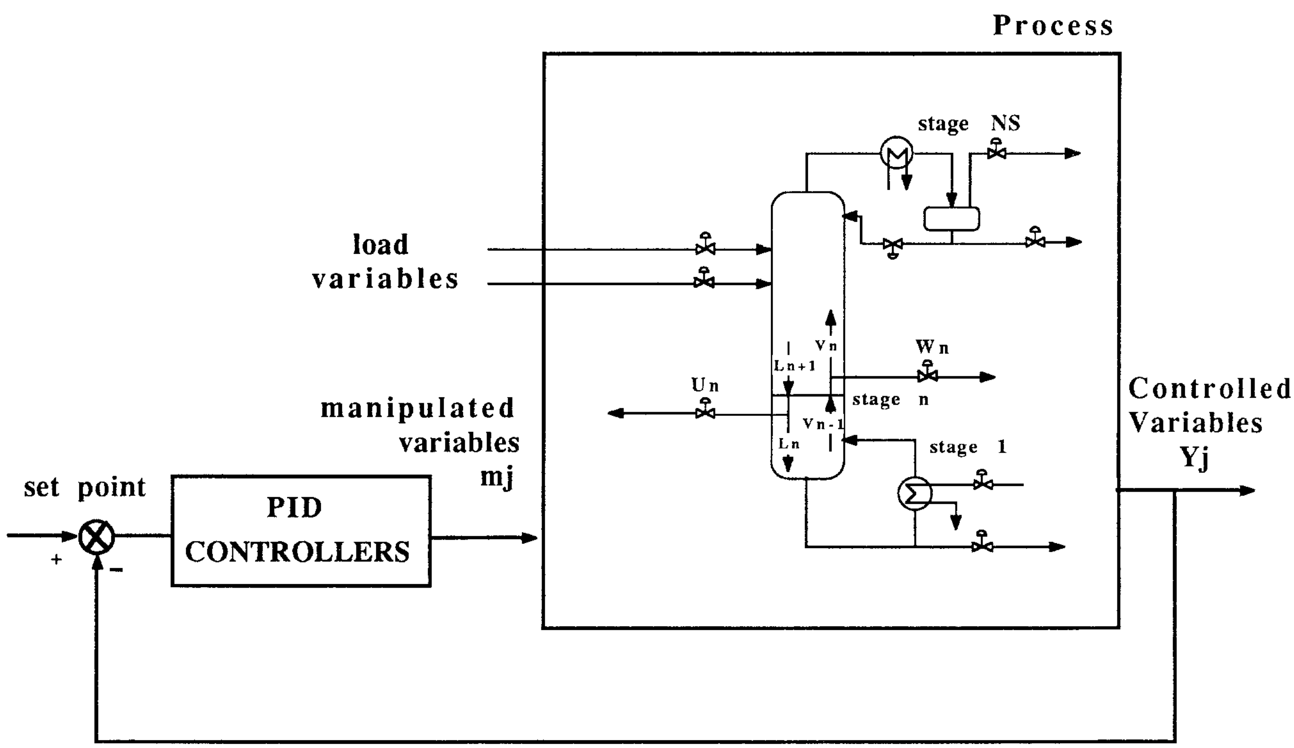

Figure 2. MIMO control structure for a distillation column.

decade many papers have been published on the dynamic behavior of chemical plants.

Since feedforward control has the theoretical potential of perfect control, it seems to be worthwhile to join all the different efforts to simplify and improve this control strategy for industrial application. However, the main problem still remains the availability and reliability of a rigorous model for the different units. Where a rigorous model is available and can be proved reliable for practical purposes, the design of feedforward schemes becomes viable and can be incorporated in the more conventional control strategies.

In the following the design of feedforward control schemes for a distillation column on the basis of rigorous dynamics is addressed. In order to emphasize the previous analysis and to schematize the theoretical and practical frame of the problems, this paper discusses and analyzes the following main points:

- model definition
- experimental validation
- feedforward control design

Finally, an industrial application shows the practical implementation and the real economic value of a feedforward control strategy to optimize the process behavior.

A Global Nonlinear State-space Model

For feedback control purposes, distillation columns have been reported in literature mainly in terms of simple models that describe the effect of input flow on output product composition, level, temperature, and so on (Rosenbrock, 1962; Weingard et al., 1972; Skogestad and Morari, 1988). Usually, the resulting model is a linear relationship between a set of input and output variables, easy to manipulate but including all the weaknesses related to its simplifications. Therefore, the continuously rising computation power of supermini computers, together with the resulting low cost of computation, has expanded tremendously the practical interest in rigorous dynamic analysis. During the last decade, the dynamic behavior of distillation columns has been investigated, with a constant trend toward using more and more sophisticated models to obtain deeper insight and a better understanding of both the phenomena that occur and expected performance (Boston et al., 1981; Gallun and Holland, 1982; Gani et al., 1986). Following this general approach, and with reference to the schematic representation of the controlled column shown in Figure 2, it is easy to derive the model equations, Eqs. 1–8, reported in Table 1.

Every stage includes liquid and vapor side streams, one feed flow rate, and one thermal duty. The overall column is constituted by NS real or theoretical stages numbered from the reboiler up to the condenser. In the formulation of the model, the following simplifications have been assumed:

a. Liquid and vapor phases at each stage are well mixed
b. Capacitance terms due to downcomers are neglected
c. Vapor holdup over the plate is disregarded

Let N be the number of components; then a set of $(2N + 5)$ equations can be written for each stage. This set applies identically for all the stages.

Table 1. Model Equations

$N \times NS$, Material balances

$$\frac{dM_{i,n}}{dt} = L_{n+1}\frac{M_{i,n+1}}{\Sigma_i M_{i,n+1}} + V_{n-1}y_{i,n-1} - (V_n + W_n)y_n - (L_n + U_n)\frac{M_{i,n}}{\Sigma_i M_{i,n}} + F_n Z_{i,n} \tag{1}$$

NS, Energy balances

$$\frac{dE_n}{dt} = HL_{n+1}L_{n+1} + HV_{n-1}V_{n-1} - HV_n(V_n + W_n) - HL_n(L_n + U_n) + HF_nF_n + Q_n \tag{2}$$

$$\text{where} \begin{cases} Q = 0 \quad \text{for} \quad n \neq 1 \quad \text{and} \quad n \neq NS \\ \text{or} \\ Q_n = U_{eff.}A_{ex}\,\text{LMTD} \quad \text{for} \quad n = 1 \quad \text{and} \quad n = NS \end{cases}$$

$N \times NS$, Vapor-liquid relationships

$$y_{i,n} = \epsilon_{i,n}K_{i,n}\left(T_n, P_n, \frac{M_{i,n}}{\Sigma_i M_{i,n}}, y_{i,n}\right)\frac{M_{i,n}}{\Sigma_i M_{i,n}} + (1 - \epsilon_{i,n})y_{i,n-1} \tag{3}$$

$2NS$, Fluid dynamic relationships

$$M_n = f(L_n, V_n, \text{geometry}) \tag{4}$$

$$P_n = P_{n+1} + \Delta P_{n+1}(L_n, V_n, \text{geometry}) \tag{5}$$

NS, Boiling points

$$\Sigma_i y_{i,n} - 1 = 0 \tag{6}$$

NS, Definitions of the energy function

$$E_n = HL_n \Sigma_i M_{i,n} \tag{7}$$

PID control equations

$$m_j = m_j^0 + K_{cj}\left\{(Y_j^{set} - Y_j) + \frac{1}{\tau_{I_j}}\int (Y_j^{set} - Y_j)\,dt + \tau_{D_j}\frac{d(Y_j^{set} - Y_j)}{dt}\right\} \quad \text{for} \quad j = 1 \ldots NC \tag{8}$$

N, number of components; NS, number of stages; NC, number of controllers.

The condenser and reboiler require further specifications to describe their thermal duties and/or fluid dynamic relationships. Details about modeling and numerical aspects of the resulting system can be found in previous papers (Rovaglio et al., 1986; Ranzi et al. 1987).

Finally, indicating with NC the number of the controlled variables and defining the appropriate control configuration, NC general PID control equations can be added to the global system. These manipulated variables, if necessary together with $2NC$ upper and lower constraints, become internal variables of the overall differential/algebraic system.

This system constitutes the nonlinear state-space model and allows one to describe servo or regulator problems for a distillation system. In this way, the rigorous simulation of a controlled distillation column is reduced to the solution of the global system of differential/algebraic equations. Once again, it is worthwhile observing that the use of implicit integration methods based on backward differentiation formulas is strongly suggested and/or becomes necessary in order to avoid an excess of computing time. As discussed elsewhere (Ranzi et al., 1988) this global approach, very effective for the previously mentioned season, has an inherent weakness in requiring continuity of the different functions and derivatives of the whole system.

This limitation, really effective in the case of the analysis of start-up and shut-down procedures (Gani and Ruiz, 1987), does not influence the numerical procedure in the case of lower or upper constraints on the manipulated variables. Based on our experience, not only limited to the presented applications, the LSODES package (Hindmarsh, 1983), properly modified for the solution of algebraic equations, has proved to be both robust and reliable.

Experimental Validation of the Model

The distillation process studied is a 26-stage deethanizer column. The feed is a saturated liquid introduced on the 17th tray. The column has a partial condenser with total liquid reflux, while the distillate leaves the reflux drum as a vapor flow. The column and its control scheme are represented in Figure 3. Main geometric characteristics are reported in Table 2. There are four controlled variables:

- temperature on stage
- reboiler level
- top pressure
- accumulator level

The corresponding manipulated variables are:

- reboiler hot-oil flow rate
- bottom flow rate
- vapor distillate
- reflux flow rate

The system is equipped with a multiloop control scheme consisting of one PID for composition control (TC) and three PI for liquid-levels (LC) and pressure control (PC). Control parameters operating on the industrial column are reported in Table 3.

The comparison between experimental data and predicted results refer to an increase of about 15% of the feed flow rate. Operating conditions and feed composition are reported in Table 4, while a direct comparison between the real disturbance acting on the column and the simulated one is shown in Figure 4. The temperature transients for the third tray (stage 4) and the reflux drum are reported respectively in Figures 5a and 5b. These

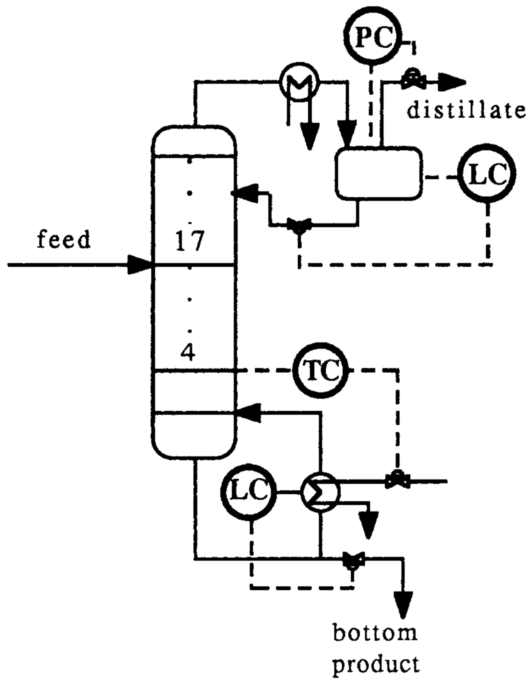

Figure 3. Deethanizer column with conventional control scheme.

comparisons show a very good agreement between experimental and predicted values; also, the bottom temperature, although not reported here, is confirmed to be practically constant (about 108°C). A more significant comparison can be found in Figures 5C and 5D where vapor distillate and bottom flow rates are shown. Unfortunately, the last variable was available in a 24 h recording while the considered disturbance acts only in the last two hours. However, a very similar trend can be recognized easily: the simulation results show the same overresponse and the same time constant as the operating ones.

For the sake of clarity, it must be emphasized that the reported comparisons were obtained on the basis of several minor assumptions related to the practical difficulty of finding detailed and complete operating data describing the dynamic transient of the whole system. Nevertheless, neither parameter

Table 2. Geometry of Deethanizer Column

		Lower Section*	Upper Section**
Tray layout			
Tower diameter		2.43 m	1.35 m
Valves (V1 ballast)		303	140
Tray spacing		0.61 m	0.61 m
Efficiency		0.60	0.75
Reflux accumulator geometry			
Length	3.6 m		
Diameter	0.9 m		
Kettle reboiler geometry			
Free volume	5.8 m^3		
Diameter	1.0 m		
Length	5.5 m		

*Split flow configuration, 16 valve trays.
**Crossflow configuration, 8 valve trays.

Table 3. Control Parameters

Loop	Gain	Time, min Constants	
		Integral	Derivative
Temperature	2	4	0.25
Bottom level	2	10	0
Accumulator level	2	10	0
Pressure	20	3	0

adjustments nor special techniques were employed for the purpose of tuning the rigorous model Eqs. 1–8, to fit the experimental data. Therefore, the reported results arise only from the predictive capacity of the model itself. Moreover, the general agreement observed between the operating and simulated trends, also obtained by maintaining the original operating settings for the control parameters, clearly indicates that rigorous dynamics of distillation systems can be of real value to verify and improve control design procedures.

Feedforward Design Procedure

On the basis of the global process within which the column operates, and from the economic point of view, the operational target of the implemented control strategy can be defined as keeping the ratio between the ethane mole fraction and the total C_3 mole fraction in the bottom product at a maximum permissible value. This allows maximizing the liquid propane product yield and, at the same time, minimizing the reboiler duty.

As can be observed in Figure 6, the operating control system shown in Figure 3 does not satisfy such a control objective. In fact, imposing a ramp disturbance, which modifies in a period of half an hour the feed composition through the following mole fraction variations:

$$C_3 \text{ from } 0.3040 \text{ to } 0.3640$$
$$C_4 \text{ from } 0.3460 \text{ to } 0.2860$$

the column shows a large variation (around 75%) in terms of the molar ratio C_2/C_3 in the bottom product. Similar large variations still exist even with higher control settings.

In an effort to obtain performance beyond that available with conventional techniques, a feedforward control scheme was investigated as a possible solution. In a feedforward control system, the principal load sources are measured and introduced

Table 4. Operating Conditions

Feed flow rate, tray 17	10.77 kmol/min
Feed temp.	300.0 K
Feed composition, mol fract	
Methane	0.1016
Ethane	0.1130
Propylene	0.0290
Propane	0.3000
i-butane	0.1104
n-butane	0.3460
Partial condenser	
Total liquid reflux	2.10 kmol/min
Vapor distillate	3.39 kmol/min
Condenser pressure	30.40 atm
Inlet temp., cooling water	295 K
Inlet temp., hot oil	505 K

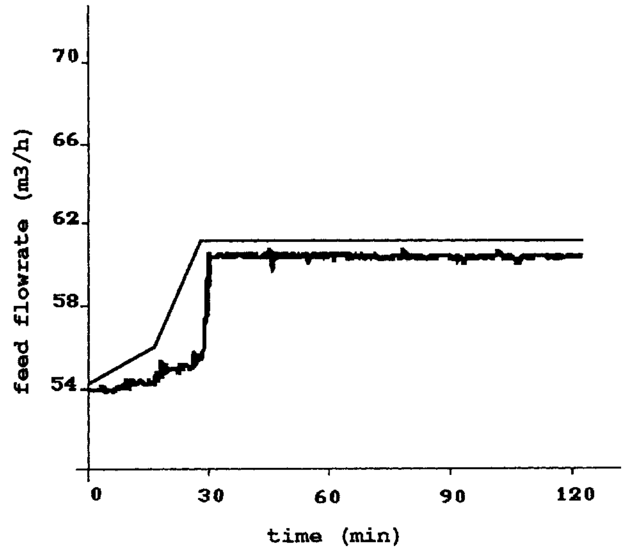

Figure 4. Disturbance imposed on feed flow rate.
— simulation data.
Irregular line, experimental data

into a computer. A computer program calculates the exact value of the manipulated variable that will cancel the effects of load variations on the process.

The exact mathematical formulation of the control problem to be solved can be strictly related to the model predictive control technique (Garcia and Prett, 1983) and consists of a general multiobjective optimization algorithm subject to multiple constraints, with the ability to distinguish and to handle hard and soft constraints. As a matter of fact, on the basis of the global model given in Table 1 and on the basis of the global solution of this system, it is possible to assume one or more particular set points as new internal variables by adding to the system one or more specification equations that completely define the required operation targets.

For the examined column, the objctive function can be formulated as follows: Compute the set point of the temperature loop, Figure 7, necessary to maintain constant the molar ratio C_2/C_3 in the bottom product. This control objective allows superimposing the feedforward control strategy on a specified feedback control scheme. In this way it will be possible to augment the specification requirements without decreasing the robustness and the performance of the implemented control scheme.

As already mentioned, every feedforward control strategy has to be related to specified disturbances acting on the system and specifically monitored. Usually, these disturbances can be defined as the most frequent or as the process variations that mainly determine a large offset from the control objective. In the case studied, both these features are summarized in feed composition disturbances.

Figures 8 and 9 show the simulated system responses for the composition disturbance already specified for Figure 6 and compare the column performances when:

- only the feedback control structure is implemented, curves (*i*)
- the feedforward-feedback structure is introduced, curves. (*ii*)

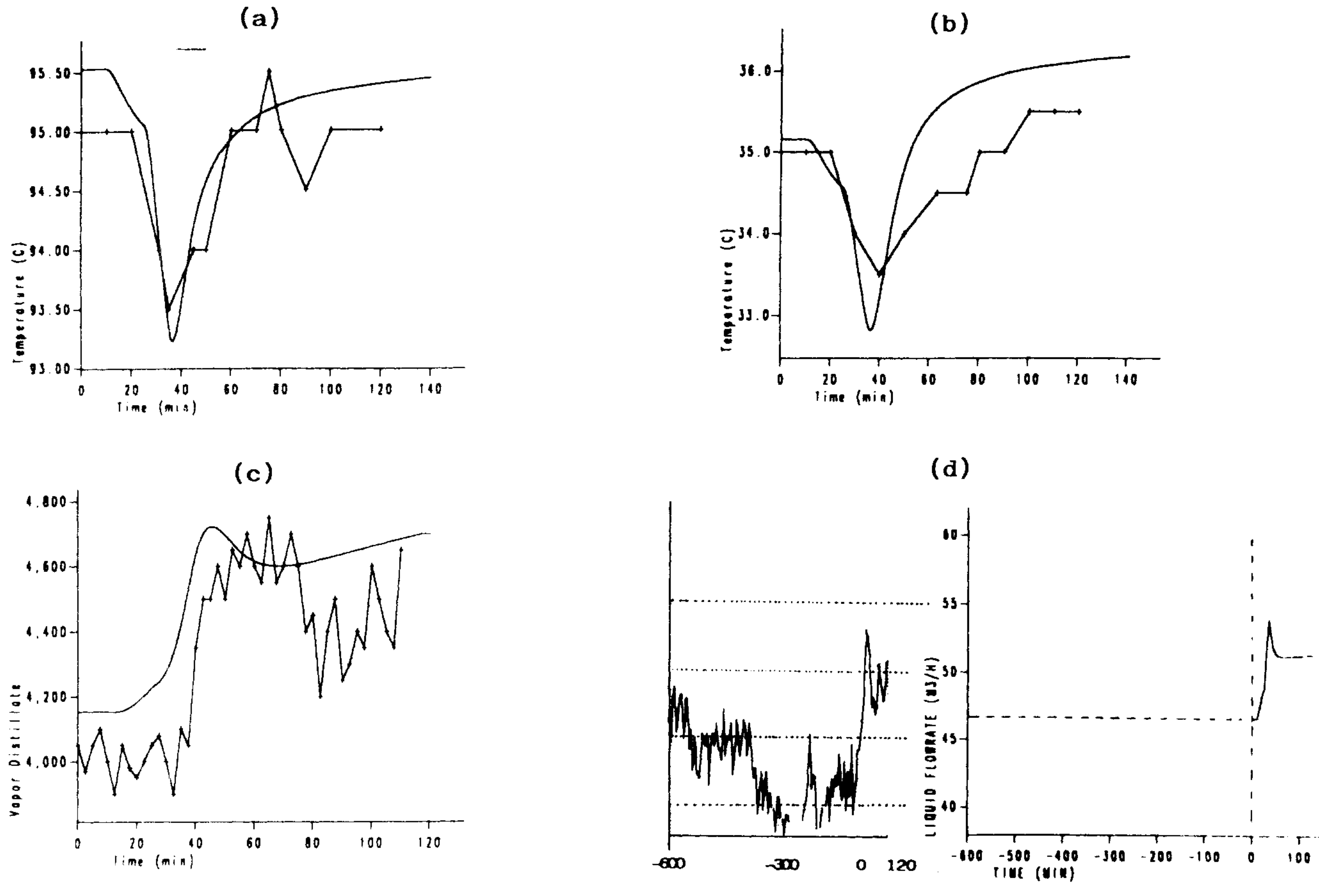

Figure 5. Dynamic evolution of some variables.

---- predicted data; + + + + experimental data

a. Temperature on tray 4
b. Temperature on reflux drum
c. Vapor flow rate from reflux drum
d. Bottom flow rate, m^3/h

From the selected variables it can be observed that, with a rigorous constant value for the C_2/C_3 molar ratio, the column behaves as more stable from the fluid dynamic point of view. The main difference is concentrated in the bottom of the column in terms of compositions and temperature variations. However, in all the different variables, it seems to be confirmed that the presence of a "perfect" feedforward control drastically reduces the typical time constant of the system. As matter of fact, even in the case of larger variations—see the temperature and propane molar fractions on the reboiler in Figure 9—the column reaches new steady state in a period practically coinciding with the evolution of the assumed disturbance.

Practical Implementation

If a sufficiently powerful computer is available for process control (for example a supermini computer), then all the previous considerations can be directly implemented as a control strategy for the examined column. Assuming it is possible to measure continuously the feed composition, it would be easy and fast to compute on-line the exact value for the temperature set point on the basis of the global rigorous model and the corresponding objective function. Unfortunately, in most cases the available hardware does not allow on-line solution of optimization problems with detailed dynamic process representation. For this reason, the previous rigorous synthesis of a feedforward control scheme may have to be reviewed and restructured on the basis of a simplified control law.

In order to derive this simplified law, several steady-state simulations of the column were run, varying the relative composition of the feed, the different target ratios of ethane to propane in the bottom product. Regressing the resulting pilot tray temperatures with the parameters mentioned, the correlation in Eq. 9 was obtained:

$$T_{set}\,(\text{tray 4}) = K_1 - K_2\frac{C_{2b}}{C_{3b}} + \frac{K_3 + K_4 C_{2b}/C_{3b}}{1 + K_5\dfrac{C_{3f}}{C_{4f}} - K_6\dfrac{C_{2f}}{C_{4f}}} \tag{9}$$

where

C_{2b}/C_{3b} = required ratio of ethane to propane in the bottom-product
C_{3f}/C_{4f} = ratio of propane to butane in the feed
C_{2f}/C_{4f} = ratio of ethane to butane in the feed
$K_1, K_2, K_3, K_4, K_5, K_6$ = numeric constants, defined as a function of temperature units and on the basis of composition (mole, volume, or weight fractions)

Components lighter than ethane and heavier than butane were assumed constant in obtaining the simplified correlation, Eq. 9. Similarly, the overhead temperature was assumed constant. This may be considered an oversimplification when the cooling fluid temperature is subject to sizable variations between day

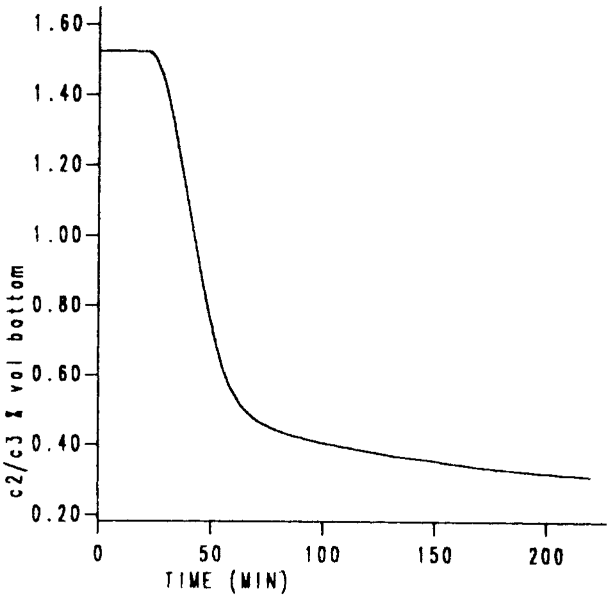

Figure 6. C_2/C_3 molar ratio in bottom product for feedback control only.

and night operation, and could be corrected by making coefficients K_3 to K_6 dependent on the overhead temperature. It is probably worthwhile to stress here that the importance of the overhead temperature in this type of column is due to its effect on the material balance. In fact, a higher overhead temperature would drive more propane to be lost in the distillate vapor product, and the ratio of C_3 to C_4 in the bottom section would consequently be reduced at constant feed composition. The

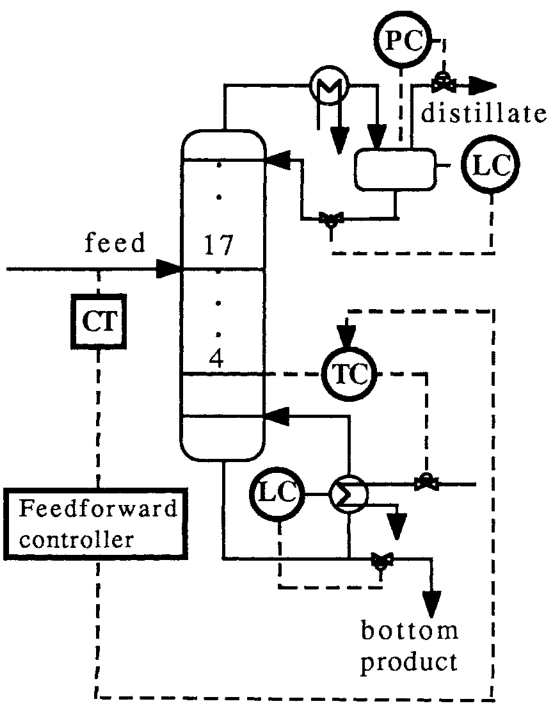

Figure 7. Feedforward control scheme for deethanizer column.

possible convenience of adjusting only coefficients K_3 to K_6 for changes in overhead temperature comes from the fact that these coefficients account for the effect of material balance, somehow converting the feed composition into a bottom composition.

Generally speaking, in devising a simplified correlation such as Eq. 9 the engineer should look carefully for the most physically significant variables in the sytem. Operating data should be studied to assess the range of variability of the main external factors and to identify those that cannot be controlled by the operation of the column itself. Subsequently, simulations of column performance under the variable external conditions should be run to check the sensitivity of the column to the factors themselves.

With Eq. 9, the real set point available for the operator is the value of the ethane to propane ratio required in the bottom product. The constant K_1, although fixed by the regression, is the first candidate for adjustment in the field during the tuning of the correlation to account for discrepancies between simulation and real operation.

In the context of the proposed simplified law, the practical importance of the rigorous dynamic model is to check the need for dynamic compensation that may have to be added to the simple Eq. 9. For this purpose, in Figure 10 the following responses are compared:

- set-point variation proposed by the simplified law, Eq. 9, curve (*i*)
- set-point variation defined by the solution of the rigorous model, curve (*ii*)

Both refer to the composition disturbance already specified for Figure 6. It can be observed that the curves practically coincide and this can be explained by the dominance of disturbance evolution with respect to process dynamics.

As a matter of fact, from the classical theory (Stephanopoulos, 1984) a feedforward control law can be defined in terms of the main transfer functions as follows:

$$G = \frac{G_d}{G_p G_m} \tag{10}$$

where G_d is the disturbance transfer function and G_m refers to measuring device. Assuming at this point $G_m = 1$, it is easy to understand the relative importance of different dynamic elements that compose Figure 10. Equation 10 seems to be practically reduced to:

$$G_c = G_d \tag{10'}$$

However, the relationship obtained, Eq. 10, cannot be generalized and depends strictly on the disturbance considered and on the inherent process characteristics.

For example, Figure 11 shows the set point variation defined by the rigorous model, Eqs. 1–8, and on the basis of the same control objective (constant C_2/C_3) when a feed flow rate disturbance is considered. Observing the figure, it is clear in this case that the suggested set point evolution cannot be related only to the disturbance dynamics, and process contribution G_p reveals its effects.

Although for the disturbance related to feed composition the process dynamics appear to be practically negligible, for a real check of the system performances the dynamic description of the measuring device must be included into the model. In fact, in the case of an analyzer the dynamic characterization can be reduced to a pure time delay, which defines a sampled description of the

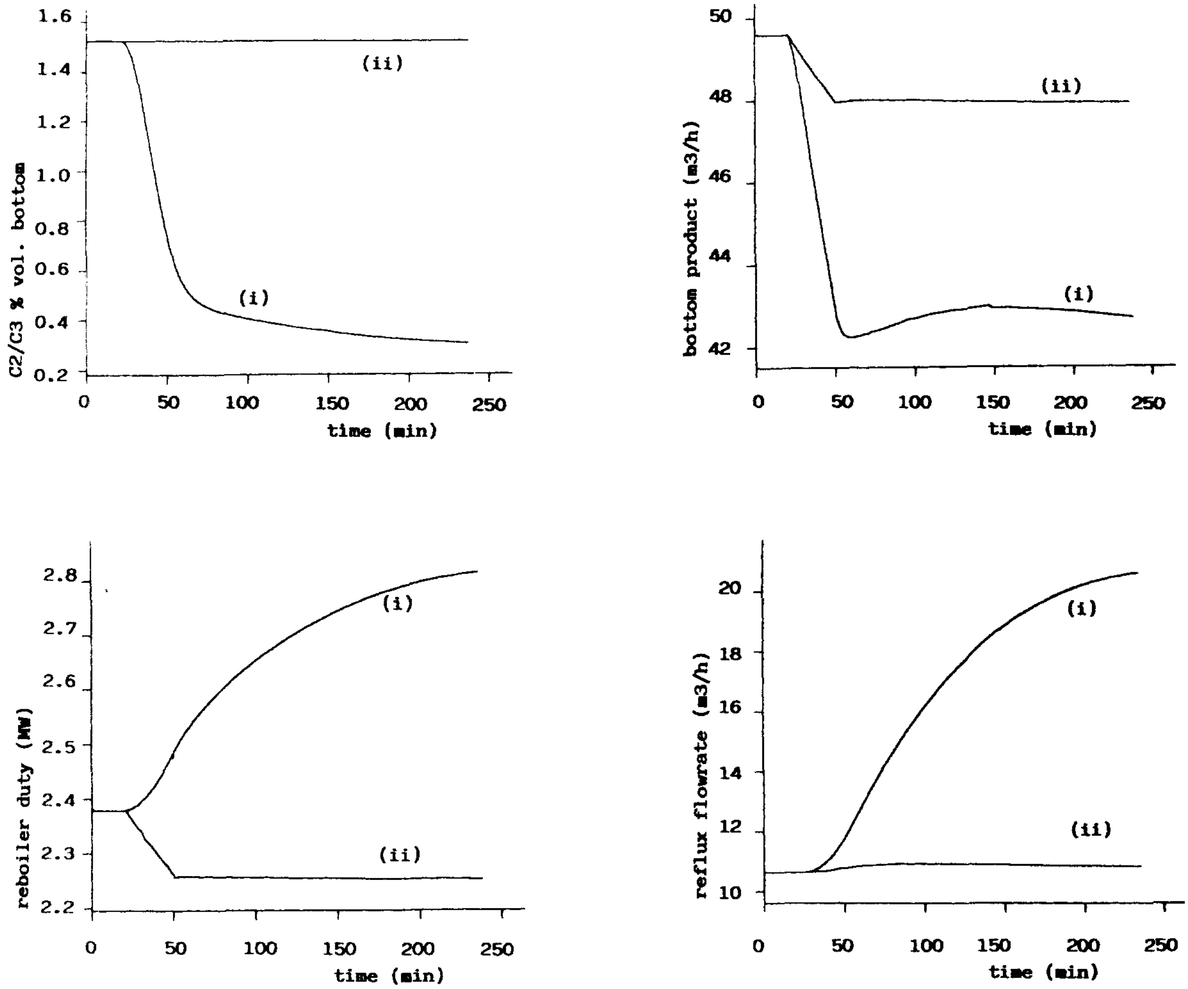

Figure 8. Control objective, heating, and fluid dynamic behavior of column.

i. Feedback control only
ii. Feedback-feedforward control

feed composition. However, this delay cannot be neglected. Figure 12 shows the set point variation proposed by the simplified control law, Eq. 9, when a sampling period of 5 min is introduced for the measuring device. This is known as zero-order continuous representation of a discrete-time signal.

However, the zero-order hold does not represent the only way to construct a continuous signal from its discrete-time values. As a matter of fact, for slowly varying signals (e.g., the composition), due to the almost constant slope of the changing variable over a large period, the superiority of a first-order hold is obvious.

With this description, the continuous signal can be given by linear extrapolation of two previous values:

$$T_{set}(t) = T_{set}(nP) + \frac{T_{set}(nP) - T_{set}[(n-1)P]}{P}(t - nP)$$

$$\text{for} \quad nP \leq t \leq (n+1)P \quad \text{and} \quad n = 1, 2, 3 \ldots \quad (11)$$

The coupled use of Eqs. 9 and 11 allows the definition of a good feedforward control action very similar to the ideal one. In Figure 13 the theoretical simplified feedforward control (not delayed) is compared with itself when a zero-order or a first-order hold is also applied. The system response is given in terms of control objective, and for reference it has been also reported for the case of conventional control. The real and practical importance of the choice in this case is quite evident: the best column performances will be obtained by a first-order hold until the sampling signal can be considered not really corrupted with noise, otherwise a conservative continuous representation (zero-order hold) will take place.

Finally, using the global rigorous model it is possible to check the real system performance with respect to the simple control law now derived, Eqs. 9 and 11, and/or to different process disturbances.

For example, Figure 14 shows the dynamic behavior of some selected variables when the set point variation, defined by Eqs. 9 and 11, is simulated by the rigorous model. The load change imposed on the sytem is the composition disturbance already previously mentioned. Comparing Figure 14 with Figures 8 and 9, it can be observed that the simulated process behavior obtained working with a simplified law practically coincides with a rigorous solution of the problem (curves *ii* in Figures 8 and 9). This confirms the practical value of the derived simple control law.

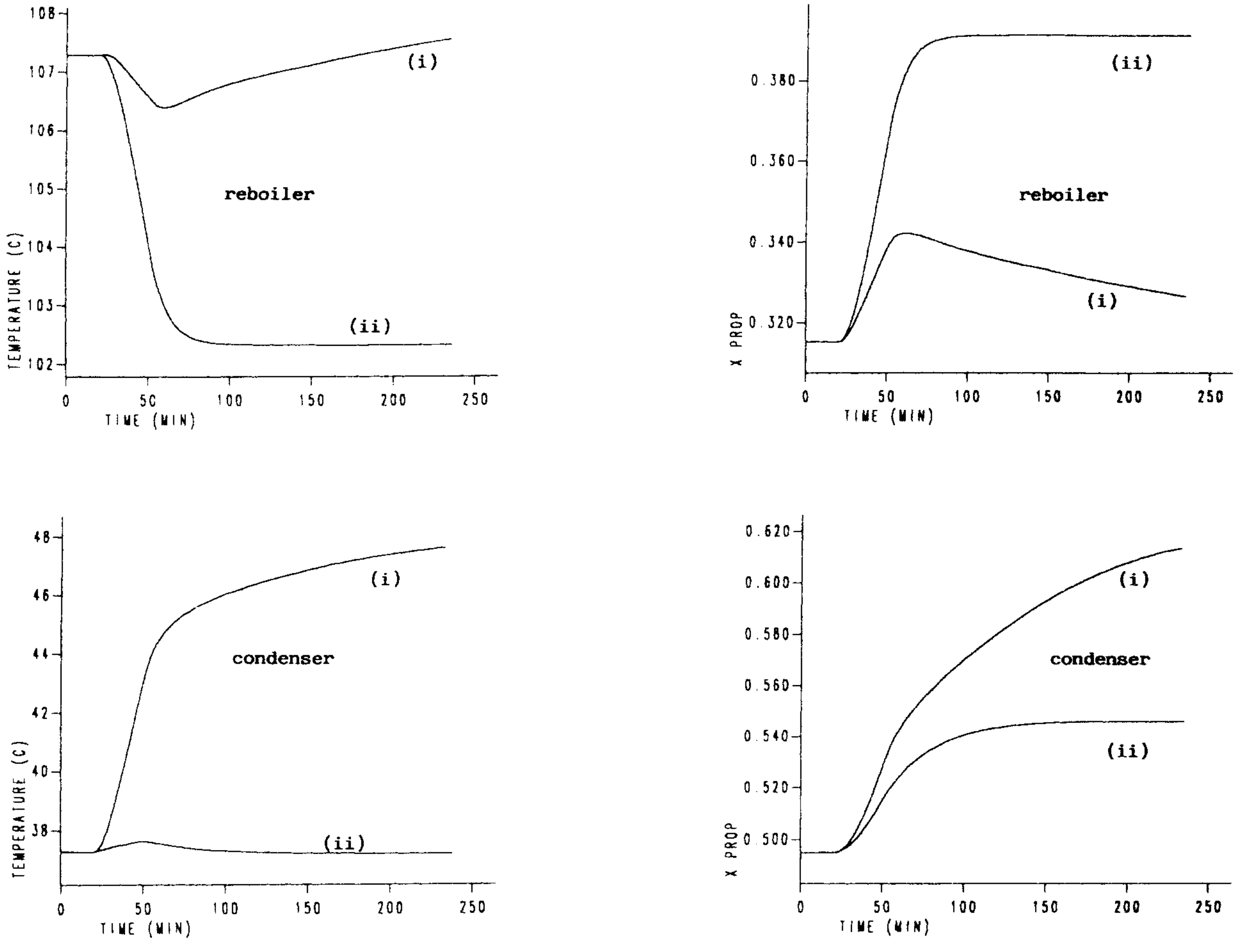

Figure 9. Temperatures and propane mole fractions vs. time.

i. Feedback control only
ii. Feedback-feedforward control

Conclusions

A general model for dynamic simulation of controlled distillation columns has been used to reproduce operating trends of some monitored variables for an industrial column. The quality of behavior reproduction has induced us to extend the use of the model to define improved control strategies on the basis of a general feedforward design procedure:

- Define the control objectives for a given process.

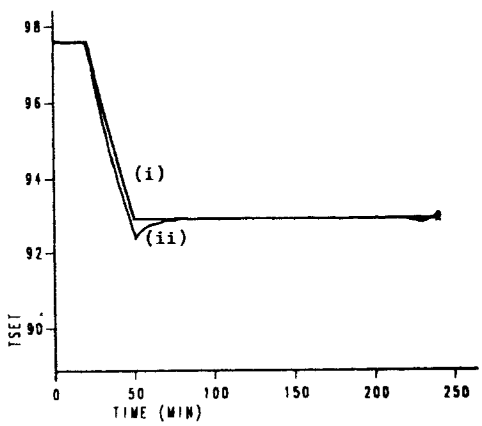

Figure 10. Proposed set point variations.

i. Simplified model
ii. Rigorous model

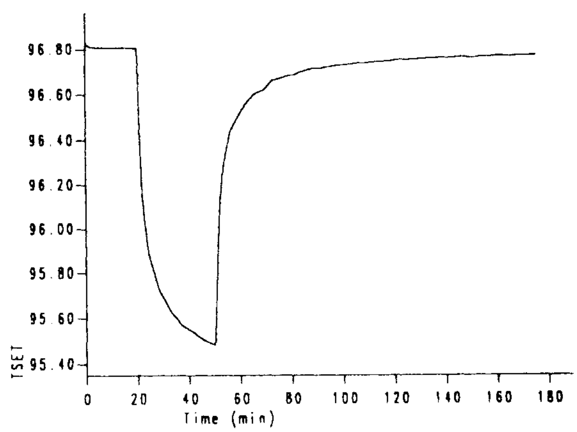

Figure 11. Set point variation proposed by rigorous model for a 10% feed flow rate disturbance.

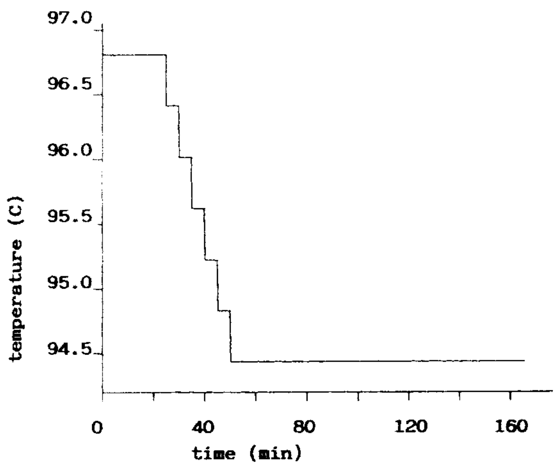

Figure 12. Setpoint variations defined by simplified control law, Eq. 9, and a zero-order hold.

• Identify the main process disturbances acting on the system.

• Identify the degrees of freedom (set points or manipulated variables) that can be used to satisfy the required control objectives.

Therefore, on the basis of the global model, it is possible to compute the rigorous trends of the controlling variables to cancel out the effects of the main disturbances, taking into account the measuring device dynamics.

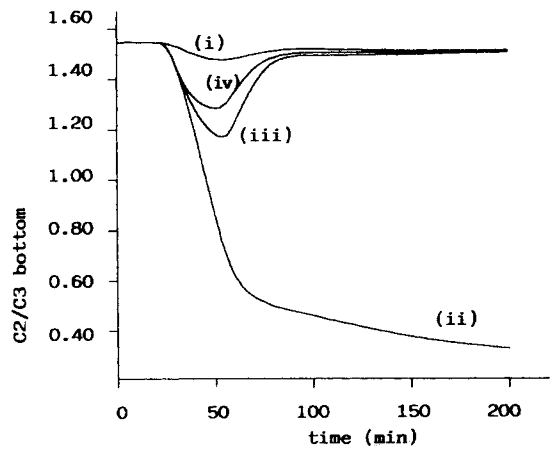

Figure 13. C_2/C_3 molar ratio in bottom product vs. time.

i. Simplified control law, Eq. 9
ii. Feedback contro only
iii. Eq. 9 and zero-order hold
iv. Eq. 9 and first-order hold

The tailored global model setup can be directly implemented on-line if a sufficiently powerful process computer is available.

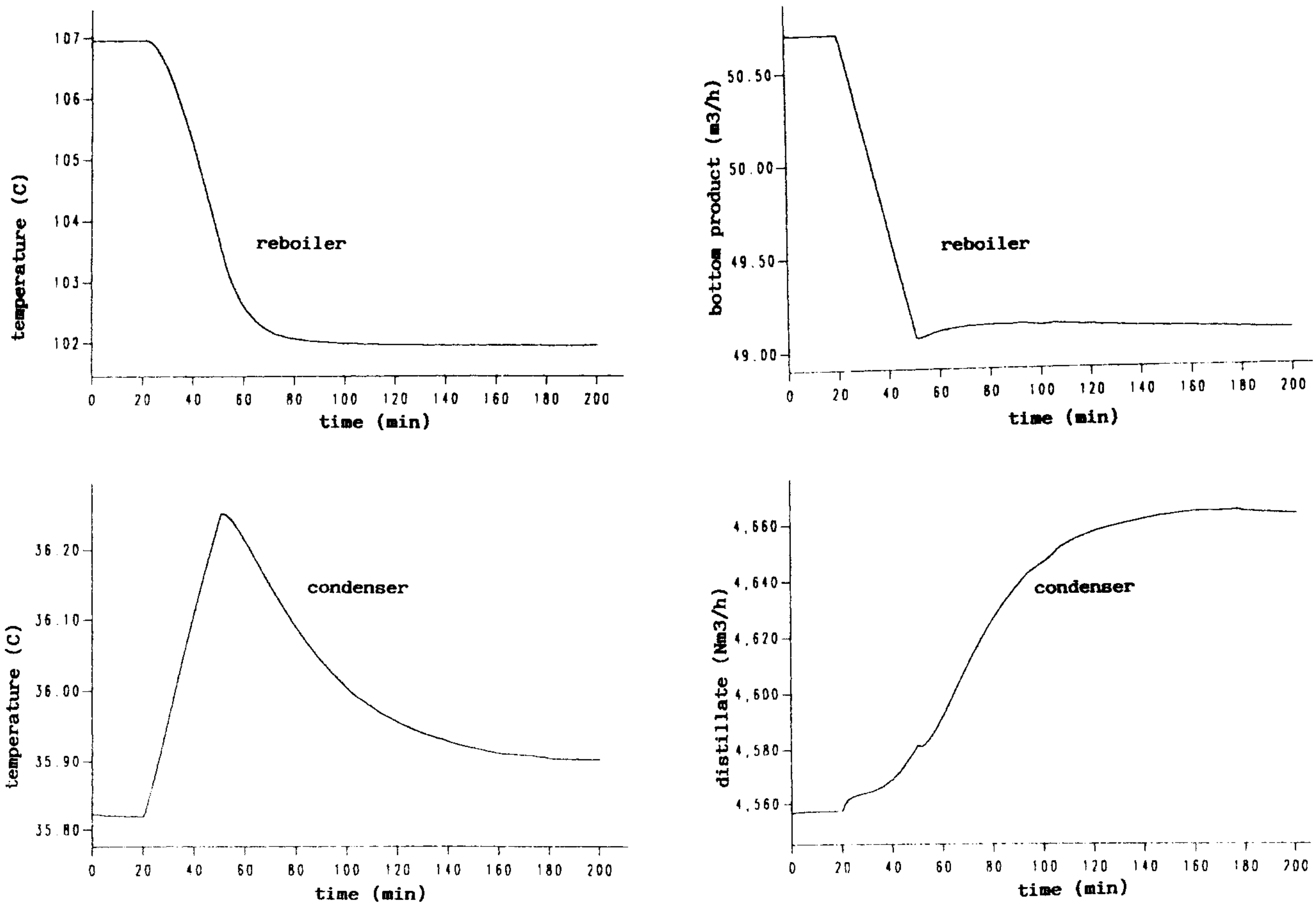

Figure 14. Dynamic behavior of column with set point variation imposed by simplified control law, Eq. 9, and a first-order hold.

Alternatively, a simplified control law can be derived on the basis of the computed rigorous behavior.

A reported example shows a simple application of such procedure to improve a deethanizer control structure against feed composition disturbances. The adopted feedforward control strategy appears to attain at the same time three main objectives:

1. Better column stability in terms of lower excursions in product flows, tray loadings, and reboiler/condenser duty
2. Increased yield in the more valuable product (bottom liquid), due to reduced quality giveaway
3. Reduced operating costs, with special reference to reboiler heat requirements.

These results seem to suggest that a feedforward control technique can be of real value from different points of view and it has to be considered as a powerful complement to the more conventional approach. Although for the sake of brevity the above analysis has been limited to feed composition disturbances (identified as the most important one for the observed process), the feedforward design procedure addressed here can be confirmed to have general applicability and to be extended easily to different process or plant situations when reliable dynamic process knowledge is available.

Notation

A_{ex} = exchange area
E = energy function for liquid holdup
F = feed flow
G = transfer function
HF = feed enthalpy
HL = enthalpy of liquid flow
HV = enthalpy of vapor flow
K = vapor-liquid equilibrium constant
K_c = controller gain
L = molar liquid flow
LMTD = logarithmic mean temperature difference
M = molar holdup
m = manipulated variable
N = number of components
NC = number of controllers
NS = number of stages
P = pressure
Q = thermal duty
T = temperature
U = liquid sidestream
U_{ef} = global heat transfer coefficient
V = molar vapor flow
W = vapor sidestream
X = liquid mole fraction
Y = controlled variable
y = vapor mole fraction

Greek letters

ϵ = tray efficiency
ΔP = pressure drop
τ_i = integral time constant
τ_D = derivative time constant

Subscripts

c = controller
d = disturbance
i = component index
j = controller index
m = measuring device
n = stage index
p = process
set = set point

Literature Cited

Badawas, P. C., "Feedforward Methods for Process Control Systems," *Practical Process Instrumentation and Control*, J. Matley, ed., McGraw-Hill, New York (1986)

Boston, J. F., H. I. Britt, and J. Jirapongphan, "An Advanced System for the Simulation of Batch Distillation Operations," *Foundations of Computer-Aided Chemical Process Design*, R. S. H. Mah, W. D. Seider, eds., Eng. Foundation New York (1981).

Congalidis, J. P., J. R. Richards, and W. H. Ray, "Feedforward and Feedback Control of a Solution Copolymerization Reactor," *AIChE J.*, **35**, 891 (1989).

Gallun, S. E., and C. D. Holland, "Gear's Procedure for the Simultaneous Solutions of Differential and Algebraic Equations with Applications to Unsteady-State Distillation Problems," *Comput. Chem. Eng.*, **6**, 231 (1982).

Gani R., C. A. Ruiz, and I. T. Cameron, "A Generalized Model for Distillation Columns. I: Model Description and Applications," *Comput. Chem. Eng.*, **10**, 181 (1986).

Gani, R., and C. A. Ruiz, "Simulation of Startup and Shutdown Behaviour of Distillation Operations," *Proc. Distillation and Absorption '87*, Brighton, England (1987).

Garcia, C. E., and D. M. Prett, "Advances in Industrial Model-Predictive Control," *Chemical Process Control-CPC III*, M. Morari, T. J. McAvoy, eds., CACHE, 245 (1983).

Hindmarsh, A. C., "ODEPACK: A Systemized Collection of ODE Solvers," Lawrence Livermore Nat. Lab., Livermore, CA (1983).

MacMullan, E. C., and F. G. Shinskey, "Feedforward Analog Computer Control of a Superfractionator," *Control Eng.*, 69 (March, 1964).

Ranzi, E., T. Faravelli, M. Rovaglio, R. Domenichini, and G. Biardi, "A General Program for Simulation of Multicomponent Distillation Column," *Proc. Distillation and Absorption '87*, Brighton, England (1987).

Ranzi E., M. Rovaglio, T. Faravelli, and G. Biardi, "Role of Energy Balance in Dynamic Simulation of Multicomponent Distillation Columns," *Comput. Chem. Eng.*, **12**, 783 (1988).

Rosenbrock, H. H., "The Control of Distillation Columns," *Trans. Inst. Chem. Eng.*, **35**, 347 (1962).

Rovaglio, M., G. Biardi, E. Ranzi, R. Binda, and P. Rizzetto, "Multicomponent Distillation Columns under Transient Conditions. I: Modeling Approach," *Ing. Chim. Ital.*, **22**, (1986).

Skogestad, S., and M. Morari, "Understanding the Dynamic Behavior of Distillation Columns," *Ind. Eng. Chem. Res.*, **27**, 1848 (1988).

Stephanopoulos, G., "*Chemical Process Control: An Introduction to Theory and Practice*," Prentice-Hall, Englewood Cliffs, NJ (1984).

Weingand, W. A., A. K. Jhawar, and T. J. Williams, "Calculation Method for the Response Time to Step Inputs for Approximate Dynamic Models of Distillation Columns," *AIChE J.*, **18**, 1243 (1972).

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