AN EXPERIMENTAL APPLICATION OF THE INFERENTIAL CONTROL SCHEME TO THE BINARY DISTILLATION COLUMN CONTROL

Won Kyoo Lee* and Kwang Eun Kim**

Department of Chemical Engineering, Ohio State University, U.S.A.*, and Yukong Limited, Korea**

(Received 21 October 1982. • accepted 15 October 1983)

Abstract— The inferential control scheme based on a linear estimator was applied to control of top composition of a pilot scale packed distillation column. Experimental and simulation studies were used to evaluate the inferential control scheme and to compare its performance with a conventional single temperature feedback control with proportional plus integral actions. Both digital simulation and experimental verification showed that the top composition control achieved with the inferential control scheme was superior to that achieved with the conventional control scheme.

INTRODUCTION

One of the major questions in process control system design is the selection of process measurements. In most processes, it is not feasible to measure all the process variables of interest because of sensor cost or time delays caused by the need for chemical analysis. An approach called inferential control has been developed by Brosilow and coworkers[1-4] to address the measurement limitation problem, especially when unmeasured disturbances are present. This inferential control system uses selected measurement of secondary process outputs, such as temperatures, in a linear combination to estimate and control the effect of unmeasurable disturbances on primary process outputs, such as product quality.

Distillation columns are a promising application area for the inferential control scheme since distillation columns with many components and large numbers of trays would create special difficulties in the selection of process measurements. The primary objective of distillation column is to deliver products of consistent quality. However, the composition of products from distillation columns is quite difficult to control. Direct measurement of product composition is often expensive and sometimes unreliable, and its associated large measurement lags often do not permit the design of an effective feedback control system.

In many industrial distillation columns, tray temperature near the end of the column is used to approximately control product composition because it has the advantage of being inexpensive, reliable and responsive measurement. Pressure compensation is used when required. Other schemes[5] involving controlling one or two temperature differentials have been proposed to improve sensitivity without influence from pressure. A multiple temperature feedback control, using an average temperature calculated from several tray temperatures, has been proposed for columns with very sharp temperature profile[2].

Brosilow and coworkers[1-4] have proposed a linear static estimator using a linear combination of selected tray temperatures, and steam and reflux flow rates to estimate product compositions in multicomponent columns. Simulation studies on an industrial column have been reported. Rencently Shah[6] reported that a static nonlinear composition estimators were significantly better than Brosilow's linear estimator based on digital simulation and experimental studies when nonlinear regions of operation were encountered.

In this paper the inferential control scheme with a linear estimator is employed in experimental and simulation studies to control the top composition of a methanol-water column. The resulting control behavior is contrasted with that achieved using conventional single temperature feedback control with proportional plus integral actions.

THEORETICAL BACKGROUND

Input disturbances frequently arise from changes in the operation of units upstream from the process.

Because of the disturbances, the process moves from one steady state to another. Inferential control of Brosilow and coworkers[1,2,4] uses selected temperature and flow measurements to estimate and control the effect of unmeasurable input disturbances on the product compositions.

Composition Estimator

For small input disturbances, the process can be assumed to behave linearly. The unmeasured product composition, y, and the measurement of tray temperature, T, can be related to the unmeasurable input disturbance, u, in the Laplace domain as follows:

$$T(s) = A^T(s)\,u(s) \tag{1}$$

$$Y(s) = B^T(s)\,u(s) \tag{2}$$

where T and Y are treated as perturbation variables, and A and B are constant matrices. When the number of disturbances u is less than the number of measurements T, one can solve for u from Equation 1.

$$u(s) = [A(s)\,A^T(s)]^{-1}\,A(s)\;T(s) \tag{3}$$

$$\alpha(s) \triangleq [A^T(s)\,A(s)]^{-1}\,A^T(s)\;B(s) \tag{4}$$

then the least-square estimator of $\hat{Y}$ is given by

$$\hat{Y}(s) = \alpha^T(s)\;T(s) \tag{5}$$

An underlying assumption in the above result is that the inputs u are Gaussian distributed. However, even if u is not gaussian, α is still the best linear estimator[7].

The estimation error Y – $\hat{Y}$ is given by

$$Y(s) - \hat{Y}(s) = W^T(s)\;u(s) \tag{6}$$

where $W(s) = B(s) - A(s)\,\alpha(s)$ (6)

The selection of the secondary measurements should aim at minimizing the number of measurements necessary to obtain accurate setimate which are insensitive to modeling errors. The secondary measurements are thus chosen to minimize the production error in the steady state which is given by

$$E = \frac{\| W(o) \|}{\| B(o) \|} \tag{7}$$

The projection error is a measure of how accurate the estimate will be and ranges from 0 and 1.

Dynamics of Interferential Control Systems

A typical process block diagram for distillation column is given to the right of the dotted line and an inferential control system to the left in Figure 1. This inferential control system is to counter the effect of the unmeasured disturbances d(s) on the product quality. The control effort is based on an inference $\hat{d}(s)$ of d(s). To isolate the effect of the unmeasured disturbances on the process outputs, the effect of all measured inputs on the secondary measurements are substracted from the measurements signal before it enters the estimator.

The Laplace domain description of the process as given in Figure 1 is taken to be

$$y(s) = B^T(s)\;u(s) + C(s)\;m(s) \tag{8}$$

$$T(s) = A^T(s)\;u(s) + P(s)\;m(s) \tag{9}$$

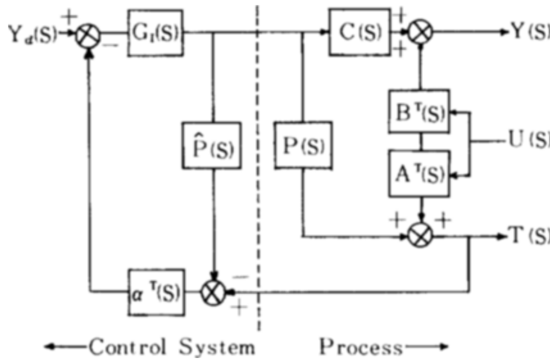

Fig. 1. Inferential control system

where A(s), B(s), C(s) and P(s) are process transfer function matrices. The job of the estimator $\alpha(s)$, satisfying the relationship given in Equation 6, is to combine its input signals $A^T(s)$ u(s) in such a way as to obtain an estimate $\hat{d}(s)$ of the effect of the disturbances on the product quality.

The response of the process output to the unmeasured disturbances is denoted as d(s) satisfying

$$d(s) = B^T(s)\;u(s) \tag{10}$$

The estimate of the effect of the unmeasured disturbances on the process output is denoted as $\hat{d}(s)$ and from Figure 1 is given by

$$\hat{d}(s) = \alpha^T(s)\;[T(s) - \hat{P}(s)\;m(s)] \tag{11}$$

from Equations 6, 9 and 10

$$\begin{aligned}\hat{d}(s) &= \alpha^T(s)\;\{A^T(s)\;u(s) + [P(s) - \hat{P}(s)]\;m(s)\}\\ &= d(s) + W^T(s)\;u(s) + \alpha^T(s)\\ &\qquad [P(s) - \hat{P}(s)]\;m(s)\end{aligned} \tag{12}$$

Clearly, smaller values of W(s) and P(s) – $\hat{P}(s)$ yield a better approximation of $\hat{d}(s)$ to d(s)

The response of the control effort to a disturbance is given by

$$\begin{aligned}m(s) &= G_I(s)\;[y_d(s) - \hat{d}(s)]\\ &= G_I(s)\;y_d(s) - G_I(s)\;\{d(s) + W^T(s)\;u(s) +\\ &\qquad \alpha^T(s)\,[P(s) - \hat{P}(s)]\;m(s)\}\end{aligned} \tag{13}$$

Rearranging of Equation 13 gives

$$\begin{aligned}m(s) &= \{I + G_I(s)\,\alpha^T(s)\;[P(s) - \hat{P}(s)]\}^{-1}\\ &\qquad G_I(s)\;[y_d(s) - d(s) - W^T(s)\;u(s)]\end{aligned} \tag{14}$$

Finally, the process output is given by

$$\begin{aligned}y(s) &= B^T(s)\;u(s) + C(s)\;m(s)\\ &\quad C(s)\;F(s)\;G_I(s)\;[y_d(s) - W^T(s)\;u(s)]\\ &\qquad + [I - C(s)\;F(s)\;G_I(s)]\;d(s)\end{aligned} \tag{15}$$

$$F(s) = \{I + G_I(s)\;\alpha^T(s)\;[P(s) - \hat{P}(s)]\}^{-1}$$

When $\hat{P}(s)$ = P(s) over the frequency range of the system, then F(s) = I, and the system is stable provided that the original process is stable and the controller $G_I(s)$ is stable. The appropriate choice for $G_I(s)$ in order to obtain a fast response to disturbances and set point changes is given according to Equation 15.

$$G_I(s) = [C(s)\ F(s)]^{-1} = F^{-1}(s)\ C^{-1}(s) \qquad (16)$$

However, it will generally not be possible to implement Equation 16 exactly because the elements of the transfer matrix C(s) will be lags whose numerator polynomials are at least one degree lower than their denominator polynomials. This means that $C^{-1}(s)$ will contain elements which will not be realizable, such as pure differentiators. Further some of the elements of $C^{-1}(s)$ can be unstable, especially if one or more of the elements of G(s) were nonminimum phases. As is common in such situations, the pure derivative is replaced by a lead-lag network, and then $G_I = K_I \frac{\tau_0 s+1}{\tau_1 s+1}$ will be a stable approximation to $[C(s)\ F(s)]^{-1}$. However, the steady state gains $G_I(s)$ should be chosen so that

$$G_I(o)\ F(o)\ C(o) = I \qquad (17)$$

Implementation of the inferential control system of Figure 1 requires the implementation of m(2n + m) transfer function matries if $\alpha(s)$, P(s) and $G_I(s)$ are full. With dim [y(s)] = m and dim [T(s)] = n, both the estimator matrix $\alpha(s)$ and the compensator maxtix P(s) have dimension m × n, and the dimension of the controller $G_I(s)$ is m × n.

DESCRIPTION OF PROCESS AND CONTROL SYSTEMS

The laboratory scale distillation column used as the subject of all simulated and experimental investigations reported in this paper is a 10 cm diameter column packed to the depth of 225 cm with 1.27 cm Raschig rings, separating a methanol-water mixture at atmospheric pressure. The distillation column was normally operated with a feed rate of 12 g-mole/min of 31 mole % methanol as given in Table 1, and was interfaced with HP 2647A intelligence graphic terminal for the inferential control system as shown schematically in Figure 2.

Table 1. Steady-State Operating Conditions

Feed flow rate	12 g-mole/min (300 ml/min)
Distillate flow rate	3.54 g-mole/min (142 ml/min)
Bottom flow rate	8.46 g-mole/min (152 ml/min)
Feed composition	31 mole % methanol
Top composition	95.6 mole % methanol
Bottom composition	3.4 mole % methanol
Reflux ratio	2.3
Column pressure	1 atm
Feed temperature	67°C

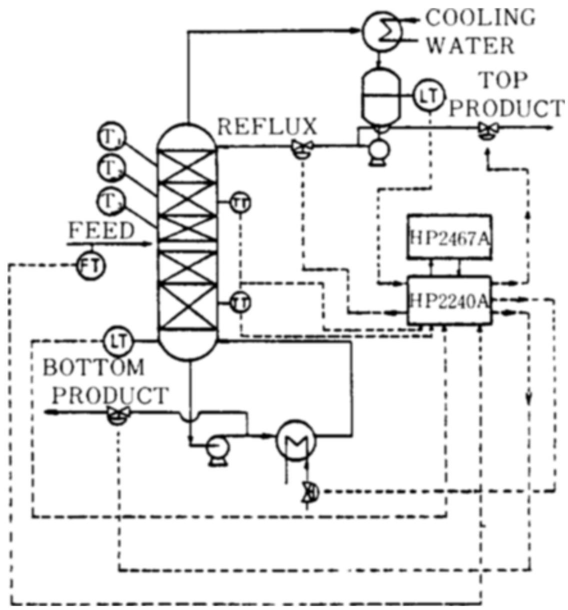

Fig. 2. Schematic diagram of distillation column

Standard data acquisition of temperatures and flow rates was done at 15-second intervals by HP 2240 A measurement and control processor. The primary control objective was to maintain top composition at a constant value of approximately 95.6% methanol in spite of disturbances in the feed flow rate and the feed composition. Top composition was obtained by drawing samples at every 10 minutes to confirm predictions from temperature measurements. As can be seen from Figure 2, top composition is controlled by manipulating reflux

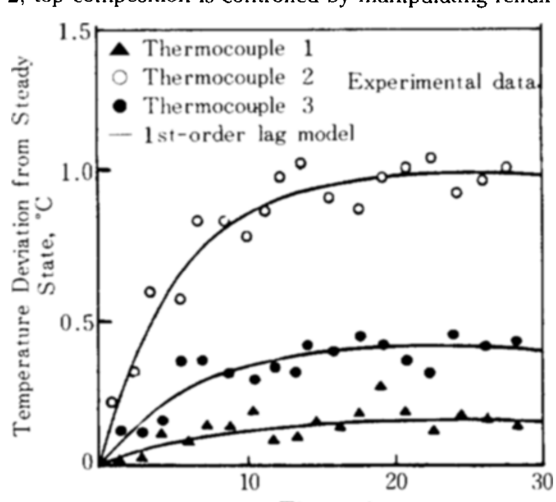

Fig. 3. Open loop responses of temperature to 40 ml step increase in feed flow rate

flow as is common is industrial practice.

In order to employ the inferential control scheme, process transfer function matrices as defined in Equations 8 and 9 must be available. For the operating conditions given in Table 1, appropriate open-loop responses for step change in feed flow were fit by a first-order lag for each transfer function. To select temperature measurements to be used to infer the overhead methanol composition, transfer functions A(s) for three different temperatures in the rectifying section and a transfer function B(s) relating the effect of feed flow and top composition are determined as follows:

$$A^T(s) = [a_1(s)\ a_2(s)\ a_3(s)] = \left[\frac{0.01}{3s+1}\ \frac{0.025}{5s+1}\ \frac{0.005}{6.7s+1}\right] \quad (18)$$

$$B^T(s) = \frac{-0.01}{36s+1} \quad (19)$$

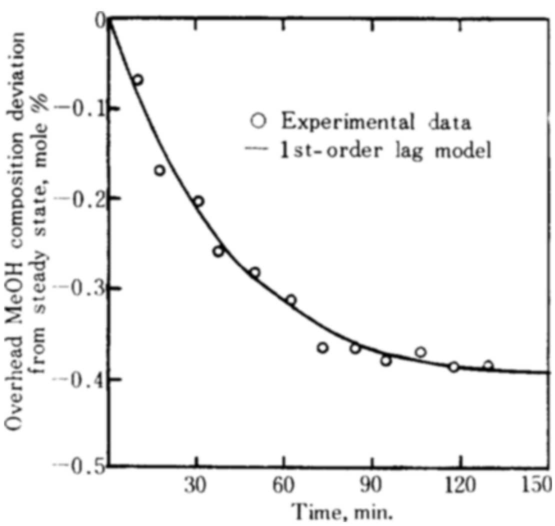

Fig. 4. Open loop response of overhead methanol composition to 40 ml step increase in feed flow rate

A comparison of simulated and experimental response data for open loop operation is shown Figures 3 and 4, respectively. With $A^T(s)$ and $B^T(s)$ the projection error in the top methanol composition using three temperature measurements are estimated as shown in Figure 5. As can bee seen from Figure 5, the projection error is minimized by selecting the middle part of the rectifying section as a location on which to measure the temperature. Since there is only a single measurement, the transfer function A(s) and the estimator $\alpha(s)$ become scalar transfer functions.

$$A^T(s) = \frac{0.025}{5s+1} \quad (20)$$

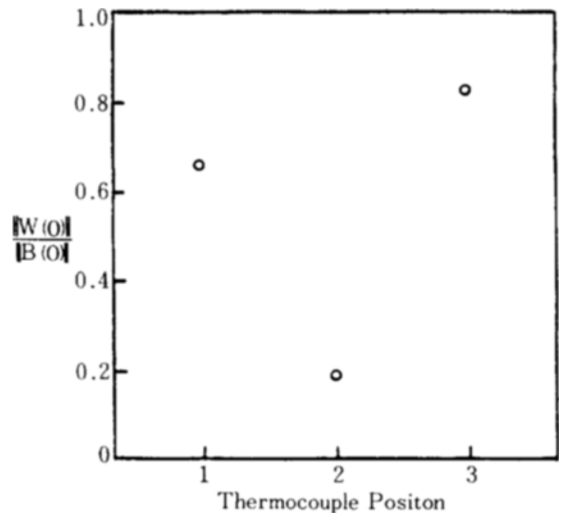

Fig. 5. Projection error in overhead methanol composition for a single temperature measurement

From Equation 6 $\alpha(s)$ is given by

$$\alpha(s) = \frac{B(s)}{A(s)} = -0.4\frac{5s+1}{36s+1} \quad (21)$$

Two other process transfer functions P(s) and C(s) are also scalar transfer functions because there is but a single measured temperature. The transfer function P(s) shown in Figure 1 is an approximation of the response of the temperature on the middle part of the rectifying section to changes in the control effort (that is, the reflux flow), and it is given below

$$\hat{P}(s) = \frac{-0.00135}{3.7s+1} \quad (22)$$

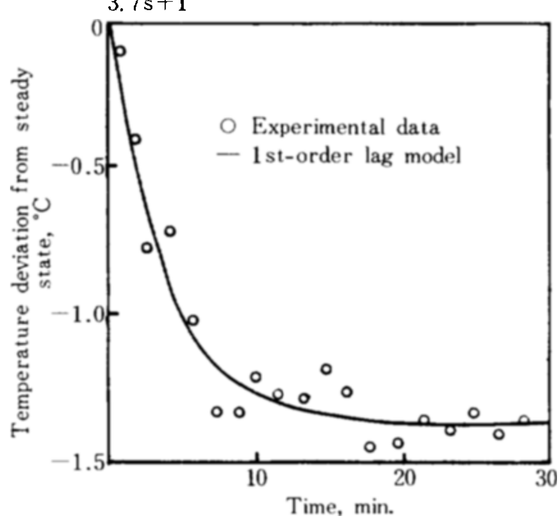

Fig. 6. Open loop response of temperature to 30 ml step increase in reflux flow rate

The transfer function relating the effect of reflux flow and top methanol composition is obtained

$$C(s)=\frac{0.0005}{41s+1} \tag{23}$$

Experimental response data for open loop operation are compared with simulated results in Figures 6 and 7.

For perfect control, given a perfect estimate, the controller transfer function $G_I(s)$ should be the inverse of the transfer function between the top methanol composition and the reflux flow. However, as can be from Equation 20, this controller would have proportional plus pure derivative actions which will not be realizable. As is common in such situations, the pure derivative action was replaced by a lead-lag network and G_I is taken as

$$G_I(s)=C^{-1}(s)\frac{1}{\tau_I s+1}=2000\frac{41s+1}{\tau_I s+1} \tag{24}$$

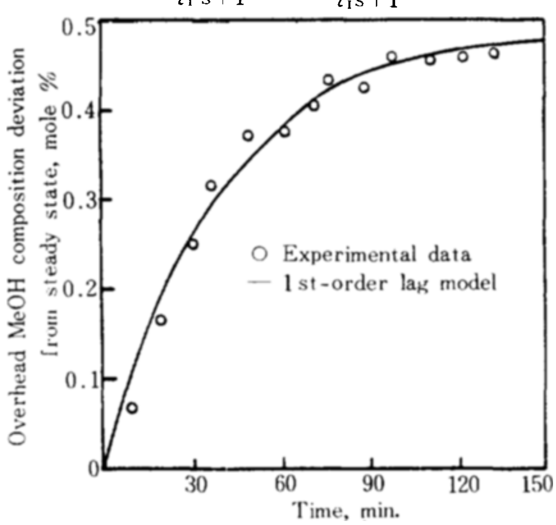

Fig. 7. Open Loop Response of Overhead Methanol Composition to 30 ml Step Increase in Reflux flow Rate

Figure 8 shows a block diagram of the inferential control system given by Equations 19 through 24, and then the single temperature feedback control system with a proportional plus integral controller is shown as an alternate control system in Figure 9.

RESULTS AND DISCUSSION

Simulation Results

A digital computer simulation using HP 2647A intelligence graphic terminal was conducted to evaluate the performance of the inferential control scheme compared to the single temperature feedback control. Figures 10 and 12 show the responses of the top methanol composition under inferential control for a step change in

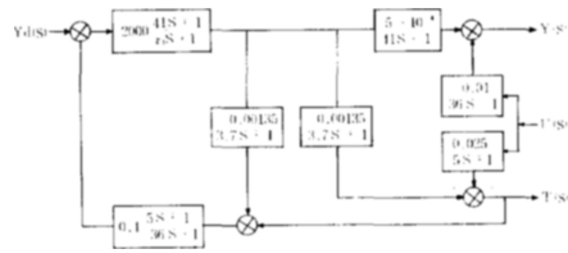

Fig. 8. Block Diagam of the Inferential Control System

the feed flow. As illustrated in Figures 10 and 12, the inferential control gave no steady-state error while the single temperature feedback control resulted in a constant offset in the top methanol composition regardless of values of proportional gain and integral time. However, as can be expected, the single temperature feedback control did a good job of maintaining the temperature were obtained regardless of values of lag time constants used under the inferential control shown in Figures 11 and 13, Simulated results showed that the inferential control system would perform significantly better than the temperature feedback control system, when both systems use the same temperature measurement.

Experimental Results

To verify the simulation results and to examine the practicality of the inferential control scheme for industrial application, experimental studies were carried out on the laboratory scale, packed distillation column using HP 2647A intelligence graphic terminal to implement the control algorithms.

The control behaviors of the top methanol composition for a 10% increase in feed flow rate are shown in Figures 10 and 12 for inferential control and single temperature feedback control. These experiments verified the superiority of the inferential control scheme. The single temperature feedback control system gave a final steady state error of 0.02 mole % in the top

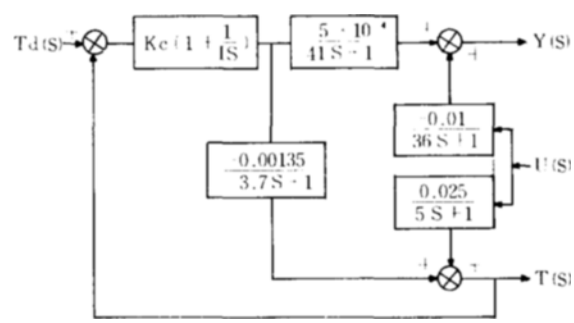

Fig. 9. Block Diagram of the Single Temperature Feedbak Control System

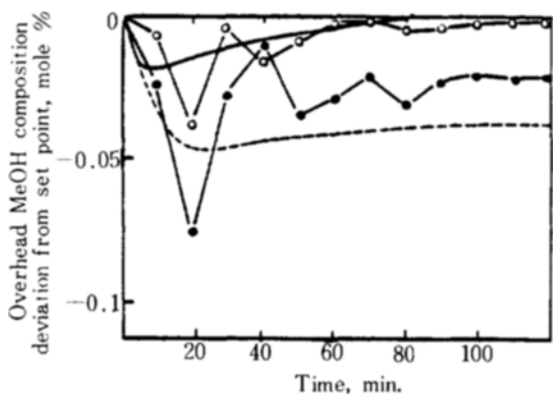

Fig. 10. Overhead Composition Responses to a 10% Step Increase in Feed flow Rate
Inferential Control (Lag= 3 min.)
○ experimental result
— simulated result
PI-Control (Kc= −2000, I=10 min.)
● experimental result
··· simulated result

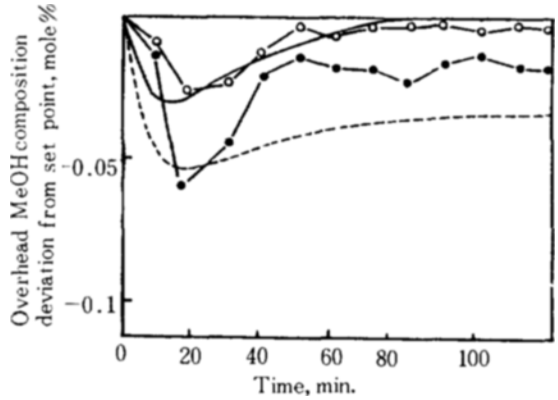

Fig. 12. Overhead Composition Responses to a 10% Step Increase in Feed flow Rate
Inferential Control (Lag= 5 min.)
○ experimental result
— simulated result
PI-Control (Kc= −1000, I= 5 min.)
● experimental result
··· simulated result

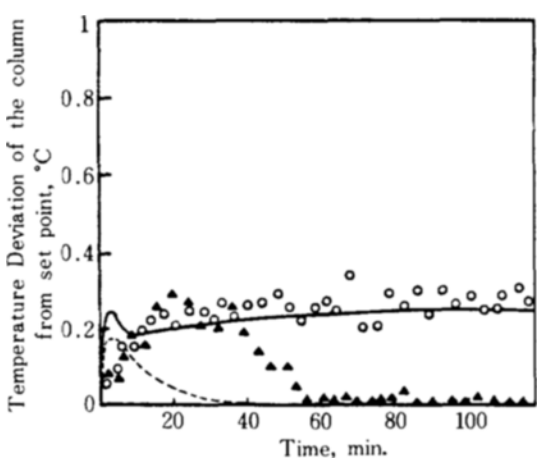

Fig. 11. Temperature Respones to a 10% Step Increase in Feed flow Rate
Inferential Control (Lag= 3 min.)
○ experimental result
— simulated result
PI-Control (Kc = −2000, I=10min.)
▲ experimental result
··· simulated result

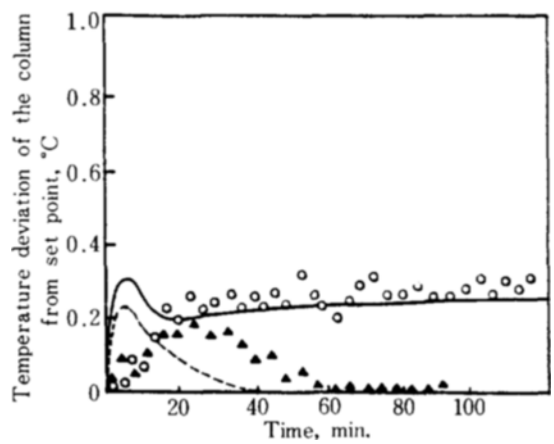

Fig. 13. Temperature Responses to a 10% Step Increase in Feed flow Rate
Inferential Control (Lag= 5 min.)
○ experimental result
— simulated result
PI-Control (Kc= −1000, I= 5 min.)
▲ experimental result
··· simulated result

methanol composition while the control temperature was maintained constant as illustrated in Figures 11 and 13. These experimental and simulated responses are in good agreement, but it can be seen from Figures 10 through 13, in general, the simulated behavior of the column exhibits a shorter response time than was observed experimentally. This implies that the process has higher order dynamics that are not included in the first-order model.

CONCLUSIONS

The inferential control scheme was successfully used to control top composition in a pilot scale, methanol-water column. Experimental results demonstrated that the inferential control scheme based on a linear estimator was superior to the single temperature feedback control for regulatory control of composition in the face of feed flow disturbance. Moreover in the experimental tests, the temperature feedback control scheme gave a steady-state top composition deviation as indicated in the simulation study.

NOMENCLATURE

A(s), B(s), C(s), D(s): process transfer function matrices
d(s): unmeasured input disturbances
$G_f(s)$: overall transfer function matrix
E: measured input disturbances
Y(s): output composition
$\alpha(s)$: composition estimator

REFERENCES

1. Brosilow, C.B., and Tong, M.: AIChE J., **24** (1978), 492.
2. Joseph, B., Brosilow, C.B., Howell, J.C. and Kerr, W.R.D.: Hydrocarbon Processing, **55** (March, 1976), 127.
3. Joseph, B. and Brosilow, C.B.: AIChE J., **24** (1978), 485.
4. Weber, R. and Brosilow, C.B.: AIChE J., **18** (1972), 614.
5. Boyd, D.M.: Chem. Engr. Progress, **71** (1975), 55.
6. Shah, M.K.: Ph.D. Dissertation, Lehigh Univ., Penn., U.S.A., (1971).
7. Rhodes, I.B.: IEEE Trans. on Automatic Control, AC-16 (1971), 688.