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THE UNIVERSITY OF ALBERTA

SELF-TUNING CONTROL OF A BINARY DISTILLATION COLUMN

by

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A THESIS

SUBMITTED TO THE FACULTY OF GRADUATE STUDIES AND RESEARCH

IN PARTIAL FULFILMENT OF THE REQUIREMENTS FOR THE DEGREE

OF MASTER OF SCIENCE

IN

PROCESS CONTROL

CHEMICAL ENGINEERING

EDMONTON, ALBERTA

FALL 1980

UNIVERSITY OF ALBERTA

FACULTY OF GRADUATE STUDIES AND RESEARCH

The undersigned certify that they have read, and recommend to the Faculty of Graduate Studies and Research, for acceptance, a thesis entitled SELF-TUNING CONTROL OF A BINARY DISTILLATION COLUMN submitted by Ronald K. Tong, B.Eng., in partial fulfilment of the requirements for the degree of Master of Science in Process Control (CHEMICAL ENGINEERING).

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(Supervisor)

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C Kiparissides

Date September 24 19 80.

DEDICATION

To My Wife Catherine

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ABSTRACT

The performance of the basic self-tuning regulator (STR) for control of top product composition of a binary distillation column is studied experimentally and by simulation. The self-tuning regulator provides satisfactory control behavior for step changes in feed flow rate and set point even though the regulator is not formulated for such changes. A PI controller, once well-tuned provides excellent performance and outperforms the STR, when used with zero initial parameter estimates.

In the case of bottom product composition control, simulations have shown that due to the long time delay in the bottom composition analysis, the PI controller is not able to provide satisfactory control. The STR, with the advantage of an inherent predictor in the model formulation, manages to achieve a much improved control performance once reasonable parameter estimates were obtained with the multiple pass method. In this work, the principle STR design parameters β_0 and $\underline{P}(0)$ have been shown to significantly effect the performance of the regulator.

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1. INTRODUCTION

1.1 Introduction

The distillation column, despite its extensive use in industry, remains among the most difficult unit operations to control. The physical complexity of the distillation process and the inherent system nonlinearity combine to provide a major challenge for those responsible for designing control systems.

With the availability of process control computers, different advanced control schemes have constantly been evaluated as part of a continuing research activity in hope of achieving better efficiency, higher throughput, lower operating cost or any attractive combination of these. The conventional three term controller, even properly tuned for the system's dynamic behavior is supposedly a compromise between disturbance minimization and set point changes. It has often been criticized for not yielding an optimal control and for its inability to handle nonlinear dynamic behavior.

To design a suitable controller, an accurate model is usually required. Among advanced control techniques, those based on linear state space models have not been very successful due to the difficult problem of representing the dynamic behavior of the process system using a reduced state space model (1). For feedforward control, even if a good model is obtained, the process operating conditions may change and the model becomes inadequate (2,3).

An adaptive control strategy, developed by Astrom and Wittenmark appears to overcome these difficulties . It allows on-line tuning of the model parameters. This strategy has been termed as the self-tuning regulator (4,5).

The basic self-tuning regulator is designed for the regulator problem. It is designed to control single input, single output systems with unknown but constant parameters. The regulator can also be applied to systems with slowly varying parameters. The algorithm combines a recursive least squares estimator with a minimum variance regulator computed from the estimated model.

1.2 Objective of Thesis:

This work is directed toward asse: ing the effectiveness of the basic self-tuning regulator for control of a binary distillation column. By applying the technique, a better understanding of its relative advantages, disadvantages and applicability in the process industry will be achieved.

In this investigation, the performance of the self-tuning regulator for control of top product composition is studied experimentally and by simulation. The control of bottom composition has been studied by simulation for load and set point changes. The resulting control performance is compared with that of a well-tuned conventional proportional plus integral (PI) controller.

1.3 Outline of Thesis

The thesis is organized in four parts:

The first section, consisting of Chapter 2, presents the mathematical formulation of the self-tuning regulator. A literature survey is not included since detail surveys have been given by Chang (6) and Lieuson (7).

Chapter 3, which constitutes the second section, provides the necessary process information for the application of the adaptive control schemes.

The third section, Chapters 4 and 5 give the results of control of top composition and bottom composition respectively.

The last section, Chapter 6 summarizes the overall conclusions concerning the utilization of the selftuning regulator.

Note: The work of this thesis was completed during 1977. However, the thesis was submitted for examination in September 1980.

2. THE SELF-TUNING REGULATOR

2.1 Theoretical Formulation

The self-tuning regulator of Astrom and Wittenmark (4) was derived using the following single-input, single-output linear stochastic model of the process as a starting point:

$$y(t) + a_1 y(t-1) + \ldots + a_n y(t-n) = b_1 u(t-k-1) + \ldots + b_n u(t-k-n) + \lambda[e(t) + c_1 e(t-1) + \ldots + c_n e(t-n)] \quad (2.1)$$

where:

y is the normalized output deviation variable

u is the normalized input deviation variable

t denotes the sampling instant

$(e(t), t = 0,1,2,\ldots)$ is a sequence of independent normal (0,1) random variables which represent the net effect of external disturbances.

k is a nonnegative integer which expresses the process time delay as a multiple of the sampling interval.

$\{a_i\}$, $\{b_i\}$, $\{c_i\}$ and λ are time invariant parameters.

The stochastic model in Equation (2.1) has received considerable attention not only in the stochastic control theory literature but also by workers in time series analysis (8,9). This model can also be expressed in backward shift operator notation as:

$$A(q^{-1})y(t) = B(q^{-1})u(t-k-1) + C(q^{-1})e(t) \quad (2.2)$$

where q^{-1} denotes the backward shift operator; i.e. $q^{-1}x(t) = x(t-1)$ and polynomials A, B and C are defined by:

$$A(q^{-1}) = 1 + a_1 q^{-1} + a_2 q^{-2} + \ldots + a_n q^{-n} \tag{2.3}$$

$$B(q^{-1}) = b_1 + b_2 q^{-1} + \ldots + b_n q^{-n+1} \tag{2.4}$$

$$C(q^{-1}) = 1 + c_1 q^{-1} + \ldots + c_n q^{-n} \tag{2.5}$$

The minimum variance regulator minimizes the loss function, $V = E\{y^2(t)\}$ where E denotes the expectation operator. Astrom (4) has derived the following minimum variance control law:

$$u(t) = -G(q^{-1})/B(q^{-1})F(q^{-1}) * y(t) \tag{2.6}$$

where the polynomials

$$F(q^{-1}) = 1 + f_1 q^{-1} + \ldots + f_k q^{-k} \tag{2.7}$$

$$G(q^{-1}) = g_0 + \ldots + g_{n-1} q^{-n+1} \tag{2.8}$$

are related to $A(q^{-1})$ and $C(q^{-1})$ by:

$$C(q^{-1}) = A(q^{-1})F(q^{-1}) + q^{-k-1} G(q^{-1}) \tag{2.9}$$

Since the model parameters $\{a_i\}$, $\{b_i\}$, $\{c_i\}$ and λ are unknown and constant, the self-tuning strategy consists of estimating the model parameters on-line and basing the control calculation at each sampling instant on the estimated parameters. A wide variety of STR algorithms can be derived depending on the particular estimation algorithm and control strategy that are selected (5). In this investigation, a recursive linear least squares estimator is used in conjunction with a minimum variance controller as originally suggested by Astrom and Wittenmark (4).

In principle, the least squares calculations could be based on updating the parameters of the model in Equation (2.1) but this would involve a complicated estimation procedure which is not suitable for on-line computation. By assuming $C_i = 0$ for $i = 1,2,\ldots n$ the parameter estimation problem reduces to a conventional linear least squares problem which can be easily formulated as a recursive algorithm. The on-line control calculations can be simplified by incorporating the identity in Equation (2.9) into the model structure, resulting in the following "predictive" model:

$$y(t) + \alpha_1 y(t-k-1) + \ldots + \alpha_n y(t-k-n) = \beta_0 [u(t-k-1) + \ldots + \beta_L u(t-L-k-1)] + \varepsilon(t) \qquad (2.10)$$

where $L = n + k - 1$ and the disturbance $\varepsilon(t)$ is a moving average of order k of the driving noise, e(t).

If the coefficients in the predictive model are known and constant, then the minimum variance regulator for the predictive model in Equation (2.10) is:

$$u(t) = 1/\beta_0 [\alpha_1 y(t) + \ldots + \alpha_n y(t-n+1)] - \beta_1 u(t-1) - \ldots - \beta_L u(t-L) \qquad (2.11)$$

In the self-tuning regulator, $\{\alpha_i\}$ and $\{\beta_i\}$ are estimated on-line and the estimates $\{\hat{\alpha}_i\}$ and $\{\hat{\beta}_i\}$ are used in the minimum variance controller of Equation (2.11). However, it may not be possible to accurately estimate all of the parameters during closed-loop operation since the inputs are correlated with the disturbance $\{\varepsilon(t)\}$. To avoid this difficulty, Astrom and Wittenmark (4) suggested that β_0 should be held at some specified value. For constant β_0, the STR control algorithm is:

$$u(t) = 1/\beta_0 \, [\hat{\alpha}_1(t)y(t) + \ldots + \hat{\alpha}_n y(t-n+1)] - \hat{\beta}_1(t)u(t-1) - \ldots - \hat{\beta}_L(t)u(t-L) \qquad (2.12)$$

which can be written as:

$$u(t) = -1/\beta_0 \, \underline{\theta}(t)\underline{\Psi}(t) \qquad (2.13)$$

where

$$\underline{\Psi}^T(t) = [-y(t), -y(t-1), - \ldots - y(t-n+1), +\beta_0 u(t-1), + \ldots, +\beta_0 u(t-L)] \qquad (2.14)$$

and

$$\hat{\underline{\theta}}^T(t) = [\hat{\alpha}_1(t), \ldots, \hat{\alpha}_2(t), \hat{\alpha}_n(t), \hat{\beta}_1(t), \ldots, \hat{\beta}_L(t)] \qquad (2.15)$$

The parameter estimates, $\hat{\underline{\theta}}(t)$ are calculated using recursive least squares identification with exponential weighting of past data (10). The algorithm is given by:

$$\hat{\underline{\theta}}(t) = \hat{\underline{\theta}}(t-1) + \underline{K}(t)\,[y(t) - \beta_0 u(t-k-1) - \underline{\Psi}^T(t-k-1)\hat{\underline{\theta}}(t-1)] \qquad (2.16)$$

$$\underline{K}(t) = \underline{\underline{P}}(t-1)\underline{\Psi}(t-k-1)[1 + \underline{\Psi}^T(t-k-1)\underline{\underline{P}}(t-1)\underline{\Psi}(t-k-1)]^{-1} \qquad (2.17)$$

$$\underline{P}(t) = 1/\mu \{ \underline{P}(t-1) - \underline{K}(t)[1 + \underline{\Psi}^T(t-k-1)\underline{P}(t-1)\underline{\Psi}(t-k-1)\underline{K}^T(t)] \} \quad (2.18)$$

where

- $\hat{\underline{\theta}}(t)$ is the estimated value of $\underline{\theta}(t)$
- $\underline{K}(t)$ is the gain vector { (n+1) x 1 }
- $\underline{P}(t)$ is proportional to the covariance matrix { (n+1) x (n+1)}
- μ is the exponential forgetting factor

In the estimatation phase, $\underline{K}(t)$ is calculated first, followed by $\underline{P}(t)$ and $\hat{\underline{\theta}}(t)$ as discussed in Appendix A.

2.2 Parameter Selection

Before implementing the self-tuning regulator, the constants and initial values listed below must be specified:

- model order n
- time delay k
- scaling factor β_0
- exponential forgetting factor μ
- initial parameter estimates $\underline{\theta}(0)$
- initial covariance matrix $\underline{P}(0)$
- constraint limits on control output u_t

Astrom and Wittenmark (4) and Chang (6) have presented guidelines for selecting appropriate values for the above constants. These recommendations have been used in this work.

2.2.1 Model Order, n:

In order that the process be adequately represented, the self-tuning regulator must have enough parameters so it could converge to the optimal minimum variance regulator. For the zero initial estimate case, if a higher order model is used, it will need more sampling intervals before the information vector ($\underline{\psi}(t)$) and the parameter vector ($\underline{\theta}(t)$) have non-zero elements. Consequently, this will result in poorer control during the initial transient period.

2.2.2 System Time Delay, k:

The choice of k, the nonnegative integer to represent the time delay of the process in terms of an integral number of sampling intervals is extremely important. It directly affects the number of parameters to be estimated. Overestimating this parameter will result in a sluggish control performance whereas underestimating k will cause cycling or bang bang control.

2.2.3 Scaling Factor, β_0:

The value selected for β_0 should be close to the actual value for the system to be controlled. From the consideration of stable operation, if the β_0 value used is much lower than the true value, it may cause the parameters to diverge (4). The effect of the choice of the β_0 on the controller performance is investigated.

2.2.4 Exponential Forgetting Factor, μ :

The use of a weighting factor less than 1 allows the weighting out of previous data. This feature enables tracking of slow time varying parameters. However, in this study, due to the short duration of the runs, the parameters are not considered to be time varying, so as a result, μ is set to 1.

2.2.5 Initial Parameter Estimates, $\underline{\theta}(0)$:

The initial estimates will frequently be set to zero when there is no a priori information. The use of other than zero initial values for the second pass or subsequent runs is investigated in this study.

2.2.6 Initial Covariance Matrix, $\underline{\underline{P}}(0)$:

The initial covariance matrix used reflects the confidence in the initial parameter estimates. A higher $\underline{\underline{P}}(0)$ value should increase the convergence rate of the parameters at each sampling interval. The effect on controller performance of this parameter is investigated in this study.

2.2.7 Limits on control signal amplitude, u_L :

Control signal limits are implemented to safeguard against run away conditions. However, too tight a limit may slow the convergence rate of the parameters.

2.3 APPLICATION OF THE STR ALGORITHM

In the past years, some of the industrial processes that have been successfully controlled by the self-tuning regulator include a paper machine (11), an ore crusher (12), a heat exchanger (13) and a super tanker (14). At the University of Alberta, Chang (6) has studied the self-tuning regulator control of a double effect evaporator and Sastry et al.(15) investigated top product composition control on a distillation column.

For implementing the self-tuning regulator, limited a priori knowledge of the process dynamics and noise characteristics of the system to be controlled are needed.

The moderate computation requirement of the algorithm makes it very attractive for industrial use. As the algorithm is adaptive, it is ideal for use in dealing with processes that have unknown process dynamics, nonlinearity, or changes in operating conditions. The regulator is best suited for a situation where the control objective is to minimize output deviation.

A possible drawback, however, would be the regulator's extreme sensitivity to parameter variation for nonminimum-phase systems (4). The necessity of extending the regulator to include feedforward compensation for measurable disturbances and set point following was recognized by Wittenmark (16). Clarke and Gawthrop (17) have recommended the use of a cost function that incorporates system input, output and set point. In Clarke's algorithm, control output is reduced at the expense of extra output variance.

In this study, the basic self-tuning regulator, as proposed by Astrom is evaluated on a pilot scale binary distillation column.

3. PROCESS CHARACTERIZATION

3.1 Process Equipment

The experimental studies were performed on a pilot scale distillation column interfaced to an IBM 1800 data acquisition and control computer. A schematic diagram of the equipment is shown in Figure 3.1. The 22.86 cm. diameter column, containing eight bubble cap trays (four caps per tray) has a 30.48 cm. tray spacing. The column is equipped with a total condenser and thermosyphon type reboiler. The column operates at atmospheric pressure, with the feed at 48% by weight methanol entering the fourth tray. Top product composition is 96.30% by weight methanol and the bottom product composition is 5.50% by weight methanol. The top composition is measured by means of an in-line capacitance probe. The bottom composition is measured with a Hewlett-Packard model 5720A gas chromatograph using an in-line liquid sampling system. The chromatograms are analyzed on a two hundred fifty-six second cycle. All measurements are transmitted to the IBM 1800 for data logging and/or control. Further specific details concerning the column are given in the theses by Svrcek (18) and Pacey (19).

The experimental operating conditions used in this work, which differ somewhat from those of previous work, are given in Table 3.1.

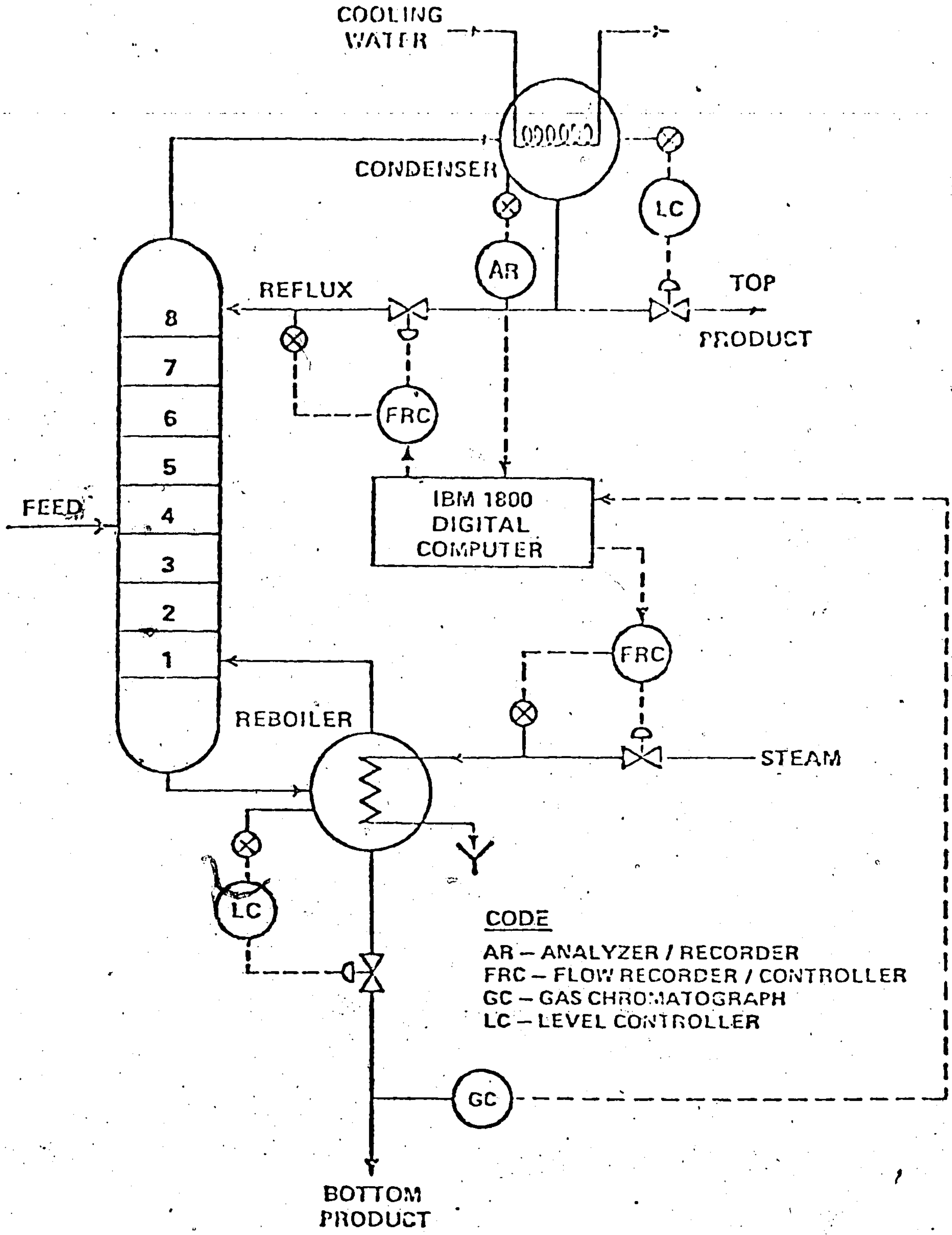


Figure 3.1 Schematic Diagram of the Distillation Column

Table 3.1 Experimental Operating Conditions (Run E1)

Feed Flow	18.0 g/s	Feed Composition	48.0 wt%
Reflux Flow	10.3 g/s	Top Product Comp.	96.3 wt%
Steam Flow	11.5 g/s	Bottom Product Comp.	5.5 wt%
Top Product Flow	8.4 g/s	Feed Temperature	74.1 C
Bottom Product Flow	9.6 g/s	Reflux Temperature	64.2 C

3.2 Process Model

A nonlinear ordinary differential equation model, based on the mass and energy balance relationships for each of the trays, reboiler and condenser is utilized for simulating the dynamic behavior of the column (20). The initial analysis results in 30 differential equations containing 140 variables but the number of unknown variables can be reduced to 20 by specifying certain inputs and outputs and by making the following assumptions:

1. Constant heat loss from each stage
2. Constant tray efficiency
3. Constant liquid holdup
4. Negligible vapor holdup

Details of the model development and numerical solution are contained in the thesis of Simonsmeier (20). The steady state operating conditions used for simulation , which were similar to those used experimentally are given in Table 3.2

Table 3.2 Operating Conditions For Simulation

Feed Flow	18.0 g/s	Feed Composition	48.0 wt%
Reflux Flow	10.3 g/s	Top Product Comp.	95.2 wt%
Steam Flow	11.75 g/s	Bottom Product Comp.	5.5 wt%
Top Product Flow	8.6 g/s	Feed Temperature	74 C
Bottom Product Flow	9.4 g/s	Reflux Temperature	64 C

3.2.1 Dynamic Testing

The objective of the experimental work was the control of either only the top composition or bottom composition. Control was evaluated for load and set point changes, i.e. for both regulatory and servo control.

The main disturbances to a column are usually changes in feed flow rate and the feed composition. The feed flow rate tends to vary much faster and more often than does the feed composition and in fact, both Shinskey (21) and Lupfer (22) suggest that feedback control designed on the basis of feed flow rate disturbances should be sufficient to compensate for the gradual feed composition changes. Hence feed flow rate was chosen as the system load disturbance.

Steam flow has normally been reserved for control of bottom composition. By following typical industrial practice, reflux flow rate rather than top product flow rate was chosen as the manipulated variable for top product composition control. Top product flow was used to control the condenser level.

In order to implement the self-tuning regulator algorithm, it is necessary that a predictive model be specified. Time delay in the process and the number of parameters in the regulator thus have to be determined in advance. In addition, high and low limits for the control signal must be chosen. To provide reliable information for choice of these values, dynamic testing of the column by using open loop step disturbance runs were performed. Feed flow rate step changes of ± 20%, from the steady state value, were introduced to study the response of the column, while the two manipulated variables, steam and reflux flows were changed ± 5% and ± 10% respectively. The responses for the step changes are plotted in Figures 3.2 to 3.7. The arrow on the abcissa scale in each plot indicates the time at which the disturbance is introduced.

Examination of the responses plotted in Figures 3.2 to Figure 3.7 shows that reasonable characterization of the different open loop responses is possible using a first order plus time delay transfer function although the values of the gain and time constant are found to be strongly dependent upon the direction and magnitude of the disturbances. The initial and final values of top and bottom composition for the different step changes in the disturbances are summarized in Table 3.3

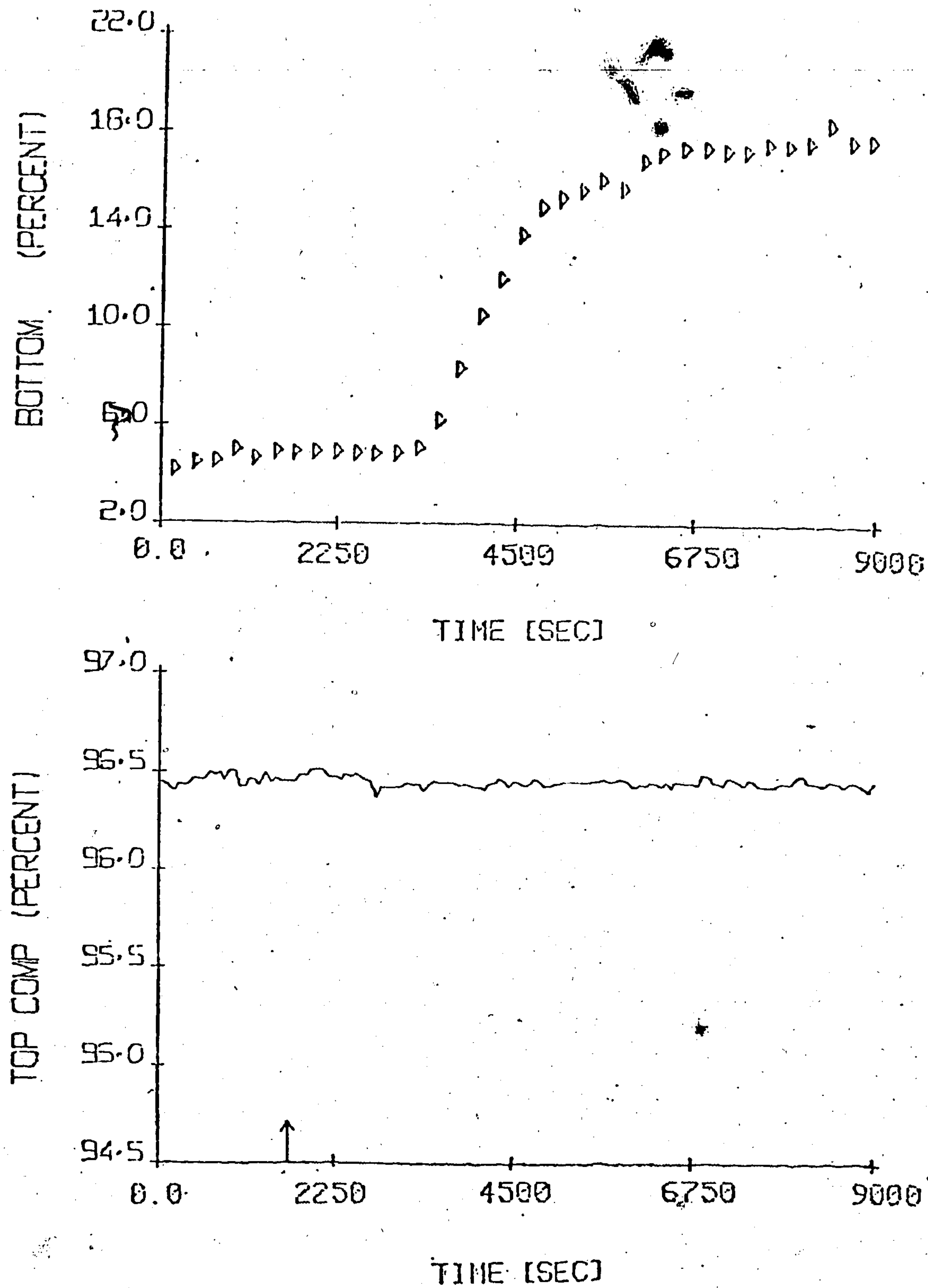


Figure 3.2 Dynamic Response of the Terminal Compositions to a Feed Flow Rate Step Change of +20%

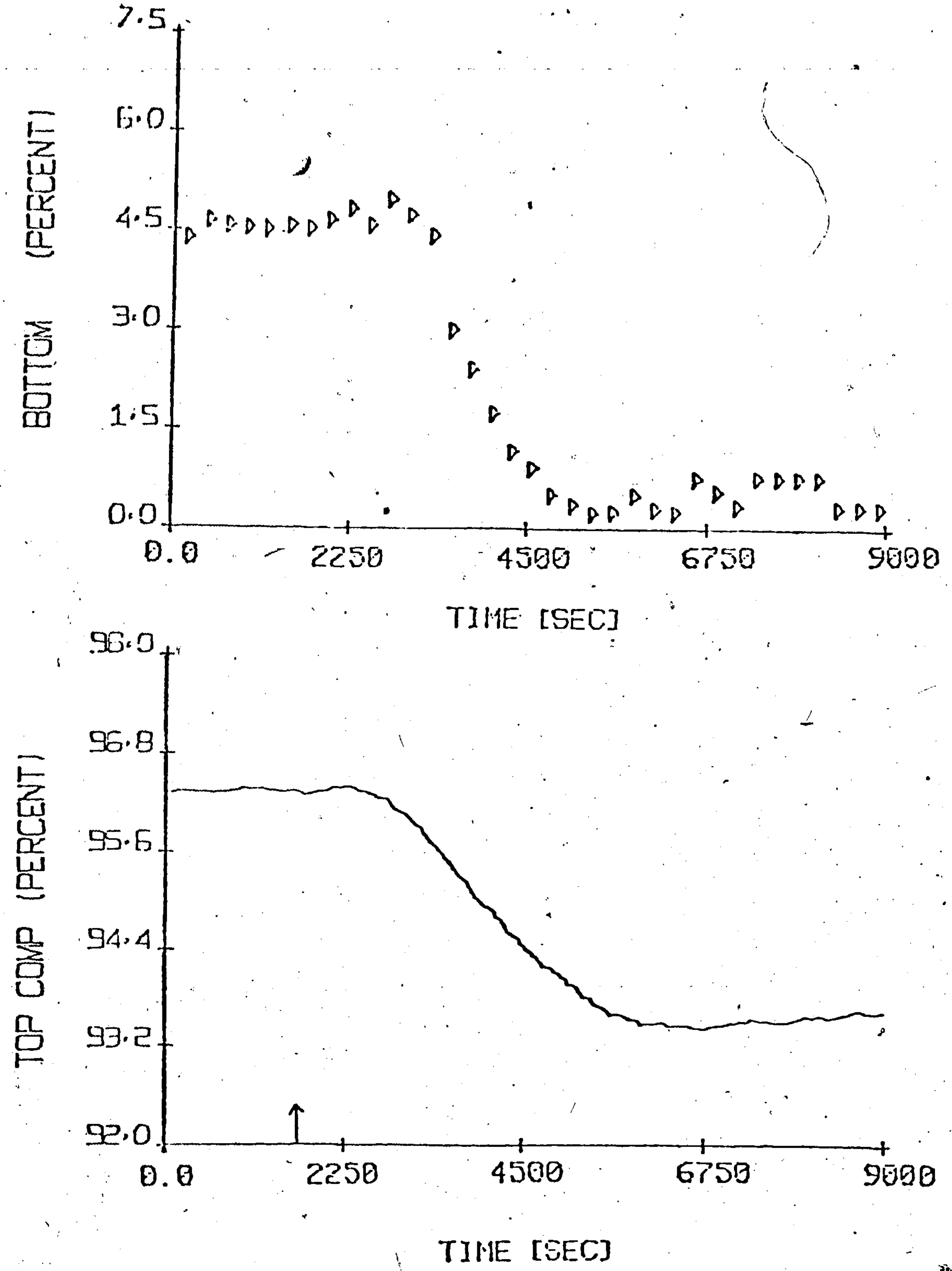


Figure 3.3 Dynamic Response of the Terminal Compositions to a Feed Flow Rate Step Change of -20%

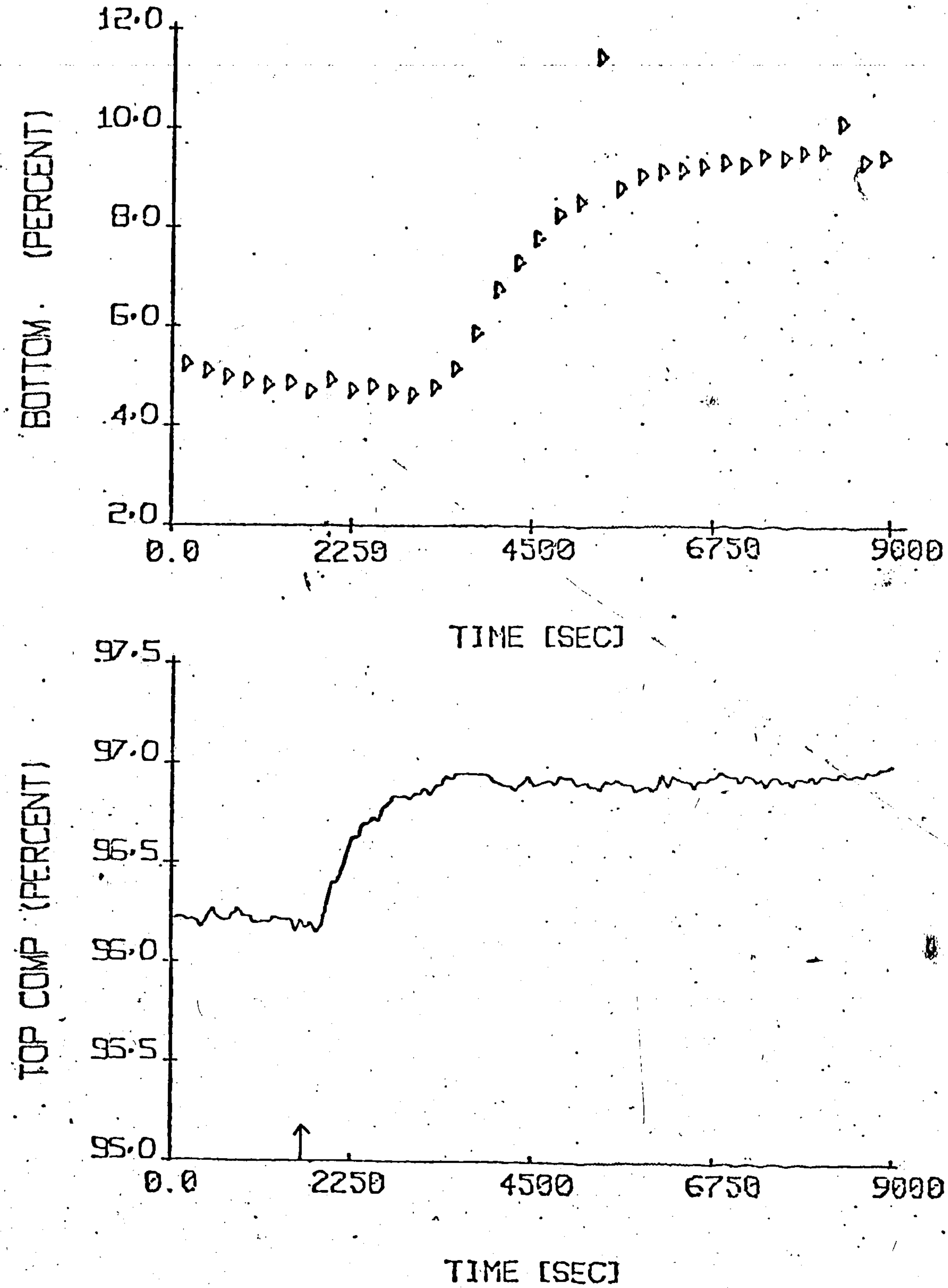


Figure 3.4 Dynamic Response of the Terminal Compositions to a Reflux Flow Rate Step Change of +10%

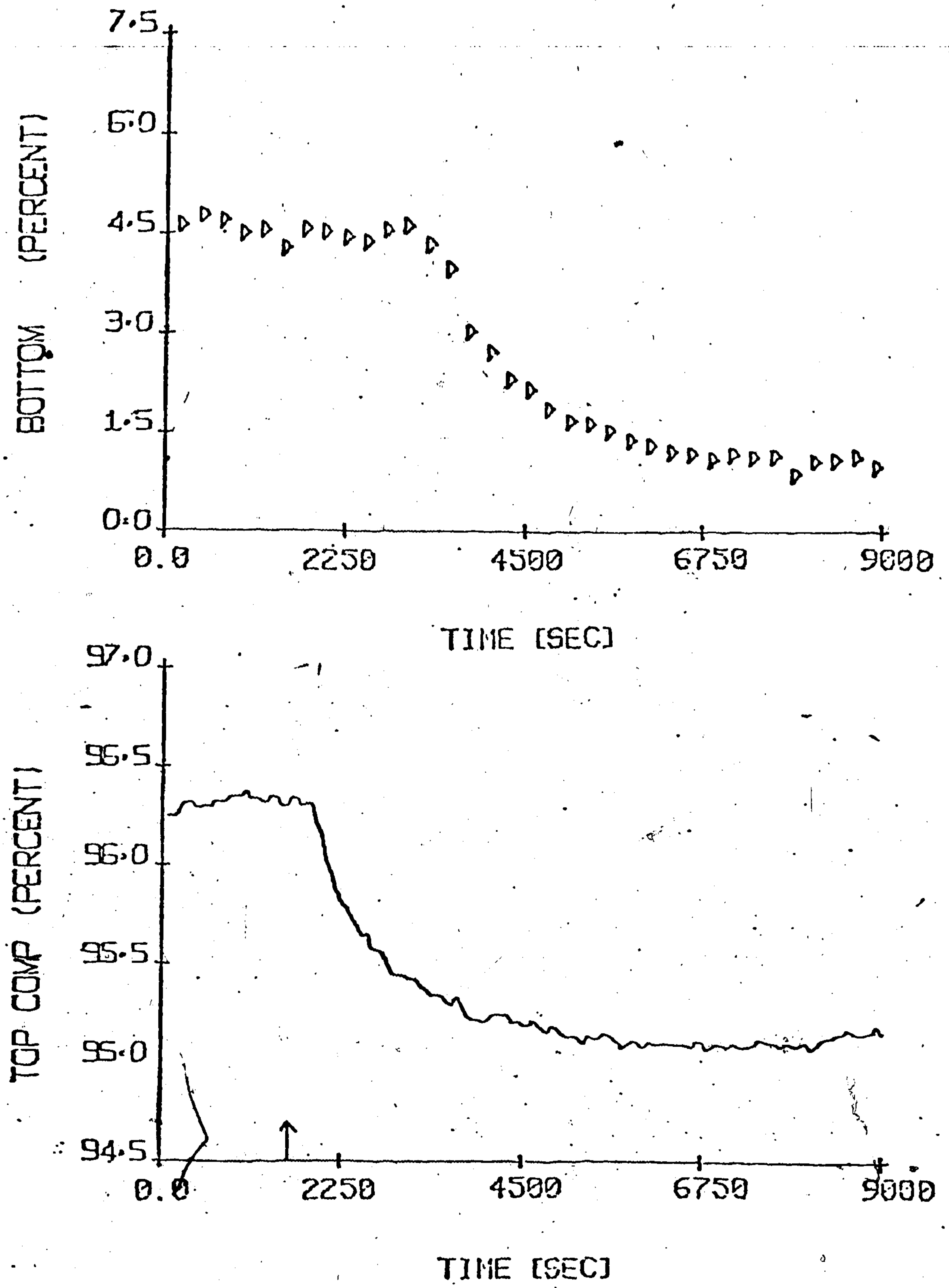
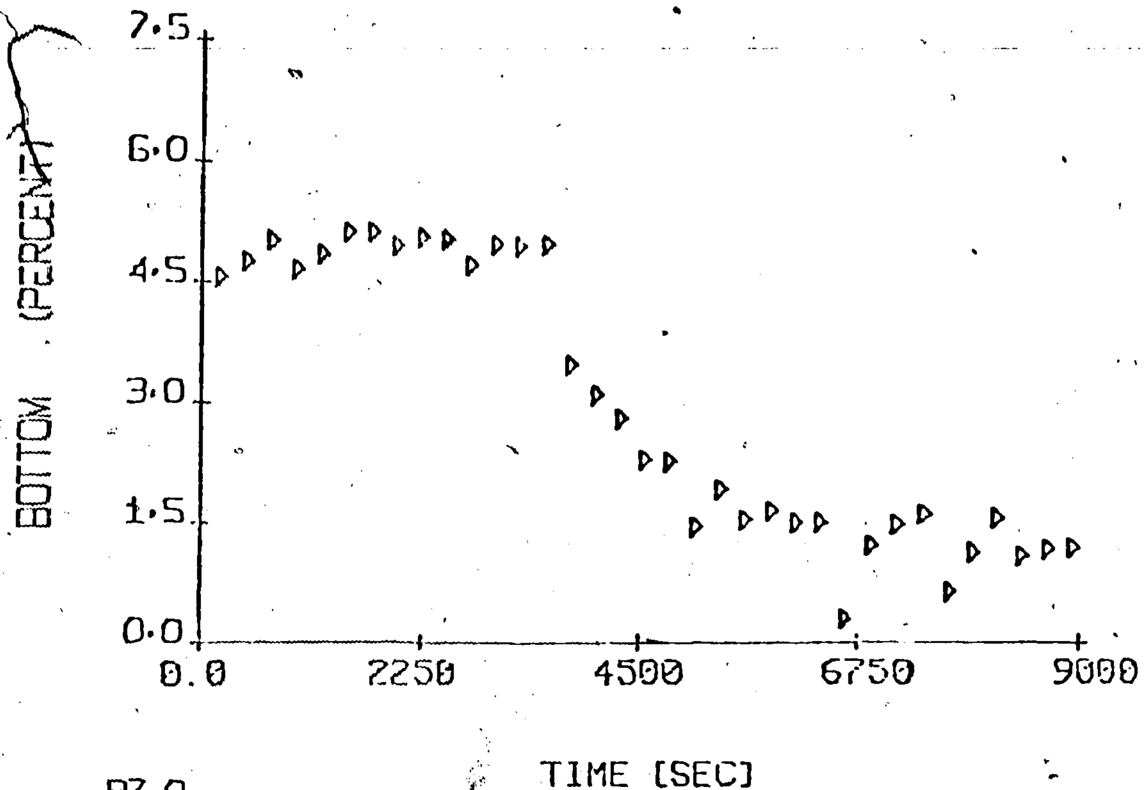


Figure 3.5 Dynamic Response of the Terminal Compositions to a Reflux Flow Rate Step Change of -10%



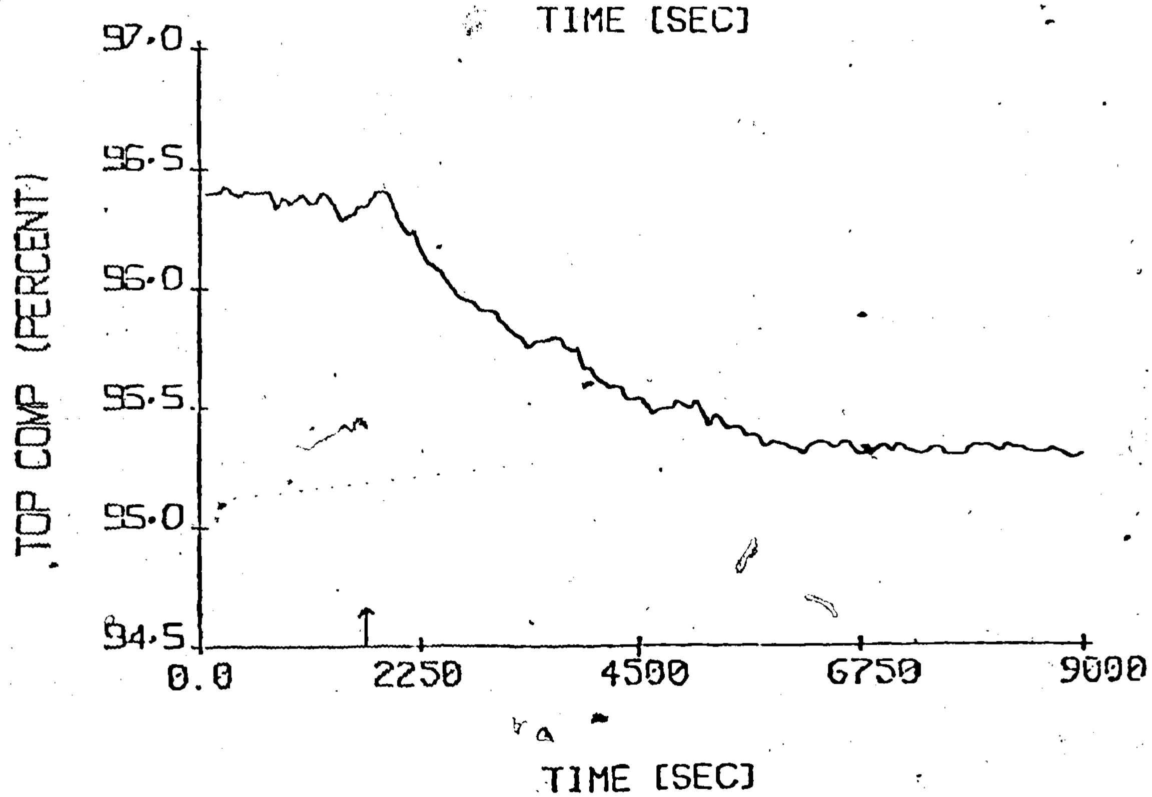


Figure 3.6 Dynamic Response of the Terminal Compositions to a Steam Flow Rate Step Change of +5%

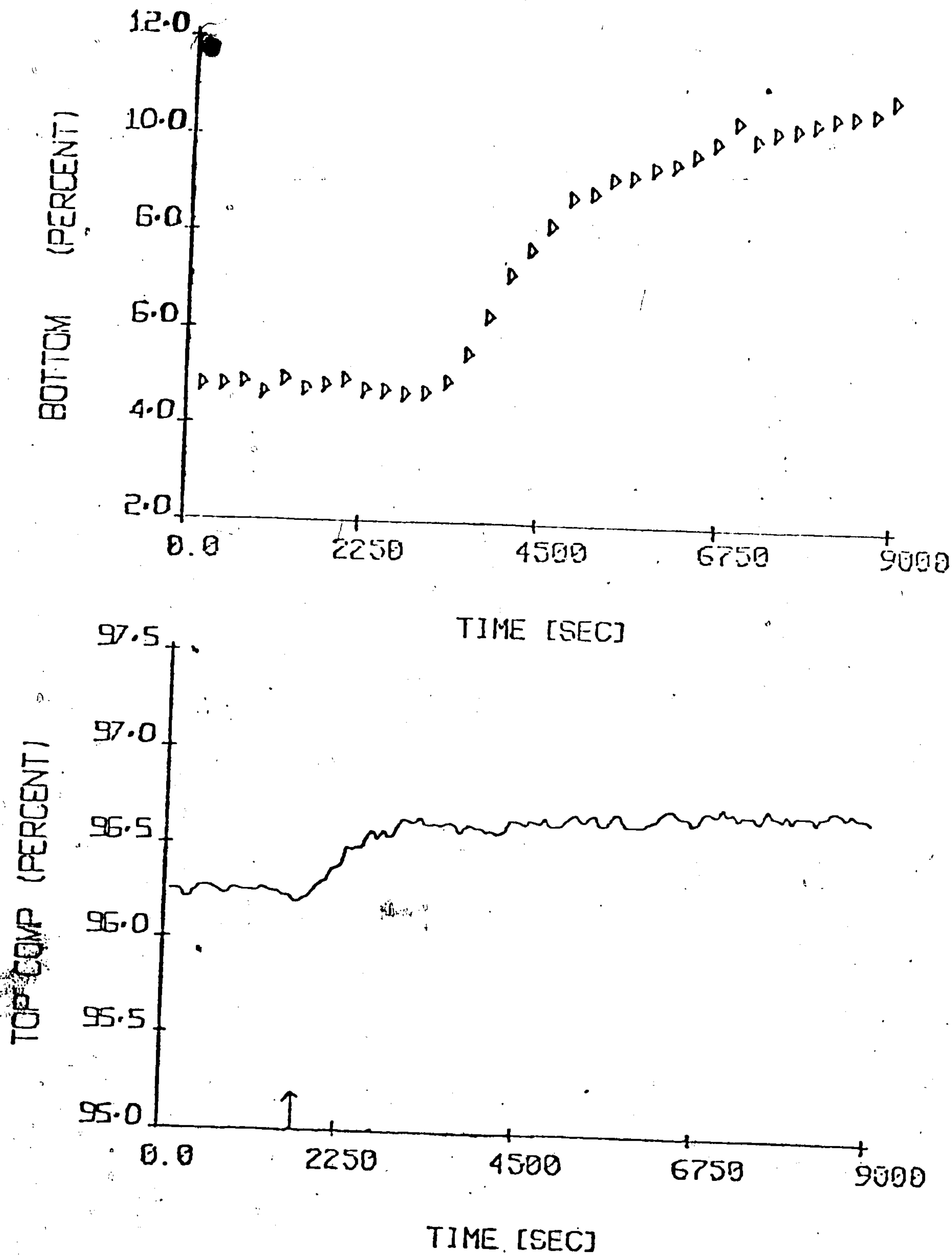


Figure 3.7 Dynamic Response of the Terminal Compositions to a Steam Flow Rate Step Change of -5%

Table 3.3 Summary Of Product Composition Changes For Step Changes In Feed, Reflux and Steam Flow Rates

Run	Type Of Disturbance	Initial Values XD	XB	Final Values XD	XB
E2	+20% Feed	96.4	4.5	96.4	17.8
E3	-20% Feed	96.4	4.5	93.6	0.3
E4	+10% Reflux	96.2	5.2	97.0	9.5
E5	-10% Reflux	96.3	4.7	95.3	1.1
E6	+5% Steam	96.4	4.5	95.3	1.1
E7	-5% Steam	96.2	4.8	96.7	10.8

3.2.2 Nonlinear Model

The simulated responses of the top and bottom composition obtained from the nonlinear model for feed flow rate step changes of ± 20% are compared to the experimental column responses in Figures 3.8 and 3.9. The output deviation responses from the nonlinear model agree reasonably closely with the experimental column outputs.

In examining the column output responses, it can be seen that the system exhibits highly nonlinear behavior. For the 20% decrese in feed flow rate, the top composition decreases by approximately three weight percent methanol, whereas for the same magnitude feed change in the opposite direction there is no effect on the top composition at all. Since the top composition proved to be insensitive to increases in feed flow rate, no control runs were performed

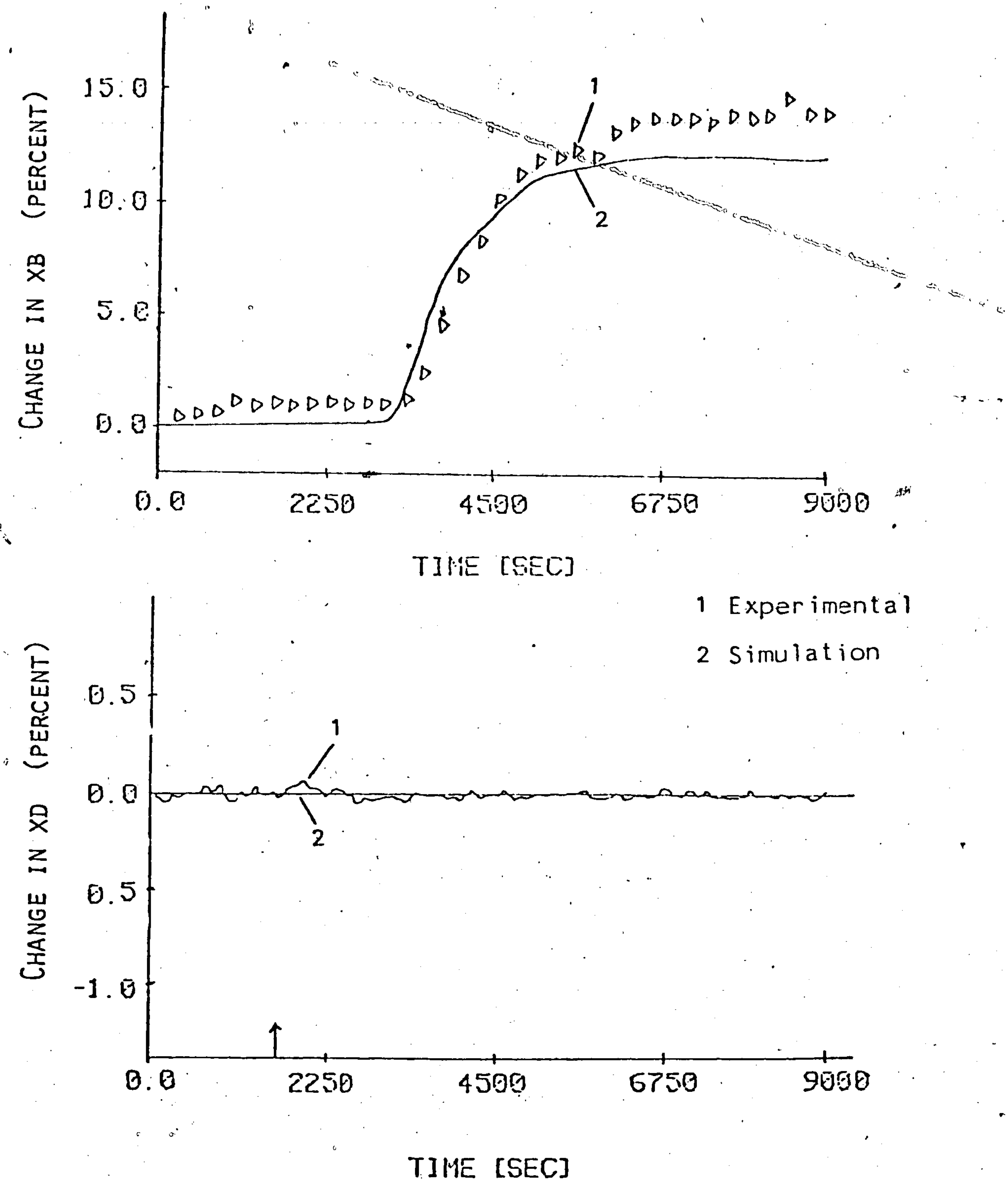


Figure 3.8 Comparison of Predicted and Experimental Dynamic Response of the Terminal Compositions to a Feed Flow Rate Step Change of + 20%

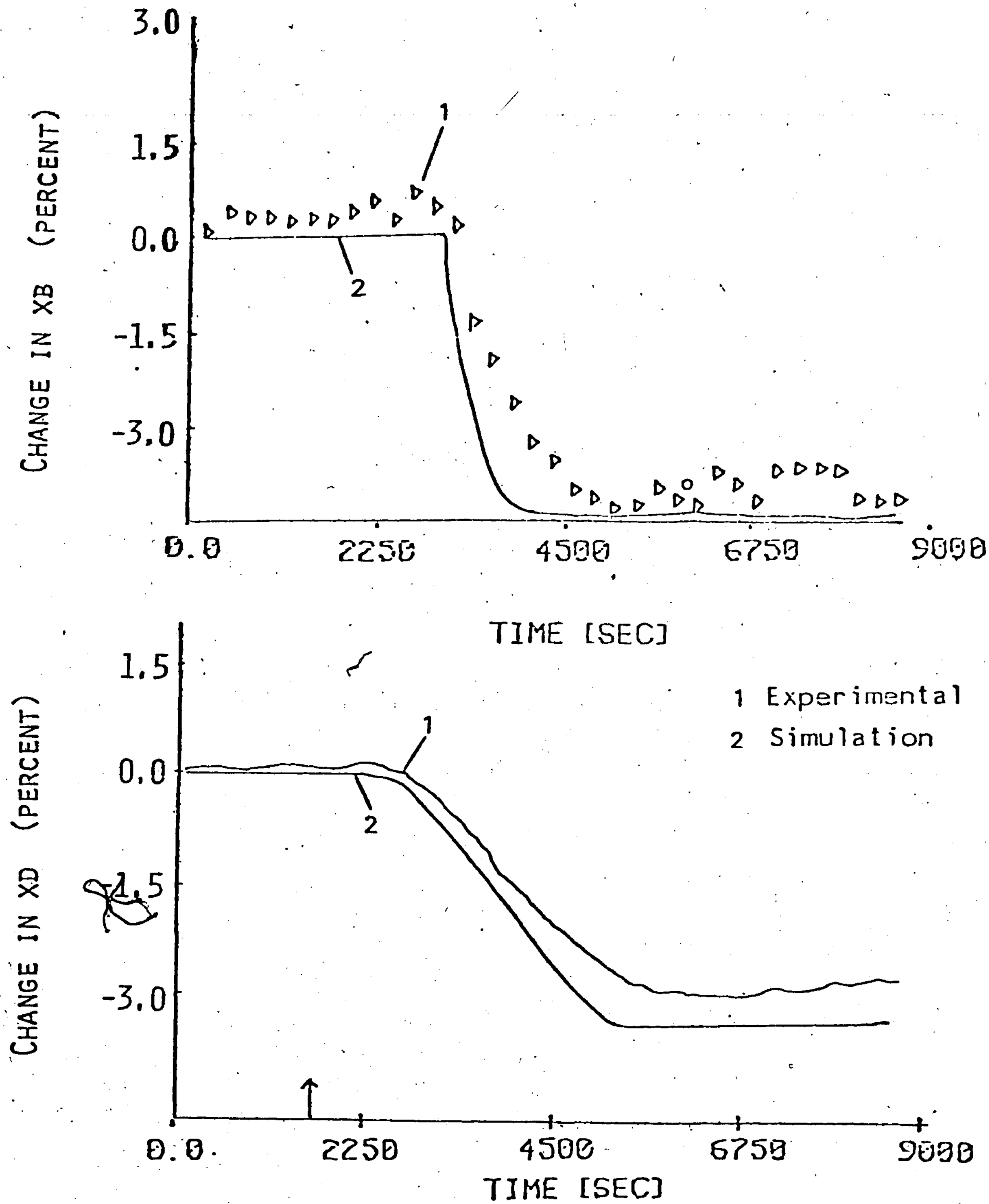


Figure 3.9 Comparison of Predicted and Experimental Dynamic Response of the Terminal Compositions to a Feed Flow Rate Step Change of -20%

for the step increase in feed flow rate.

3.2.3 Transfer Function Model

As mentioned in section 3.2.1, a first order plus time delay transfer function provides a reasonably good representation of column dynamic behavior for the conditions used. This is important since the choice of k, which represents the time delay in the STR model, is crucial. Too large a value of k can give an output variance which is larger than necessary. Also, it can prove to be difficult to get the estimates to converge at times. Too small a k makes it very difficult to get good control and may even make it impossible to stabilize the system (16).

3.2.3.1 Top Composition Time Delay

Analysis of the experimental responses in section 3.2.1, indicates that the time delay for the top product composition, (XD vs Reflux) is nearly constant at one minute. Therefore with a control cycle of sixty-four seconds, k was chosen to be one.

3.2.3.2 Bottom Composition Time Delay

The dynamic responses of the bottom composition to steam flow rate step changes of ± 5% provide a good indication of the time delay, (XB vs Steam). The total time delay is estimated to be between five to six sampling intervals. The system time delay is due to

(i) The gas chromatograph analysis time

(ii) The process time delay

(iii) The transport time delay within the sampling system

The analysis of the system used to establish the total time delay is given in Appendix B.

3.3 Composition Measurements

The accuracy of terminal composition measurements and the reliability of the measuring devices are vital to the control system.

The use of the capacitance probe provides continuous measurement for top product composition. This continuous signal is sampled at 64 second intervals for data logging and control purposes. This method of analysis provides good accuracy for solutions of high methanol content which is the case for the top product.

A Hewlett-Packard model 5720A gas chromatograph with an automatic sampling valve is used for bottom composition measurements since the capacitance method of analysis is not satisfactory for streams with a low methanol content (18).

3.3.1 Process and Measurement Noise

High frequency noise was found in both terminal compositions as can be seen from Figures 3.10 and 3.11. The noise is in part due to the measuring devices, but it is mainly caused by the noisy manipulated variables, steam and

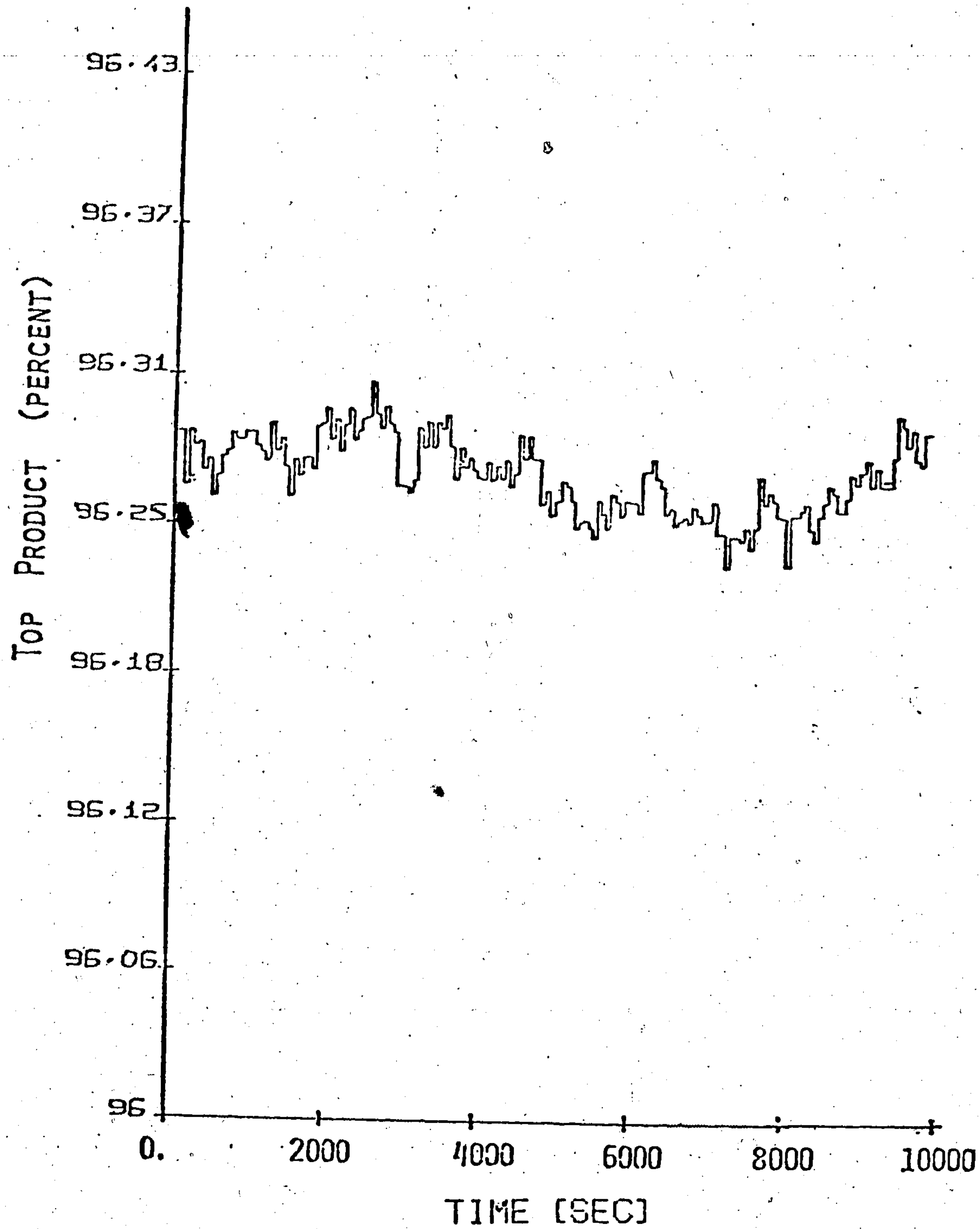


Figure 3.10 Experimental Steady State Top Product Composition Noise Display

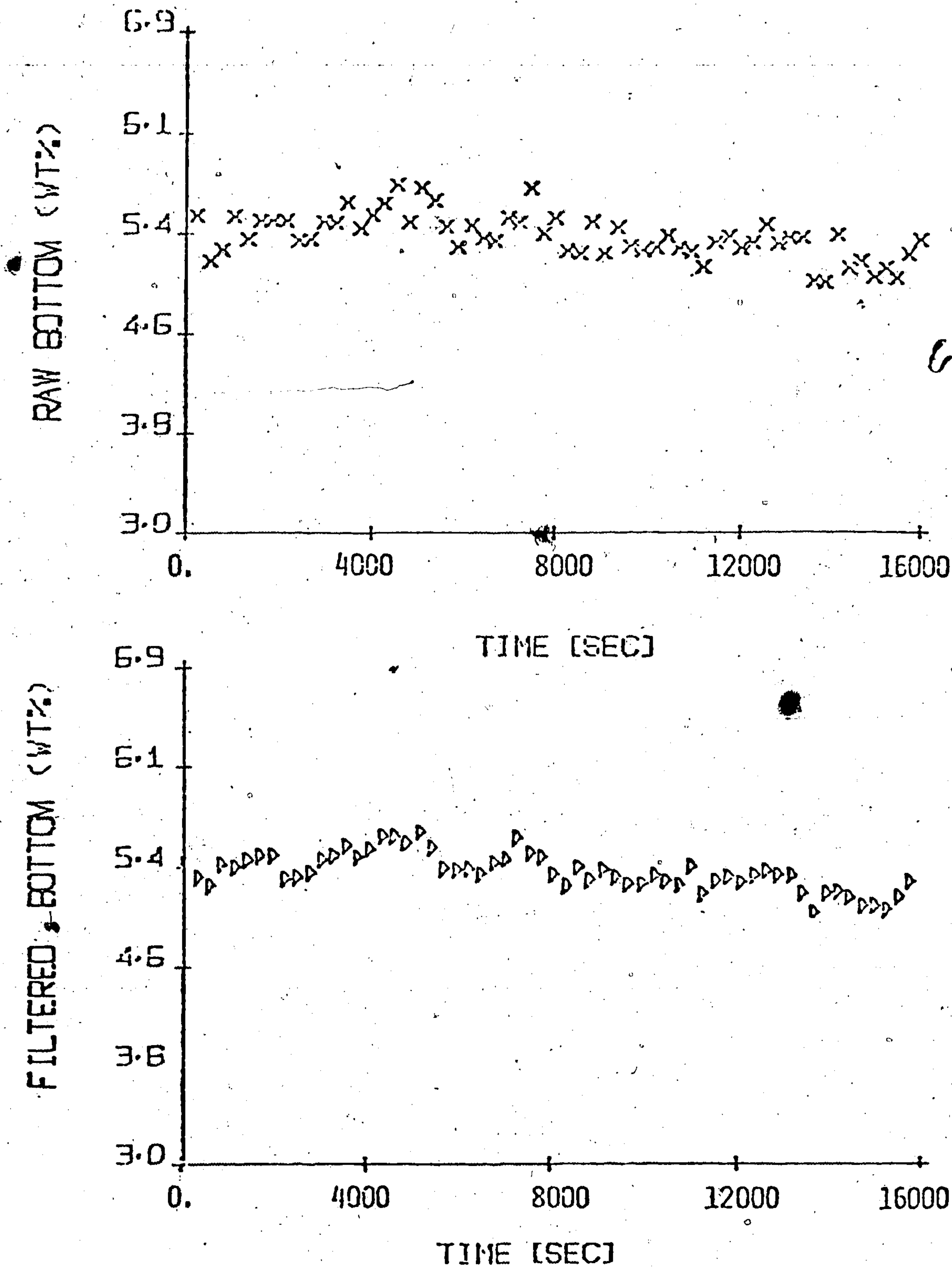


Figure 3.11 Experimental Steady State Bottom Product Composition Noise Display

reflux flows as shown in Figures 3.12 and 3.13.

The high noise level in the steam flow rate and the bottom composition measurement together with a time delay that is larger than the system time constant have a detrimental effect on the performance and stability of the bottom composition control system..

A summary of all the measured variables of the distillation column with their standard deviation is given in Table 3.4.

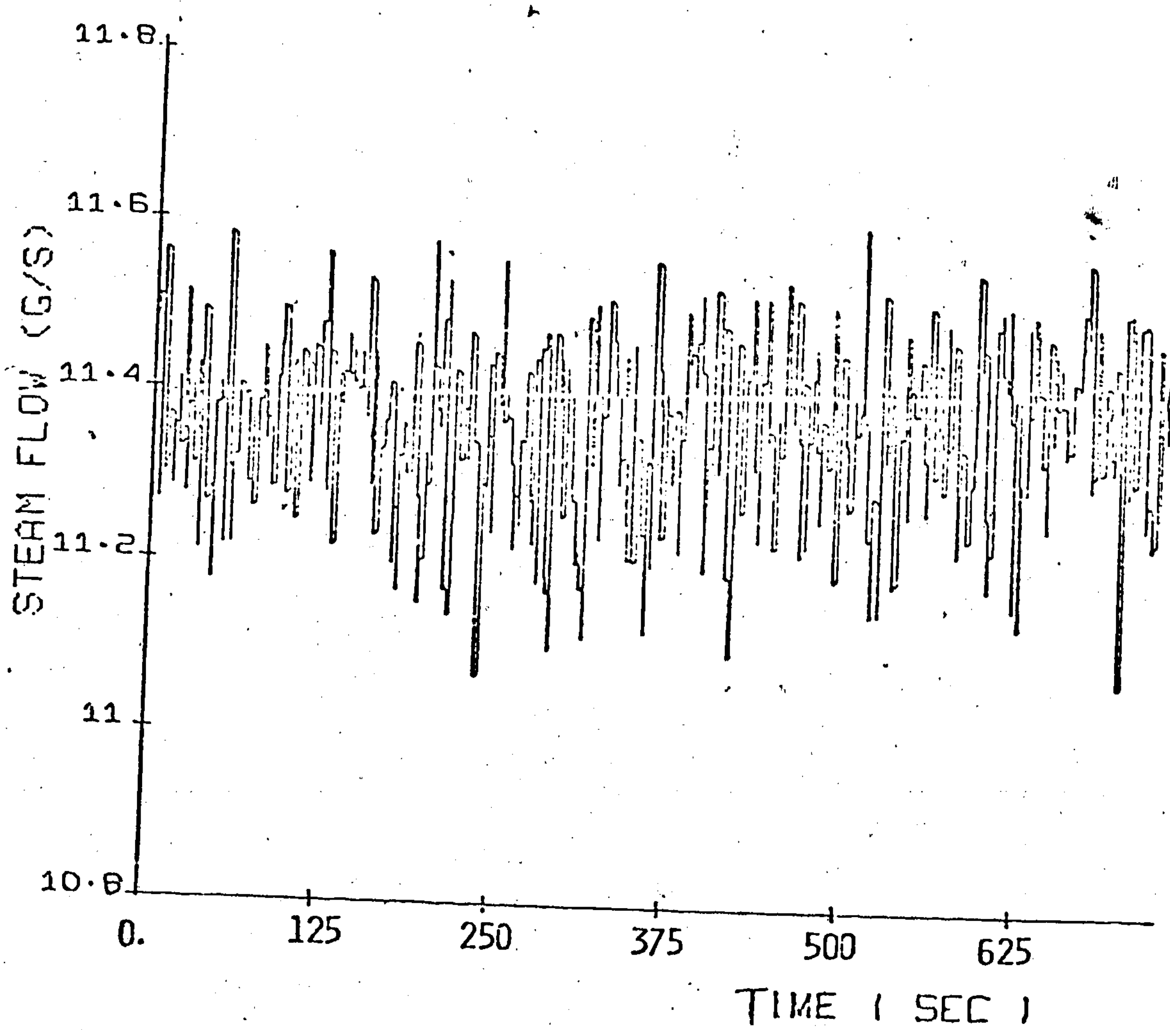


Figure 3.12 Experimental Steady State Steam Flow Noise Display

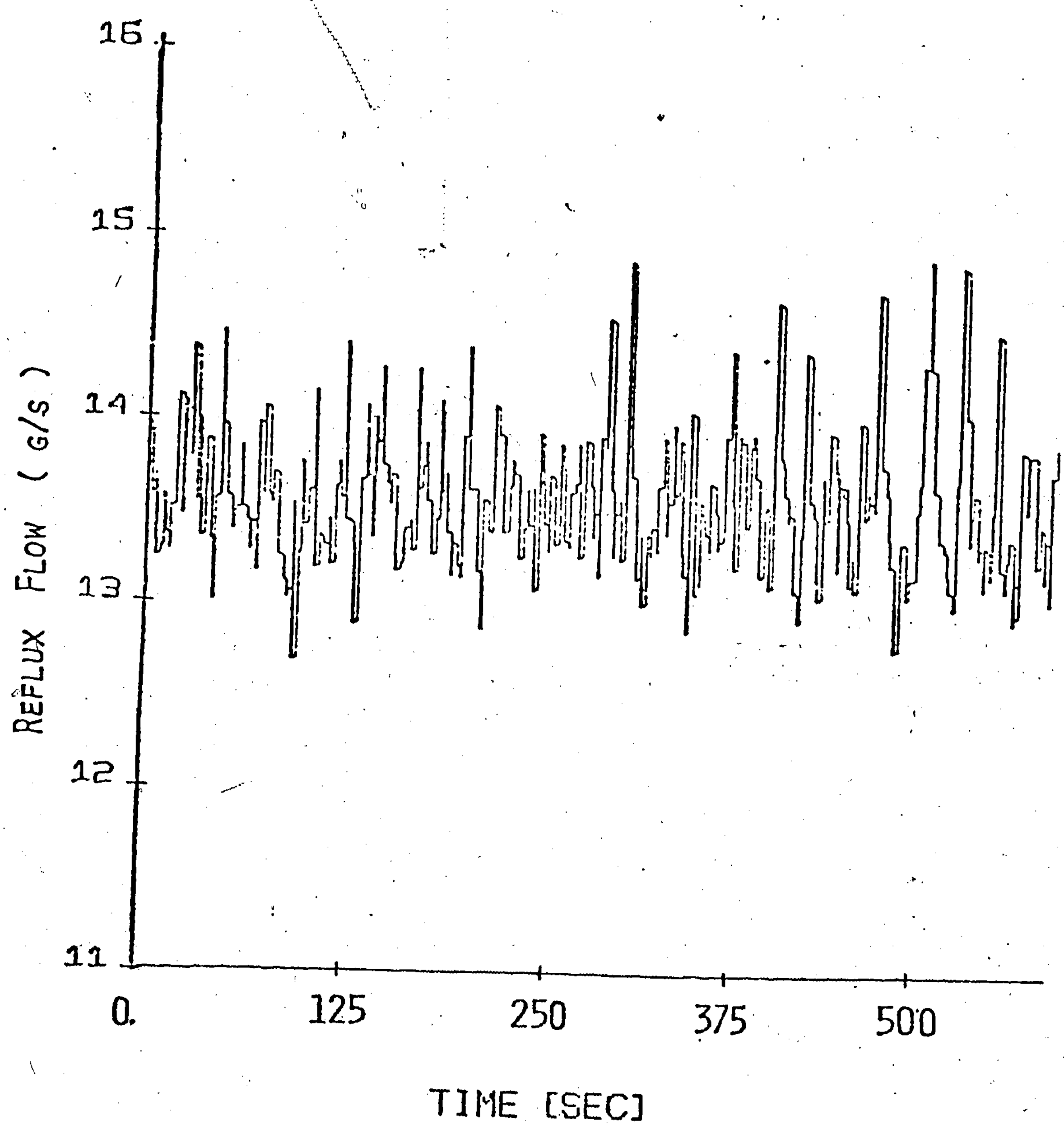


Figure 3.13 Experimental Steady State Reflux Flow Noise Display

TABLE 3.4 SUMMARY OF THE DISTILLATION COLUMN MEASURED VARIABLES (WITH STANDARD DEVIATION BASED ON 50 SAMPLES PER VARIABLE AT STEADY STATE CONDITIONS)

FEED FLOW	= 18.012	G/SEC	DEV= 0.0373
REFLUX FLOW	= 10.304	G/SEC	DEV= 0.1301
STEAM FLOWW	= 11.534	G/SEC	DEV= 0.5275
BOTTOM PROD	= 9.099	G/SEC	DEV= 0.2163
TOP PROD	= 8.750	G/SEC	DEV= 0.0610
COOL WATER	=472.685	G/SEC	DEV= 1.9260
DIST COMP	= 96.583	WT% MEOH	DEV= 0.1523
BOTTOMS COMP	= 5.400	WT% MEOH	DEV= 0.0000
FEED COMP	= 49.250	WT% MEOH	DEV= 0.0000
PRESSIRE	= 0.491	KPA	DEV= 0.0585
COND LEVEL	= 16.913	CM	DEV= 0.0449
REB'R LEVEL	= 59.479	CM	DEV= 0.3310
DIFF PRESS	= 5.013	KPA	DEV= 0.0041
REB'R O'HEAD	= 94.5	DEG C	DEV= 0.2145
PLATE 1 TEMP	= 85.3	DEG C	DEV= 0.2091
PLATE 2 TEMP	= 79.0	DEG C	DEV= 0.1661
PLATE 3 TEMP	= 75.1	DEG C	DEV= 0.1945
PLATE 4 TEMP	= 74.1	DEG C	DEV= 0.1881
PLATE 5 TEMP	= 70.5	DEG C	DEV= 0.2120
PLATE 6 TEMP	= 67.9	DEG C	DEV= 0.2030
PLATE 7 TEMP	= 66.2	DEG C	DEV= 0.1860
PLATE 8 TEMP	= 64.5	DEG C	DEV= 0.1861
COND TEMP	= 61.9	DEG C	DEV= 0.1794
STEAM TEMP	= 89.7	DEG C	DEV= 0.2476
COND'T TEMP	= 104.4	DEG C	DEV= 0.2887
REFLUX TEMP	= 48.5	DEG C	DEV= 0.1984
FEED TEMP	= 40.7	DEG C	DEV= 0.2053
BOTTOMS TEMP	= 50.9	DEG C	DEV= 0.2142
REB'R TEMP	= 93.9	DEG C	DEV= 0.2300
FEED INLET	= 72.9	DEG C	DEV= 0.2535
REFLUX INLET	= 64.4	DEG C	DEV= 0.7267
COL O'HEAD	= 64.9	DEG C	DEV= 0.1798
WATER INLET	= 9.4	DEG C	DEV= 0.1782
WATER OUTLET	= 21.3	DEG C	DEV= 0.2021

4. TOP PRODUCT COMPOSITION CONTROL

4.1 System Description

A binary solution of methanol and water was fed to the fourth tray of the eight tray distillation column. The high methanol concentration overhead vapor was condensed in the total condenser. Condenser level was controlled by top product flow with the reflux flow chosen to control the top product composition.

Most of the column's instrumentation is pneumatic with analog controllers in each of the flow control loops. The column was operated in an analog supervisory control mode as interfaced to the IBM 1800 process computer.

4.2 Experimental Evaluation

4.2.1 System Configuration and Implementation

All process measurement signals were transduced to the computer for logging and/or control purposes. The control law calculation was scheduled to run every 64 seconds with the required inputs being the top product composition and reflux flow rate. The IBM 1800 Direct Digital Control (DDC) system has a basic scan frequency of one second. Each point in the data base was scanned at the specified poll time which could be any integer multiple of the basic scan frequency. The raw data information was processed, converted, and put in the data base. At the scheduled interval, the user written control law algorithm calculated the error difference between the target composition set point and the measured composition signal and then calculated the required control action of the manipulated variable, reflux flow rate. Limit checking was performed before the calculated value was passed for transmission as the set point to the reflux flow controller. Figure 4.1 is a schematic representation of the system which shows the interface between the process and the computer. The source listing of the complete set of control programs is given in Appendix C.

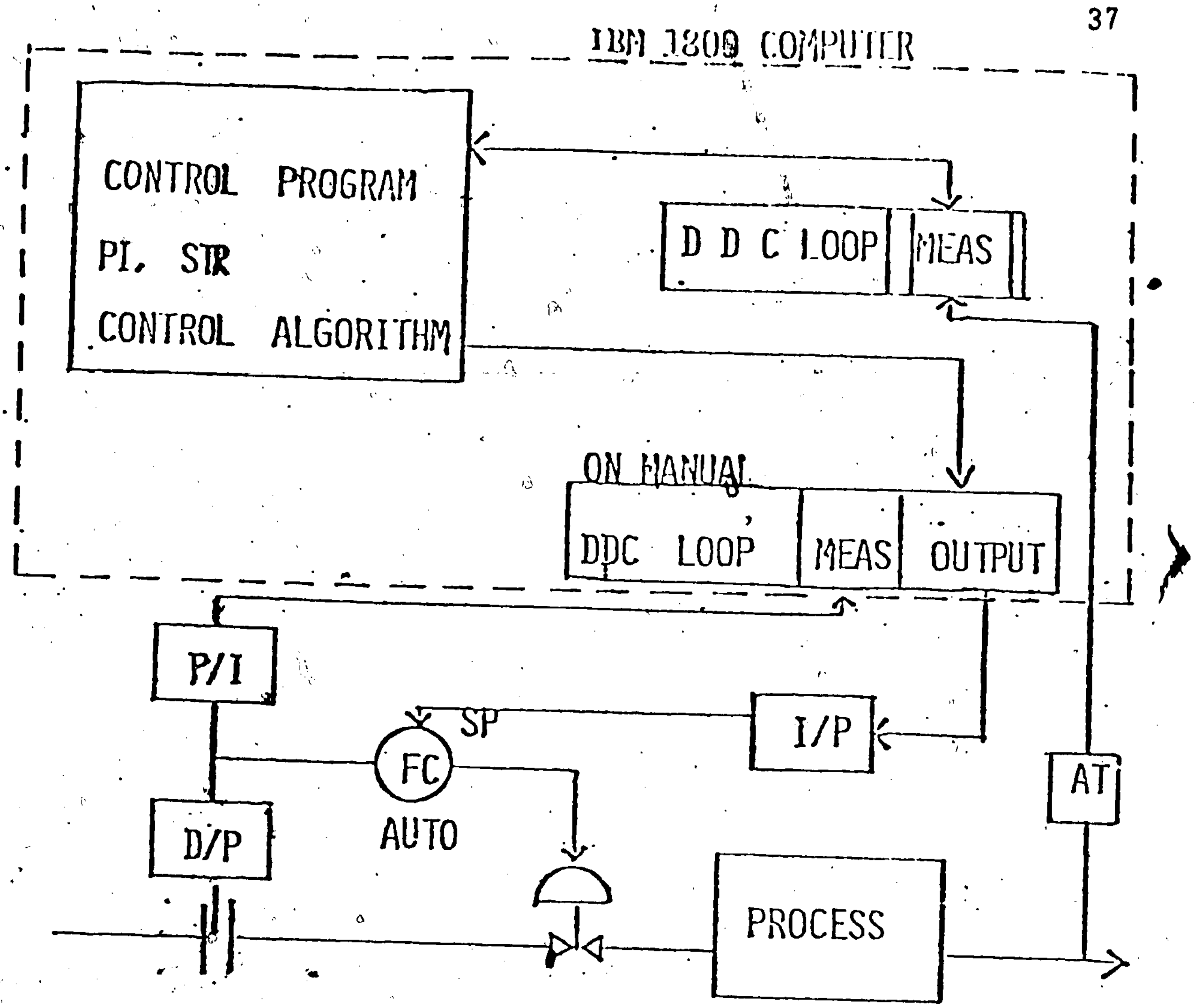
4.2.2 Control Strategy

4.2.2.1 Proportional Plus Integral Control

In order to achieve a valid comparison between the STR and PI control performance, the control performance achieved using a well-tuned PI controller was used as a reference in evaluating STR control behavior. The controller's setting was determined using the Ziegler-Nichols tuning method on the actual column. Tuning details can be found in Appendix D.

4.2.2.2 Self-Tuning Regulator

With the real time data base and the information vector of previous input and output signals, model parameters α_i and β_i were estimated at every sampling interval. At the same instant, these estimated parameters were used for calculation of the control signal to minimize output



D/P	DIFFERENTIAL PRESSURE TRANSMITTER
P/I	PRESSURE TO CURRENT CONVERTER
I/P	CURRENT TO PRESSURE CONVERTER
FC	ANALOG FLOW CONTROLLER
SP	SET POINT
MEAS	MEASUREMENT
AT	ANALYZER TRANSMITTER

Figure 4.1 Process & Computer Interface For Analog Supervisory Control

variance of the system. Parameters required for operation of the algorithm that were selected for use with the base run were:

model order	$N = 2$
scaling factor	$\beta_0 = 0.01$
initial P(0) value	$\underline{P}(0) = 3{,}000\ \underline{I}$
time delay	$k = 1$
initial parameter values	$\underline{\theta}(0) = 0.$
exponential forgetting factor	$\mu = 1.0$

These values were determined by digital simulation prior to the experimental testing.

Note that a model order of two was selected to allow for variation in the gain and time constant of the system despite the fact that a first order plus time delay model was adequate for characterizing a particular response. This was considered necessary because the gain and time constants were found to be strongly dependent upon the direction and magnitude of the disturbances.

4.2.3 Results

Tests were performed for a feed flow step decrease of twenty percent and step changes in composition set points of +1.0% and -1.5% from the steady state values. The performance of both the PI and STR control algorithms was evaluated for these changes.

When the column reached steady state at the required operating condition, the control task was initiated via the teletype terminal. It was at this instant that run time, time and magnitude of the disturbance, control option, and controller constants were specified. For all experimental runs, a run time of 180 time units was used. Each time unit spanned 64 seconds. The disturbance or change in set point was introduced at the 30th time unit from the start of the run. The teletype input for a PI control test run for a 1% set point increase is given in Figure 4.2.

4.2.3.1 Proportional Plus Integral Control Tests

Ziegler-Nichols tuning of the top product composition control yielded a controller setting with a proportional gain of K_p=-45, and a reset time T_R=5.84 minutes. A summary of the excellent results is given in Table 4.1 with the controlled responses plotted in Figures 4.3 to 4.5.

Table 4.1 Summary Of Top Product Experimental PI Control Tests

Run	K_p	T_R	ISE	Figure	Disturbance
E8	-45.	5.84	0.0000	4.3	Feed Flow -20%
E9	-45.	5.84	0.0002	4.4	Set Point + 1%
E10	-45.	5.84	0.0014	4.5	Set Point - 1.5%

4.2.3.2 Experimental Self-Tuning Regulator Tests

i) Effect of β_0:

Earlier studies (5) had indicated that the choice of β_0 was not crucial, though it should be selected in a manner to

```
>ØRKTØ1

   RUN TIME(IMTS),PRINT CNTRL(KPRT),BOTMS SAMPLING(ISMP)

>180 5 4

   T COMP=96.160 RFLX=10.330 B COMP= 5.540 STM=11.050 FEED=17.980
   KTYP--T COMP SP,2 B COMP SP,3 FEED CHANGE AND MVR-- 0 STR  1 MVR

>1 0
   DISTURBANCE FROM(KDIS) AND VALUE(DIST)

>30 97.16
   KCONT=1-T STR B OPN,2-T PI B OPN,3-T OPN B STR,4-T OPN B PI
         5-T STR B PI,6-BOTH PI,7-BOTH STR,8-T PI B STR

>2
   SPECIFY KPT,KIT

>-45. -7.7

          **PI ON TOP AND BOTTOM COMP OPEN**
      KPT=-45.0000 KIT= -7.7000

      CONTROL ACTION INITIATED AT 17 59 28
```

Figure 4.2 Teletype Input For A PI Controller Test For A 1% Step Increase In Set Point

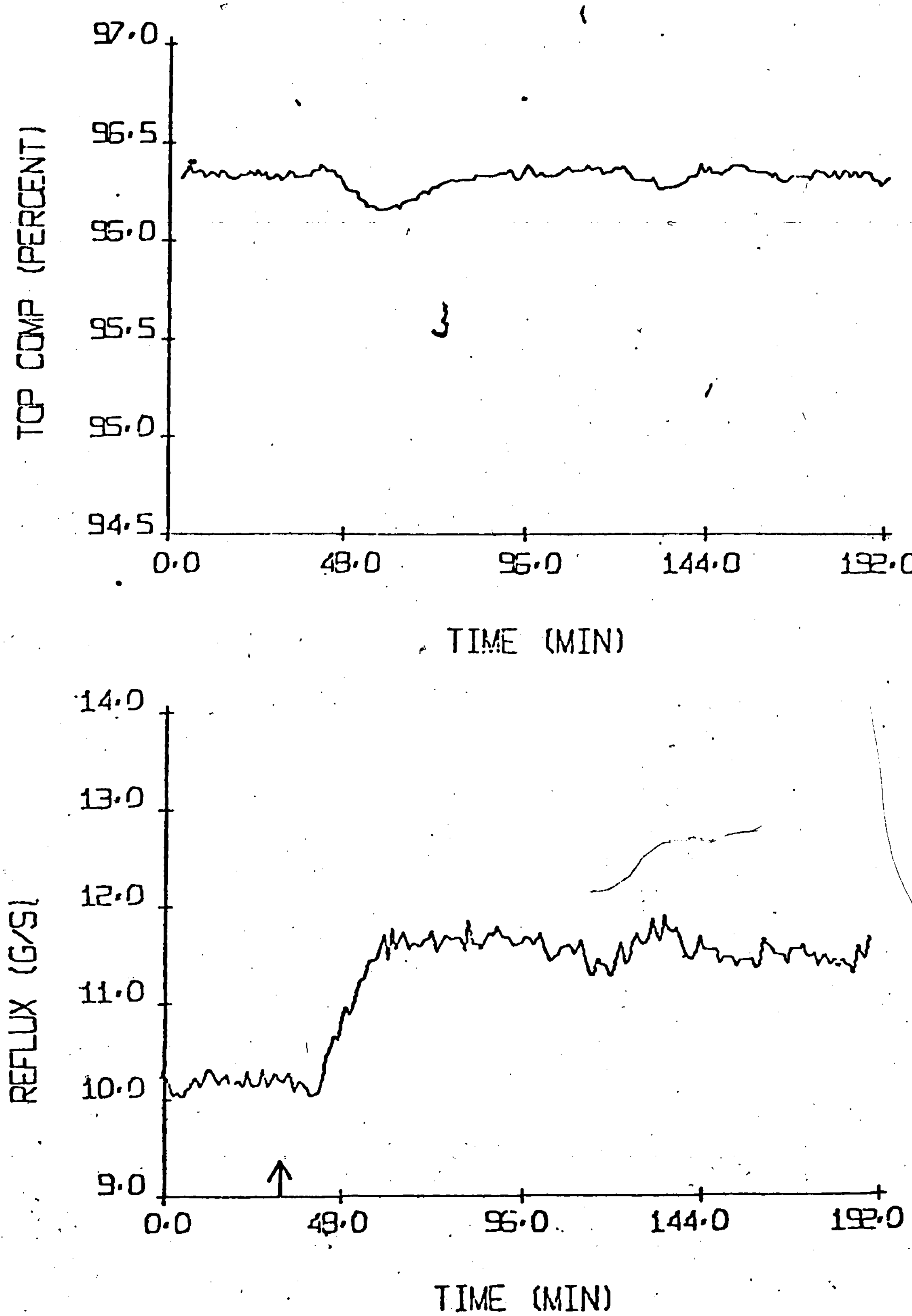


Figure 4.3 Experimental PI Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate

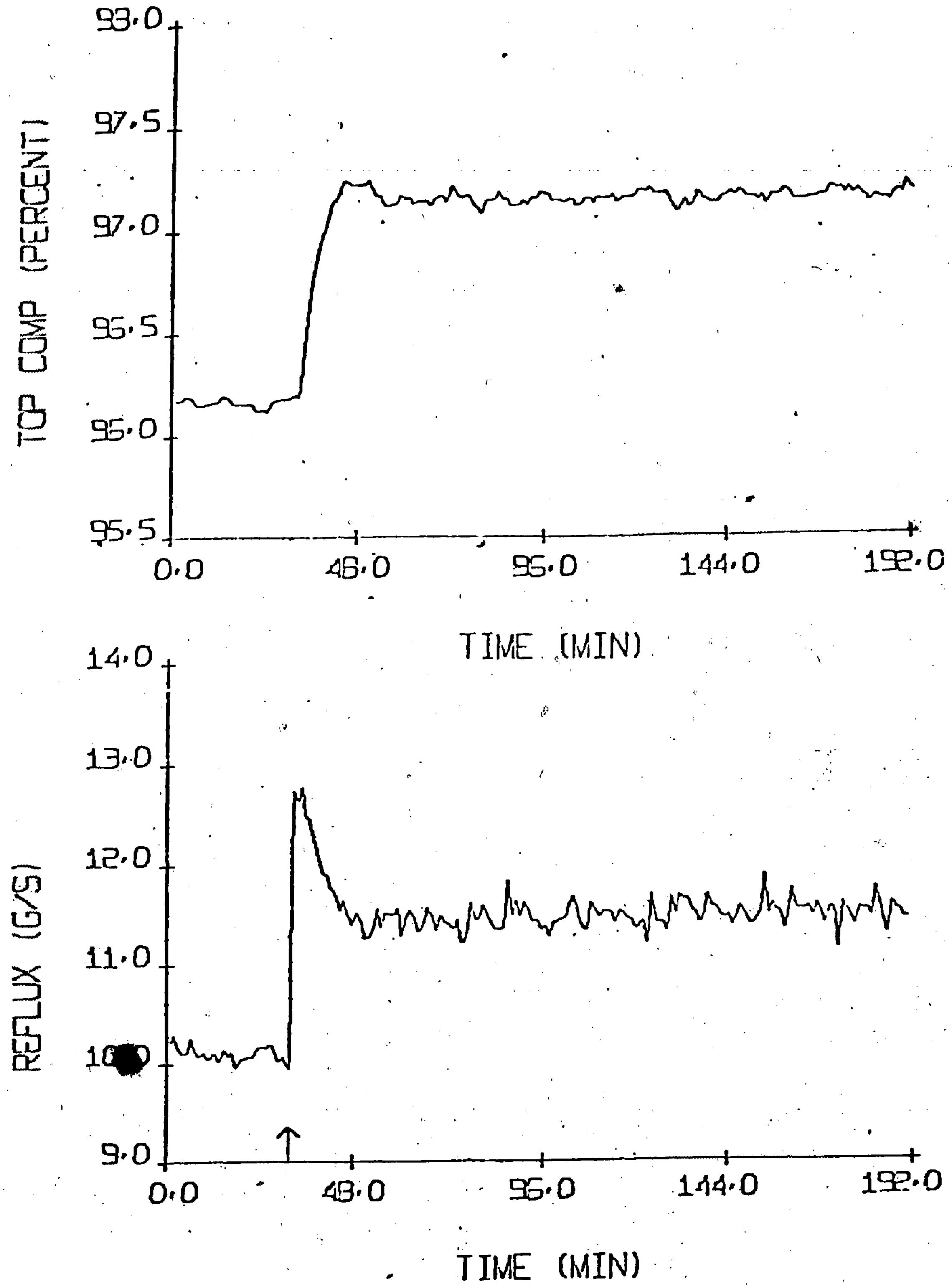


Figure 4.4 Experimental PI Control Of Top Product Composition for a +1% Step Change In Set Point

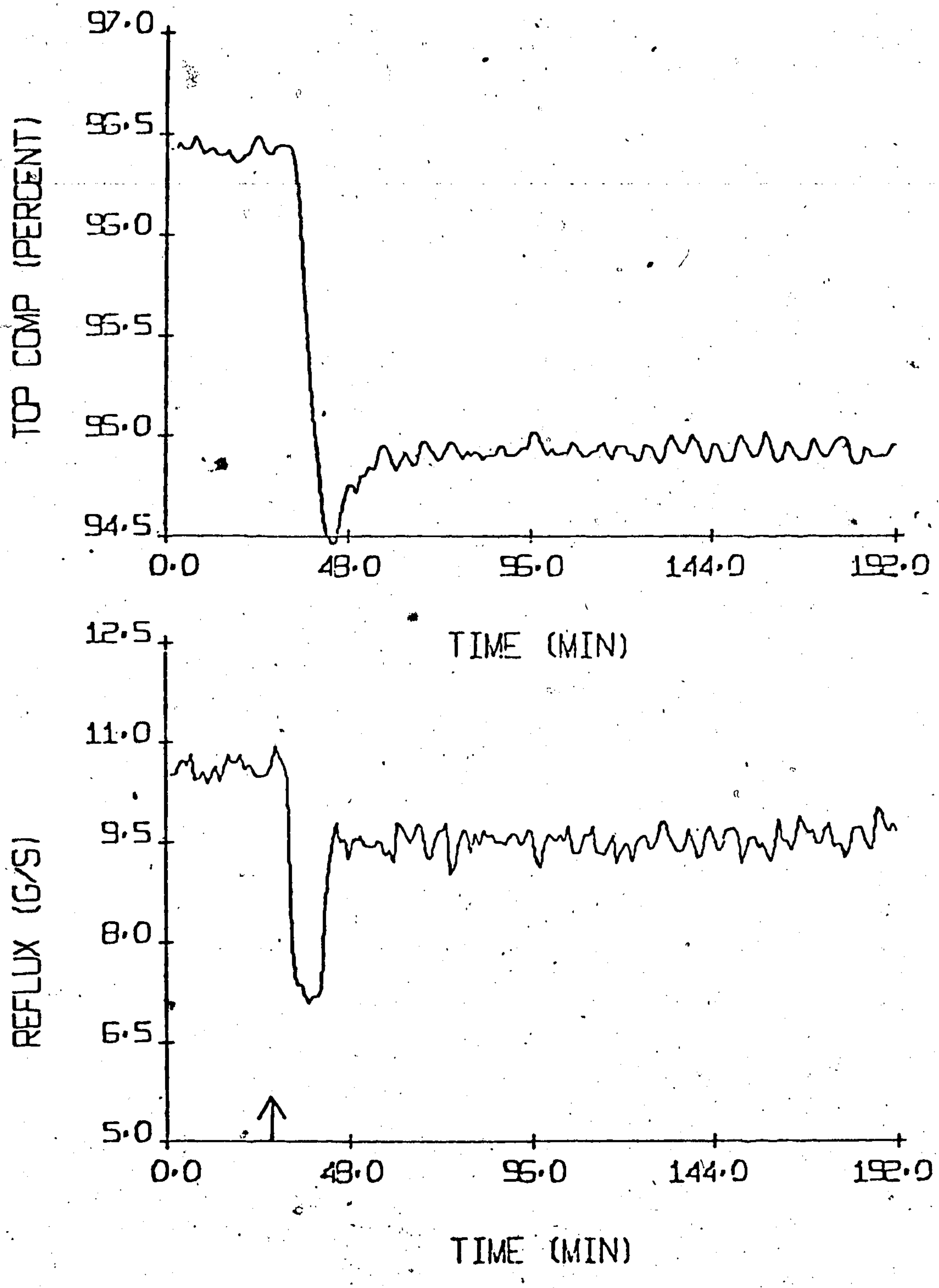


Figure 4.5 Experimental PI Control Of Top Product Composition For a -1.5% Step Change In Set Point

avoid instability. However, the results of this study indicate that the choice of the β_o value significantly influences the control behavior. The tests listed in Table 4.2 were made to study the effect of β_o on control performance. The output responses are plotted in Figures 4.6 to 4.11.

Table 4.2 Effect Of β_o Value On Experimental STR Control Of Top Product Composition

M = 2 k = .1 L = 3 $\underline{\theta}(0) = 0.0$

Run	β_o	$\underline{P}(0)$	ISE	Figure
Feed Flow -20%				
E11	0.05	250 I	$33.\times10^{-4}$	4.6
E12	0.01	250 I	$22.\times10^{-4}$	4.7
E13	0.005	250 I	$14.\times10^{-4}$	4.8
Set Point +1%				
E14	0.05	3,000 I	$11.\times10^{-4}$	4.9
E15	0.01	3,000 I	$7.\times10^{-4}$	4.10
E16	0.005	3,000 I	$6.\times10^{-4}$	4.11

The results in all runs showed that the control behavior was significantly improved by lowering the β_o value. The maximum deviation of the composition from the set point value for the feed flow rate changes, decreased as the value of β_o was decreased. The maximum deviation of the composition decreased from 1.3 to 0.8 weight percent with β_o changed from a value of 0.05 to 0.005 as shown in Figures 4.6 and 4.8. In the case of the set point changes, the

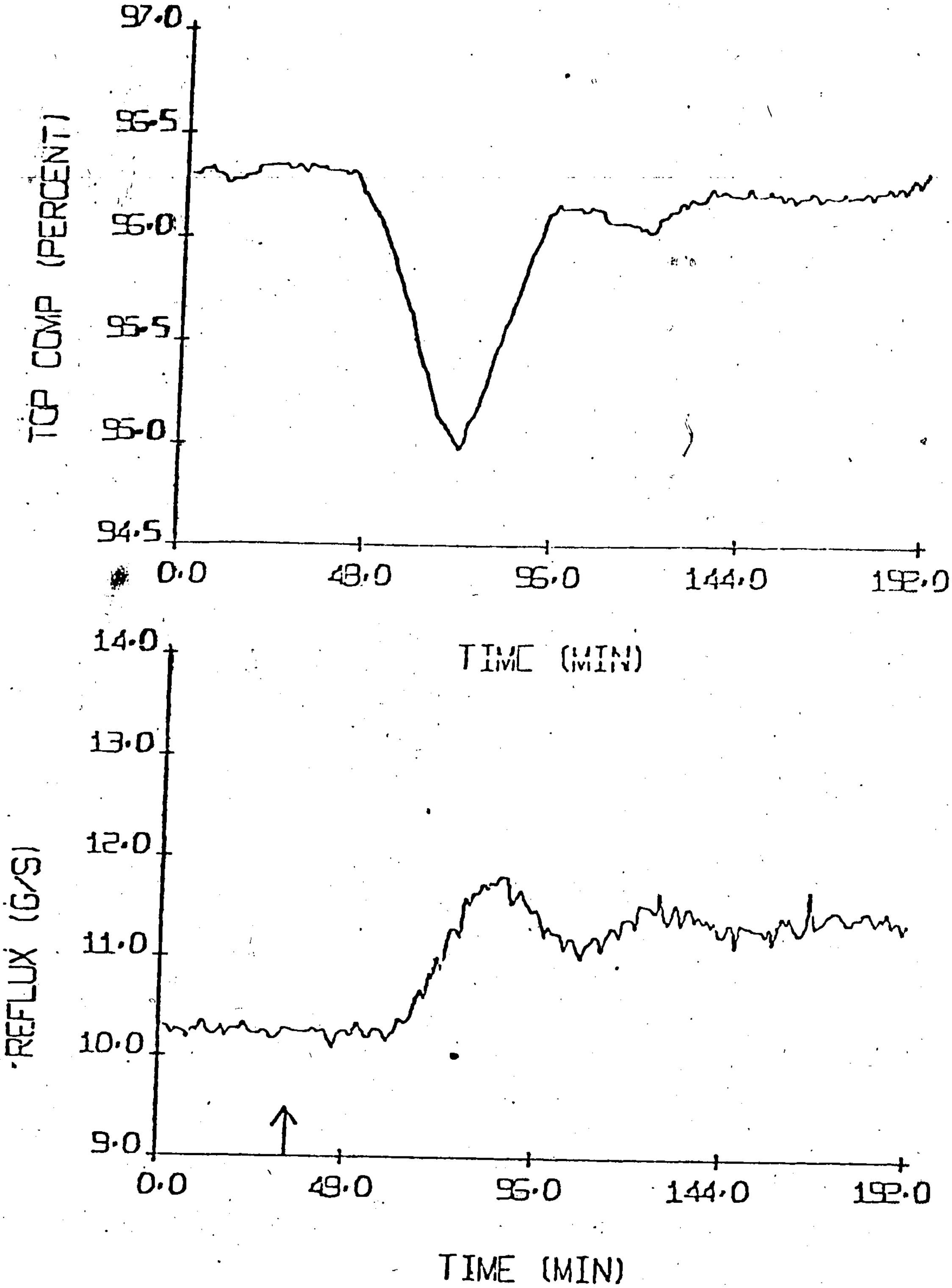


Figure 4.6 Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate, $\beta_o = 0.05$, $\underline{P}(0) = 250\,\underline{I}$.

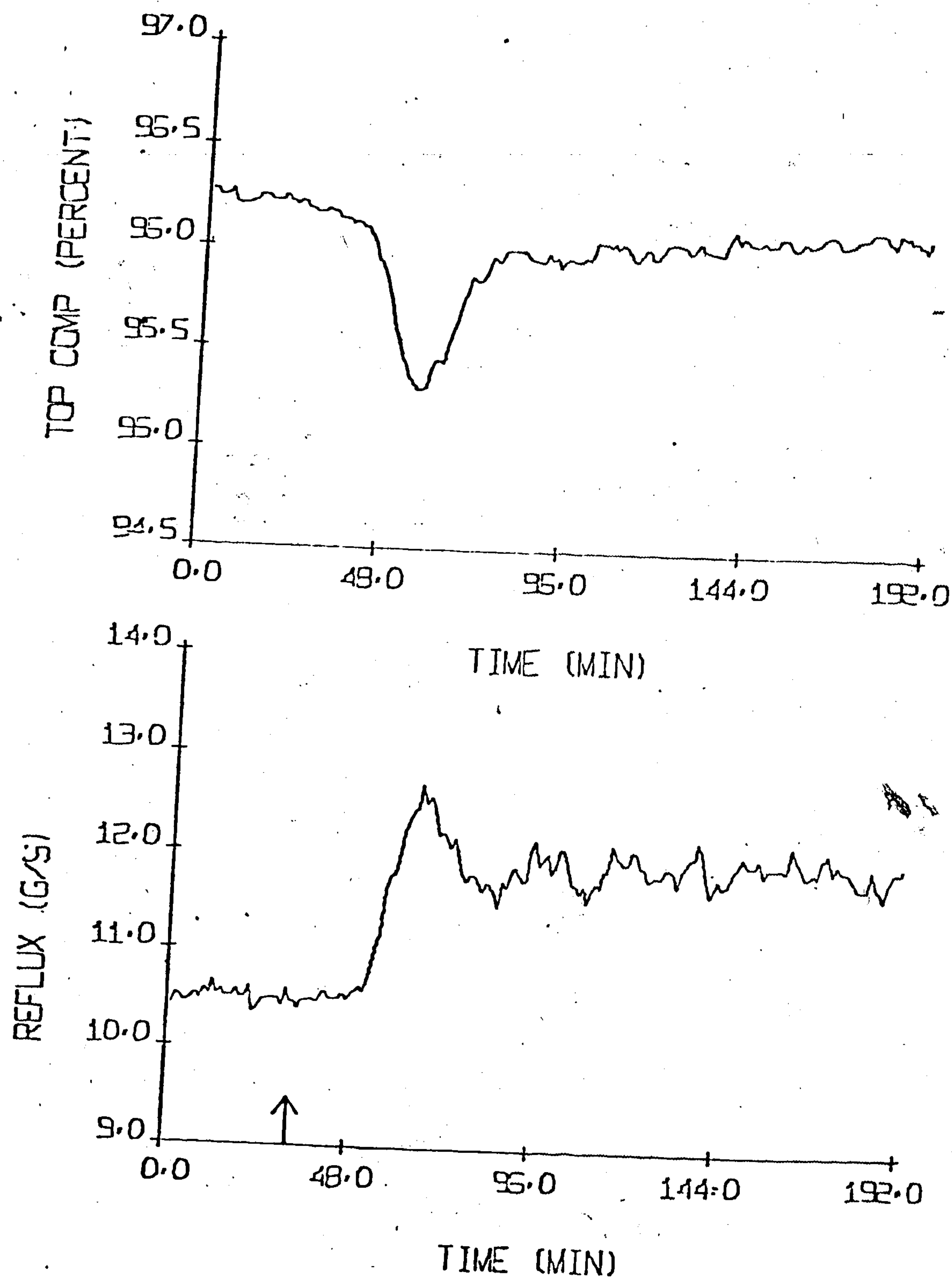


Figure 4.7 Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate, β_0 = 0.01, $\underline{P}(0)$ = 250 $\underline{\underline{I}}$.

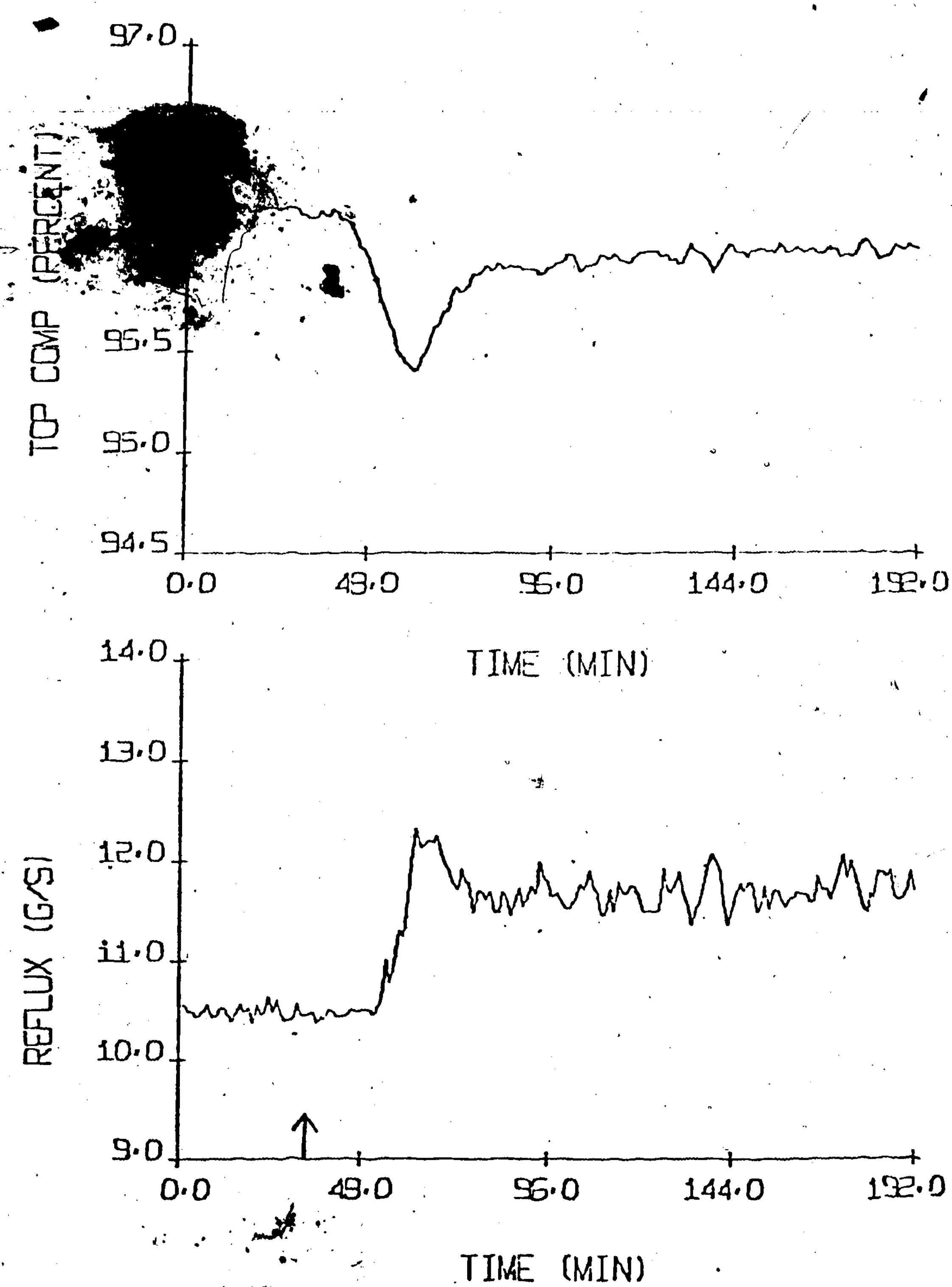


Figure 4.8 Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate, β_0 = 0.005, $\underline{P}(0)$ = 250 $\underline{I}$.

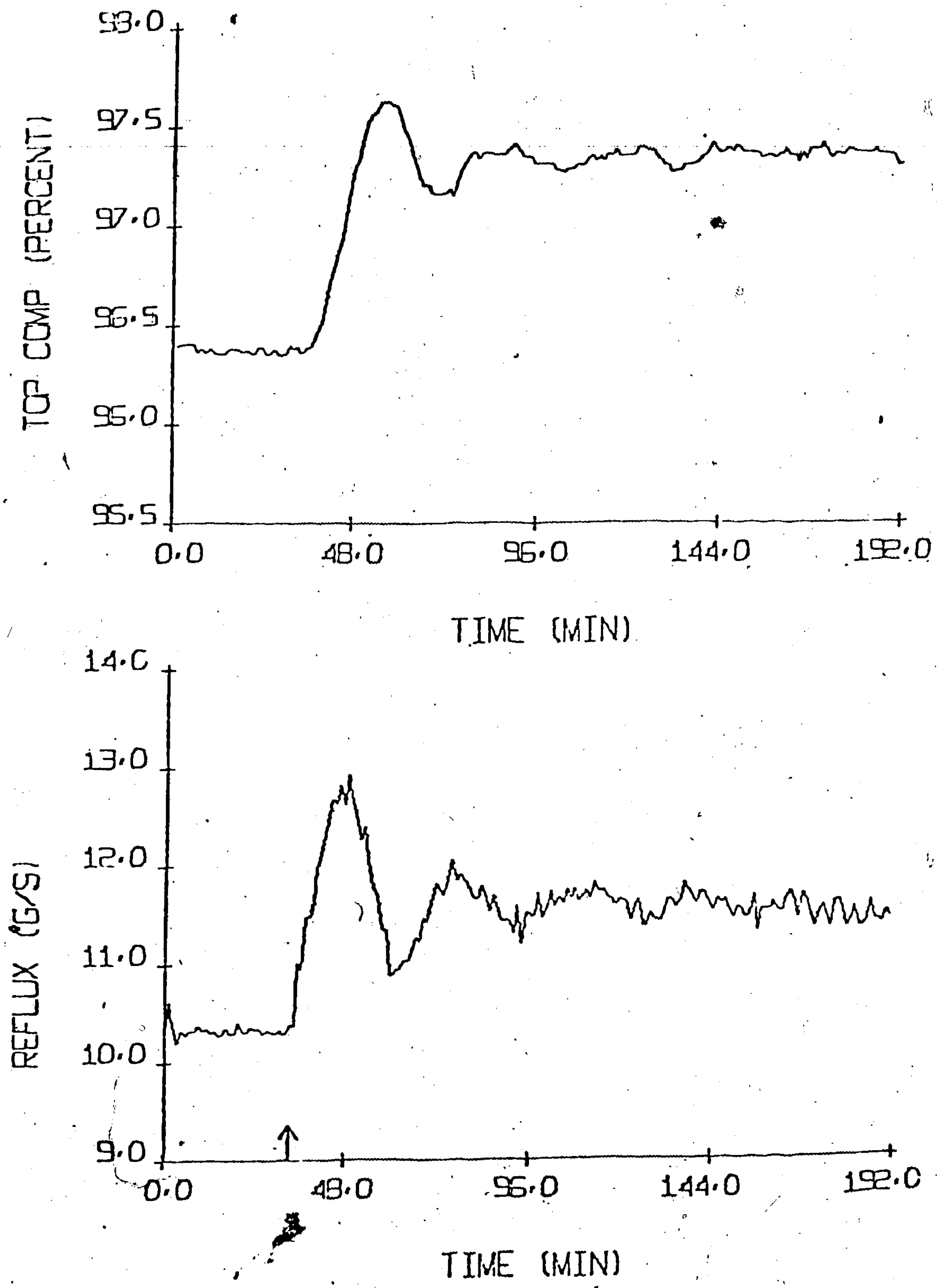


Figure 4.9 Experimental STR Control Of Top Product Composition For a +1% Step Change In Set Point, $\beta_0 = 0.05$, $\mathbf{P}(0) = 3000\,\mathbf{I}$.

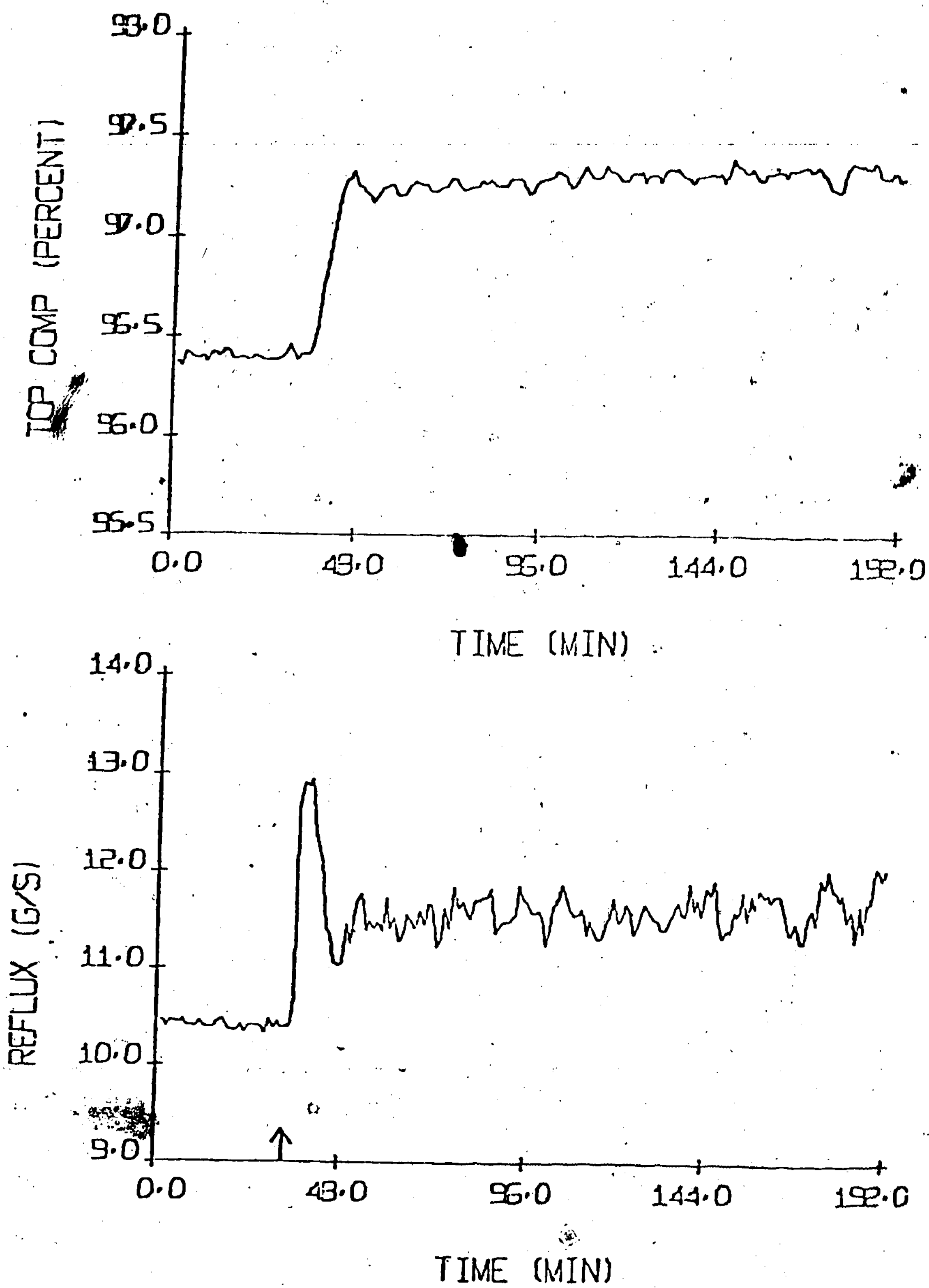


Figure 4.10 Experimental STR Control Of Top Product Composition For a +1% Step Change In Set Point, β_o = 0.01, $\underline{P}(0)$ = 3000 $\underline{\underline{I}}$.

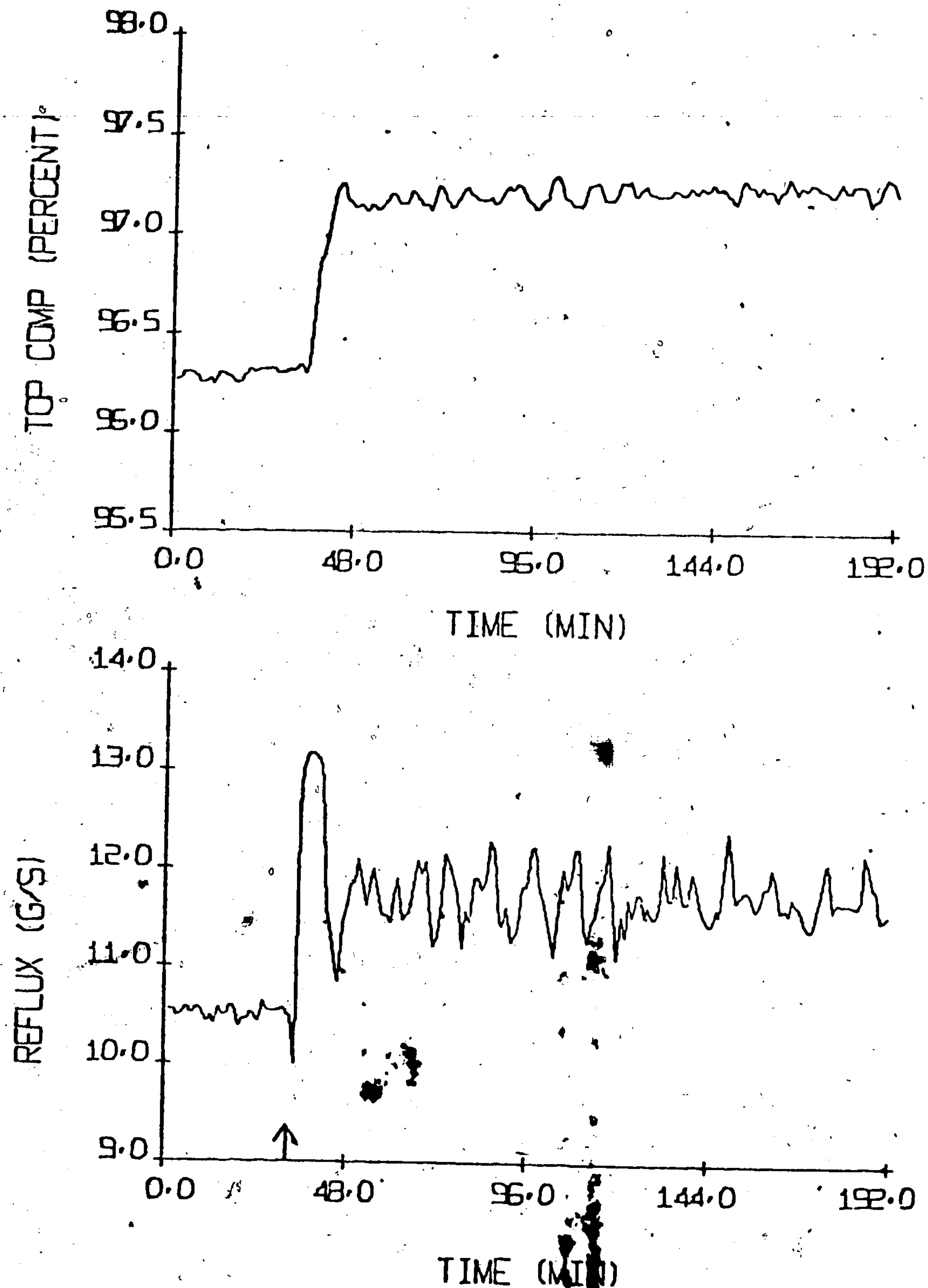


Figure 4.11 Experimental STR Control Of Top Product Composition For a +1% Step Change in Set Point, $\beta_0 = 0.005$, $\underline{P}(0) = 3000\ \underline{I}$.

overshoot of 0.3% observed in Figure 4.9 was totally eliminated by lowering β_o from 0.05 to 0.005 as seen in Figure 4.11. For both the servo and regulatory control tests performed, the best control behavior was obtained using the lowest β_o value. The control performance was not as sluggish for the smaller value of β_o and gave the smallest integral square error (ISE) value.

In examining the results listed in Table 4.2, it can be seen that an increase in the $1/\beta_o$ value reduces the ISE for all runs. This behavior is not surprising since $1/\beta_o$ acts as a gain term for the control input calculation (cf Equation 2.11) in the STR algorithm. An increase of the $1/\beta_o$ value has prompted the parameters to converge faster. The estimated parameters for the change in feed flow rate shown in Figure 4.12, took 46 time units after the disturbance was introduced to reach their steady state values with β_o = 0.05, and 40 time units with β_o = 0.005 in Figure 4.14. Figures 4.12 to 4.14 are included to demonstrate typical parameter convergence.

However, the reflux flow and top product composition displayed the most oscillatory behavior for the lowest value used, giving a "bang bang" like performance. This excessive control behavior was believed to be caused by the high $1/\beta_o$ gain effect that induced the high sensitivity of the regulator to the noise in the system.

In examining the output responses plotted in Figures 4.6 to 4.11, it can be seen that steady state error (offset)

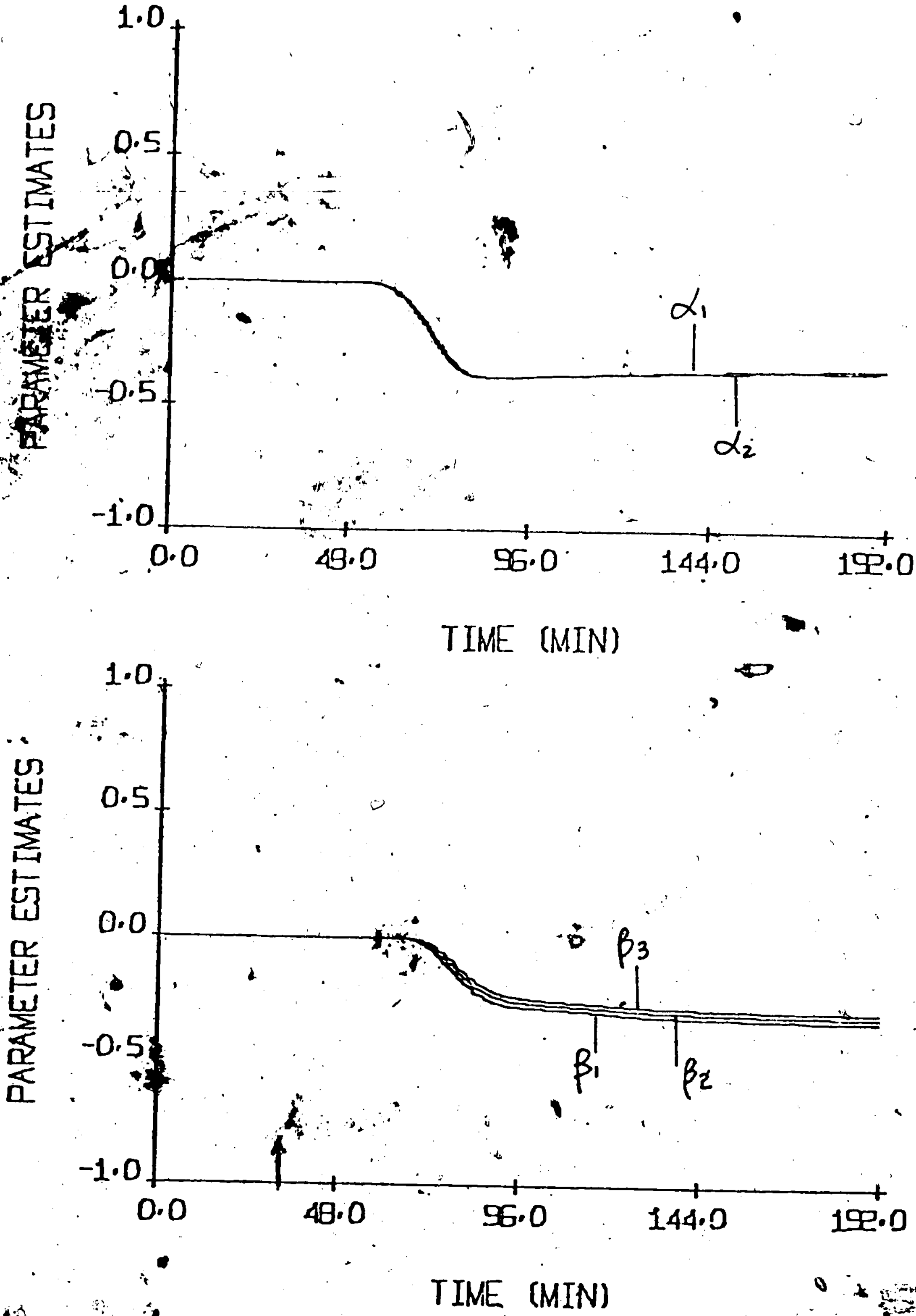


Figure 4.12 Estimated Parameters For Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate With $\beta_0 = 0.05$, $\underline{P}(0) = 250\ \underline{I}$.

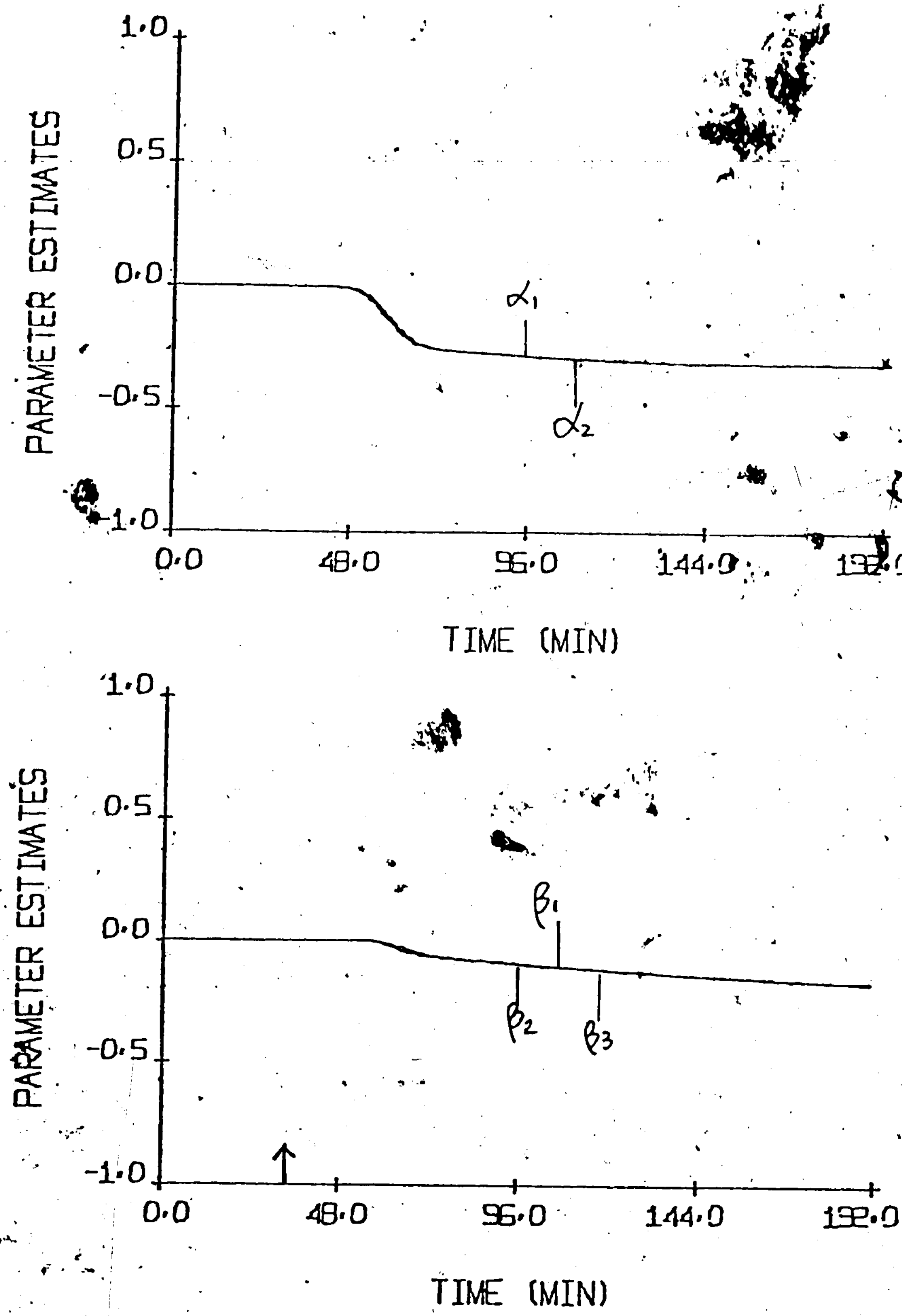


Figure 4.13 Estimated Parameters For Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate With $\beta_0 = 0.01$, $\underline{P}(0) = 250\,\underline{I}$.

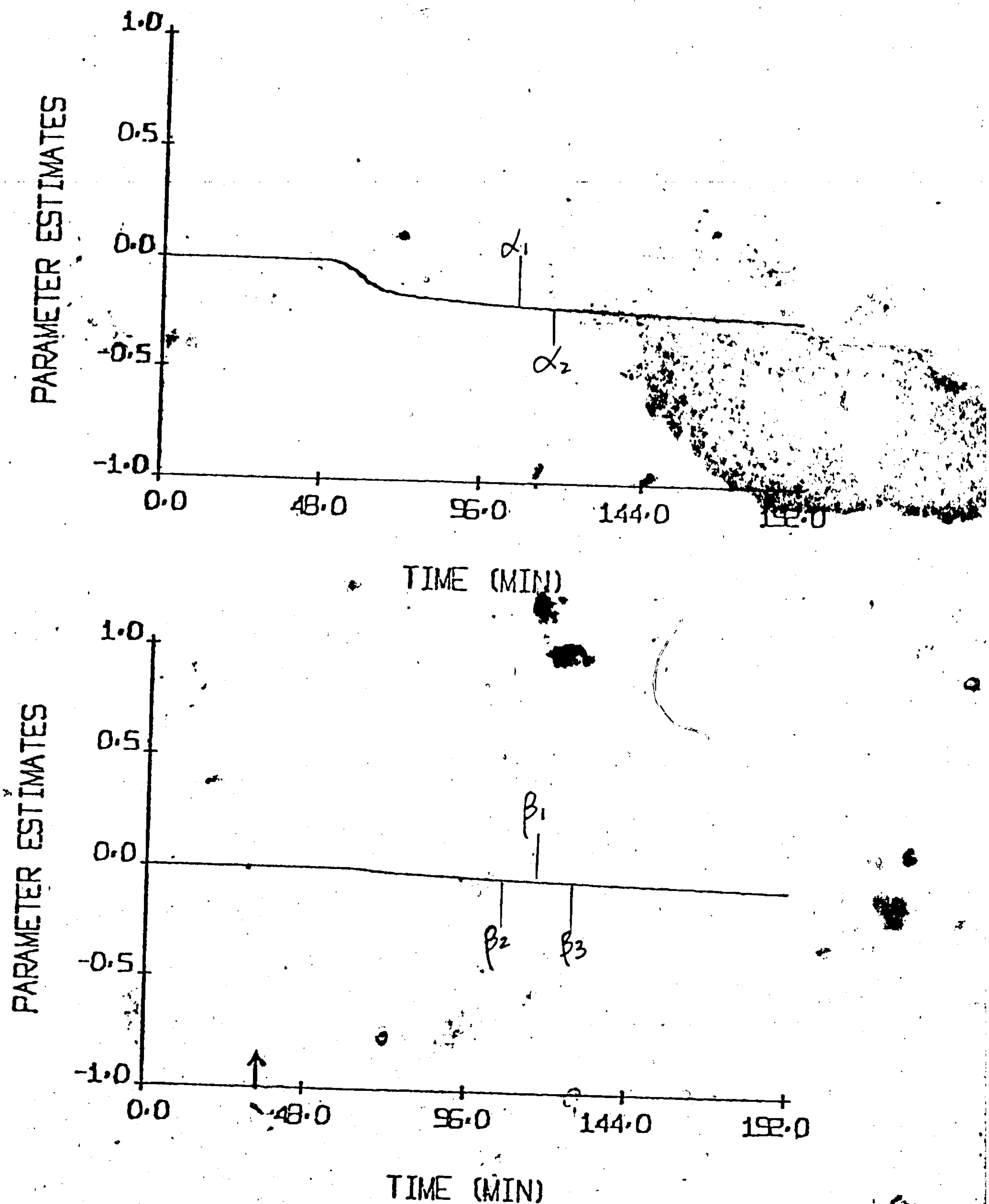


Figure 4.14 Estimated Parameters For Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate With β_a = 0.005, $\underline{P}(0) = 250\ \underline{I}$.

resulted, with the offset increasing with the lower values of β_0 value used. Consider the STR algorithm as given by Equation 2.12

$$u(t) = \frac{1}{\beta_0} [\hat{\alpha}_1 (t)y(t) + \ldots + \hat{\alpha}_n y(t-n+1)],$$
$$-\hat{\beta}_1 (t)u(t-1) - \ldots - \hat{\beta}_L (t)u(t-L)$$

in which u(t) and y(t) are defined as the normalized deviation variables. At steady state, the values of y(t) and u(t) will be the same as their past values. Hence, by assuming the system has reached steady state at the end of the run, this equation can be rearranged into the form

$$y = u * \beta_0 (1.0 + \Sigma \beta_i) / \Sigma \alpha_i$$

Now as $\Sigma \beta_i$ value approaches -1.0, the final offset, y, will approach zero. Table 4.3 is a summary of the final estimated α_i and β_i parameters for the control tests shown in Figures 4.6 to 4.11.

Table 4.3 Effect Of β_o Value On Estimated Parameter Values

Run	β_o	Figure	α_i	$\Sigma\alpha_i$	β_i	$\Sigma\beta_i$
Feed Flow -20%						
E11	0.05	4.12	-0.3571 -0.3490	-0.7061	-0.3383 -0.3193 -0.3013	-0.9589
E12	0.01	4.13	-0.3287 -0.3243	-0.6530	-0.1870 -0.1832 -0.1796	-0.5498
E13	0.005	4.14	-0.2487 -0.2452	-0.4939	-0.0765 -0.0750 -0.0736	-0.2251
Set Point +1%						
E14	0.05	-	-0.[illegible]256 -0.4824	-1.1080	-0.5140 -0.3044 -0.14[illegible]9	-0.9683
E15	0.01	-	-0.4999 -0.2924	-0.7923	-0.3155 -0.2757 -0.2744	-0.8656
E16	0.005	-	-0.4699 -0.2600	-0.7299	-0.2058 -0.2095 -0.2269	-0.6422

These results show that as β_o is decreased, the difference between the $\Sigma\beta_i$ value and -1.0 increases. The values changed from -0.9589 to -0.2251 in Runs E11 to E13 for a 20% step decrease in feed flow rate and from -0.9683 to -0.6422 in Runs E14 to E16 for a 1% increase in set point. In the case of $\Sigma\alpha_i$, the absolute value decreased as β_o was lowered, thus also leading to an increase in the offset. The combined effect of $\Sigma\alpha_i$ and $\Sigma\beta_i$ outweighed the effect of β_o as shown by the larger offset that resulted.

For example, in the case of Run E12, the values were $\Sigma\beta_i$ = -0.5498; $\Sigma\alpha_i$ = 0.6530; β_o = 0.01; giving

y = (12 - 10.4)/(10.4)*(0.01)* (1 - 0.5498)/-0.6530 = 0.00106

so for the set point of 96.2%, the final value of y will be 96.1%. This calculated value is in agreement with the final value in Figure 4.7. This trend is consistent for all of the tests.

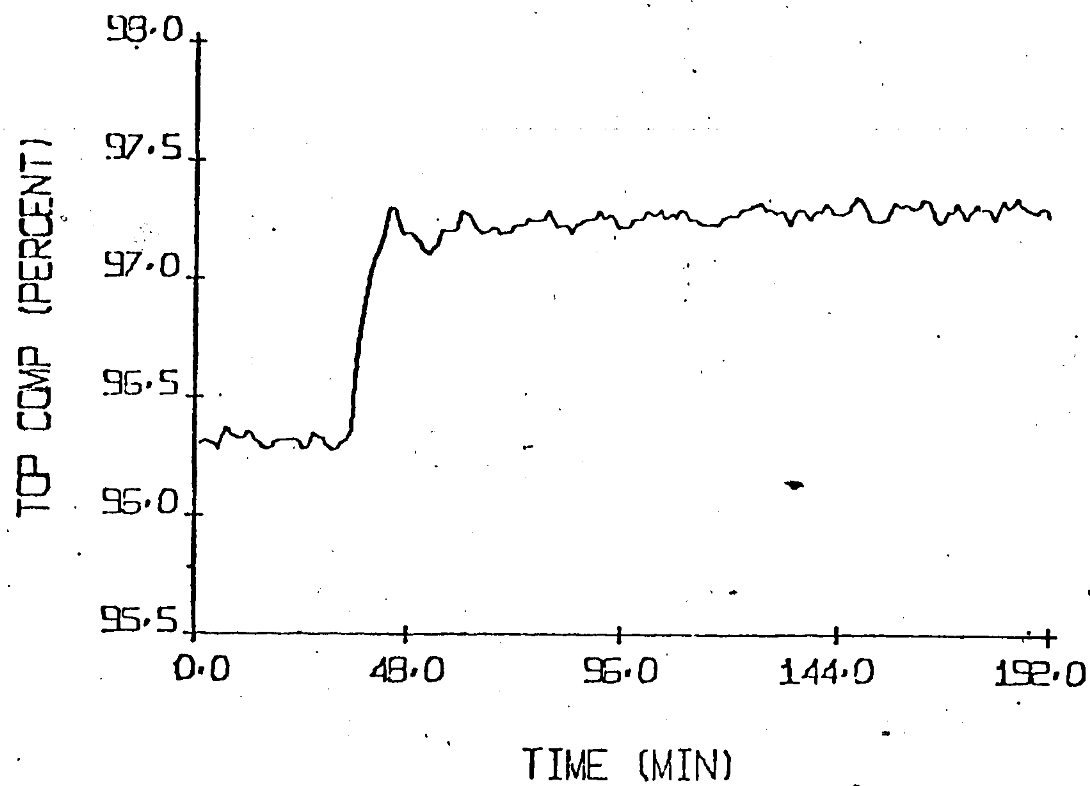
For performance evaluation, since an ISE index was used, it should be noted that the duration of the run influences the accumulated error due to steady state offset. For the tests performed, one could achieve faster response and reduced ISE by lowering β_o at the expense of increasing offset.

ii) Effect of $\underline{\underline{P}}(0)$ value:

The $\underline{\underline{P}}(0)$ value set at the beginning of the run was determined by the amount of confidence which could be placed in the initial parameter estimates. If the parameter estimates were considered to be close to the true values, the $\underline{\underline{P}}(0)$ value was set to a low value. Tests performed using different values of $\underline{\underline{P}}(0)$ are summarized in Table 4.4. The output responses are plotted in Figures 4.15 to 4.20.

Table 4.4 Effect Of Initial Value Of Covariance Matrix, $\underline{P}(0)$ On The Experimental STR Control Of Top Product Composition

Run	β_0	$\underline{P}(0)$	ISE	Figure
Set Point +1%				
E15	0.01	3,000 $\underline{I}$	7.3×10^{-4}	4.10
E17	0.01	10,000 $\underline{I}$	6.0×10^{-4}	4.15
E18	0.01	30,000 $\underline{I}$	5.1×10^{-4}	4.16
Set Point -1.5%				
E19	0.01	10,000 $\underline{I}$	11.9×10^{-4}	4.17
E20	0.01	30,000 $\underline{I}$	14.4×10^{-4}	4.18
Feed Flow -20%				
E13	0.005	250 $\underline{I}$	14.0×10^{-4}	4.8
E21	0.005	500 $\underline{I}$	8.7×10^{-4}	4.19
E22	0.005	1,000 $\underline{I}$	4.7×10^{-4}	4.20



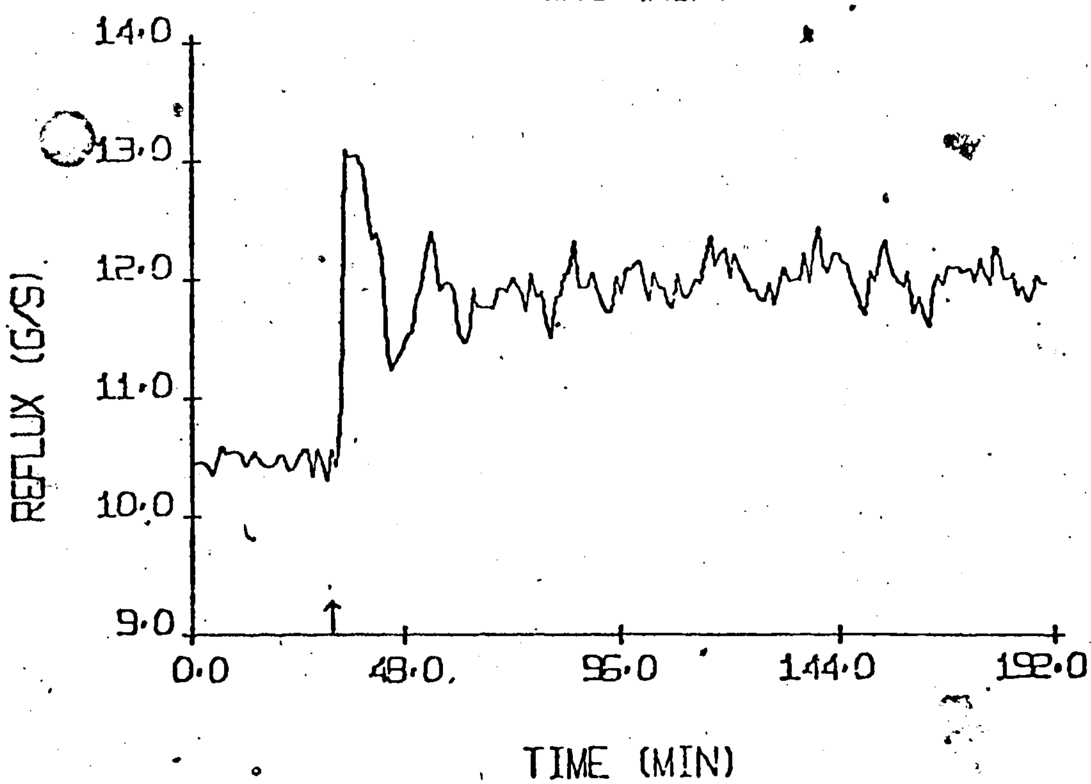


Figure 4.15 Experimental STR Control Of Top Product Composition For a +1% Step Change In Set Point, $\beta_o = 0.01$, $\underline{P}(0) = 10000\ \underline{\underline{I}}$.

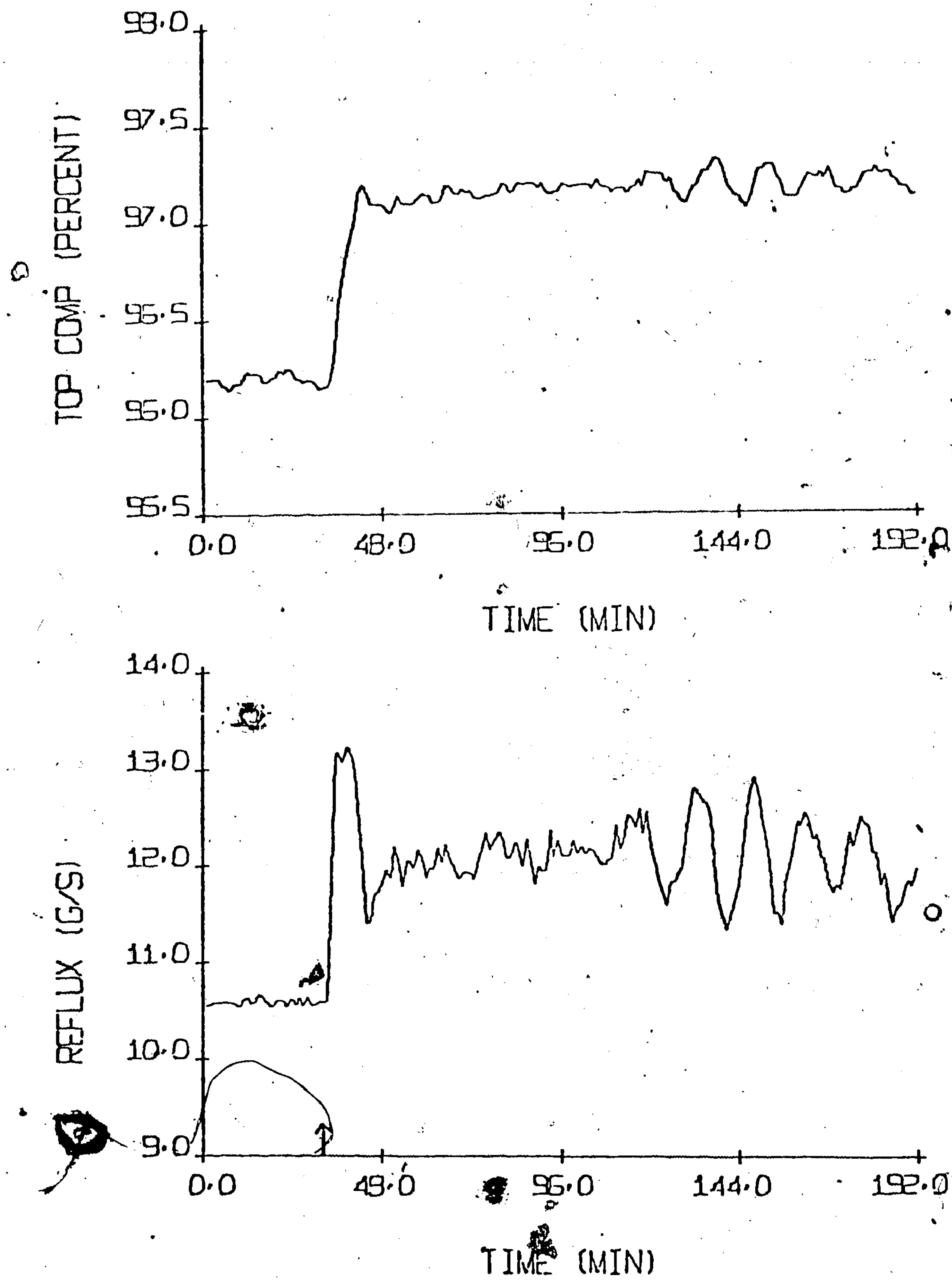
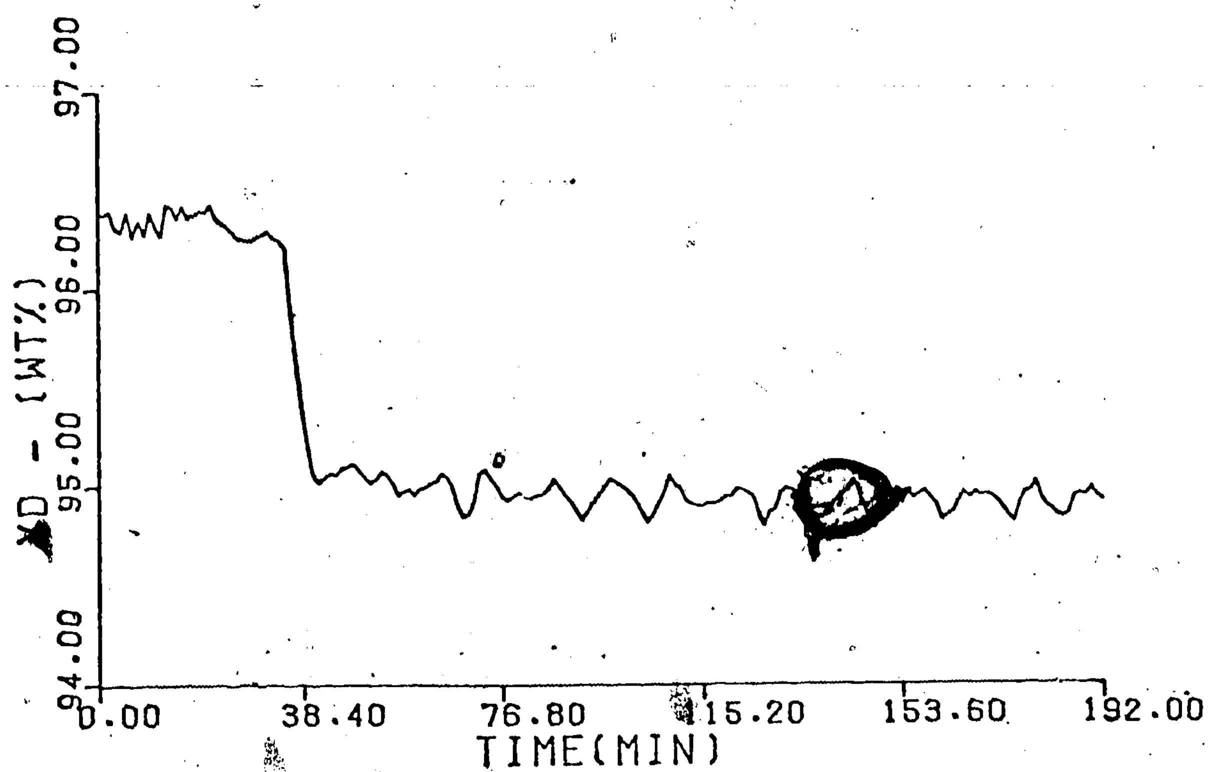


Figure 4.16. Experimental STR Control Of Top Product Composition For a +1% Step Change In Set Point, $\beta_0 = 0.01$, $\underline{P}(0) = 30000\ \underline{I}$.



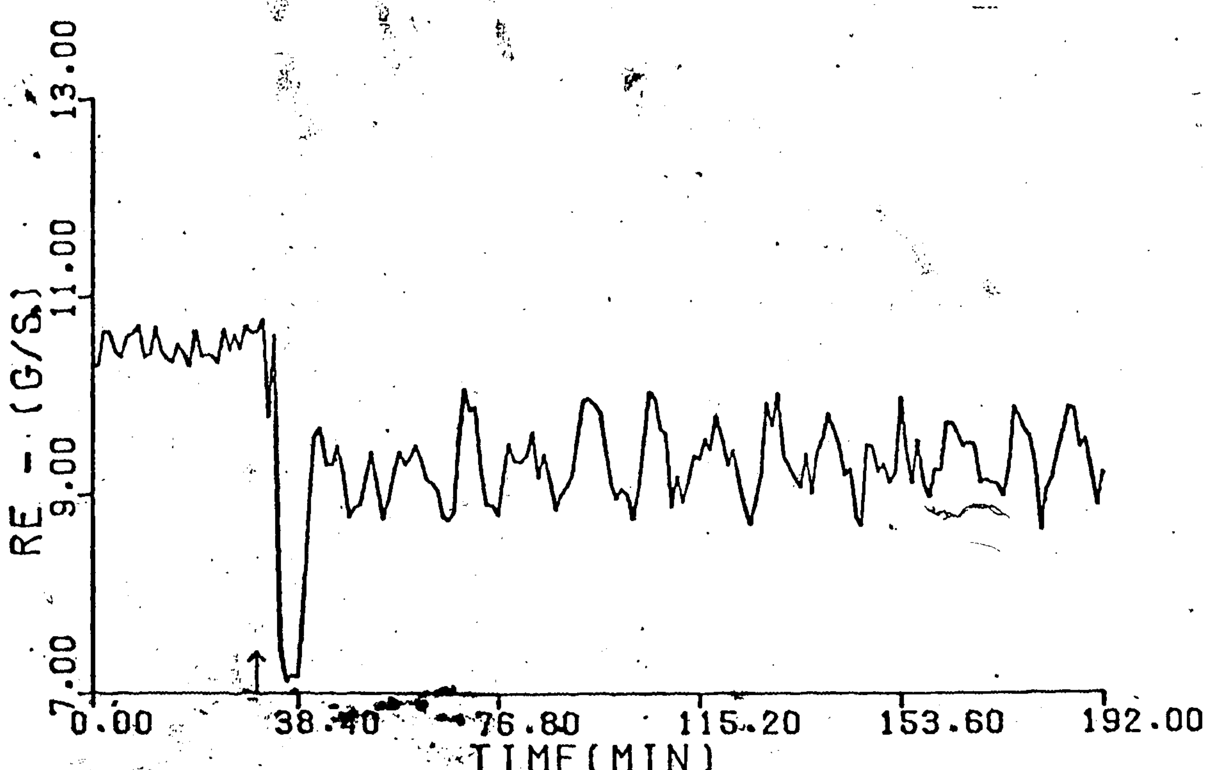


Figure 4.17 Experimental STR Control Of Top Product Composition For a -1.5% Step Change In Set Point, β_0 = 0.01, $\underline{P}(0)$ = 10000 $\underline{I}$.

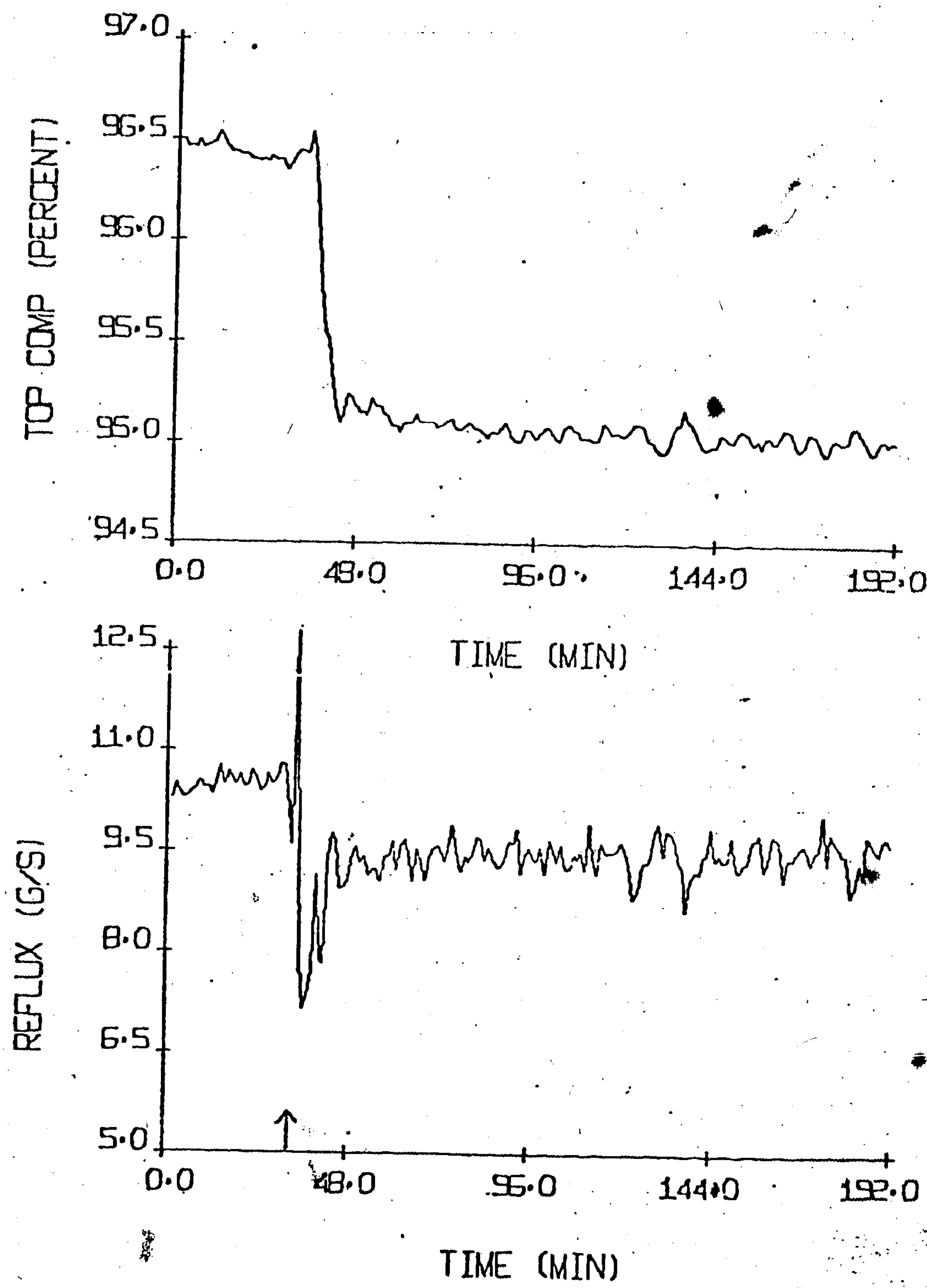


Figure 4.18 Experimental STR Control Of Top Product Composition For a -1.5% Step Change In Set Point, β_o = 0.01, $\underline{P}(0)$ = 30000 $\underline{I}$.

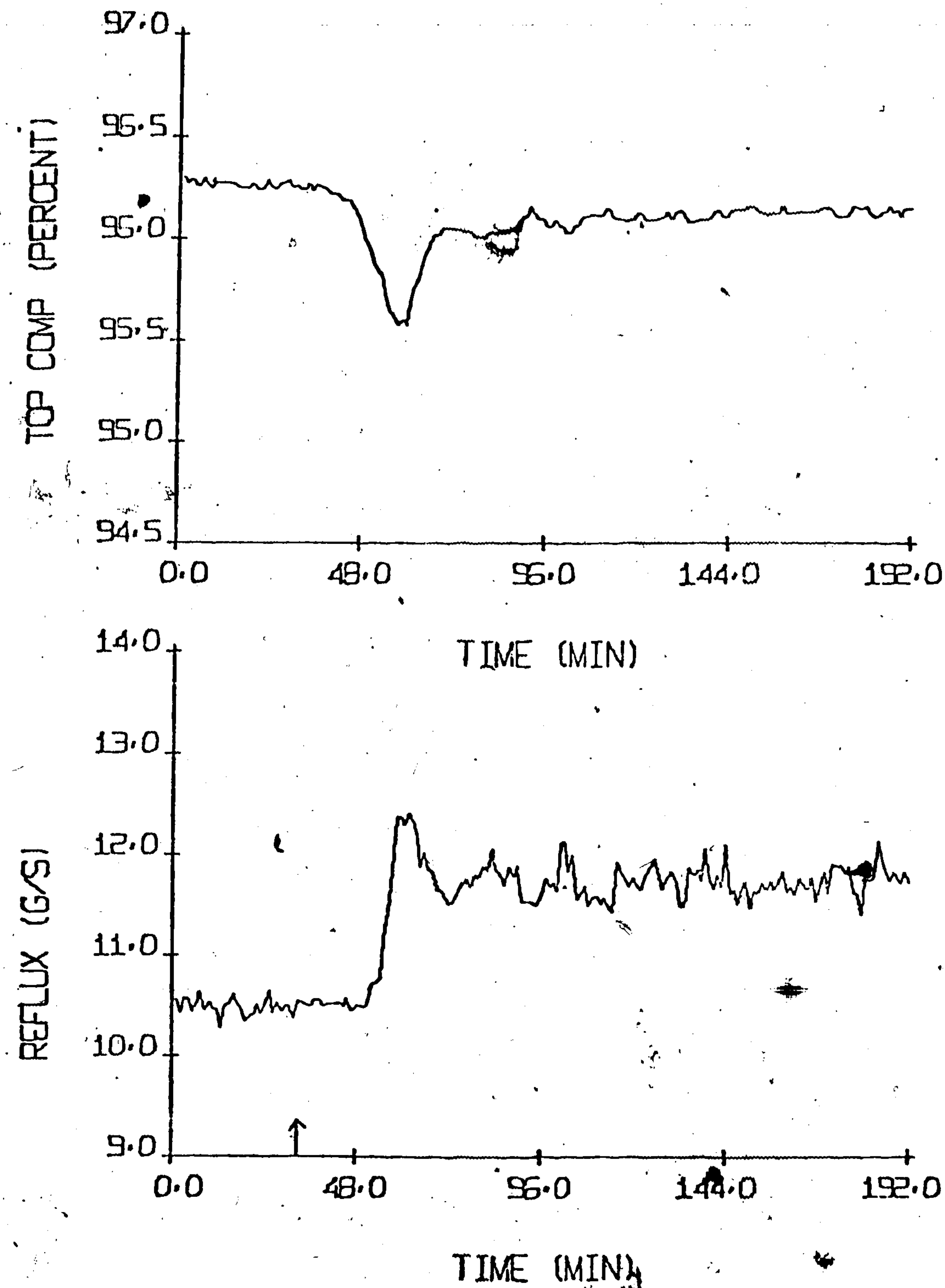


Figure 4.19 Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate, β_0 = 0.005, $\underline{P}(0)$ = 500 $\underline{I}$.

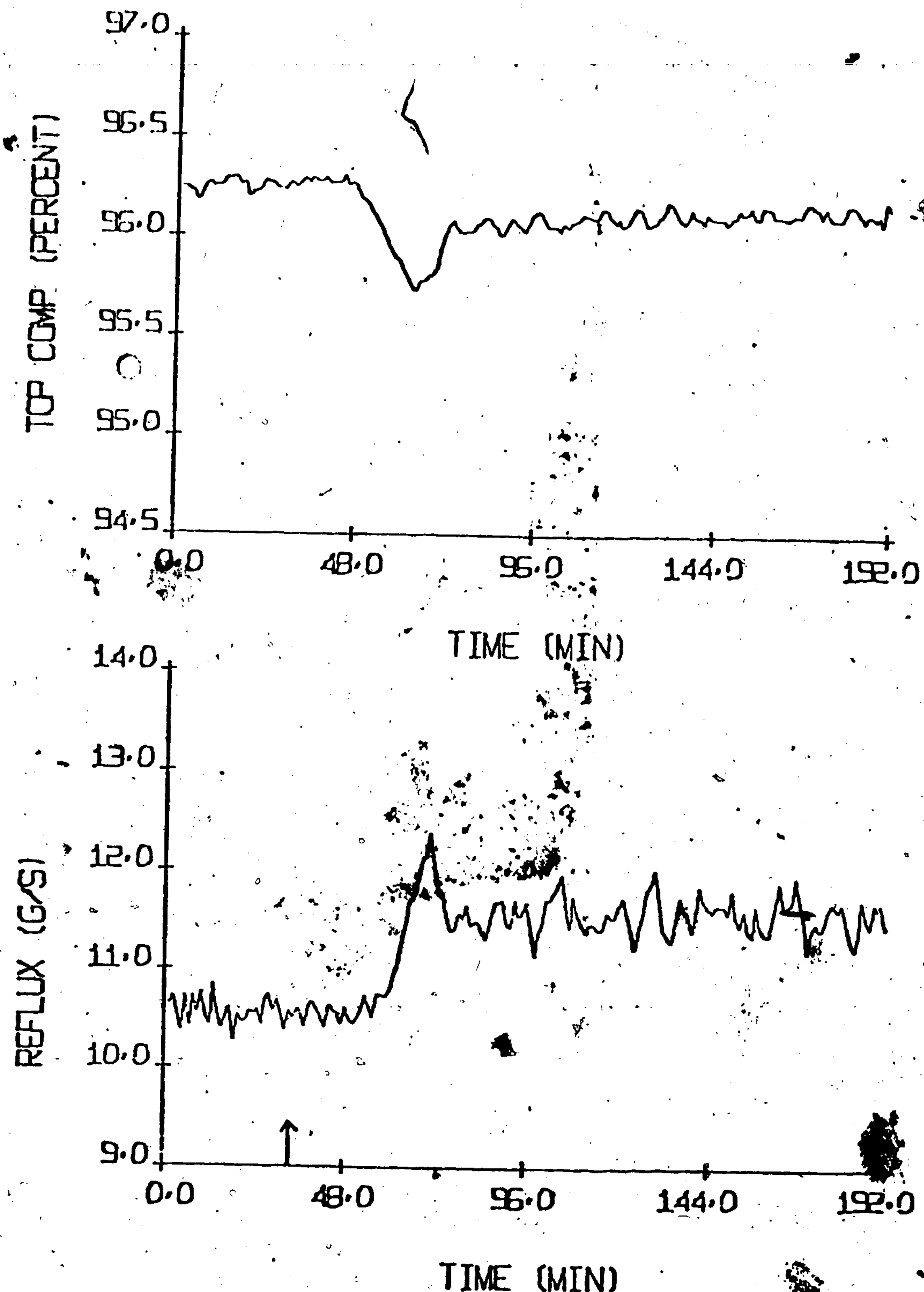


Figure 4.20 Experimental STR Control Of Top Product Composition For a -20% Step Change In Feed Flow Rate, $\beta_0 = 0.005$, $\underline{P}(0) = 1000\ \underline{I}$.

From the results obtained, the sustained oscillations that have been evident in all STR runs became more pronounced when a higher $\underline{P}(0)$ value was used. This is obvious from the responses plotted in Figures 4.15 - 4.20. When $\underline{P}(0)$ = 30,000 $\underline{I}$ was used, as can be seen in Figure 4.18, the manipulated variable (reflux) banged between its high and low limits at the time the set point change was introduced. This behavior resulted because of large fluctuations in the estimated parameter values as shown in Figure 4.21. For a 1% step increase in set point, with $\underline{P}(0)$ = 30,000 $\underline{I}$, the control behavior became oscillatory as shown by the response in Figure 4.16. The poor control is to be expected due to the changes in the parameter values (cf Figure 4.22). The parameters drift away from each other, unlike the adaption pattern of the parameters of the other tests, and never reach steady values.

Changing the value of $\underline{P}(0)$ showed that the effect on controller performance, to a large extent, was almost identical to the gain term in a proportional controller. In addition, the steady state offset was found to decrease as the $\underline{P}(0)$ value was increased. From the α_i and β_i values given in Table 4.5, it can be seen that as the value of $\underline{P}(0)$ increased, $\Sigma\beta_i$ moved closer to -1.0 and the absolute value of $\Sigma\alpha_i$ increased, so for the same β_0 value, the offset would certainly be smaller. The ISE values were lowered by increasing $\underline{P}(0)$ for all tests except the 1.5% step decrease in set point run which was caused mainly by the initial

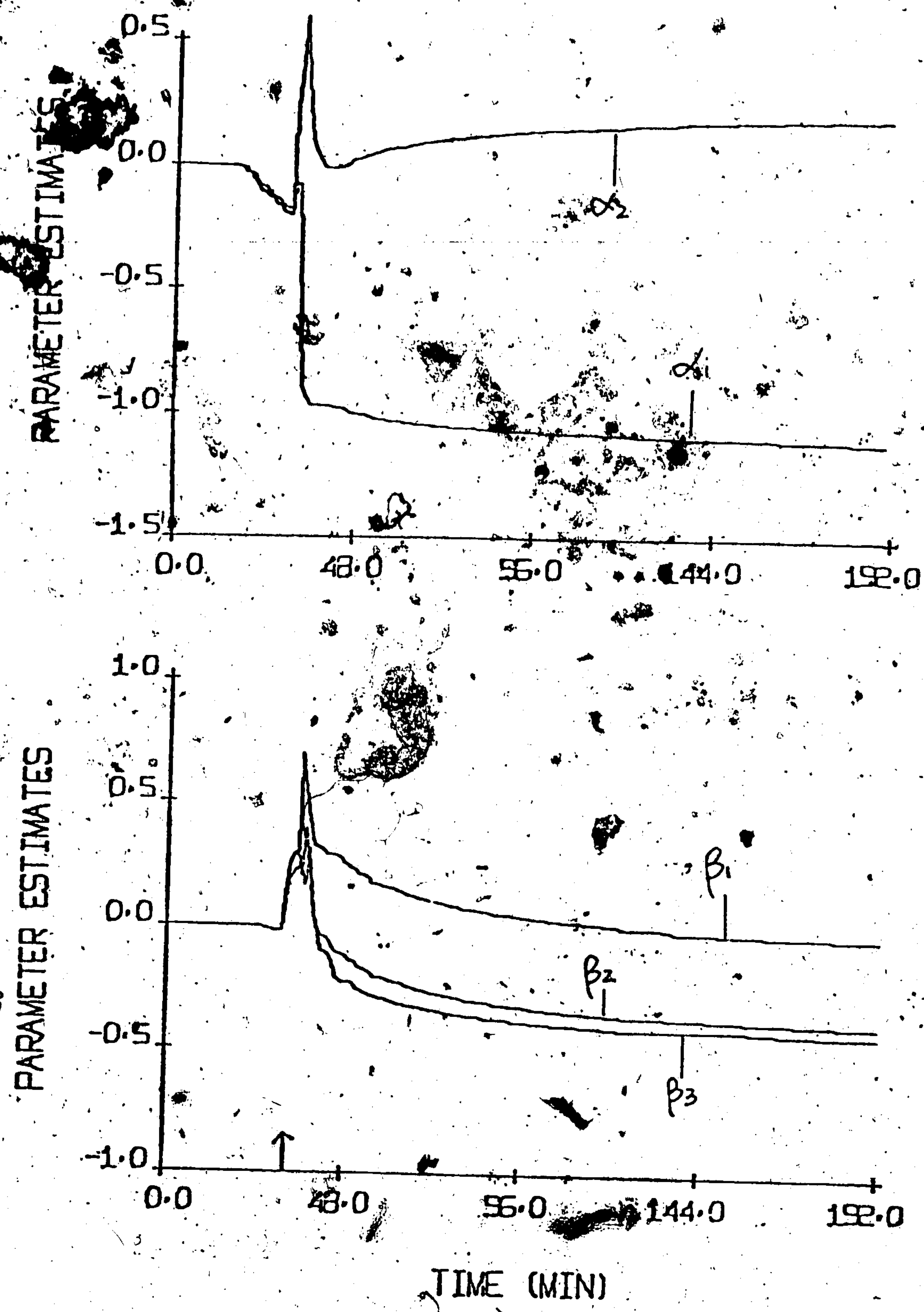


Figure 4.21 Estimated Parameters For Experimental STR Control Of Top Product Composition For a -1.5% Step Change In Set Point, β_0 = 0.01, $\underline{P}(0)$ = 3000 $\underline{\underline{I}}$.

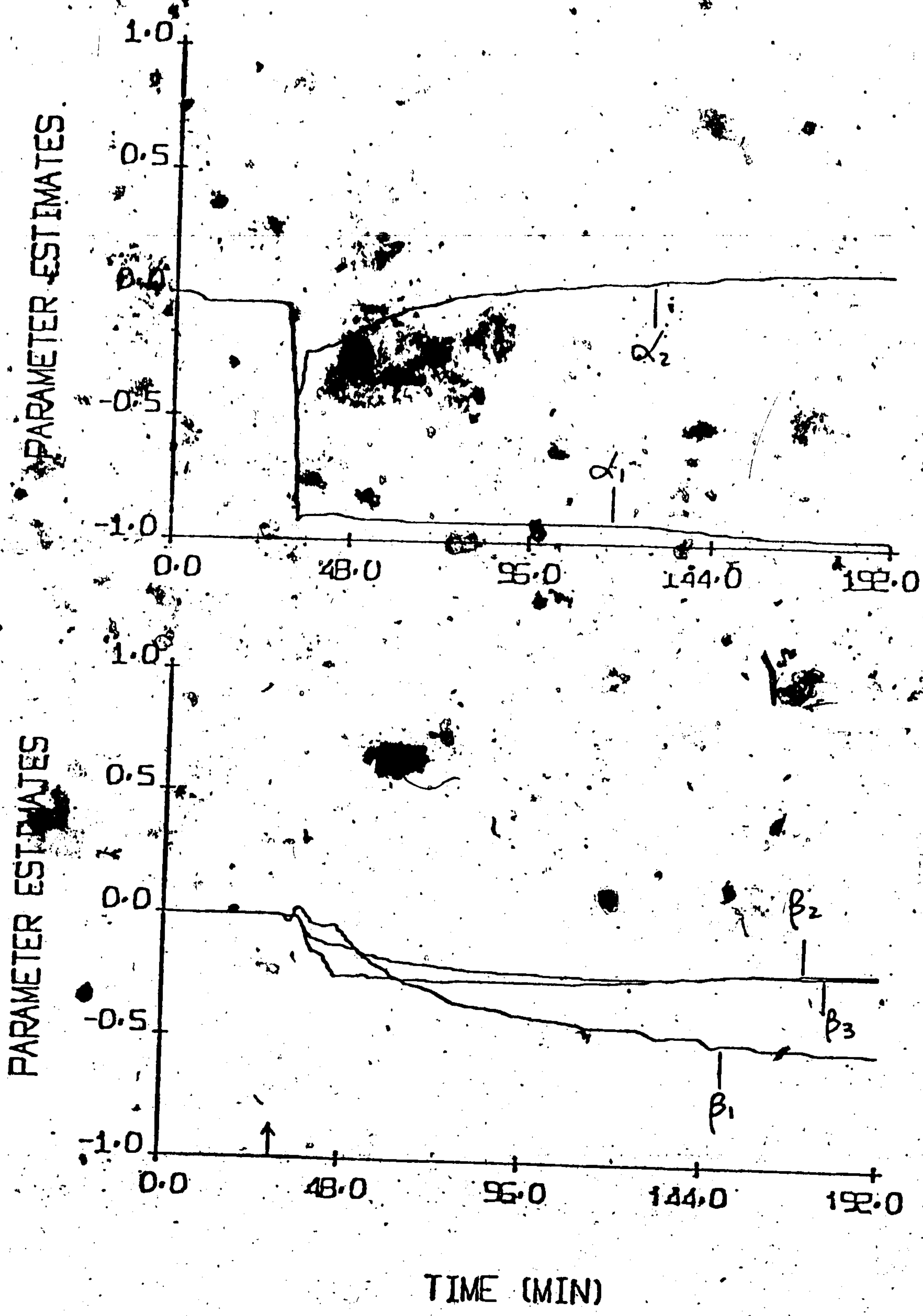


Figure 4.22 Estimated Parameters For Experimental STR Control Of Top Product Composition For a +1% Step Change In Set Point, $\beta_o = 0.01$, $\underline{P}(0) = 30000\ \underline{I}$.

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Table 4.5 Effect Of $\underline{P}(0)$ Value On The Estimated Parameters

Run	$\underline{P}(0)$	Figure	Final Estimated Parameter Value α_i	$\Sigma\alpha_i$	β_i	$\Sigma\beta_i$
Set Point +1%						
E15	3,000 $\underline{I}$	4.10	-0.4999	-0.7923	-0.3155	-0.8656
			-0.2924		-0.2757	
					-0.2744	
E17	10,000 $\underline{I}$	-	-0.6945	-0.8613	-0.3454	-0.9400
			-0.1668		-0.2969	
					-0.2977	
E18	30,000 $\underline{I}$	4.22	-0.9795	-0.8634	-0.5376	-0.9464
			0.1161		-0.1979	
					-0.2100	
Set Point -1.5%						
E19	10,000 $\underline{I}$	-	-0.8636	-0.8324	-0.2964	-0.8376
			0.0312		-0.2439	
					-0.2973	
E20	30,000 $\underline{I}$	4.21	-1.0953	-0.8748	-0.0348	-0.8497
			0.2205		-0.3866	
					-0.4283	
Feed Flow -20%						
E13	250 $\underline{I}$	4.8	-0.2487	-0.4939	-0.0765	-0.2251
			-0.2452		-0.0750	
					-0.0736	
E21	500 $\underline{I}$	-	-0.2899	-0.5733	-0.1132	-0.3298
			-0.2834		-0.1083	
					-0.1083	
E22	1,000 $\underline{I}$	-	-0.2980	-0.5868	-0.1213	-0.3568
			-0.2888		-0.1185	
					-0.1170	

4.3 Simulation Study

4.3.1 System Configuration and Implementation

The simulations were performed using the University of Alberta computing services Amdahl 470V/6 computer system. The nonlinear distillation process model developed by Simonsmeier (20) was incorporated into the simulation program to generate the distillation column dynamic behavior. The PI and STR control routines are identical to those used in the experimental study with implementation of the control action on the model performed in a manner analogous to that of the experimental study. The source listing of the complete set of control programs used in the simulations is given in Appendix E.

Prior to any simulation run, an input data file was set up to provide the initial steady state operating conditions for the column and the control data parameters required. A run time of 169 time units, with each time unit being 64 seconds, was used for all the simulations. The disturbance was always introduced after 20 time units of simulation time had elapsed. The program integration interval was fixed at 8 seconds. The integration interval was selected as a compromise between reducing the computer execution time and accurately calculating the transient response of the column.

4.3.2 Proportional Plus Integral Control Tests

The Ziegler-Nichols tuning constants of K_P = -45., T_R = 5.84 min. determined experimentally, were used for the simulation runs. The results are summarized in Table 4.6.

Table 4.6 Summary Of Simulation Results For Top Product Composition PI Control

Run	Disturbance	K_P	T_R	ISE	Figure
S3	Feed Flow -20%	-45.	5.84	1.0×10^{-4}	4.23
S4	Set Point +1.0%	-45.	5.84	21.0×10^{-4}	4.24
S5	Set Point -1.5%	-45.	5.84	59.0×10^{-4}	4.24

Operating Conditions

Feed Flow	18.00 g/s	Top Composition	95.7%
Reflux Flow	11.00 g/s	Bottom Composition	5.6%
Steam Flow	11.95 g/s		

For a 20% step decrease in feed flow rate, the PI controller gave excellent control behavior as can be seen from the response in Figure 4.23. The maximum deviation was only 0.07 wt % from the set point value so further improvement in the control performance by further tuning will likely not result. For step changes in set point, satisfactory control responses were also obtained. As can be seen from the control behavior in Figures 4.24 and 4.25, that resulted by using the same controller settings as for the disturbance in feed flow rate, a larger overshoot was observed for the decrease in set point as compared to the increase. This is due to the nonlinear dynamic characteristics of the column.

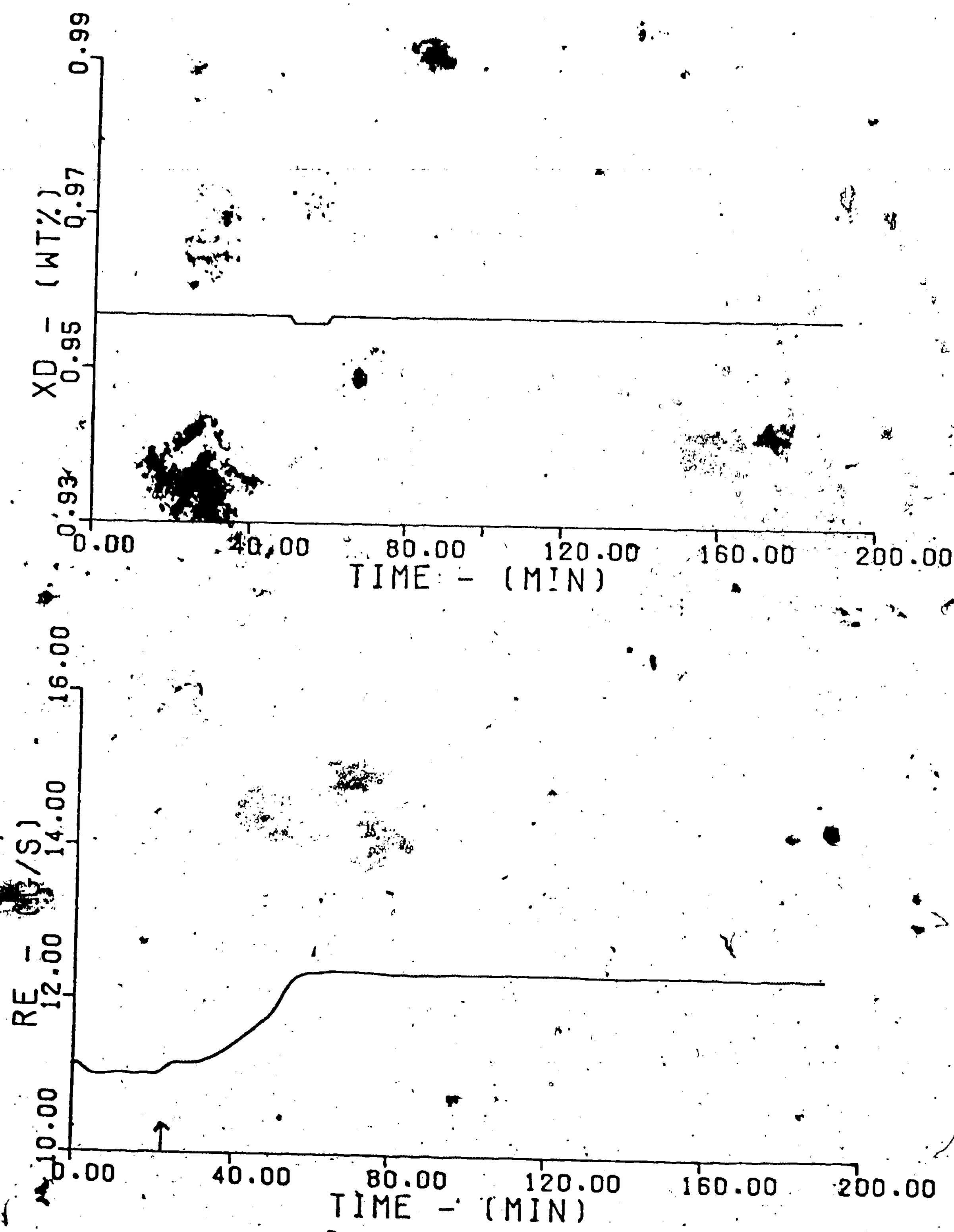


Figure 4.23 Simulation Of PI Top Product Composition Control For a 20% Step Change In Feed Flow Rate

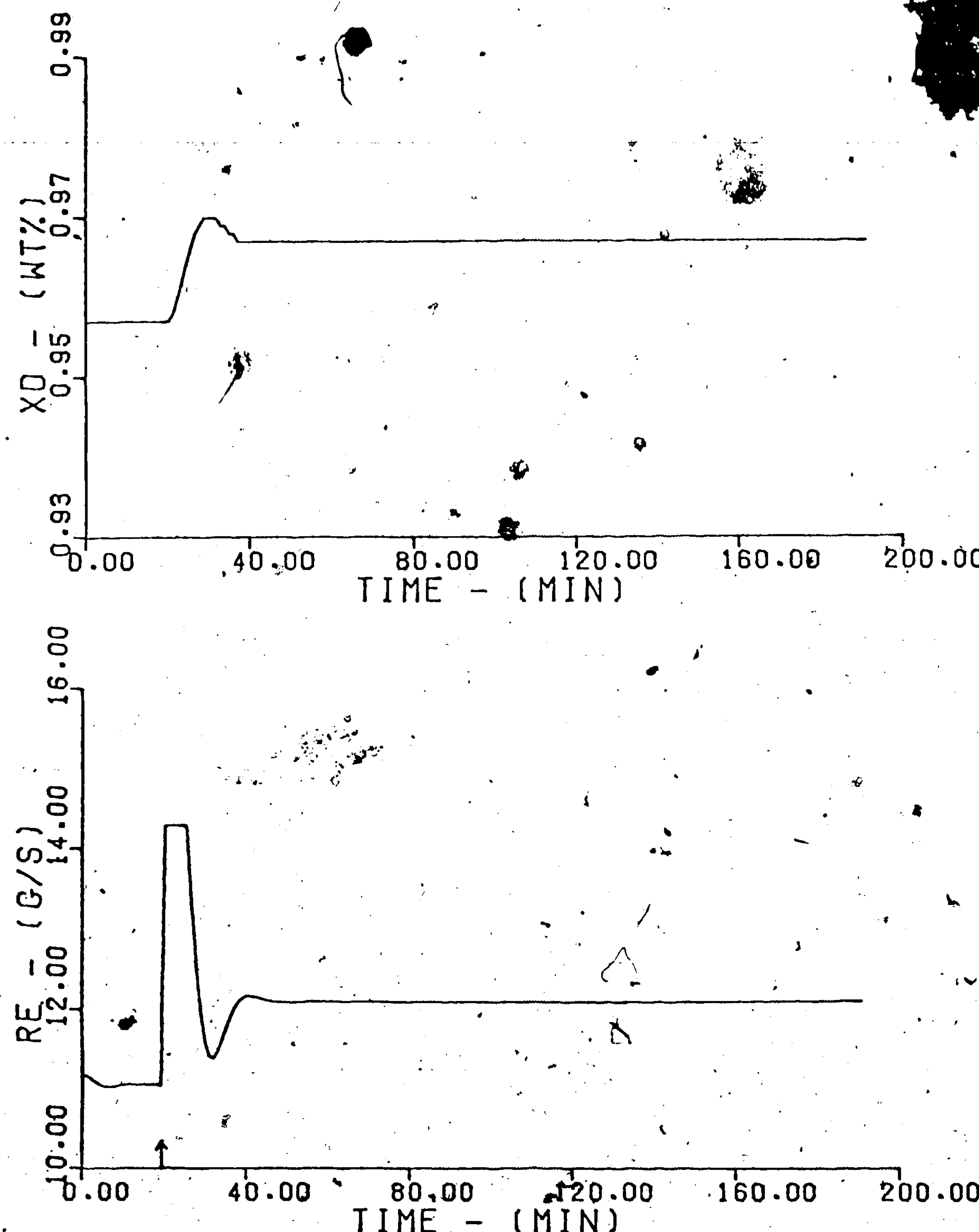


Figure 4.24 Simulation Of PI Top Product Composition Control For a +1% Step Change In Set Point

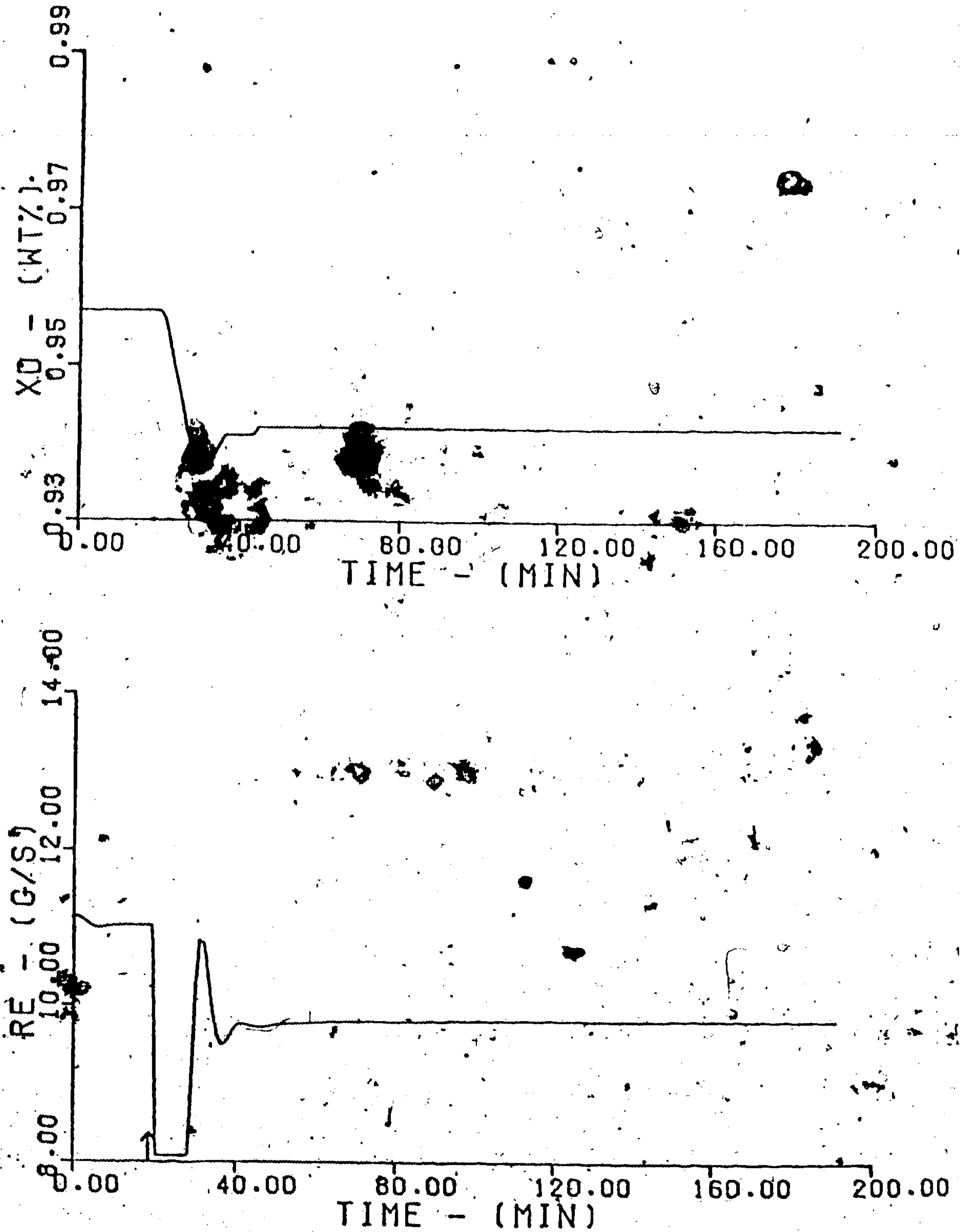


Figure 4.25 Simulation Of PI Top Product Composition Control For a -1.5% Step Change In Set Point

4.3.3 Self-Tuning Regulator Tests

Preliminary simulations were carried out to determine a suitable set of numerical values for the STR parameters. The operating conditions were set identical to those used in an earlier study by Sastry et al (15). The following parameters values as used by Sastry, were found to be satisfactory.

$\beta_0 = 0.01$

$\mu = 1.00$

$\underline{P}(0) = 3,000\ \underline{I}$

$\underline{\theta}(0) = 0.0$

The output responses for the STR runs using this set of parameters are plotted in Figures 4.26 to 4.28, and the results tabulated in Table 4.7.

Table 4.7 Summary Of Simulated STR Top Product Composition Control Results

M = 2 K = 1 L = 2

$\beta_0 = 0.01$ $\underline{P}(0) = 3,000\,\underline{I}$

Run	ISE	Final XD	Final Parameters α_i	Final Parameters β_i	Figure	Disturbance
S8	108.6×10^{-4}	95.95%	-0.4333 -0.4055 Σ-0.8388	-0.2410 -0.2318 Σ-0.4728	4.26	Feed Flow -20%
S9	33.7×10^{-4}	96.95%	-0.4529 -0.2356 Σ-0.6885	-0.0739 -0.0592 Σ-0.1331	4.27	Set Point + .1%
S11	69.1×10^{-4}	94.57%	-0.5680 -0.1060 Σ-0.6740	-0.0816 -0.1216 Σ-0.2032	4.28	Set Point -1.5%

<u>Operating Conditions</u>

Feed Flow	18.00 g/s	Top Composition	= 96.02%
Reflux Flow	17.90 g/s	Bottom Composition	= 1.49%
Steam Flow	15.40 g/s		

During the initial period of operation, the parameters for a 1.5% step decrease in set point control exhibited a large fluctuation causing the manipulated variable (reflux flow) to be driven to its control limit at 6.0 g/s before levelling off to a new steady state level. This behavior was similar to that observed in the subsequent experimental work. The control performance was reasonably satisfactory, considering zero initial parameter estimates were used, except that offsets were observed in the product composition for either step changes in load or set point, as was the case with the experimental results.

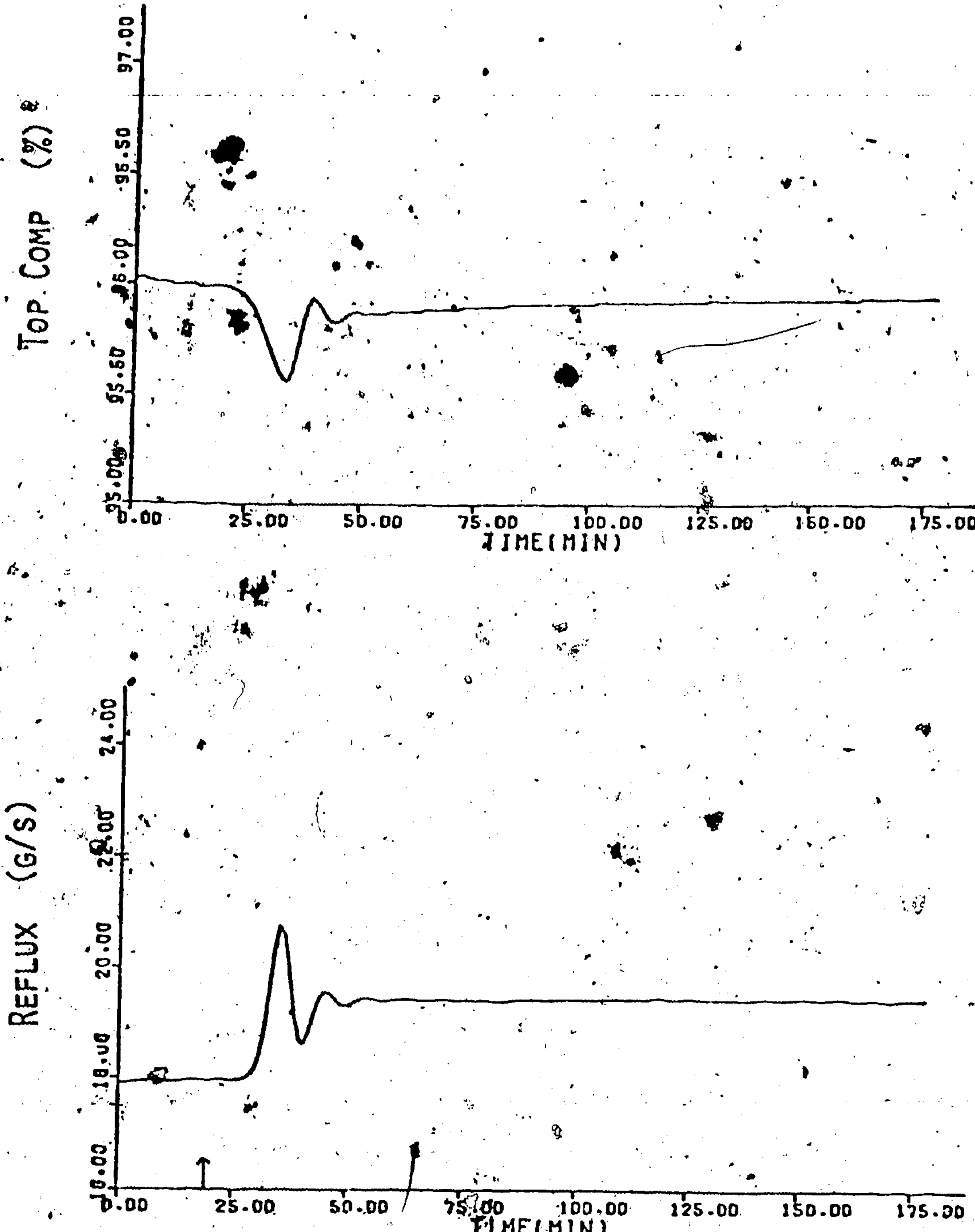


Figure 4.26 Simulation Of STR Top Product Composition Control For a -20% Step Change In Feed Flow Rate

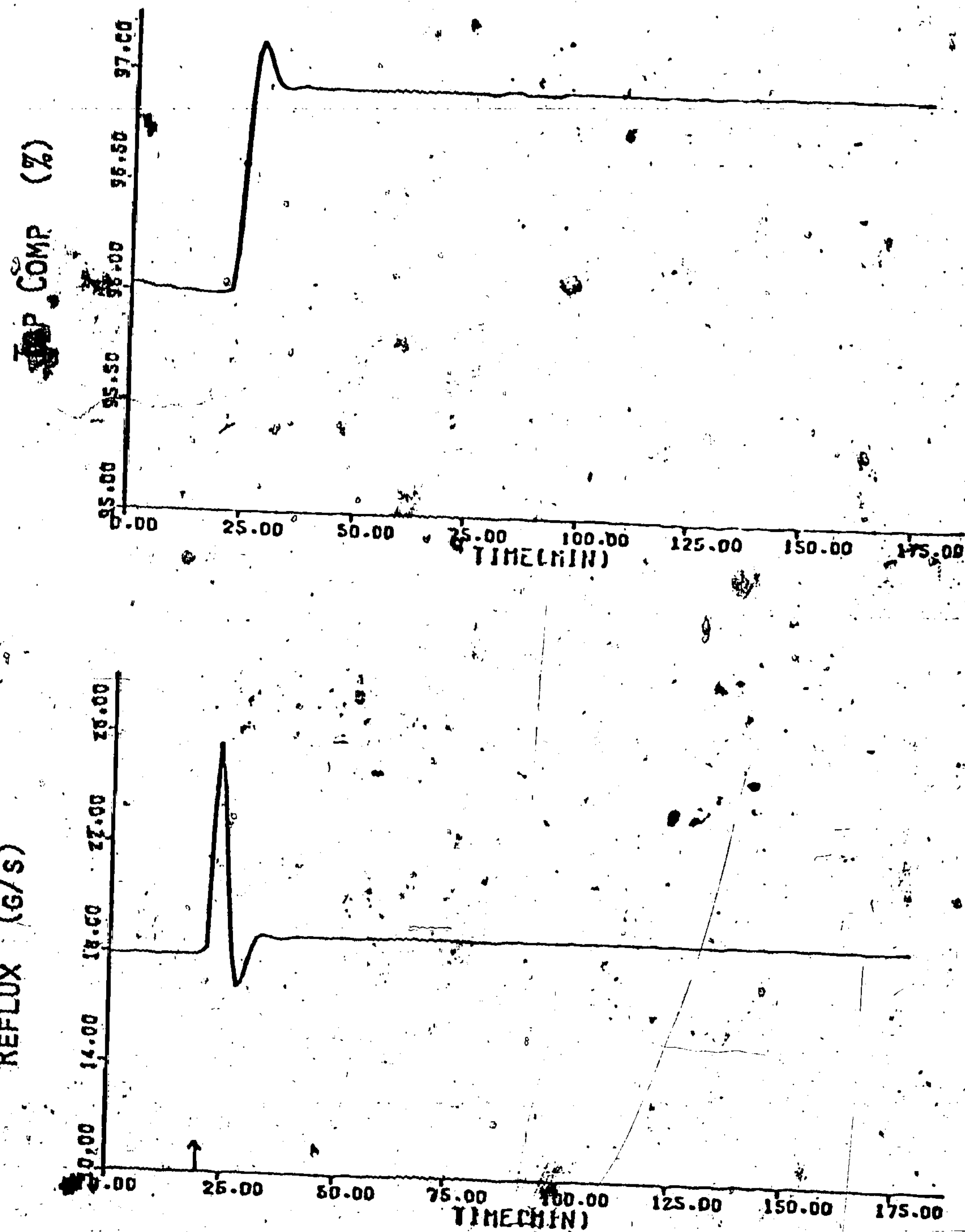
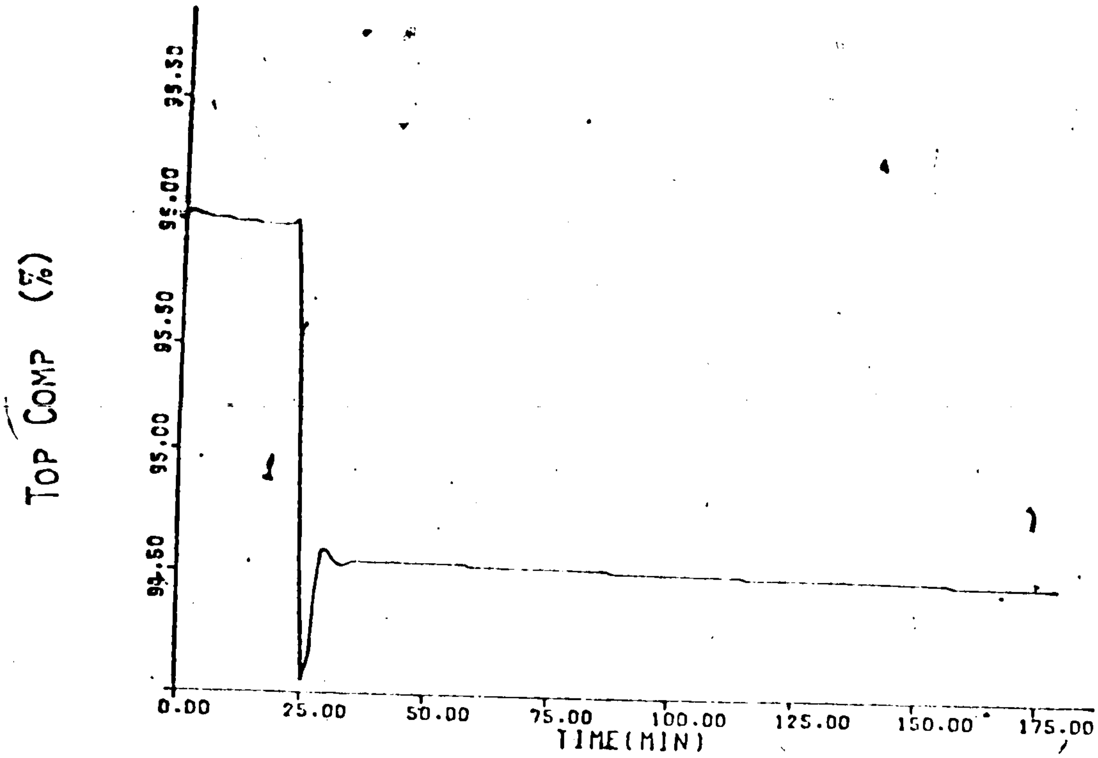


Figure 4.27 Simulation Of STR Top Product Composition Control For a +1% Step Change In Set Point



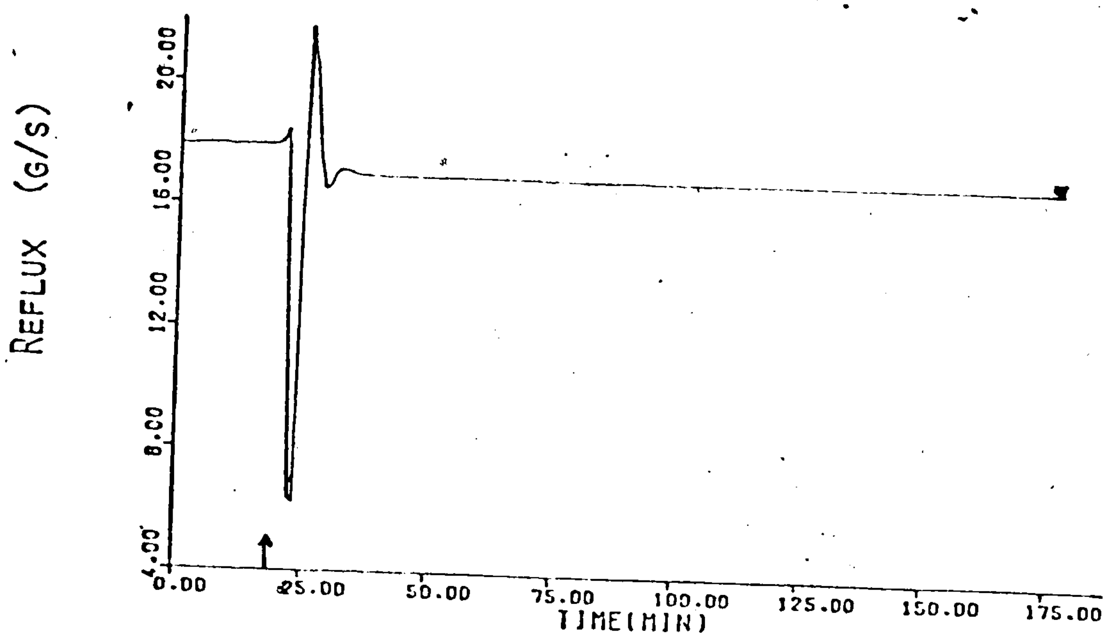


Figure 4.28 Simulation Of STR Top Product Composition Control For a -1.5% Step Change In Set Point

In the search for a $\underline{P}(0)$ value that could be used for both load disturbances and set point changes, tests were performed to study its effect and the results are summarized in Table 4.8. It was found that the $\underline{P}(0)$ value could be increased to 40,000 $\underline{I}$ for the set point changes without causing unstable responses. However ~~for the~~ feed flow rate decrease of -20%, a value of $\underline{P}(0)$=3,000 $\underline{I}$ appeared to be the upper limit to avoid oscillatory behavior.

Table 4.8 Summary Of The Effect Of $\underline{P}(0)$ Value On The Simulated STR Control Of Top Product Composition

$M = 2 \quad k = 1 \quad L = 2$

Run	$\underline{P}(0)$	Final XD	ISE	Final Parameters α_i	$\Sigma\alpha_i$	β_i	$\Sigma\beta_i$
Feed Flow -20%							
S 6	5,000 $\underline{I}$	-	10.8×10^{-4}	Oscillatory			
S 7	4,000 $\underline{I}$	-	12.7×10^{-4}	Oscillatory			
S 8	3,000 $\underline{I}$	95.95%	108.4×10^{-4}	-0.4333 -0.4055	-0.8388	-0.2410 -0.2328	-0.4738
Set Point +1%							
S 9	3,000 $\underline{I}$	96.95%	33.7×10^{-4}	-0.4529 -0.2356	-0.6885	-0.0739 -0.0592	-0.1331
S10	40,000 $\underline{I}$	96.98%	26.0×10^{-4}	-0.8745 -0.0324	-0.9069	-0.3089 -0.2217	-0.5306
Set Point -1.5%							
S11	3,000 $\underline{I}$	94.57%	69.1×10^{-4}	-0.5680 -0.1060	-0.6740	-0.0816 -0.1216	-0.2032
S12	40,000 $\underline{I}$	94.52%	62.0×10^{-4}	-0.9994 +0 2437	-0.7557	-0.3342 -0.4098	-0.7440

The simulation results indicated that the effects of the $\underline{\underline{P}}(0)$ value on offset and integral square error were consistent with that found in the experimental runs. Both the ISE value and offset decreased with a larger $\underline{\underline{P}}(0)$ value. The effect on offset could again be explained by the fact that for an increase of $\underline{\underline{P}}(0)$ from 3,000 (Run S9) to 40,000 $\underline{\underline{I}}$ (Run S10) that $\Sigma\beta_i$ had increased from -0.1231 to -0.5306 and $\Sigma\alpha_i$ from -0.6885 to -0.9069.

Another STR parameter, the number of β parameters, (i.e. the L value) was evaluated. Use of a value of 3 instead of 2 was investigated, but only minor improvement resulted as can be seen from the results in Table 4.9.

Table 4.9 Summary Of The Effect Of The Number Of β Parameters (L) On The Simulated STR Control Of Top Product Composition

M = 2 K = 1

β_0 = 0.01 $\underline{P}(0)$ = 40,000$\underline{I}$

Run	L	Final XD	ISE	α_i	$\Sigma\alpha_i$	β_i	$\Sigma\beta_i$
Set Point +1.0%							
S10	2	96.98%	26.0×10^{-4}	-0.8745 -0.0324	-0.9069	-0.3089 -0.2217	-0.5306
S13	2	96.99%	25.6×10^{-4}	-0.8745 -0.0098	-0.8608	-0.2541 -0.2188 -0.2374	-0.7203
Set Point -1.5%							
S12	2	94.52%	62.0×10^{-4}	-0.9994 +0.2437	-0.7557	-0.3342 -0.4098	-0.7440
S14	3	94.51%	61.3×10^{-4}	-0.9686 +0.2699	-0.6987	-0.2415 -0.3284 -0.2662	-0.8861

The ISE and offset were reduced somewhat, but the final estimated values of $\Sigma\alpha_i$ and $\Sigma\beta_i$ parameters adapted in such a fashion that the effect on offset of their summation values cancel each other. These results suggest that when a higher than adequate number of β parameters is used, only marginal improvement may result and then at the expense of additional computation effort.

4.4 Discussion

The top product control system is fast and stable. It has a short time delay and is relatively noise free. The operating environment is unchanging. These combined factors allow for good control behavior using a conventional controller so it is not surprising that there is no advantage in using an advanced control scheme like the self-tuning regulator which is mainly designed to deal with processes that have a time delay, are subject to a changing environment, stochastic disturbances and system noise.

For control of top product composition, the conventional proportional plus integral controller once well tuned performed superbly in both simulation and experimental tests. The responses for a -1.5% step change in set point as shown in Figures 4.5 and 4.25 may have to be detuned to reduce the overshoot, but overall PI control provides a control performance that is satisfactory. In both simulation and experimental tests, PI clearly outperformed STR, yet in a previous study by Sastry el al (15), in a comparison between the STR and a moderately tuned PI controller, it was found that the STR provided the better performance. It should be noted that a comparison between well tuned PI control and STR control, with zero initial parameter estimates, is somewhat unfair since in general the STR begins with very poor initial parameter estimates while the PI controller has previously been tuned, usually for the disturbance of interest. In addition, using the integral square error as the performance index gives a severe measure of STR performance since the initial parameter estimation period gives rise to a relatively high integral square error before the model acquires a reasonable set of parameters.

The results have shown that the choice of β_o and the $\underline{\underline{P}}(0)$ value significantly influence control performance. Pretuning of these parameters would be advantageous before using the STR for control testing. The bang bang type of control performance that prevails with a high $\underline{\underline{P}}(0)$ value and/or a small β_o value suggests that by adding control weighting as proposed by Gawthrop and Clarke (17) the control performance should improve. The problem of offset could be resolved by increasing the $\underline{\underline{P}}(0)$ value, or replacing u(t) by u(t) - u(t-1) in equations (2.10 and 2.11) (5).

5. BOTTOM PRODUCT COMPOSITION CONTROL

5.1 Description of Simulation Study

The column model used in the simulation program was identical to the one used in the top product control study except for a program counter incorporated into the program. This counter delayed the bottom composition calculated by the model for 4 time units before the control routine was allowed access to the information. For the bottom product composition control, a simulation run time of 45 time units was employed with each time unit being 256 seconds.

The counter was implemented in the program in order to simulate the equivalent time delay present in the bottom product G.C. analysis system. The effect of changes in steam flow were not delayed but were input directly to the model to immediately influence the bottom composition. The control of the bottom dynamic behavior of the column was complicated by this excessive transport time delay.

The control law calculation was scheduled to run every 256 seconds with the required inputs being the bottom product composition and steam flow rate. The operating sequence of events in the cycle can be summarized as follows:

1. Bottom composition is updated by the simulated gas chromatograph analysis.
2. Values for steam flow and bottom composition are acquired by the control program.

3. The control signal is calculated using the specified control routine.

4. The control signal is cascaded to the steam flow analog controller.

5.2 Proportional Plus Integral Control

The initial controller settings were determined experimentally using the Ziegler-Nichols tuning method. Details of the tests performed are given in Appendix F. The constants were further tuned for a 20% step decrease in feed flow rate. A summary of the different PI control tests is provided in Table 5.1.

Table 5.1 Summary Of Simulation Results for PI Control Of Bottom Product Composition

Run	K_P	T_R	ISE	Figure	Disturbance
B1	0.00	0.	3.297	[illegible]	Feed Flow -20%
B2	0.06	42.	1.5[illegible]6	5.1	Feed Flo[illegible] 20%
B3	0.08	42.	1.167	5.1	Feed Flow -20%
B4	0.06	42.	1.925	5.2	Feed Flow +20%
B5	0.08	42.	1.575	5.2	Feed Flow +20%
B6	0.06	42.	0.082	5.3	Set Point - 2%
B7	0.08	42.	0.081	5.3	Set Point - 2%
B8	0.06	42.	0.082	5.4	Set Point + 2%
B9	0.08	42.	0.090	5.4	Set Point + 2%

Operating Conditions

Feed Flow	18.00	g/s
Reflux Flow	11.00	g/s
Steam Flow	11.95	g/s

Top Composition = 95.7%
Bottom Composition = 5.6%

5.2.1 Load Disturbance Control

All feed flow rate disturbances were introduced 2.5 time units from the start of the simulation run. For a 20% step decrease in feed flow rate as shown in Figure 5.1, the bottom composition did not start decreasing until 4 time units after the disturbance was introduced due to the time delay incorporated into the output from the column dynamic model. It was essentially after this elapsed time delay that the manipulated variable (steam flow) was gradually lowered to raise the composition back to its set point. The output responses were sluggish, due to the long system time delay and time constant. The open loop response has been included to provide a contrast to the control runs. Clearly, of the different simulations performed, the preferred controller setting was $K_P = 0.08$, $T_R = 42$ minutes as shown by the lower ISE value of 1.167 as listed in Table 5.1.

For a feed flow rate step increase of 20%, this same controller setting again provided the better control performance as shown by the responses plotted in Figure 5.2. As can be seen, the system is easier to control for an increase in feed flow rate than a decrease in feed flow rate. This is due to the different gains and time constants that exist due to the nonlinear dynamic response of the column to positive and negative feed flow rate disturbances.

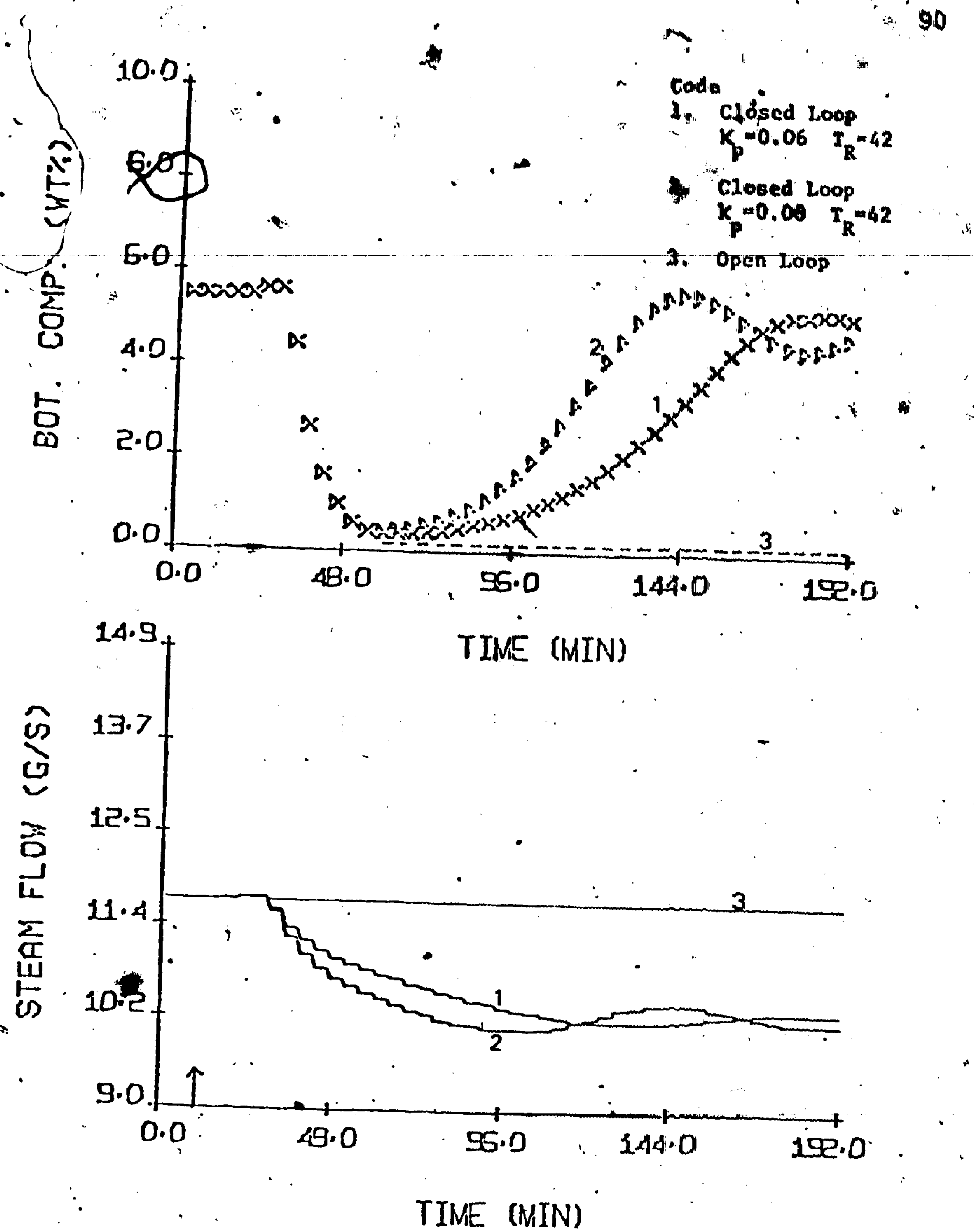


Figure 5.1 Simulated Behavior Of Bottom Product Composition Under Open Loop And Under Closed Loop PI Control For a 20% Step Decrease In Feed Flow Rate

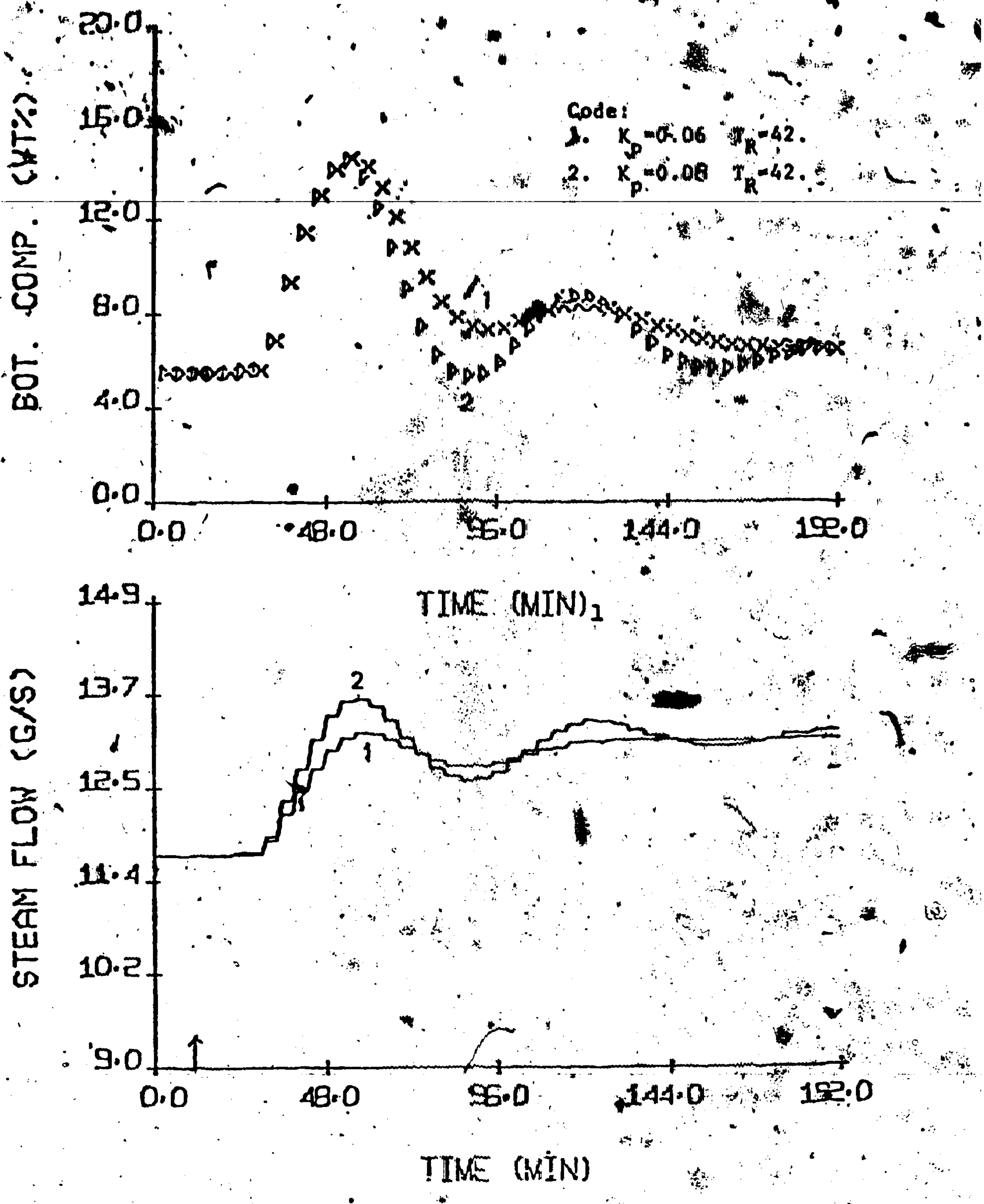


Figure 5.2 Simulated Behavior For PI Control Of Bottom Product Composition For a 20% Step Increase In Feed Flow Rate

5.2.2 Set Point Control

Servo control performance was evaluated for a -2% step change in set point with the resulting manipulated and controlled variable responses, for the two selected settings, given in Figure 5.3. As can be seen, both choices of settings provided reasonably satisfactory control with almost identical ISE values as shown in Table 5.1.

For a +2% step change in set point, the control performance using K_p = 0.08, T_R = 42. resulted in an overshoot and a slightly oscillating behavior. By lowering K_p to the value of 0.06, an improved control performance was obtained (cf Figure 5.4).

On the basis of the simulation results listed in Table 5.1, the control performance obtained using the controller setting of K_p = 0.08, T_R = 42. will be used as the base case for evaluation of the self-tuning regulator for both servo and regulatory control.

5.3 Self-Tuning Regulator

5.3.1 System Identification

Preliminary simulations were carried out for control of bottom composition using the STR with zero initial estimates, but difficulties were encountered in trying to obtain a converged set of parameter estimates. This is illustrated by the results of Run B10. For a feed flow disturbance of -20%, using β_0 = 10.0, $\underline{P}(0)$ = 50.$\underline{I}$ and $\underline{\theta}(0)$ = 0.0 for a total run time of 700 minutes, the responses and

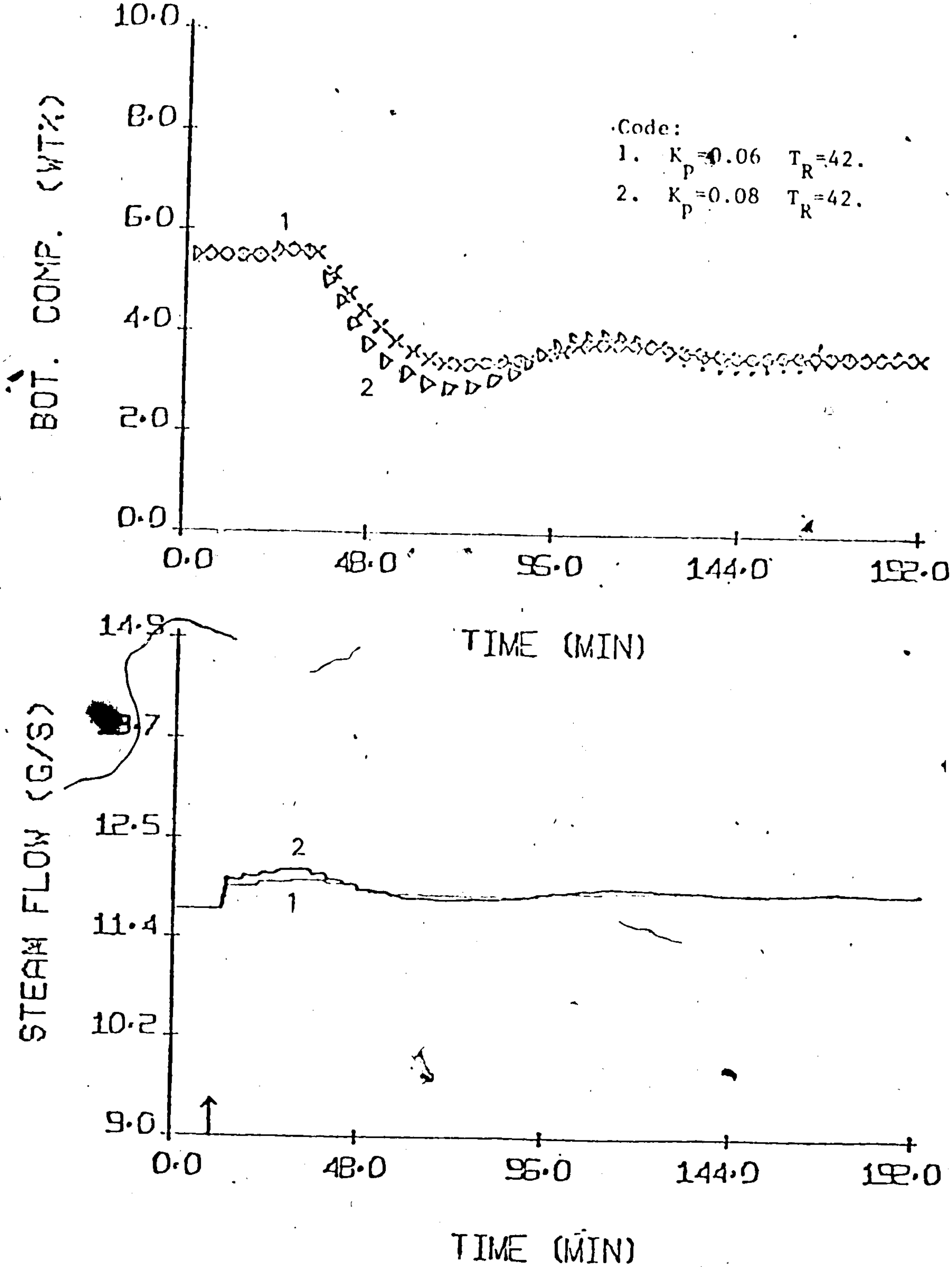


Figure 5.3 Simulated Behavior For PI Control Of Bottom Product Composition For a 2% Step Decrease In Set Point

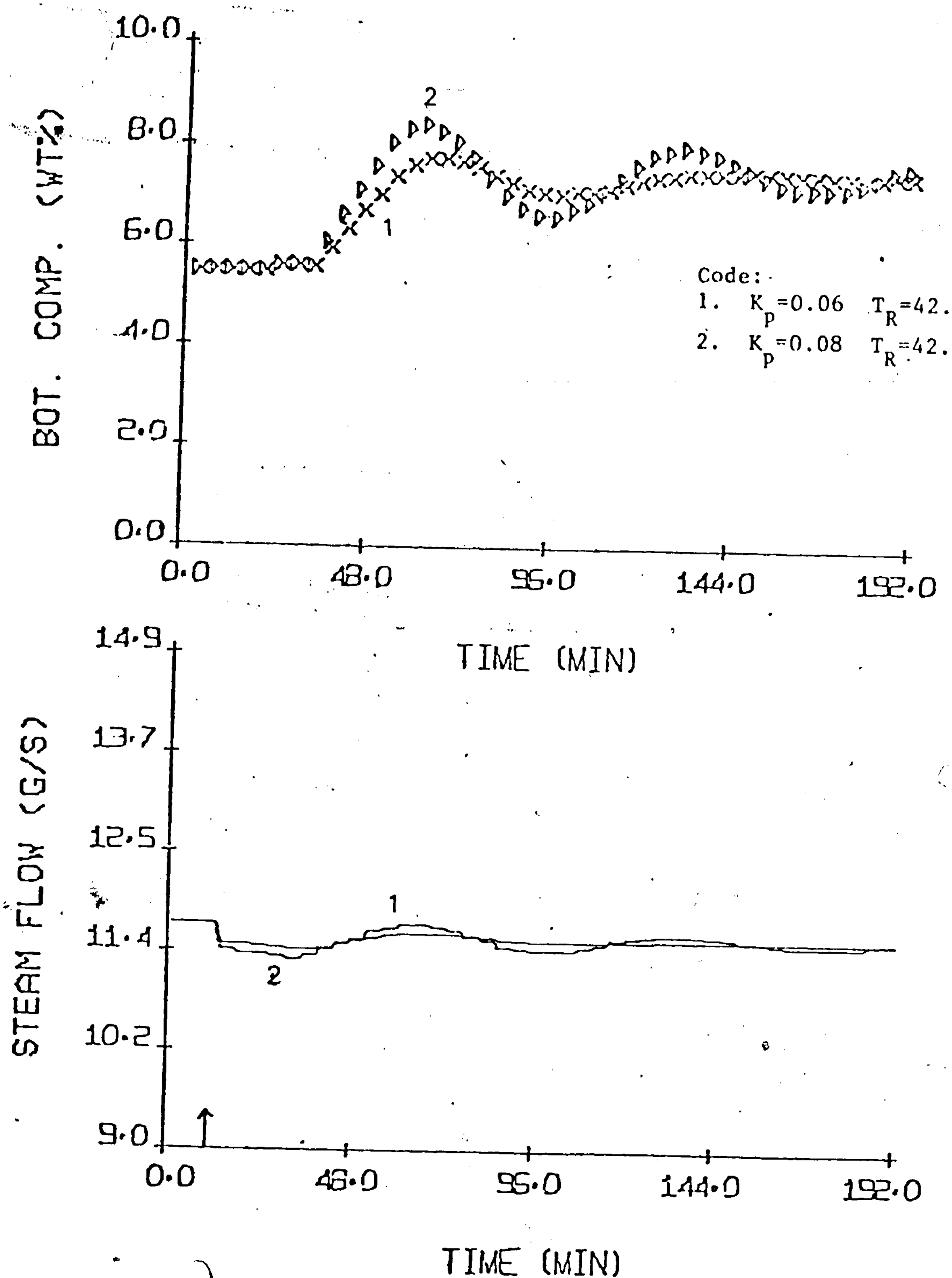


Figure 5.4 Simulated Behavior For PI Control Of Bottom Product Composition For a 2% Step Increase In Set Point

parameter behavior are given in Figures 5.5 and 5.6 respectively. As can be seen, the system tends to go unstable with the bottom composition response oscillating with an increased magnitude.

For the model chosen with seven parameters, five β_i and two α_i, must be estimated. As Figure 5.6 shows, the estimated α_i parameters were diverging from one another, cycling at the same frequency as the oscillations found with the steam flow and bottom composition in Figure 5.5. Since the β_i parameters exhibited large initial fluctuations and then levelled off to some steady state values gradually, the transient behavior evident in Figure 5.5 must be attributed to the behavior of α_i estimates.

This problem in system identification is believed to be caused by the large time delay relative to the time constant. Instead of using zero initial parameter estimates that failed to yield good control, a multiple pass method was used. By performing parameter estimation runs prior to the actual control runs, it is expected that better performance due to a better set of initial estimated parameters, will result. For Run B11, a square pulse disturbance in feed flow rate was introduced into the system. This is accomplished by introducing a 20% step decrease in feed flow rate after 10 minutes of simulation time had elapsed. The decrease in feed flow rate was then maintained for a duration of 170 minutes. The system response and the parameters are plotted in Figures 5.7 and

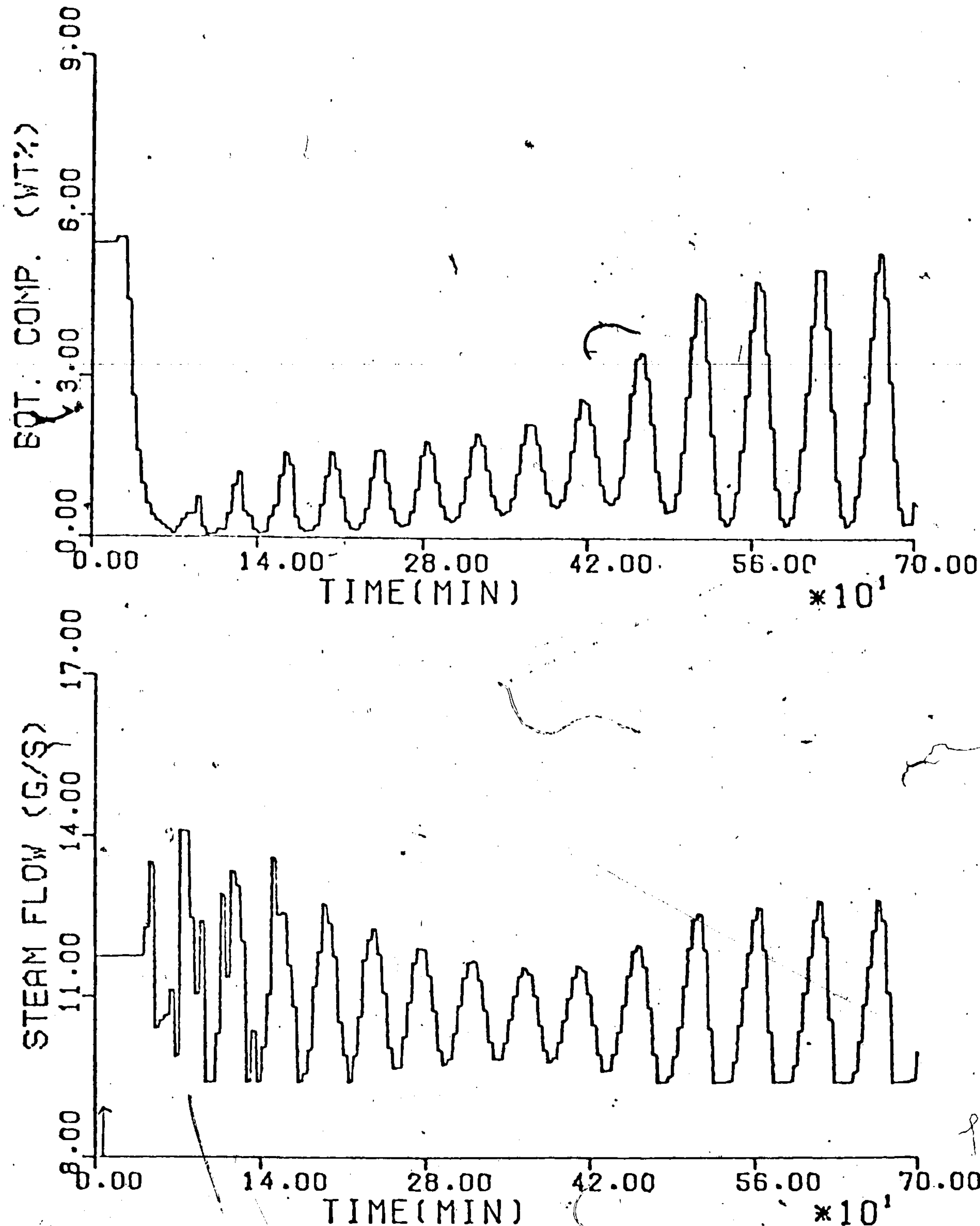


Figure 5.5 Simulated Response For STR Bottom Product Composition Control For a -20% Step Change In Feed Flow Rate, Starting With Zero Initial Parameter Estimates

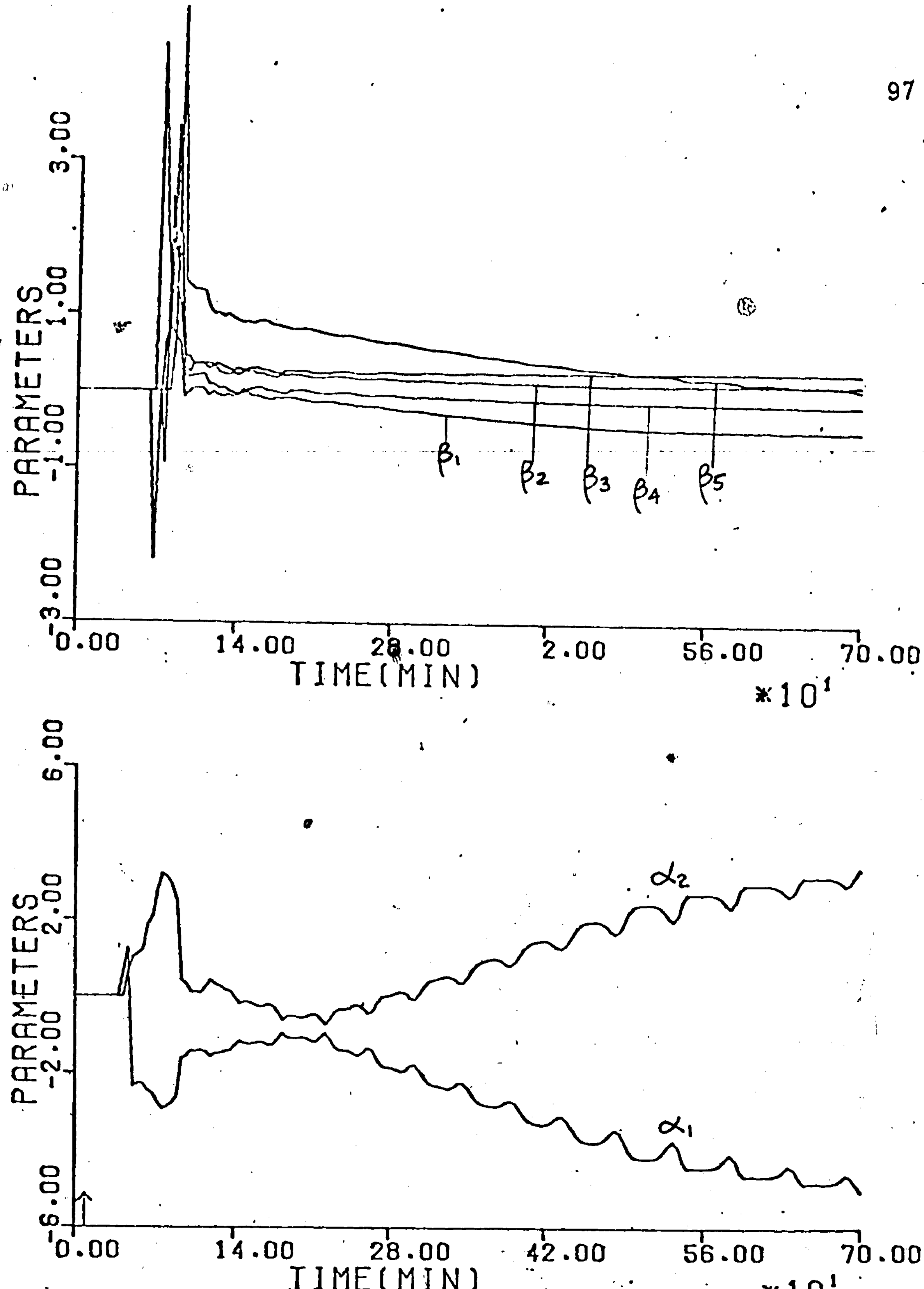


Figure 5.6 Parameter Behavior For Simulated STR Bottom Product Composition Control For a -20% Step Change In Feed Flow Rate, Starting With Zero Initial Parameter Estimates

5.8 respectively. The 2nd arrow indicates the time that the flow is returned to its initial steady state level. As would be expected, the system response was identical to that of Run B10 until the feed was returned to its normal level after 180 minutes of simulation elapsed time. The behavior of the parameters (cf Figure 5.8) shows that as the feed is returned to its normal value, the α_i parameters start converging instead of diverging as in Run B10. Despite this convergence, the bottom product composition response exhibited significant oscillation but did manage to return to its set point after about 700 minutes of time.

By performing the parameter estimation run (Run B11), a set of converged estimated parameters is obtained. Furthermore, the fact that the output returned to its set point provides some confidence in the estimated parameter values. . Results from parameter estimation runs performed using different flow patterns are discussed in Appendix G.

With the estimated parameters obtained from using the square pulse disturbance in Run B11, the parameters were further adapted using a 20% step decrease in feed flow rate in Run B1[illegible]. The results in Figure 5.9 show that the bottom composition started to move back to its set point as the estimated parameters began to reach their steady state values in 180 minutes or 45 control intervals after the disturbance was introduced. This is a significant improvement compared to Run B1[illegible] in which use of the zero initial parameter estimates resulted in an unstable system.

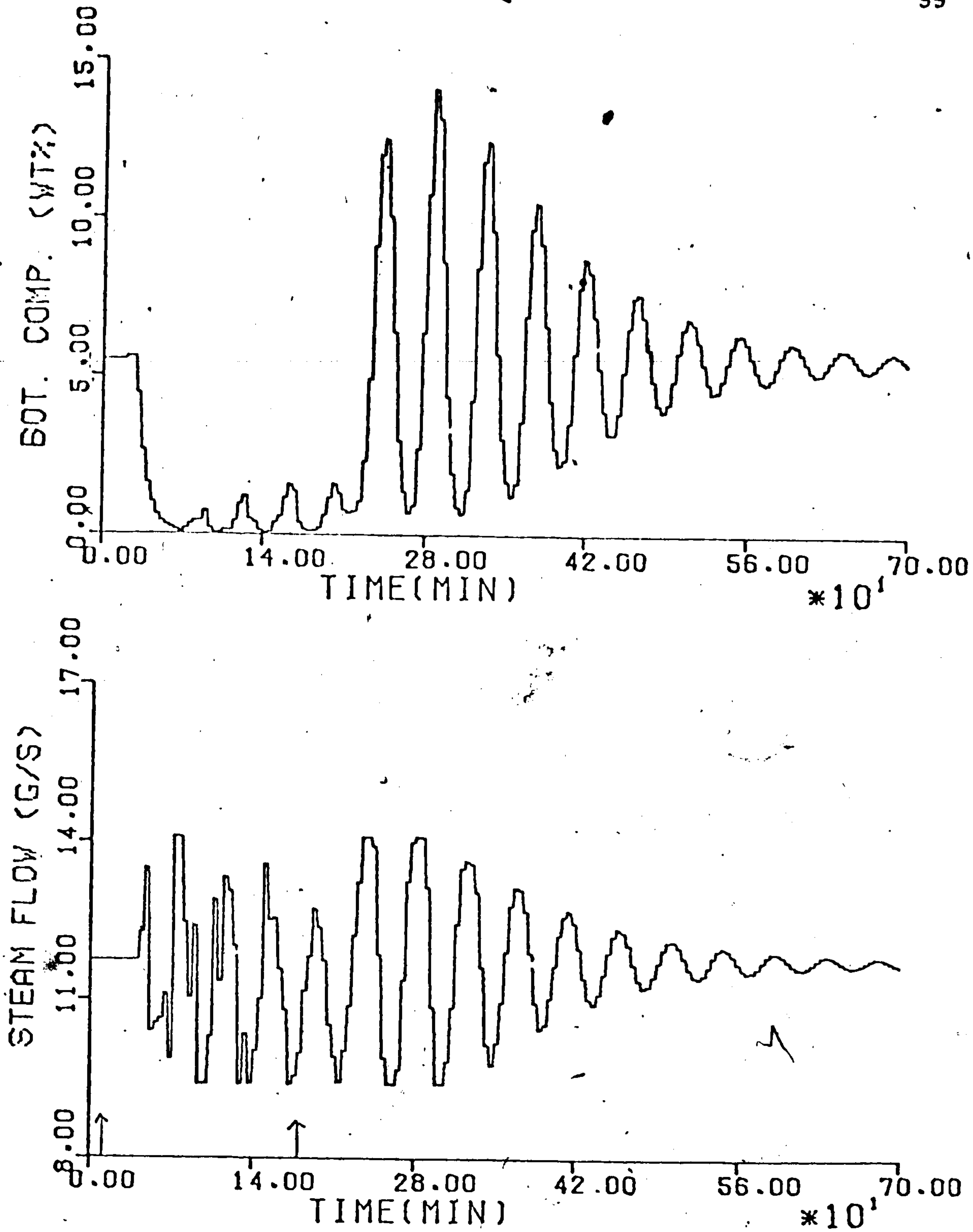


Figure 5.7 Simulated Response For STR Bottom Product Composition Control For a Pulse Disturbance In Feed Flow Rate Of -3.6 g/s In Magnitude For a Duration Of 170 Minutes, Starting With Zero Initial Parameter Estimates

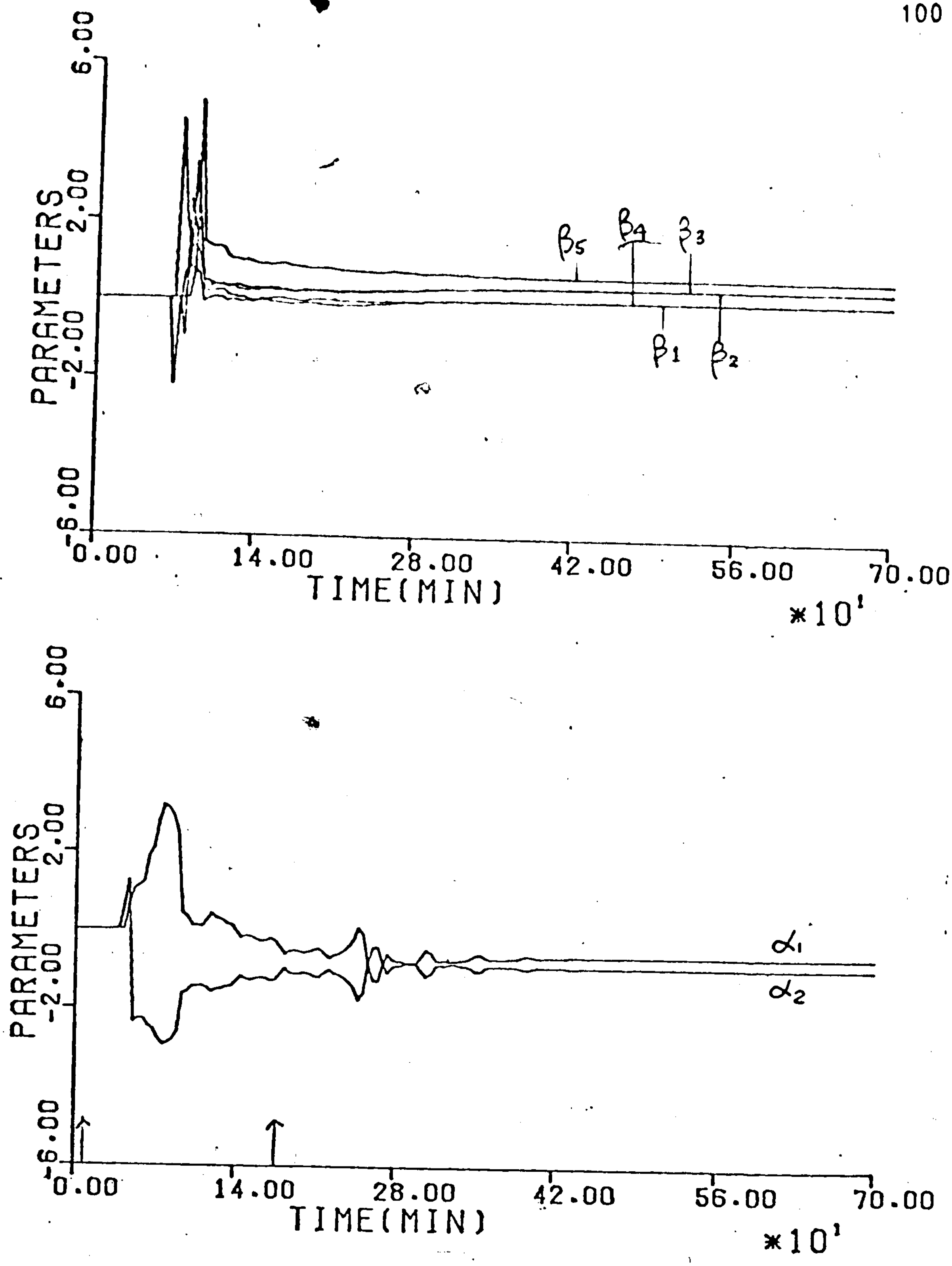


Figure 5.8 Adaption Pattern Of Parameters For a Pulse Disturbance In Feed Flow Rate Of -3.6 g/s In Magnitude For a Duration Of 170 Minutes, Starting With Zero Initial Parameter Estimates

The parameters identified in Run B12 were such that the $\Sigma \beta_i$ was close to the desired value of -1.0 which implies that reliable parameter estimates have been obtained. The behavior of the parameter estimates in Run B12 are plotted in Figure 5.10 and the results for all the parameter estimation runs are summarized in Table 5.2.

Table 5.2 Summary Of STR Parameter Identification Runs For Bottom Product Composition Control

$M = 2 \quad K = 4, \quad L = 5$

$\beta_0 = 10. \quad \underline{P}(0) = 50.$

Run	Figure	Initial Parameters	Final Parameters α_i	Final Parameters β_i	Flow Disturbance
B10	5.5 5.6	0.0	-5.0358 3.3362	-0.4895 0.1190 0.2624 -0.1594 0.0596	18.00 g/s 0 – 10 – 700 min. 14.40 g/s
B11	5.7 5.8	0.0	-0.6066 -0.8923	0.0398 0.3613 0.3830 0.0480 0.6255	18.00 g/s 0 – 10 – 180 – 700 min. 14.40 g/s
B12	5.9 5.10	Final Parameters Of Run B11	-4.1887 3.0602	0.1503 0.0120 0.1294 -0.3217 -0.959515	18.00 g/s 0 – 10 – 700 min. 14.40 g/s

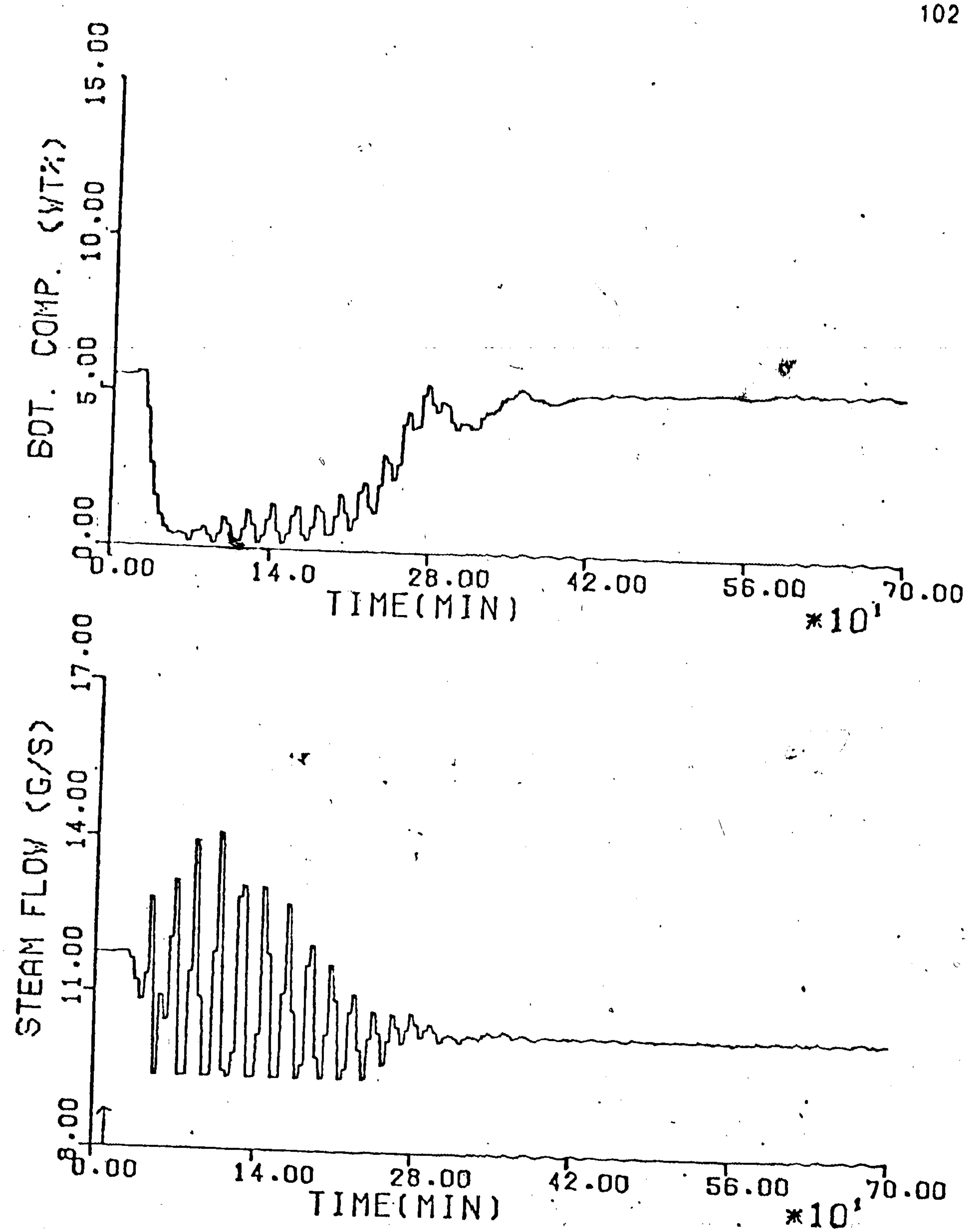


Figure 5.9 Simulated Response Of STR Bottom Product Composition Control For a -20% Step Change In Feed Flow Rate, Starting With The Initial Estimates As Those Obtained From Run B11

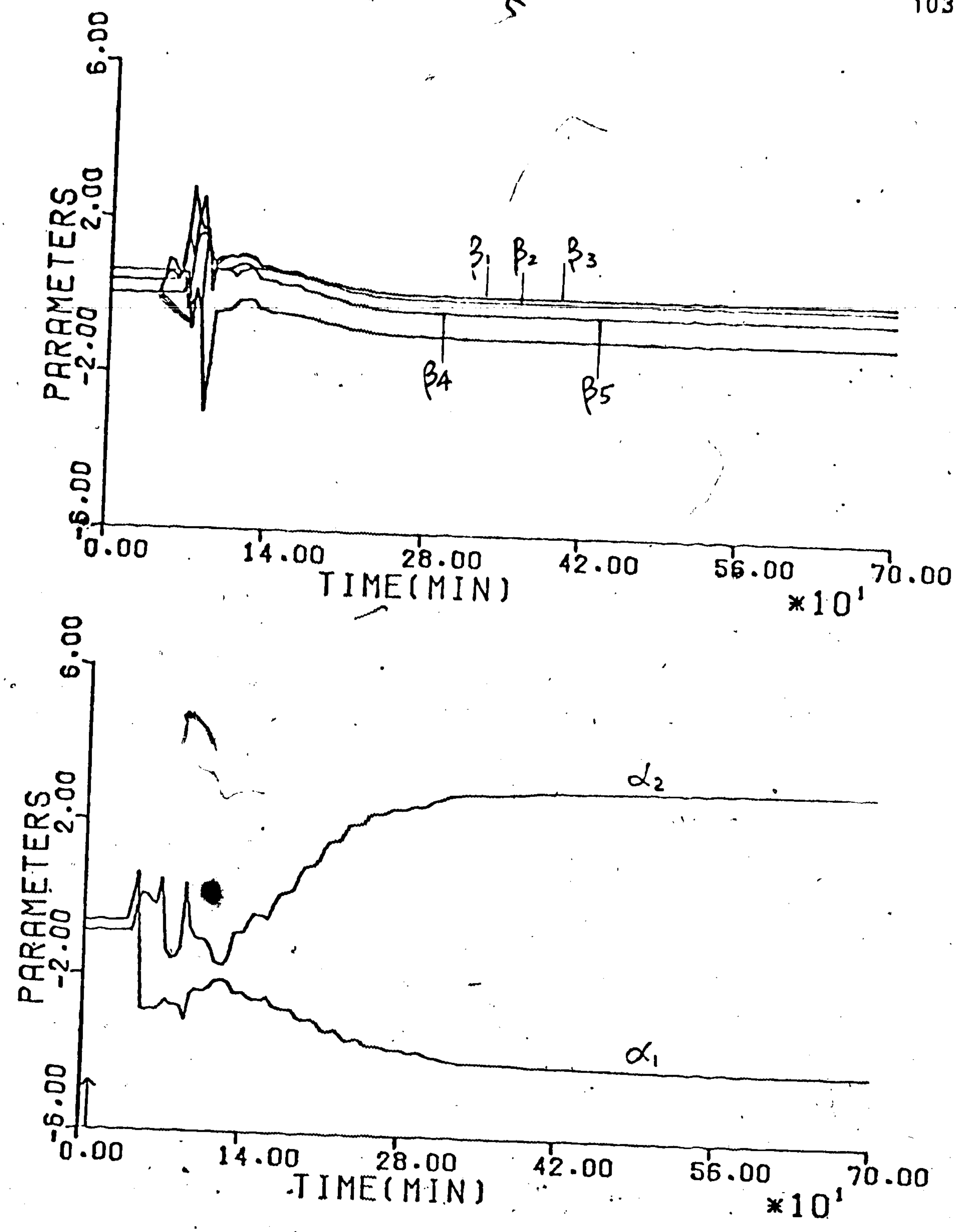


Figure 5.10 Parameter Behavior For Simulated STR Bottom Composition Control For a -20% Step Change In Feed Flow Rate, Starting With The Initial Estimates As Those Obtained From Run B11

5.3.2 Load Disturbance Control

With the final estimated parameters obtained from Run B12, the control performance for a load disturbance was tested by introducing a 20% step decrease in feed flow rate. In these tests, the parameter identification was stopped prior to the introduction of the disturbance. This approach, designated as Fixed Parameter Control (FPC), has resulted in good control performance with the output response significantly better than the well-tuned PI. To compare the performance of the two types of control, the PI and FPC responses are plotted in Figure 5.11. The FPC response is quicker and the resulting ISE is lower compared to the value obtained under PI control. The results for the step changes in feed flow rate tests are summarized in Table 5.3.

Table 5.3 Summary Of Multiple Pass FPC Control Of Bottom Composition For Step Changes In Feed Flow Rate

Run	ISE	Figure	Disturbance
B13	0.792	5.11	Feed Flow -20%
B14	0.835	5.12	Feed Flow +20%

For a +20% feed flow rate load disturbance, with the same set of parameters and using the identical approach mentioned above, excellent FPC performance was obtained. The comparison between the PI and FPC control responses is

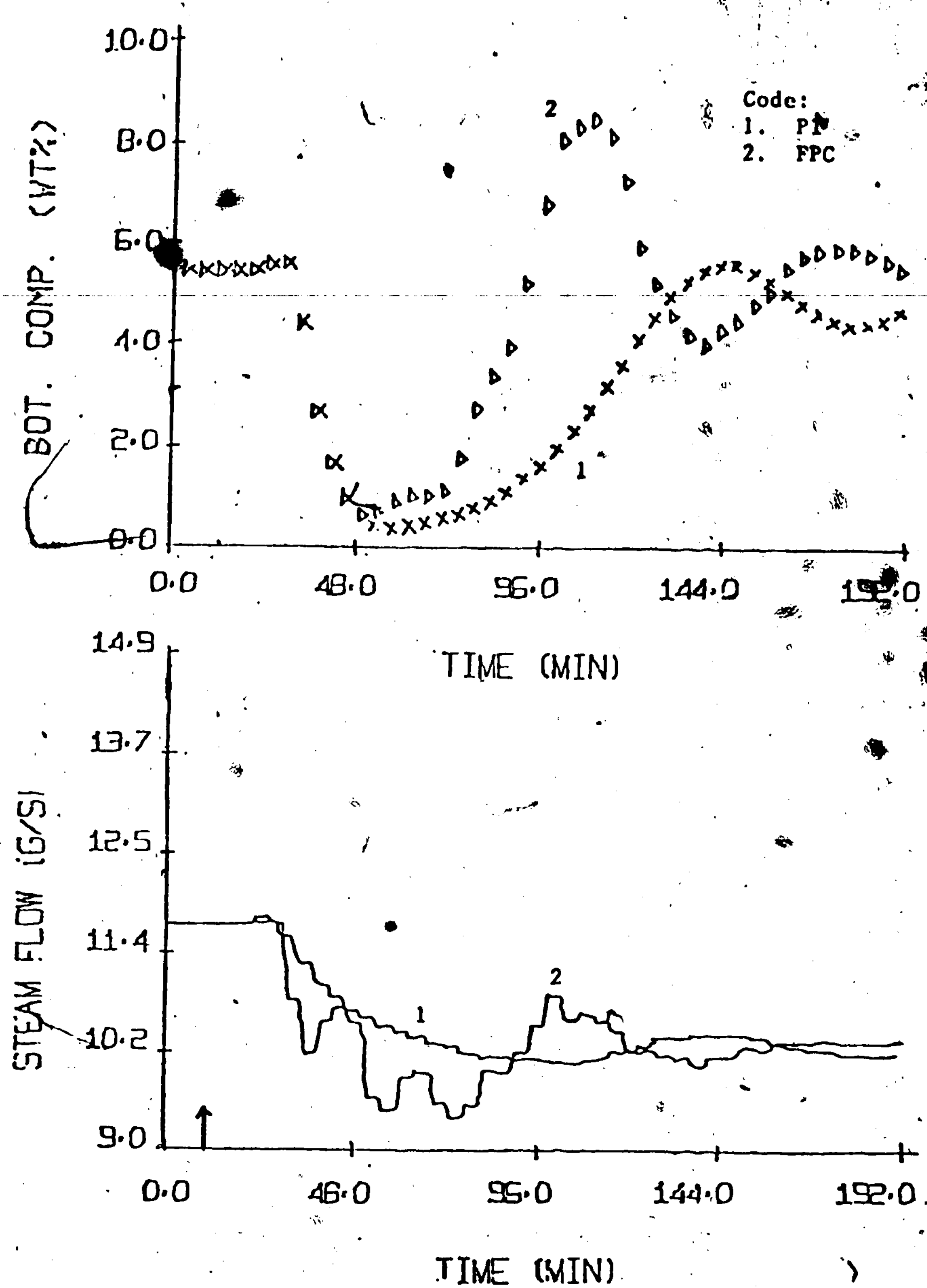


Figure 5.11 Comparison Of Simulation Responses For PI And Multiple Pass FPC Control Of Bottom Composition For a 20% Step Decrease In Feed Flow Rate

shown in Figure 5.12. The ISE value has decreased from 1.575 to 0.835.

5.3.3 SET POINT CONTROL

The control responses for a 2% step increase of set point under FPC operation using the parameters estimated in Run B12 has not performed well, but the ISE result is still comparable to that achieved using a well-tuned PI controller. It is not surprising that the FPC control response is not optimum since the parameters used were those estimated from load disturbances. From the responses plotted in Figure 5.13, it can be seen that both the input and output variables for the FPC run oscillate throughout the entire duration of the run.

To improve the control, it was decided to adapt the parameters obtained in Run B12. Allowing the parameters to adapt for a set point change has resulted in an improvement of control performance as shown by the lower ISE value. The improved control performance is obvious from the responses plotted in Figure 5.14, although a slight oscillation in the composition response still remains. However, by using the final parameters obtained in this run and adopting the FPC approach, the ISE is further reduced and the slight oscillation is eliminated as can be seen from the responses in Figure 5.15.

For the case of a 2% step decrease in the set point, using the FPC approach with the parameters obtained from a

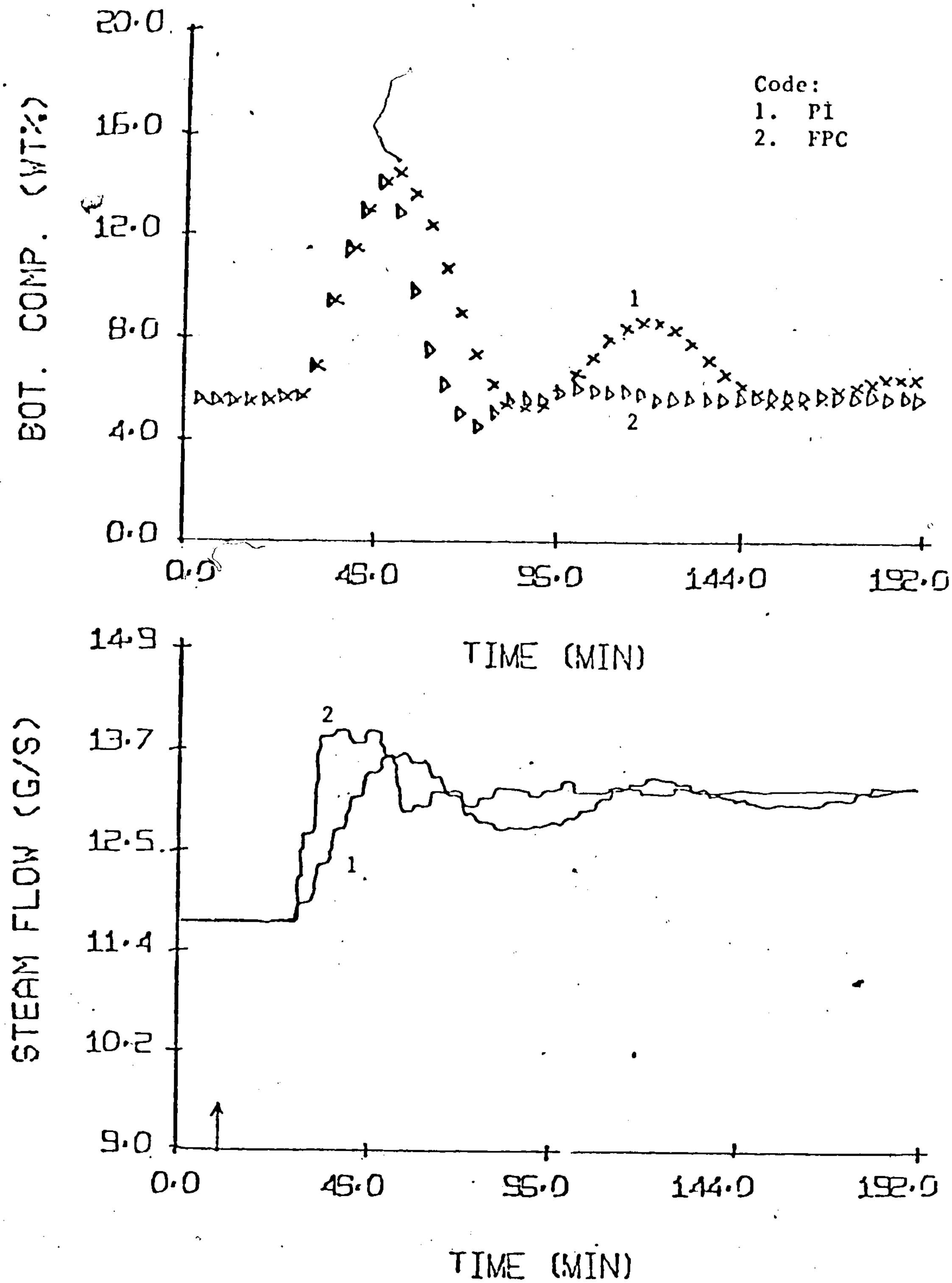


Figure 5.12 Comparison Of Simulation Responses For PI And Multiple Pass FPC Control Of Bottom Composition For a 20% Step Increase In Feed Flow Rate

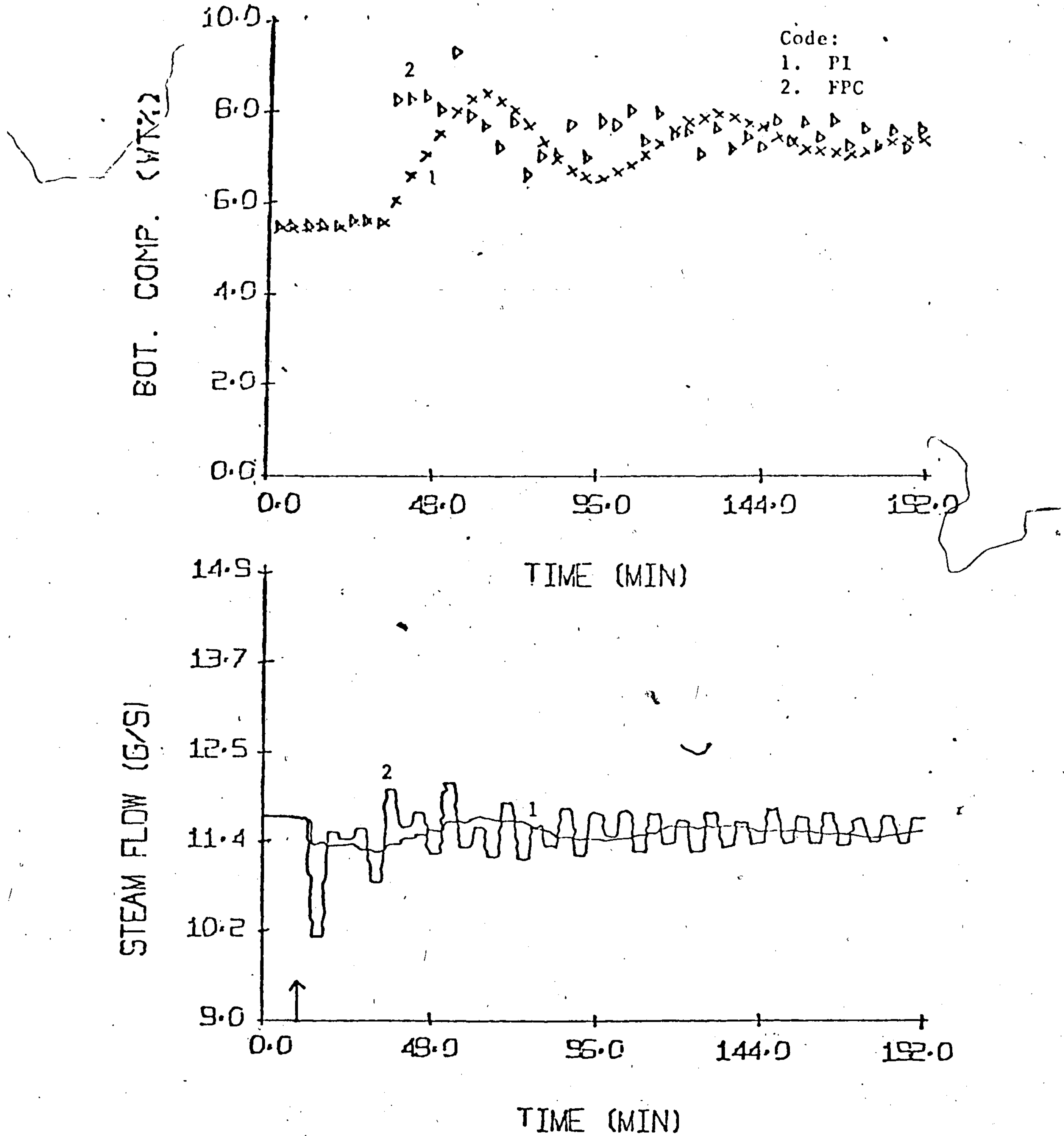


Figure 5.13 Comparison Of Simulation Responses For PI And Multiple Pass FPC Control Of Bottom Composition For a 2% Step Increase In Set Point

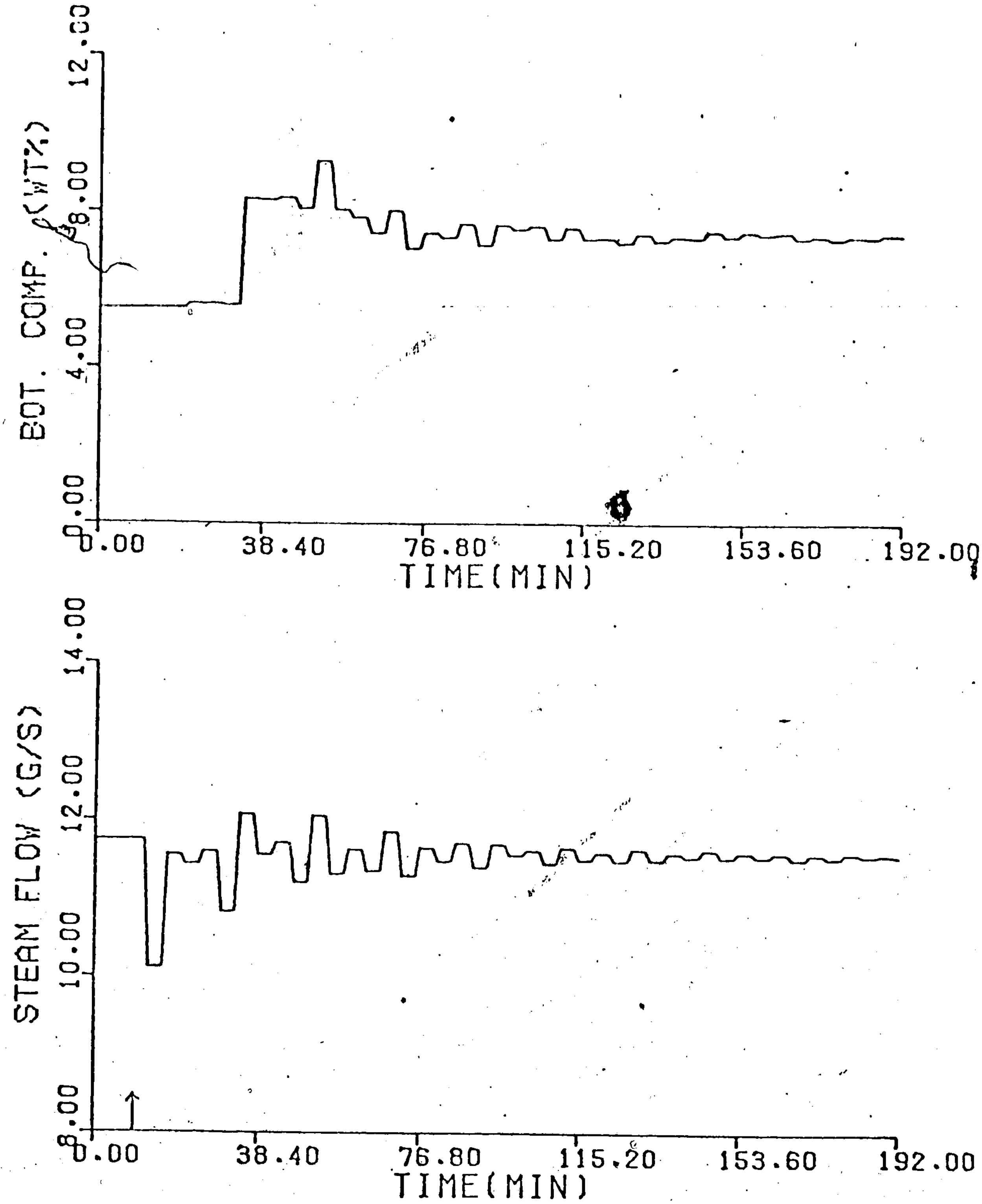


Figure 5.14 Simulated Response Of STR Control Of Bottom Product Composition For a 2% Step Increase In Set Point, Starting With Initial Parameters Values As Specified In Table 5.4

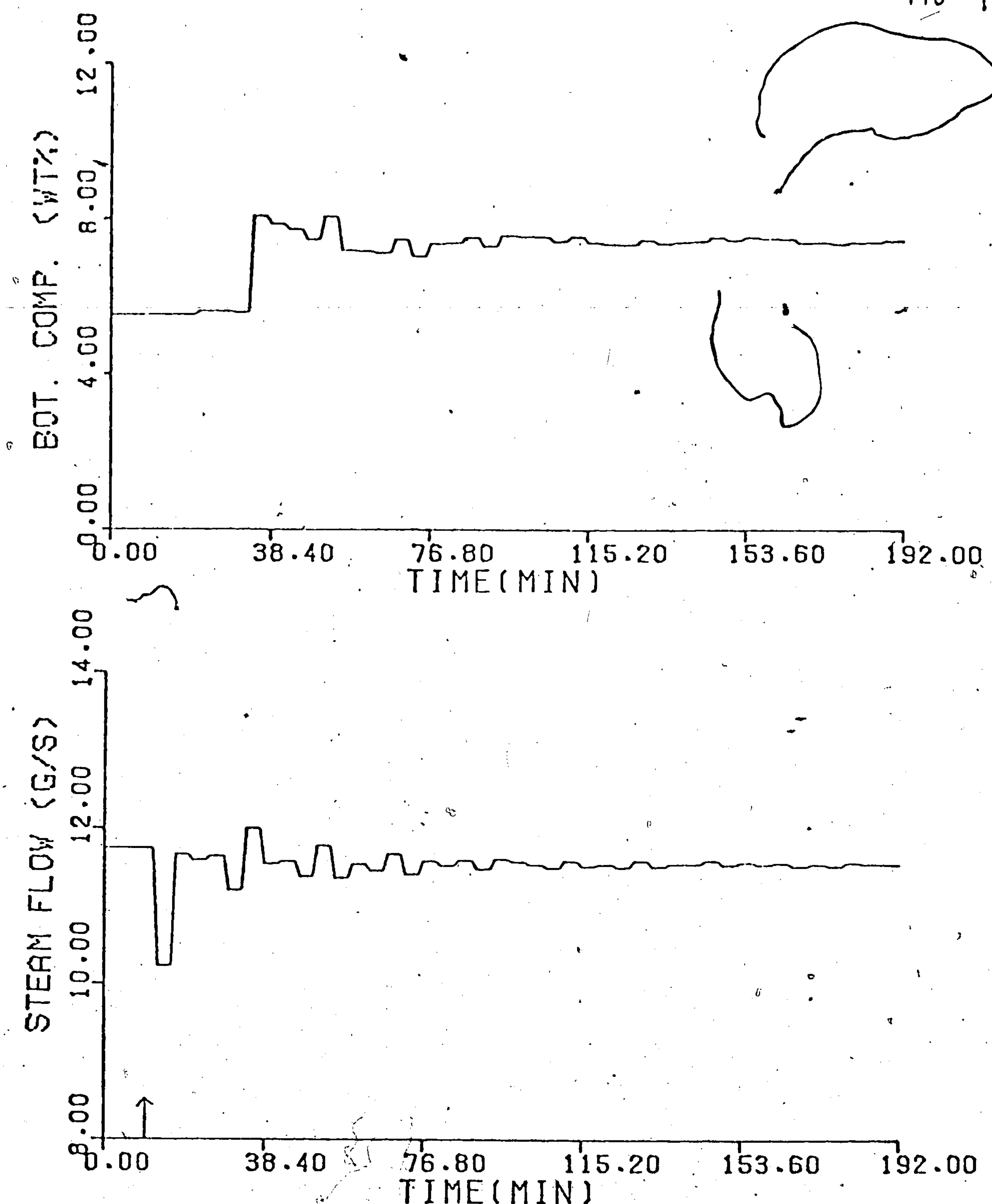


Figure 5.15 Simulated Response For FPC Control Of Bottom Product Composition For a 2% Step Increase In Set Point, With Parameters Obtained From Run B20

-20% feed flow rate disturbance, the control performance is better than that achieved using PI control. The controlled responses for both PI and FPC are plotted in Figure 5.16. By allowing the parameters to adapt as done in the case of the increase in set point, further improvement in control performance and reduction in ISE value were accomplished. The final FPC performance is shown in Figure 5.17. A summary of FPC and STR tests performed for set point disturbances is provided in Table 5.4.

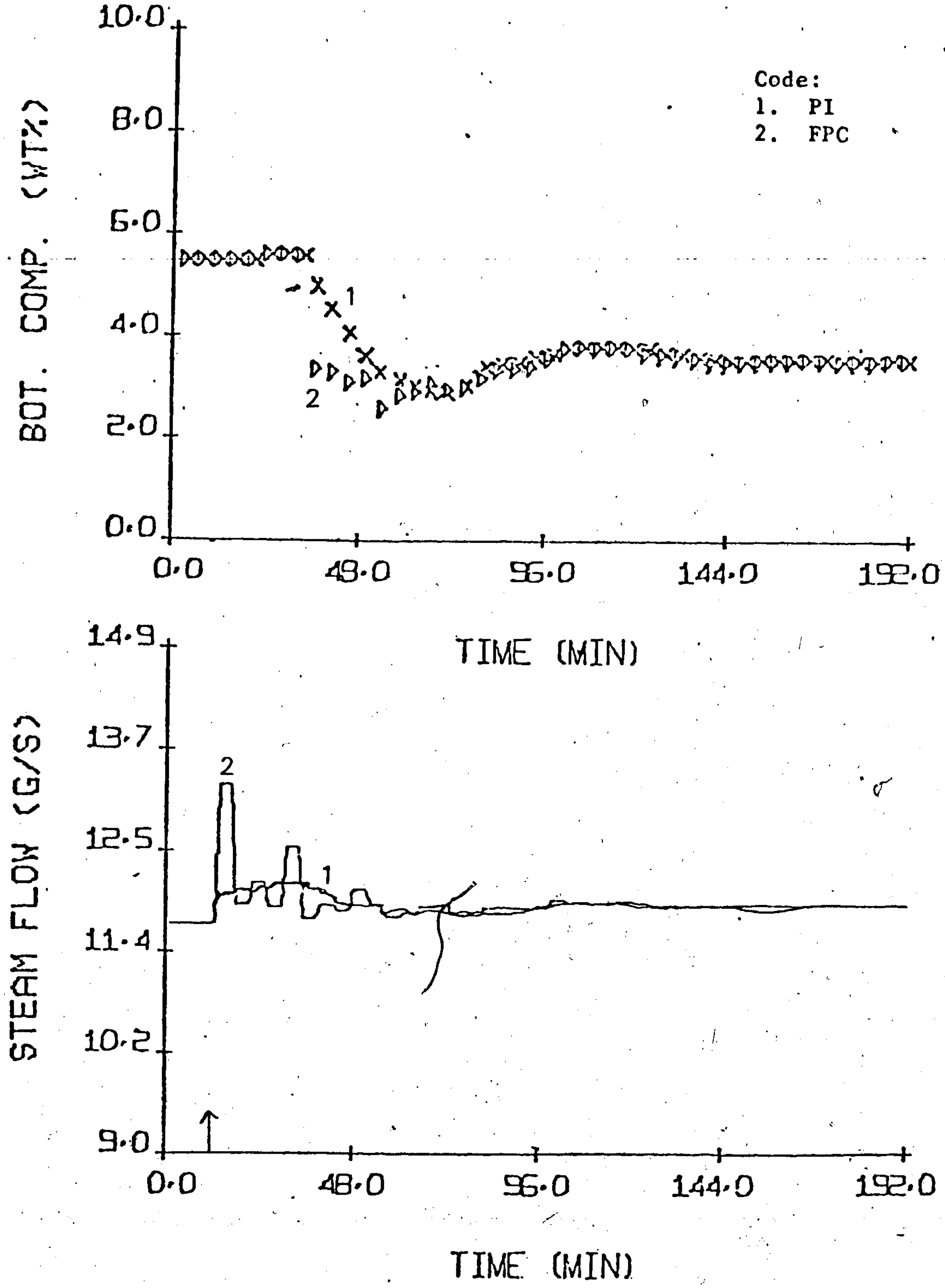


Figure 5.16 Comparison Of Simulation Responses For PI And Multiple Pass FPC Control Of Bottom Product Composition For a 2% Step Decrease In Set Point

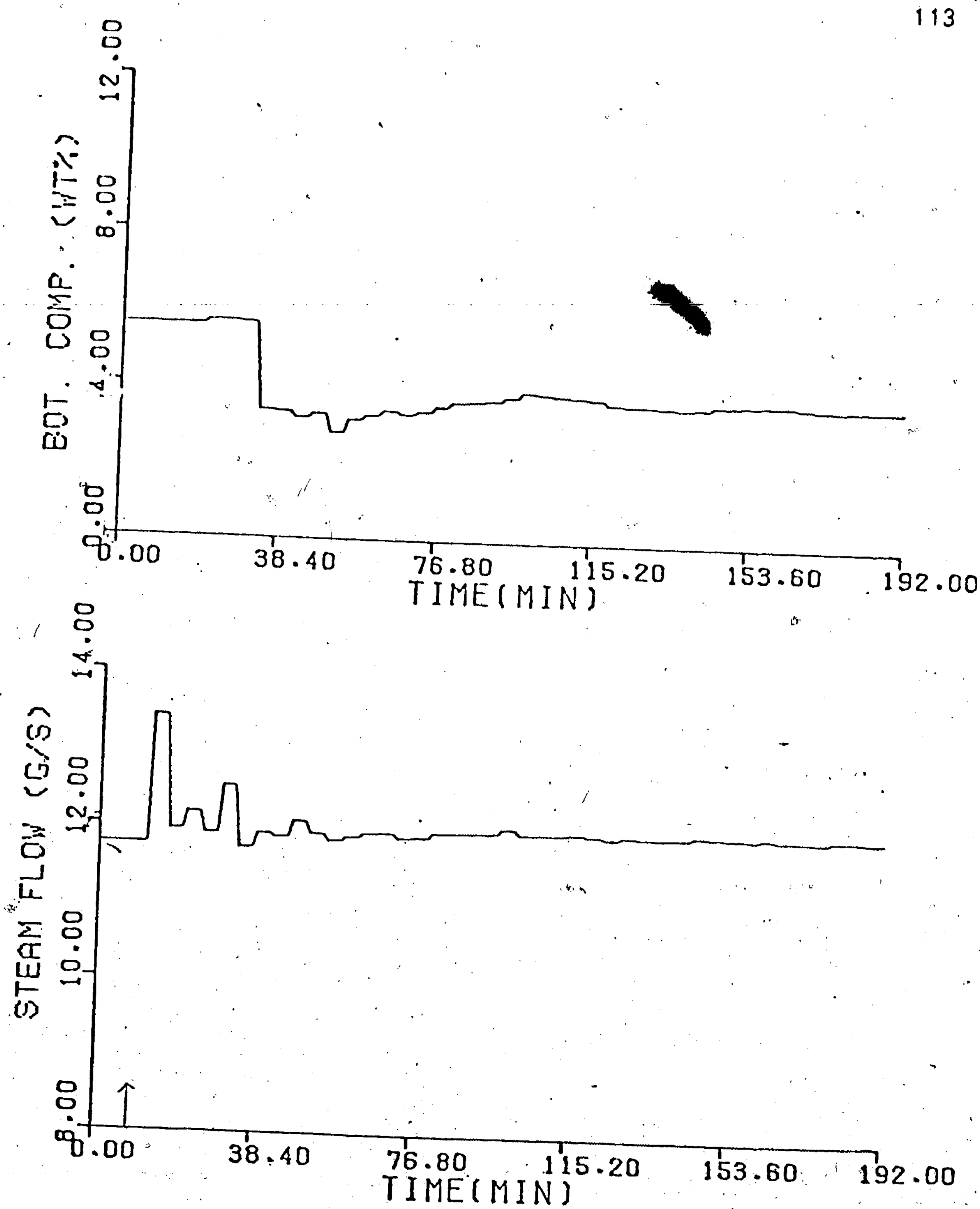


Figure 5.17 Simulated Response For FPC Control Of Bottom Product Composition For a 2% Step Decrease In Set Point, With Parameters Obtained From Run B23

Table 5.4 Summary Of FPC And STR Runs For Bottom Product Composition Control For Set Point Disturbances

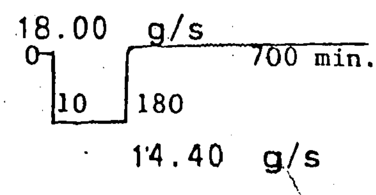
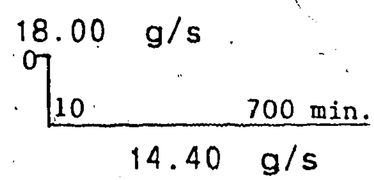
Run	ISE	Figure	Initial Parameters	Final Parameters α_i	Final Parameters β_i	Disturbance
B19	0.083	5.13	Run B12 Final Parameters	FPC		Set Point +2%
B20	0.078	5.14	Run B12 Final Parameters	-3.8792 3.2285	0.1169 0.0091 0.0784 -0.1901 -0.9639	Set Point +2%
B21	0.060	5.15	Run B20 Final Parameters	FPC		Set Point +2%
B22	0.073	5.16	Run B12 Final Parameters	FPC		Set Point -2%
B23	0.078		Run B12 Final Parameters	-4.2326 3.1423	0.1445 0.0398 0.1446 -0.2426 -0.9163	Set Point -2%
B24	0.069	5.17	Run B23 Final Parameters	FPC		Set Point -2%

5.3.4 Effect Of Time Delay Tests:

For bottom composition control, the large time delay is the main reason that the control performance is so poor. The effect of time delay on the self-tuning regulator has been discussed theoretically by Astrom (5). The variance of the output variable is expected to increase when there is added time delay in the system.

To investigate how system time delay affected control system performance, the inserted time delay was changed from 16 minutes to 4 minutes. The multiple pass method was used. The parameter estimation runs were performed using the same procedure as outlined in Section 5.4.1. Instead of having to estimate 7 parameters, due to the reduced time delay that resulted in a change of the k value from 4 to 1, only 4 parameters are now required. The runs showing the effect of the reduced time delay on the ISE and the estimated parameter values are summarized in Table 5.5.

Table 5.5 Summary Of Reduced Time Delay Runs For Bottom Product Composition Control Under FPC Operation

Run	ISE	Initial Parameters	Final Parameters α_i	β_i	Disturbance
B15		0.0	-3.3333 2.0047	-0.3171 -0.5549	
B16		Run B15 Final	-4.5801 2.9814	0.4449 -1.4525	
B17	0.1153	Run B16 Final Parameters	FPC		Feed Flow -20%
B18	0.1079	Run B16 Final Parameters	FPC		Feed Flow +20%
B25	0.0260	Run B16 Final Parameters	FPC		Set Point + 2%
B26	0.1215	Run B16 Final Parameters	FPC		Set Point - 2%

By reducing the time delay from 16 minutes to 4 minutes, the control performance for the feed flow disturbances is excellent. The ISE values decreased from 0.792 to 0.115 for a negative feed flow rate disturbance, and from 0.835 to 0.108 for the positive feed flow rate disturbance. In the case of a step increase in set point, the ISE values decreased from 0.083 to 0.026. However, for the set point decrease of 2% test, use of the FPC parameters resulted in a performance that caused the ISE value to increase from 0.073 to 0.1215. The fact that these FPC parameters were adapted for feed flow disturbances may account for the increase in ISE for the set point run.

Reducing the time delay of the experimental equipment is possible by shortening the bottom measurement system recycle line and improving the G.C. filtration system. By model formulation, the reduction of K value required in the STR can also be accomplished by an increase of system sampling interval. The results from the tests performed for different sampling intervals are given in Appendix H.

5.4 Discussion

The simulation results for proportional plus integral control have demonstrated the difficulty in satisfactorily controlling bottom product composition. The long time delay and time constant clearly dominate the output responses. The use of only a feedback control scheme has no capability to compensate for the effects of the long time delay since no corrective action is possible until the output variable deviates from the set point. Consequently, to avoid excessive control, the controller setting has to be eased off and the control behavior becomes sluggish.

When zero initial parameter estimates were used, the self-tuning regulator did not perform well either. By adopting the multiple pass method, the necessary adaption for the parameter estimates is obtained. The fact that the model has a correct time delay parameter defined provides an inherent predictor to cope with the time delay problem. As a result, the multiple pass approach yielded superior control when compared with well tuned PI control runs. The acceptable control performance of the multiple pass approach for set point disturbances using parameters that were tuned for load disturbances has demonstrated the robustness of the control scheme. If the parameters are allowed to adapt, even better performance can be obtained. Furthermore, the control performance can be greatly improved by reducing the system time delay, providing the appropriate set of system parameters are used in the multiple pass method.

6. CONCLUSIONS AND RECOMMENDATIONS

Control of top product composition using the self-tuning regulator has shown that the top product composition control system is fast and stable. This is due to the fact that the loop contains only a short time delay and is relatively noise free. A PI controller, once well-tuned, also provided excellent performance in both simulation and experimental tests. The PI controller outperformed the STR, using zero initial parameters. In contrast, the bottom product composition loop contains a long time delay so the PI controller was not able to provide satisfactory control. The STR, with the advantage of an inherent predictor in the model formulation, managed to achieve a much improved control performance once reasonable estimates were obtained with the multiple pass method.

The basic self-tuning regulator provides satisfactory control behavior for step changes in feed flow rate and set point even though the regulator is not formulated for such changes. The regulator was able to adapt the parameters for the disturbance to perform adequately. Tests using the fixed parameter controller, despite the fact that the parameters were identified for a different disturbance, showed that the controller was robust enough to provide good overall control.

On the basis of the results from the series of tests conducted on the effect of STR design parameters on control performance, the following conclusions can be stated:

a) Scaling factor, β_o :

The choice of the β_o value significantly influences the control behavior. A decrease of the β_o value has the effect of increasing the controller gain, as a result the change in controller output at each control interval becomes larger. Consequently, the parameters converge at a faster rate. In addition, the values of the final converged parameters change with different β_o values. The smaller the β_o value used, which causes an increase in the difference between $\Sigma\beta_i$ and -1.0, and a decrease in the absolute value of $\Sigma\alpha_i$, the larger the offset.

b) Initial covariance matrix, $\underline{\underline{P}}(0)$:

The use of a high $\underline{\underline{P}}(0)$ value which implies the lack of confidence in the initial parameter estimates, allows a larger change in the values of the estimated parameters at each control interval. Tests using zero initial parameter estimates showed that use of a higher value resulted in improved control performance. By starting with a high $\underline{\underline{P}}(0)$ value, showing less confidence in the parameters, there is more chance to converge to the true values. Consequently, the offset was found to decrease with an increase in the $\underline{\underline{P}}(0)$ value. However, too high a $\underline{\underline{P}}(0)$ value causes a very large fluctuation in the estimated parameters at the time the disturbance is introduced and can lead to an unstable behavior.

c) The number of β parameters, L:

Limited simulation tests performed indicate that if a higher than an adequate number of β parameters is used, only marginal improvement may be realized.

d) Initial parameters estimates, $\underline{\theta}(0)$:

Using zero initial parameters estimates and a high $\underline{\underline{P}}(0)$ value provides satisfactory top product composition control. For bottom product composition control, the STR has difficulty in identifying the system parameters when zero initial parameter estimates are used. This is due to the large time delay relative to system time constant. Parameter estimation runs were required to provide good initial system parameters for subsequent satisfactory control of bottom product composition.

Recommendations for future studies are the following:

a) The inserted time delay in the simulated bottom composition control loop reflects the existing time delay in the actual distillation column. The major problem encountered in the control of bottom composition was the long time delay. The reduction of experimental system time delay can be accomplished by redesigning the G.C. sample circulation system and an improvement and size reduction of the filtration equipment. The G.C. analysis time could likely be improved by experimenting with column packing other than Poropak Q.

b) Adding control weighting to the regulator as proposed by Clarke and Gawthrop (15) is recommended to remove the bang bang performance that is evident in this work.

c) Incremental control output [u(t) - u(t-1)] in the STR formulation to remove offset should be included.

Nomenclature

a) Alphabetic

A	Polynomial in Astrom model
a	Coefficient in polynomial A
B	Polynomial in Astrom model
b	Coefficient in polynomial B
C	Polynomial in Astrom model
c	Coefficient in polynomial C
E	Expectation operator
e	Normal random variable
F	Polynomial
G	Polynomial
H	Vector
$\underline{K}$	Gain vector
k	Time delay
L	Number of parameters
n	Order of Model
$\underline{P}$	Covariance matrix
t	sampling instant
$\underline{u}$	Control vector
V	Loss function
$\underline{y}$	Output vector

b) Greek

α_i	Coefficient of predictive model

β_i	Coefficient of predictive model
β_o	Scaling factor
ε	Moving average process of driving noise e
θ	Parameter vector
λ	Constant
μ	Exponential forgetting factor
Ψ	Information vector

c) Subscripts

i	i(th) element
K_p	Proportional controller gain value
T_R	Reset time
u_L	Control limit

d) Superscripts

T	Matrix transpose
-1	Matrix inverse
^	Estimated value

e) Abbreviation

DDC	Direct digital control
GC	Gas Chromatograph
PI	Proportional plus integral
STR	Self-tuning regulator

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APPENDIX A: RECURSIVE LEAST SQUARES ESTIMATION

In the self-tuning regulator, a recursive least squares algorithm is used to update the parameters at every sampling instant based on current and previous input and output data. The equations used are A.1, A.2 and A.3.

$$\underline{\hat{\theta}}(t) = \underline{\hat{\theta}}(t-1) + \underline{K}(t)\,[y(t) - \beta_0 u(t-k-1) - \Psi^T(t-k-1)\underline{\hat{\theta}}(t-1)] \qquad (A.1)$$

$$\underline{K}(t) = \underline{\underline{P}}(t-1)\Psi(t-k-1)\,[1+\Psi^T(t-k-1)\underline{\underline{P}}(t-1)\Psi(t-k-1)]^{-1} \qquad (A.2)$$

$$\underline{\underline{P}}(t) = \frac{1}{\mu}\{\underline{\underline{P}}(t-1) - \underline{K}(t)\,[1+\Psi^T(t-k-1)\underline{\underline{P}}(t-1)\Psi(t-k-1)]\,\underline{K}^T(t)\} \qquad (A.3)$$

The updating gain matrix $\underline{K}(t)$ in Equation A.2 is calculated first, then the covariance matrix $\underline{\underline{P}}(t)$ in Equation A.3, followed by the parameter vector $\underline{\hat{\theta}}(t)$ in Equation A.1.

The recursive least squares algorithm is derived in the following manner:

The predictive model used for estimating the self-tuning regulator parameters on-line is given by Equation A.4:

$$y(t) + \alpha_1 y(t-k-1) + \alpha_2 y(t-k-2) + \ldots + \alpha_n y(t-k-n) = \beta_0 [\,u(t-k-1) + \beta_1 u(t-k-2) + \ldots + \beta_L u(t-1-k-L)] + \varepsilon(t) \qquad (A.4)$$

where $L = n+k-1$

Defining:

$$\Psi^T(t) = [\,-y(t), -y(t-1), \ldots, -y(t-n+1), \beta_0 u(t-1), \beta_0 u(t-L)] \qquad (A.5)$$

$$\underline{\theta}(t) = [\hat{\alpha}_1, \hat{\alpha}_2, \ldots, \hat{\alpha}_m, \hat{\beta}_1, \hat{\beta}_L] \qquad (A.6)$$

allows Equation A.4 to be written as:

$$y(t) = \beta_0 u(t-k-1) + \Psi^T(t-k-1)\,\underline{\theta} + \mathcal{E}(t) \qquad (A.7)$$

Adopting the notation used by Sage & Melsa (7), let

$Z_t = y(t) - \beta_0 u(t-k-1)$

$\underline{H}_t = \Psi^T(t-k-1)$

$\underline{X} = \underline{\theta}$

$V_t = \mathcal{E}(t)$

$$Z_t = \underline{H}_t X + V_t \qquad (A.8)$$

The least squares estimate of $\underline{X}$, $\hat{\underline{X}}_{(t)}$ is as follows:

$$\hat{\underline{X}}_t = \hat{\underline{X}}_{t-1} + \underline{K}_t(Z_t - \underline{H}_t \hat{\underline{X}}_{t-1}) \qquad (A.9)$$

$$\underline{K}_t = \underline{\underline{P}}_t \underline{H}_t^T \qquad (A.10)$$

$$\underline{\underline{P}}_t = \underline{\underline{P}}_{t-1} - \underline{\underline{P}}_{t-1} \underline{H}^T (1 + \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T) \ \underline{H}_t \underline{\underline{P}}_{t-1} \qquad (A.11)$$

At time t, $\underline{H}_t$, Z_t, $\hat{\underline{X}}_{t-1}$, $\underline{\underline{P}}_{t-1}$ are available, so $\underline{\underline{P}}_t$ is calculated using $\underline{\underline{P}}_{t-1}$ and presently available $\underline{H}_t$ before $\hat{\underline{X}}_t$ can be calculated.

Another form of Equation A.10 can be written by substituting Equation A.11 into Equation A.10.

$$\begin{aligned}
\underline{K}_t &= [\underline{\underline{P}}_{t-1} - \underline{\underline{P}}_{t-1} H^T (1 + \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T)^{-1} \underline{H}_t \underline{\underline{P}}_{t-1}]\, \underline{H}_t^T \\
&= \underline{\underline{P}}_{t-1} \underline{H}_t^T - \underline{\underline{P}}_{t-1} \underline{H}_t^T (1 + \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T)^{-1} \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T \\
&= \underline{\underline{P}}_{t-1} \underline{H}_t^T [1 - (1 + \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T)^{-1} \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T] \\
&= \underline{\underline{P}}_{t-1} \underline{H}_t^T [1 - (\underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T) / (1 + \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T)] \\
&= \underline{\underline{P}}_{t-1} \underline{H}_t^T [1 + \underline{H}_t \underline{\underline{P}}_{t-1} \underline{H}_t^T]^{-1}
\end{aligned} \qquad (A.12)$$

For this study, Equation A.12 is used instead of Equation A.10. $\underline{\underline{P}}$ would be calculated at time t and used at time $t+1$ for calculating $\underline{K}_{t+1}$ to obtain $\hat{\underline{X}}_{t+1}$.

APPENDIX B: DETERMINATION OF BOTTOM COMPOSITION LOOP TIME DELAY

The chromatograph analysis of samples of the bottom product provides the composition measurement. The gas chromatograph analysis time and sample system transportation time delay plus the process time delay constitute the total system time delay.

1. Gas chromatograph analysis time:

   The gas chromatograph cycle time limits the sampling period to no shorter than 256 seconds due to the Poropak Q column packing conditions required for a reliable analysis.

2. Transportation time delay calculation:

   The product circulation system arrangement contributes a transport time delay of up to 13 minutes. The system configuration that results in this time delay is shown schematically in Figure B1. The magnitude of the time delay was calculated as follows:

   Section 1: Distance between reboiler bottom and pump (cf Figure B1) = 210.00 cm

   Size of tubing (copper type L)
   O.D. = 0.95 cm
   I.D. = 0.79 cm
   Thickness = 0.08 cm

   Based on a volumetric flow rate of 0.83 cm^3/sec., the transport time delay, excluding the effect of the heat exchanger is:

$$\text{Time} = \frac{(\text{distance}) * (\text{cross-sectional area})}{(\text{volumetric flowrate})}$$

$$= 123.00 \text{ seconds}$$

Section 2: Distance between pump and second filter = 486.41 cm

Size of tubing (copper type L)
O.D. = 0.64 cm
I.D. = 0.48 cm
Thickness = 0.08 cm

Velocity = 4.10 cm/sec
Time to travel specified distance = 118.64 seconds

Section 3: Distance between second filter and gas chromatograph

Size of tubing (plastic)
I.D. = 0.25 cm

Velocity = 14.80 cm/sec
Time to travel specified distance = 6.52 seconds

Total transportation time for all three sections = 248.16 seconds

Contribution of heat exchanger to transportation time delay:

The heat exchanger is located between the reboiler and pump as shown in Figure B1.

Exchanger manufacturer:
Heat-X-Changer Co. Inc., Brewster, New York.

Type: Single pass finned heat exchanger.

Specification:

O.D. = 6.35 cm
Length = 30.96 cm
Copper pipe fitting O.D. = 3.48 cm
Copper pipe fitting I.D. = 3.18 cm

Volume of fluid in heat exchanger to be displaced before a fresh sample reaches the gas chromatograph:

Cross-sectional area of heat exchanger = 7.92 cm^2
Volume of heat exchanger = 7.92 cm^2 x 30.96 cm
= 245.20 cm^3
Flowrate = 0.83 cm^3/sec
Time = 295.42 seconds

Contribution of filters to transportation time delay:

There are two filters in series:

First filter volume = not significant
Second filter volume = 200.00 cm^3

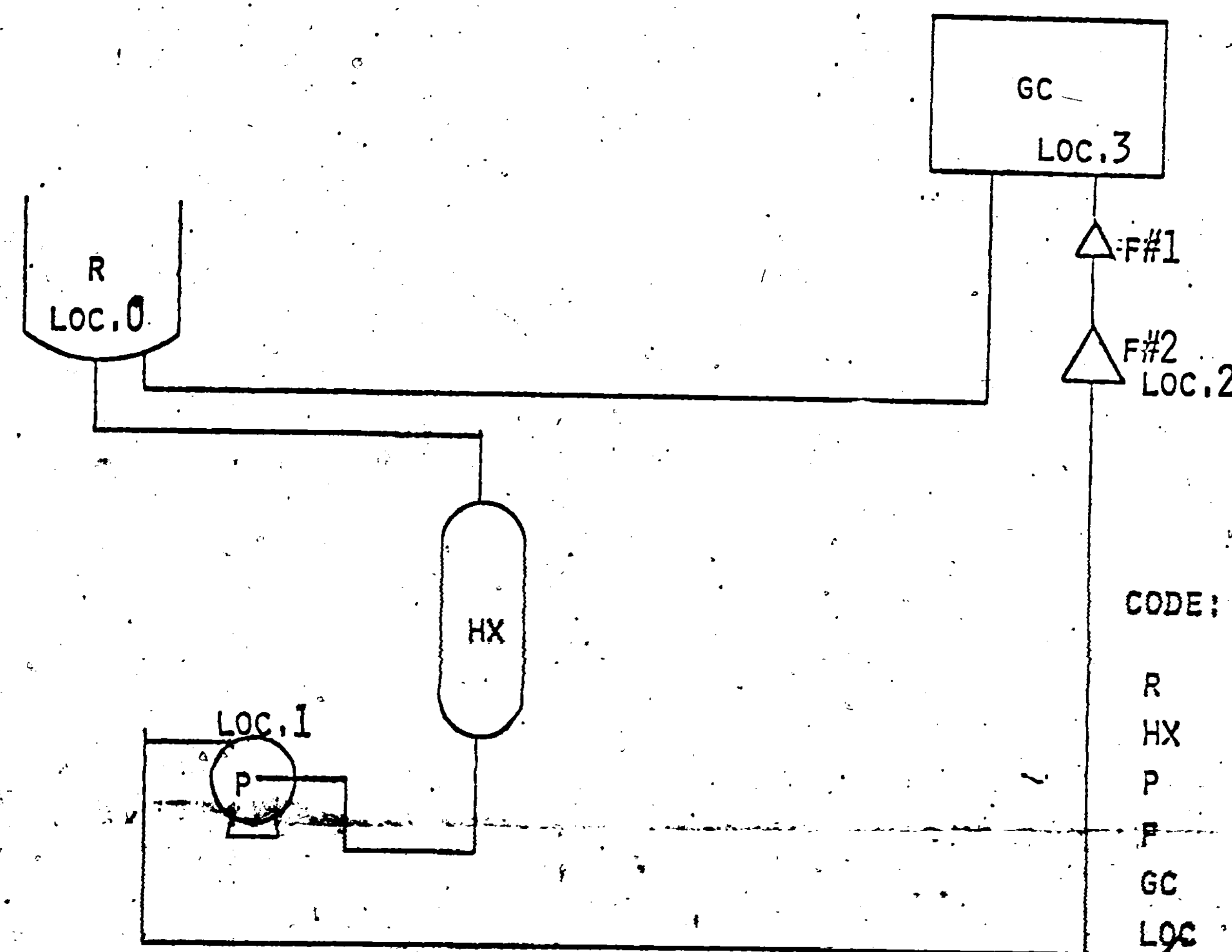
Flowrate = 0.83 cm^3/sec
Time = 240.96 seconds

Total time delay due to transportation:
(248.16 + 295.42 + 240.96) = 784.54 seconds

3. Process time delay measurement:

Experimental time delay tests with samples taken at the bottom of the reboiler and downstream of the gas chromatograph heat exchanger provide a reasonable estimate of the process time delay. Seven samples, at 256 second intervals, were taken with the steam flow rate decreased by 5% at 128 seconds of elapsed time. The results are given in Figure B2. The composition of the samples taken at the bottom of the reboiler began to increase after one sample interval and those at the heat exchanger did not change for two sample intervals due to extra transport time delay. Consequently, the process time delay was estimated as one sample interval, 256 seconds.

As can be seen, the total time delay appears to be about 21 minutes, so with a sampling interval of 256 seconds, the resulting K value is 5. This value for the time delay is consistent with that established by calculation.



CODE:

R	REBOILER
HX	HEAT EXCHANGER
P	PUMP
F	FILTER
GC	GAS CHROMATOGRAPH
LOC	LOCATION

FIGURE B1 SCHEMATIC DIAGRAM OF GAS CHROMATOGRAPH SAMPLING SYSTEM

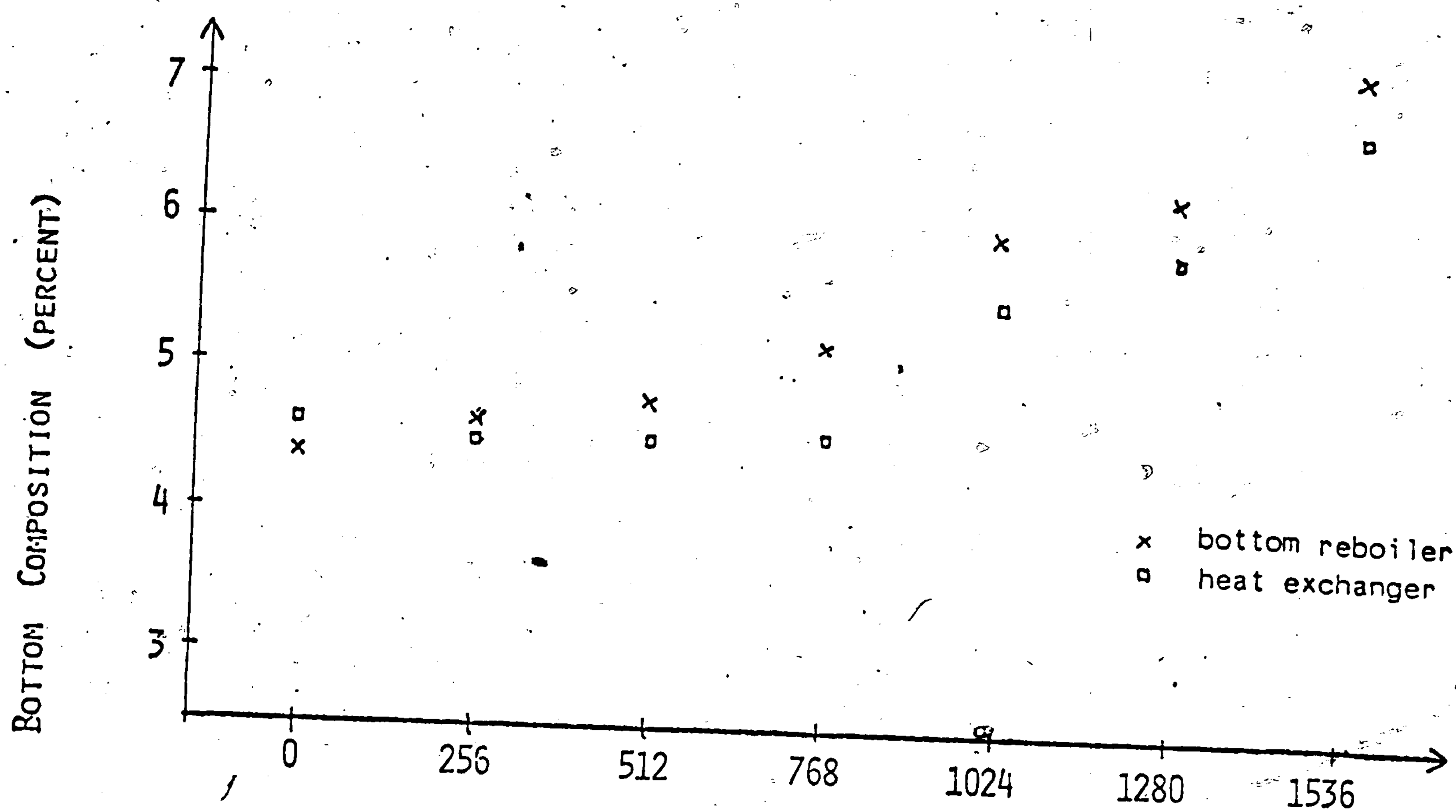


FIGURE B2 DYNAMIC RESPONSE OF BOTTOM COMPOSITION TO A STEAM FLOW RATE STEP CHANGE OF -5% WITH SAMPLES TAKEN AT REBOILER BOTTOM AND HEAT EXCHANGER OUTLET

APPENDIX C: EXPERIMENTAL CONTROL PROGRAM LISTING

```
// FOR RKT01
*IOCS(DISK)
*IOCS(PAPERTAPE)
C        STR CONTROL PROGRAM
C
C
C         ALGORITHMS AVAILABLE
C                1.STR-U METHOD
C                2.STR-DELU
C                3.PI CONTROL
C
C
C        DISTURBANCE USED
C              1.DETERMINISTIC
C              2.STOCHASTIC
C
C
      REAL LTS,MU,LDUMT(10),LDUMB(10),KPT,KIT,KPB,KIB
      DIMENSION PT(8,8),PB(8,8),X1(8),X2(8),XDUMT(10),XDUMB(10)
      DATA KPT,KIT,KPB,KIB/0.0,0.0,0.0,0.0/
      DATA MT,MB,KT,KB/2,2,1,0/
      DATA BOB,BOT,RT,RB,KSTR/0.0,0.0,0.0,0.0,0.0/
      DATA ACCMT,ACCMB,REST,RESB/0.0,0.0,0.0,0.0/
      DEFINE FILE 50(4,320,U,IFILE)
      CALL GETTY(LT)
      LUNW=LT
C  LT IN THIS PROGRAM IS LUN NUMBER BUT IN RKT02 IT IS FLOW
C
C     INITIALIZE INPUT AND OUTPUT DUMMY VECTORS
      DO 403 I=1,10
      LDUMT(I)=0.0
      LDUMB(I)=0.0
      XDUMB(I)=0.0
  403 XDUMT(I)=0.0
      IBTC=0
      IPRT=0
      KK=0
C
C     CONTROL LOOPS I.D.
C     FEED,TOP COMP.,REFLUX,BOTTOM COMP.,STEAM
      IFD=1537
      IXT=1540
      ILT=1538
      IXB=1543
      ISM=1539
C
C     ITEM KEYS 1=MEAS. 2=SET PT 3=OUTPUT
      I1=1
      I2=2
      I3=3
C
C     SET INITIAL COVARIANCE MATRIX FOR TOP: 100.
C     SET INITIAL COVARIANCE MATRIX FOR BOTTOM 10.
C     INITIAL PARAMETERS SET TO ZERO
      DO 20 J=1,8
      DO 20 I=1,8
      PB(I,J)=0.0
   20 PT(I,J)=0.0
      DO 21 I=1,8
      X1(I)=0.0
```

```
      X2(I)=0.0
      PT(I,I)=100.0
   21 PB(I,I)=10.0
C
      MU=1.0
      MVR=0
      KPAR=4
C
      WRITE(LT,11)
   11 FORMAT(' RUN TIME(IMTS),PRINT CNTRL(KPRT),BOTMS SAMPLING(ISMP)')
      CALL FFINP(LT,3,0,IMTS,0,KPTR,0,ISMP,IER)
C
C     GET CURRENT FLOWS SET PTS AND COMP. MEAS.
      IT=1
      CALL GTVLU(IXT,I1,XTS,IER,IT)
      CALL GTVLU(IXB,I1,XBS,IER,IT)
      IT=4
      CALL GTVLU(IFD,I2,FD,IER,IT)
      CALL GTVLU(ISM,I2,SFS,IER,IT)
      CALL GTVLU(ILT,I2,LTS,IER,IT)
      WRITE(LT,13) XTS,LTS,XBS,SFS,FD
   13 FORMAT(/,'T COMP='F6.3,' RFLX='F6.3,' B COMP='F6.3,' STM='F6.3,
     1' FEED='F6.3)
C
C     GET CURRENT OUTPUTS IN % FROM REFLUX AND STEAM
      CALL GTVLU(ILT,I3,LTS,IER,1)
      CALL GTVLU(ISM,I3,SFS,IER,1)
C
C     SET DISTURBANCE TYPE
      WRITE (LT,15)
   15 FORMAT('KTYP--T COMP SP,2 B COMP SP,3 FEED CHANGE AND MVR--0 STR
     1 1 MVR')
      CALL FFINP(LT,2,0,KTYP,0,MVR,IER)
C
C
C      JDEL=1 STR-U JDEL=2 STR-DELU
      WRITE(LT,27)
   27 FORMAT('ENTER JDEL 1=STR-U,2=STR-DELU   JDIS 1=DET. 2=STOC.')
      CALL FFINP(LT,2,0,JDEL,0,JDIS,IER)
C
C
C
C     GET TIME AND MAGNITUDE OF DISTURBANCE
      WRITE(LT,10)
   10 FORMAT('DISTURBANCE FROM(KDIS) AND VALUE(DIST)')
      CALL FFINP(LT,2,0,KDIS,1,DIST,IER)
      WRITE (LT,16)
   16 FORMAT('KCONT=1-T STR B OPN,2-T PI B OPN,3-T OPN B STR,4-T OPN B P
     1I')
      WRITE(LT,26)
   26 FORMAT('        5-T STR B PI,6-BOTH PI,7-BOTH STR,8-T PI B STR')
      CALL FFINP(LT,1,0,KCONT,IER)
      GO TO (224,113,225,114,223,111,222,112),KCONT
  113 WRITE(LT,22)
   22 FORMAT('SPECIFY KPT,KIT')
      CALL FFINP(LT,2,1,KPT,1,KIT,IER)
      WRITE(LT,36) KPT,KIT
      CALL MMANL(ILT)
      GO TO 333
  224 WRITE(LT,23)
```

```
 23    FORMAT('SPECIFY BOT,RT, IDENT STOP KSTR')
       CALL FFINP(LT,3,1,BOT,1,RT,0,KSTR,IER)
       WRITE(LT,99)
    99 FORMAT('SPECIFY X1(I), I=1,5')
       CALL FFINP(LT,5,1,X1(1),1,X1(2),1,X1(3),1,X1(4),1,X1(5),IER)
       WRITE(LT,39) MT,KT,BOT
       WRITE(LT,53)
    53 FORMAT('SPECIFY X1(I),I=6,8)
       CALL FFINP(LT,3,1,X1(6),1,X1(7),1,X1(8),IER)
       WRITE(LT,98) (X1(I),I=1,5)
       WRITE(LT,92)(X1(I),I=6,8)
    92 FORMAT(5X,'X1(6)=',F7.4,'X1(7)=',F7.4,'X1(8)=',F7.4)
       CALL MMANL(ILT)
       GO TO 333
 225   WRITE(LT,24)
 24    FORMAT('SPECIFY BOB,RB,KB,IDENT STOP KSTR')
       CALL FFINP(LT,4,1,BOB,1,RB,0,KB,0,KSTR,IER)
       WRITE(LT,97)
    97 FORMAT('SPECIFY X2(I),I=1,5')
       CALL FFINP(LT,5,1,X2(1),1,X2(2),1,X2(3),1,X2(4),1,X2(5),IER)
       WRITE(LT,54)
    54 FORMAT('SPECIFY X2(I),I=6,8)
       CALL FFINP(LT,3,1,X2(6),1,X2(7),1,X2(8),IER)
       WRITE(LT,41) MB,KB,BOB
       WRITE(LT,96)(X2(I),I=1,5)
       WRITE(LT,93)(X2(I),I=6,8)
    93 FORMAT(5X,'X2(6)=',F7.4,'X2(7)=',F7.4,'X2(8)=',F7.4)
       CALL MMANL(ISM)
       CALL MMANL(IXT)
       GO TO 333
 114   WRITE(LT,25)
 25    FORMAT('SPECIFY KPB,KIB')
       CALL FFINP(LT,2,1,KPB,1,KIB,IER)
       WRITE(LT,48) KPB,KIB
       CALL MMANL(ISM)
       CALL MMANL(IXT)
       GO TO 333
 111   WRITE(LT,17)
 17    FORMAT('SPECIFY KPT,KIT,KPB,KIB')
       CALL FFINP(LT,4,1,KPT,1,KIT,1,KPB,1,KIB,IER)
       WRITE(LUNW,40)
       WRITE(LUNW,42) KPT,KIT,KPB,KIB
       CALL MMANL(ILT)
       CALL MMANL(ISM)
       GO TO 333
 222   CONTINUE
       WRITE (LT,14)
 14    FORMAT('BETA O (BOT,BOB), NOISE COVAR(RT,RB),KB,IDENT STP(KSTR)')
       CALL FFINP(LT,6,1,BOT,1,BOB,1,RT,1,RB,0,KB,0,KSTR,IER)
       WRITE(LT,99)
       CALL FFINP(LT,5,1,X1(1),1,X1(2),1,X1(3),1,X1(4),1,X1(5),IER)
       WRITE(LT,53)
       CALL FFINP(LT,3,1,X1(6),1,X1(7),1,X1(8),IER)
       WRITE(LT,97)
       CALL FFINP(LT,5,1,X2(1),1,X2(2),1,X2(3),1,X2(4),1,X2(5),IER)
       WRITE(LT,54)
       CALL FFINP(LT,3,1,X2(6),1,X2(7),1,X2(8),IER)
       WRITE (LUNW,35)
       WRITE(LT,37) MT,KT,MB,KB,BOT,BOB
       WRITE(LT,98) (X1(I),I=1,5)
```

```
      WRITE(LT,92)(X1(I),I=6,8)
      WRITE(LT,96)(X2(I),I=1,5)
      WRITE(LT,93)(X2(I),I=6,8)
      CALL MMANL(ILT)
      CALL MMANL(ISM)
      GO TO 333
112   WRITE (LT,18)
18    FORMAT('SPECIFY KPT,KIT,AND BOB,RB,KB')
      CALL FFINP(LT,5,1,KPT,1,KIT,1,BOB,1,RB,0,KB,IER)
      WRITE(LT,97)
      CALL FFINP(LT,5,1,X2(1),1,X2(2),1,X2(3),1,X2(4),1,X2(5),IER)
      WRITE(LT,54)
      CALL FFINP(LT,3,1,X2(6),1,X2(7),1,X2(8),IER)
      WRITE (LT,44)
      WRITE(LT,45) KPT,KIT,MB,KB,BOB
      WRITE(LT,96)(X2(I),I=1,5)
      WRITE(LT,93)(X2(I),I=6,8)
      CALL MMANL(ILT)
      CALL MMANL(ISM)
      GO TO 333
223   WRITE(LT,19)
19    FORMAT('SPECIFY BOT,RT, AND KPB,KIB')
      CALL FFINP(LT,4,1,BOT,1,RT,1,KPB,1,KIB,IER)
      WRITE(LT,99)
      CALL FFINP(LT,5,1,X1(1),1,X1(2),1,X1(3),1,X1(4),1,X1(5),IER)
      WRITE(LT,53)
      CALL FFINP(LT,3,1,X1(6),1,X1(7),1,X1(8),IER)
      WRITE (LT,46)
      WRITE(LT,47) MT,KT,BOT,KPB,KIB
      WRITE(LT,98) (X1(I),I=1,5)
      WRITE(LT,92)(X1(I),I=6,8)
      CALL MMANL(ILT)
      CALL MMANL(ISM)
333   CONTINUE
      CALL TIME(IHR,IMT,ISEC)
      WRITE(LT,38) IHR,IMT,ISEC
      ICYC=640
      REFT=XTS
      REFB=XBS
      WRITE(50'1) KPTR,REFT,REFB,IPRT,ACCMT,ACCMB,MU,KPAR,KSTR,MVR,KK
     1,JDEL,JDIS,MT,KT,MB,KB,RT,RB,BOB,BOT,FD
      WRITE(50'2) LDUMT,LDUMB,XDUMT,XDUMB,X1,X2,LTS,SFS
      WRITE(50'3) PT,PB
      WRITE(50'4) KCONT,IMTS,LUNW,KDIS,DIST,KTYP,IBTC,ISMP,XTS,XBS,
     1LTS,SFS,KPT,KPB,KIT,KIB,REST,RESB
      ILEV=8
      IBIT=2
      ITMR=7
C  TO PUT  THE TIMER IN RESTART TABLE  **********        ******
C
      CALL SCPTB(1,ITMR,ILEV,IBIT,1,ICYC,IE)
      CALL CYCLE(ILEV,IBIT,ITMR,ICYC)
35    FORMAT (/,10X,'**STR  ON BOTH COMPOSITIONS**')
36    FORMAT(/,10X,'**PI ON TOP AND BOTTOM COMP OPEN**',/,5X,'KPT=',F8.4,
     1' KIT=',F8.4)
37    FORMAT(/,5X,' MT=',I2,' KT=',I2,' MB=',I2,' KB='I2,' BOT=',F8.4,
     1' BOB='F8.4)
38    FORMAT(/,5X,'CONTROL ACTION INITIATED AT',3I3,/)
39    FORMAT(/,10X,'**STR ON TOP COPMP., BOTM OPEN LOOP**',//,5X,'MT=',
     1I2,' KT=',I2,'  BOT=',F8.4)
```

```
40   FORMAT(/,10X,' **PI  ON BOTH COMPOSITIONS**')
41   FORMAT(/,10X,'**TOP OPEN LOOP, STR ON BOTM COMP.**',//,5X,'MB=',I2
    1,' KB=',I2,' BOB=',F8.4)
42   FORMAT(/,5X,'KPT='F8.4,' KIT=',F8.4,' KPB=',F8.4,' KIB=',F8.4)
44   FORMAT(/,10X,'**PI ON TOP AND STR ON BOT COMPS.**')
45   FORMAT(/,5X,'KPT='F8.4,' KIT=',F8.4,' MB=',I2,' KB=',I2,
    1' BOB=',F8.4)
46   FORMAT(/,10X,'**STR ON TOP AND PI ON BOT COMPS.**')
47   FORMAT(/,5X,'MT=',I2,'KT=',I2,'BOT=',F8.4,'KPB=',F8.4,'KIB=',F8.4)
48   FORMAT(/,10X,'**TOP OPEN LOOP, PI ON BOTM COMP.**',//,5X,'KPB='
    1,F8.4,' KIB=',F8.4)
98   FORMAT(/,5X,'X1(1)=',F7.4,'X1(2)=',F7.4,'X1(3)=',F7.4,'X1(4)=',
    1F7.4,'X1(5)=',F7.4)
96   FORMAT(/,5X,'X2(1)=',F7.4,'X2(2)=',F7.4,'X2(3)=',F7.4,'X2(4)=',
    1F7.4,'X2(5)=',F7.4)
     CALL EXIT
     END
*DELET      A          RKTO1 *****
*STORECII  A        1 RKTO1 RKTO1
*FILES(50,DISTO,O)
*CCEND
// END

// FOR RKTO2
*IOCS(DISK)
*IOCS(PAPERTAPE)
     REAL LTS,LT,KPT,KIT,KPB,KIB,LDUMT(10),LDUMB(10),MU,LTN
     DIMENSION PT(8,8),PB(8,8),X1(8),X2(8),XDUMT(10),XDUMB(10)
     DEFINE FILE 50(4,320,U,IFILE)
     DEFINE FILE 51(240,46,U,JFILE)
     IFD=1537
     ILT=1538
     ISM=1539
     IXT=1540
     IXB=1543
     IBM=1544
     I1=1
     I2=2
     I3=3
     READ (50'1) KPTR,REFT,REFB,IPRT,ACCMT,ACCMB,MU,KPAR,KSTR,MVR,KK
    1,JDEL,JDIS,MT,KT,MB,KB,RT,RB,BOB,BOT,FDS
     READ (50'2) LDUMT,LDUMB,XDUMT,XDUMB,X1,X2,LT,SF
     READ (50'3) PT,PB
     READ (50'4) KCONT,IMTS,LUNW,KDIS,DIST,KTYP,IBTC,ISMP,XTS,XBS,
    1LTS,SFS,KPT,KPB,KIT,KIB,REST,RESB
     KK=KK+1
     IPRT=IPRT+1
     IBTC=IBTC+1
     IF(JDIS-1)23,23,25
23   IF(KK-KDIS) 333,14,333
14   CONTINUE
     GO TO (44,22,55), KTYP
44   XTS=DIST
     GO TO 333
22   XBS=DIST
     GO TO 333
55   CALL PTVLU(IFD,I2,DIST,IER)
333  CONTINUE
     GO TO 17
25   IF(KK-30)17,27,27
```

```
   27 CALL GAUSS(333,DIST,18.,SDIST)
      CALL PTVLU(IFD,I2,SDIST,IER)
   17 CONTINUE
C  GET THE DATA FROM  PROCESS MODEL
C
      IT=1
      CALL GTVLU(IXT,I1,XT,IER,IT)
      CALL GTVLU(IXB,I1,XB,IER,IT)
      CALL GTVLU(IBM,I1, BM,IER,IT)
      IT=4
      CALL  GTVLU(IFD,I1,FD,IER,IT)
      IF(KK-IMTS) 11,12,12
   11 CONTINUE
      CALL GTVLU(ILT,I1,VLT,IER,IT)
      CALL GTVLU(ISM,I1,SM,IER,IT)
      GO TO (222,111,33,33,222,111,222,111), KCONT
  111 CALL GTVLU(ILT,I3,LT,IER,1)
      CALL PID(KPT,KIT,REST,REFT,XTS,LTS,XT,CLT,ACCMT)
      CALL PTVLU(ILT,I3,CLT,IER)
      GO TO 33
  222 CONTINUE
      OLT=LT
      CALL GTVLU(ILT,I3,LT,IER,1)
      XTN=(XT-XTS)/REFT
      IF(JDEL-1)69,69,70
   70 LTN=(LT-OLT)/LTS
      OLT=LT
      GO TO 71
   70 LTN=(LT-LTS)/LTS
   71 CALL STR(MT,KT,BOT,RT,PT,X1,LDUMT,XDUMT,XTN,LTS,LTN,CLT,ACCMT,OLT)
      CALL PTVLU(ILT,I3,CLT,IER)
   33 CONTINUE
      IF(IBTC-ISMP) 13,15,13
   15 IBTC=0
      GO TO (13,13,223,112,112,112,223,223), KCONT
  112 CALL GTVLU(ISM,I3,SF,IER,1)
      CALL PID(KPB,KIB,RESB,REFB,XBS,SFS,XB,CSF,ACCMB)
      CALL PTVLU(ISM,I3,CSF,IER)
      GO TO 16
  223 CONTINUE
      OSF=SF
      CALL GTVLU(ISM,I3,SF,IER,1)
      XBN=(BM-XBS)/REFB
      IF (JDEL-1)68,68,86
   86 SFN=(SF-OSF)/SFS
      OSF=SF
      GO TO 67
   68 SFN=(SF-SFS)/SFS
   67 CALL STR(MB,KB,BOB,RB,PB,X2,LDUMB,XDUMB,XBN,SFS,SFN,CSF,ACCMB,OSF)
      CALL PTVLU(ISM,I3,CSF,IER)
   16 CONTINUE
      GO TO 13
   12 CALL CANCL(07)
C  TO REMOVE THE TIMER FROM RESTART TABLE    ****     ******
C
      CALL SCPTB(0,7,8,2,1,600,IE)
      CALL PTVLU(IFD,I2,FDS,IER)
      CALL PTVLU(ILT,I3,LTS,IER)
      CALL PTVLU(ISM,I3,SFS,IER)
      CALL TIME(IHR,IMT,ISEC)
```

```
          WRITE (LUNW,38) IMT,ISEC,XT,LT,XB,SF,FD,ACCMT,ACCMB
          WRITE (LUNW,66) IHR,IMT,ISEC
   66     FORMAT(/,'JOB COMPLETED AT',3I3)
   C      PUT DATA ACCUMULATION LOOPS NON-OPERABLE
          CALL NONOP(1584)
          CALL NONOP(1585)
          CALL NONOP(1586)
          CALL NONOP(1587)
          CALL NONOP(1588)
          CALL NONOP(1589)
          CALL NONOP(1590)
          CALL NONOP(1591)
          CALL NONOP(1592)
          CALL NONOP(1667)
          GO TO 999
   13     CONTINUE
          CALL TIME(IHR,IMT,ISEC)
          IF(IPRT-KPTR) 412,413,412
  413     IPRT=0
          WRITE (LUNW,38) IMT,ISEC,XT,VLT,XB,SM,FD,ACCMT,ACCMB
  412     CONTINUE
   38     FORMAT(/,2I3,'XT=',F6.3,'LT=',F6.3,'XB=',F6.3,'SF=',F6.3,
         1'FD=',F6.3,2F9.7)
          WRITE(50'1) KPTR,REFT,REFB,IPRT,ACCMT,ACCMB,MU,KPAR,KSTR,MVR,KK
         1,JDEL,JDIS,MT,KT,MB,KB,RT,RB,BOB,BOT,FDS
          WRITE(50'2) LDUMT,LDUMB,XDUMT,XDUMB,X1,X2,LT,SF
          WRITE(50'3) PT,PB
          WRITE(50'4) KCONT,IMTS,LUNW,KDIS,JIST,KTYP,IBTC,ISMP,XTS,XBS,
         1LTS,SFS,KPT,KPB,KIT,KIB,REST,RESB
  999     CONTINUE
          WRITE(51'KK)IMT,XT,VLT,XB,BM,SM,FD,(X1(I),I=1,8),(X2(I),I=1,8)
          CALL EXIT
          END
// FOR PID
          SUBROUTINE PID(KP,KI,RES,REF,XSP,USP,XTV,UCT,ACCUM)
          REAL KP,KI
          UMIN=-0.5
          UMAX=0.5
          ER=(XTV-XSP)/REF
          RES=RES+ER
          SP=KP*ER+KI*RES
          ACCUM=ACCUM+ER*ER
          IF(SP-UMAX) 1,1,2
    2     SP=UMAX
          GO TO 3
    1     IF(SP-UMIN) 4,4,3
    4     SP=UMIN
    3     CONTINUE
          UCT=(1.0+SP)*USP
          RETURN
          END
// FOR STR
          SUBROUTINE STR(M,K,BO,R2,P,X,LDUM,XDUM,XTN,USP,LTN,UCT,ACCUM,OUCT)
          REAL LTN,MU,LDUM(10)
          DIMENSION P(8,8),PHI(8,8),Y(8),X(8),PY(8),AKY(8,8),PPH(8,8),
         1         ERK(8),XDUM(8 ),AK(8)
          READ (50'1) KPTR,REFT,REFB,IPRT,ACCMT,ACCMB,MU,KPAR,KSTR,MVR,KK
         1,JDEL
          DO 33 I=1,8
          DO 33 J=1,8
```

```
      PHI(I,J)=0.0
   33 PHI(I,I)=1.0
      UMAX=1.5*USP
      UMIN=0.5*USP
      L=M+K
      N=L+M
      K2=K+2
      K4=K+4
      DO 204 I=1,K2
      KIN=K2-I+2
  204 XDUM(KIN)=XDUM(KIN-1)
      DO 205 I=1,K4
      KIN=K4-I+2
  205 LDUM(KIN)=LDUM(KIN-1)
      DO 206 I=1,M
      IK1=I+K+1
  206 Y(I)=XDUM(IK1)
      DO 207 I=1,L
      IM=I+M
      IK2=I+K+1
  207 Y(IM)=LDUM(IK2)
      XDUM(1)=-XTN
      LDUM(1)=LTN*BO
      IF(MVR-1) 50,51,50
   50 IF(KK-KSTR) 52,51,51
   52 CONTINUE
C   COMPUTE THE  GAIN VALUE
C
      DO 22 I=1,N
      PY(I)=0.0
      DO 22 J=1,N
   22 PY(I)=PY(I)+P(I,J)*Y(J)
      YPY=0.0
      DO 23 I=1,N
      AK(I)=PHI(I,I)*PY(I)
   23 YPY=YPY+Y(I)*PY(I)
      YPY=YPY+R2
      DO 25 I=1,N
   25 AK(I)=AK(I)/YPY
C   COMPUTE  COVARIANCE MATRIX
C
      DO 26 I=1,N
      DO 26 J=1,N
   26 AKY(I,J)=PHI(I,J)-AK(I)*Y(J)
      DO 27 I=1,N
      DO 27 J=1,N
      PPH(I,J)=0.0
      DO 27 IJ=1,N
   27 PPH(I,J)=PPH(I,J)+P(I,IJ)*PHI(IJ,J)
      DO 24 I=1,N
      DO 24 J=1,N
      P(I,J)=0.0
      DO 28 IJ=1,N
   28 P(I,J)=P(I,J)+AKY(I,IJ)*PPH(IJ,J)
   24 P(I,J)=P(I,J)/MU
C UPDATE  PARAMETER  VECTOR
C
      YX=0.0
      DO 29 I=1,N
      X(I)=PHI(I,I)*X(I)
```

```
29      YX=YX+Y(I)*X(I)
C  COMPUTE  ERROR
C   Y(4) IS EQUAL TO BO*U(T-2)
C
        ERR=XTN-LDUM(K+1)-YX
        DO 201 I=1,N
        ERK(I)=AK(I)*ERR
201     X(I)=X(I)+ERK(I)
C  COMPUTE  ACCUMALATED  ERROR
C
        ACCUM=ACCUM+XDUM(1)**2
C   COMPUTE  CONTROL  VECTOR
C
51      CONTINUE
        YX=0.0
        DO 202 I=1,M
202     YX=YX+XDUM(I)*X(I)
        YY=0.0
        DO 203 I=1,L
        IM=I+M
    203 YY=YY+LDUM(I)*X(IM)
        LTN=(-YY-YX)/BO
        IF(KK-KPAR) 410,53,53
53      CONTINUE
        IF(JDEL-1)99,99,98
     98 UCT=OUCT+LTN*USP
        GO TO 97
     99 UCT=(1.+LTN)*USP
C  CONSTRAINTS  ON THE  CONTROLLER
C
 97     IF(UCT    -UMIN)44,44,45
45      IF(UCT    -UMAX)110,47,47
47      UCT    =UMAX
        GO TO 110
44      UCT    =UMIN
        GO TO 110
410     UCT=USP
110     RETURN
        END
        CALL GAUSS(IX,SA,AM,V)
        AA=0.0
        DO 7 I=1,12
        IY=IX*899
        IF (IY) 5,6,6
      5 IY=IY+32767+1
      6 YFL=IY
        YFL=YFL/32767.0
        IX=IY
      7 AA=AA+YFL
        V=(AA-6.0)*SA+AM
        RETURN
        END
*DELET     A          RKTO2 Y****
*STORECII  A        1 RKTO2 RKTO2        0802   020109
*FILES(50,DISTO,O)
*FILES(51,DIST1,O)
*CCEND
// END

END OF FILE
```

APPENDIX D: TOP PRODUCT CONTROLLER TUNING BY THE ZIEGLER-NICHOLS CONTINUOUS CYCLING METHOD

The distillation column was operated at the following conditions:

Feed Flow	= 18.00	g/s
Reflux Flow	= 13.70	g/s
Steam Flow	= 14.70	g/s
Top Composition	= 96.00	%
Bottom Composition	= 0.00	%
Feed Composition	= 50.00	%

Various proportional gains were tested, under closed loop control, for a 1% step increase in set point to obtain a sustained system oscillation. The results are summarized in Table D1.

Table D1 Results of Top Product Composition Ziegler-Nichols Tuning Runs

Run	Proportional Gain	System Performance
D1	-18.0	offset
D2	-12.0	offset
D3	-12.0	offset
D4	-20.0	offset
D5	-30.0	offset
D6	-50.0	offset
D7	-75.0	damped oscillation
D8	-100.0	sustained oscillation

The control and manipulated variable responses for Run D8 are plotted in Figures D1 and D2. On the basis of the Ziegler-Nichols recommended settings, the controller constants for proportional plus integral control are:

$$K_P = 0.45 K_{P\,MAX} = -45.0$$

$$T_R = P_u/1.2 = 6.2 \text{ minutes}$$

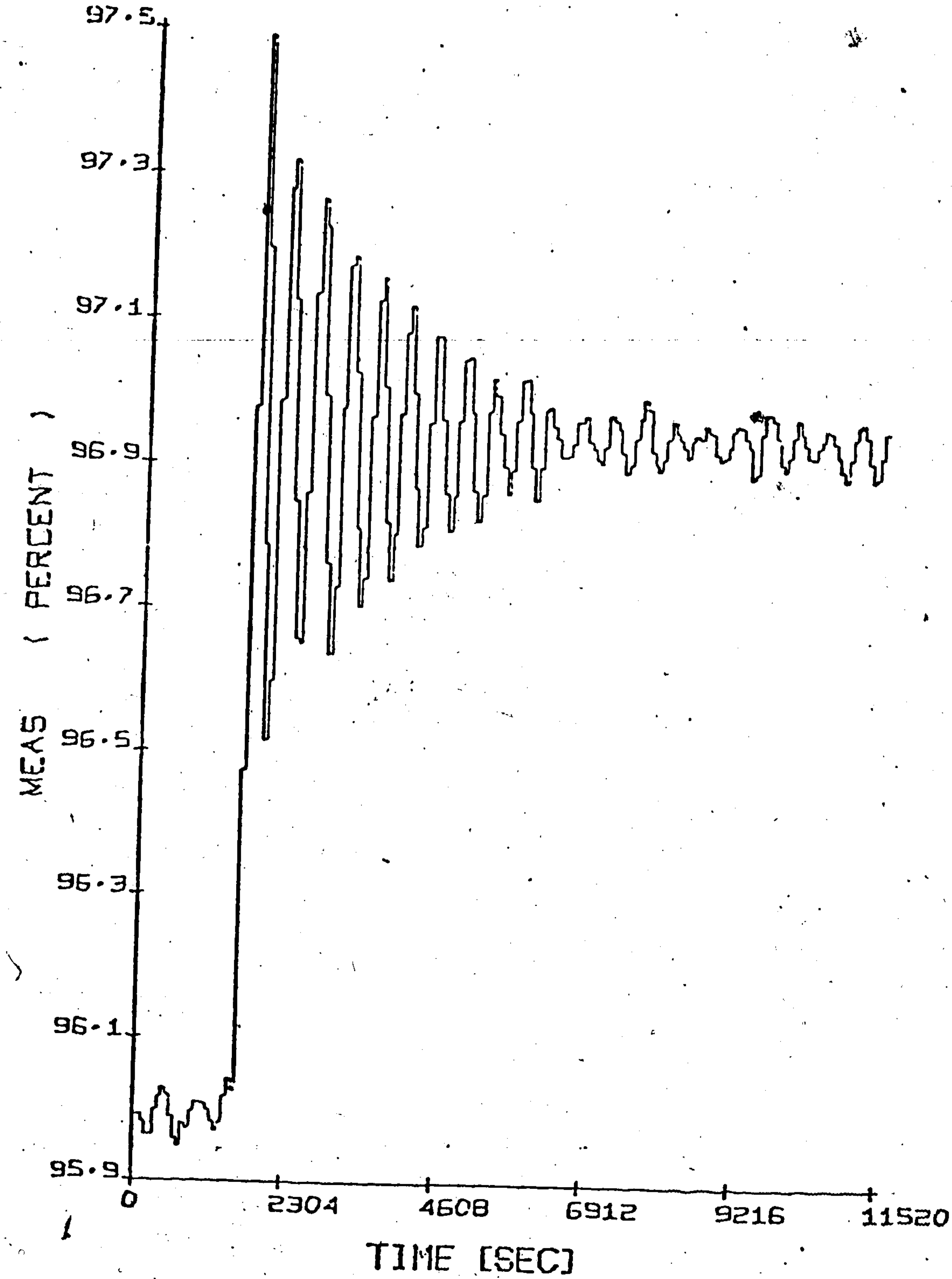


FIGURE D1 TOP PRODUCT COMPOSITION RESPONSE TO A +1% STEP CHANGE IN SET POINT USING A PROPORTIONAL CONTROLLER WITH $K_P = -100.0$

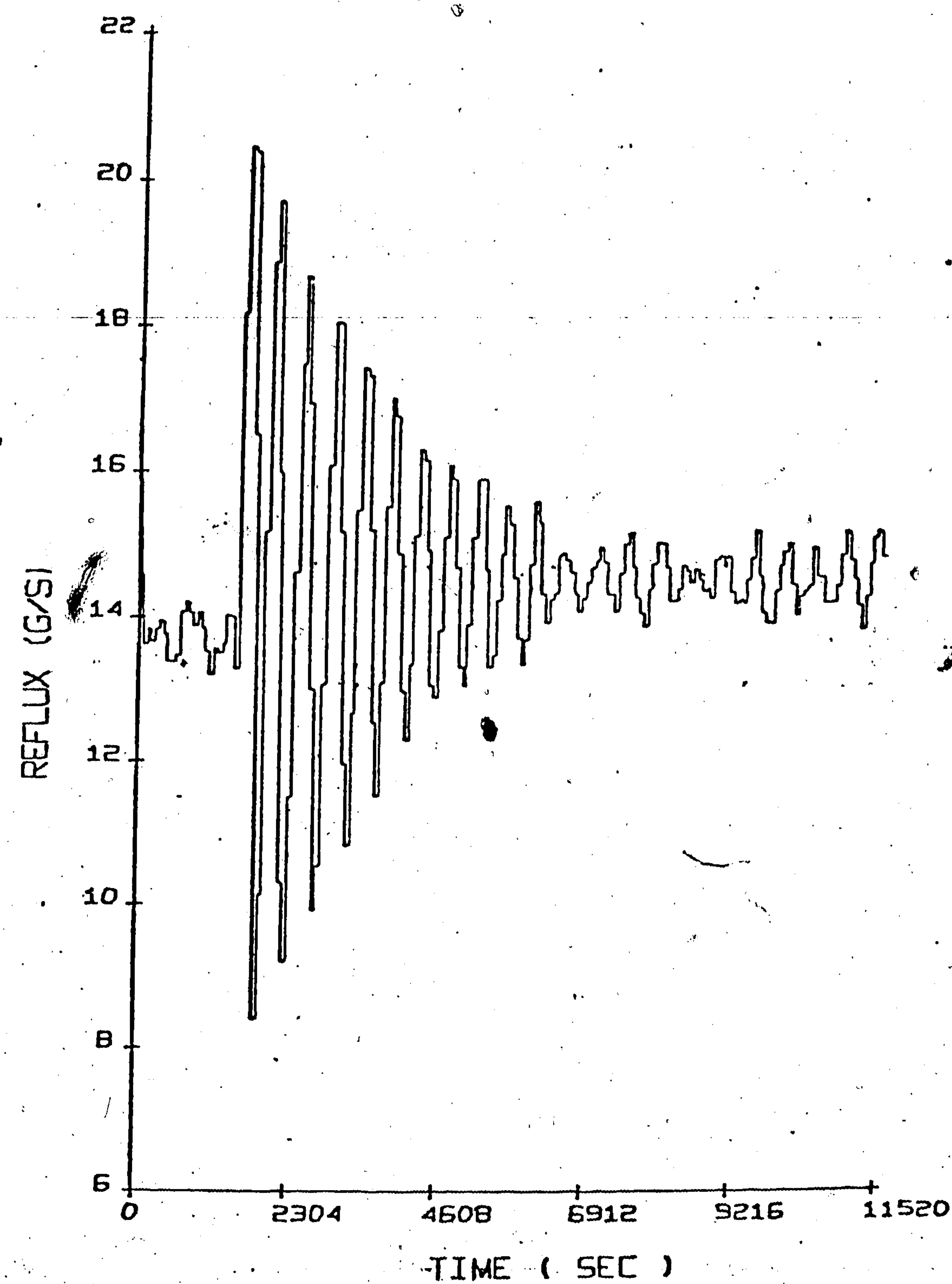


FIGURE D2 REFLUX FLOW RATE RESPONSE FOR PROPORTIONAL CONTROLLER WITH K_p = -100.0 OF TOP PRODUCT COMPOSITION CONTROL FOR A +1% STEP CHANGE IN SET POINT

APPENDIX E: SIMULATION PROGRAM LISTING

```
CCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCC
C                                                                          C
C          UNIVERSITY OF ALBERTA,  CHEMICAL ENGINEERING DEPARTMENT         C
C          8  TRAYS DISTILLATION   COLUMN CONTROL SIMULATION               C
C                                       BY RON TONG                        C
C                                                                          C
CCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCCC
C
C---------------------------------------------------------------------------
C      THIS PROGRAM USES AN EULER METHOD OF INTEGRATION
C      TO PREDICT THE TRANSIENT RESPONSE OF A 10 STAGE
C      DISTILLATION COLOUMN. THE ASSUMPTIONS ARE
C            1. CONSTANT WEIGHT HOLDUP
C            2. UNIFORM LIQUID COMPOSITION-PERFECT MIXING
C            3. NEGLIGIBLE VAPOUR HOLDUP
C      THE LIQUID AND VAPOUR FLOWS ARE CALCULATED IN SUBROUTINE
C      VFLOW, USING THE TRAY MASS BALANCES AND ENTHALPY BALANCES.
C      THE VAPOUR COMPOSITIONS ARE CALCULATED FROM THE LIQUID
C      COMPOSITIONS IN SUBROUTINE COMP. ENTHALPIES FOR LIQUID AND
C      VAPOUR STREAMS ARE CALCULATED IN SUBROUTINE ENTHALPY. THE
C      MAINLINE PROGRAM PERFORMS THE ACTUAL INTEGRATION AND
C      THE SUBROUTINES IN THE CORRECT ORDER.
C---------------------------------------------------------------------------
C      CONTROL STRATEGIES AVAILABLE
C         1. PROPORTIONAL INTEGRAL
C         2. SELF TUNING REGULATOR
C---------------------------------------------------------------------------
C
C---------------------------------------------------------------------------
C      VARIABLE      FUNCTION
C      --------      --------
C           AIE      SUM OF ERROR SQUARE TERMS IN TOP COMPOSITION
C          AIEB      SUM OF ERROR SQUARE TERMS IN BOTTOM COMPOSITION
C          AHLO      INTERCEPT IN STEAM ENTHALPHY RELATION
C          BHLO      SLOPE     IN STEAM ENTHALPHY RELATION
C           BOB      SCALING FACTOR USED IN STR BOTTOM CONTROL
C           BOT      SCALING FACTOR USED IN STR TOP CONTROL
C             D      DISTURBANCE VECTOR AT 3 DIFFERENT TIME  (G/S)
C            DA      FIRST HALF OF DATE TO BE PRINTED
C            DT      INTEGRATION TIME STEP  (SECOND)
C           DXB      BOTTOM COMPOSITION DELAYED BY GC MEASUREMENT TIME LAG
C            FL      INITIAL FEED FLOW  (G/S)
C            HF      FEED ENTHALPHY  (KJ/KG)
C           HSI      STEAM ENTHALPHY  (KJ/KG)
C            IB      BOTTOM GC MEASUREMENT TIME LAG LOOP COUNTER
C           IBD      BOTTOM G C MEASUREMENT TIME DELAY
C                    IN NUMBER OF SAMPLING INTERVALS (S.I.)
C         ICHNG      VECTOR OF TIME AT WHICH DISTURBANCE OCCURS (S.I.)
C            IN      INDEX TO CHECK PRINT TIME FOR COLUMN DYNAMICS
C            IR      REFLUX FLOW NOISE GAUSSIAN CONSTANT USED
C            IV      FEED   FLOW NOISE GAUSSIAN CONSTANT USED
C          ISPC      TIME AT WHICH SET POINT CHANGES ARE TO BE MADE (S.I.)
C           IST      INDEX TO CHECK THE SAMPLING TIME
C         ITEST      TYPE OF DISTURBANCE
C         JCONT      CONTROL CONFIGURATION KEY
C                    JCONT=1 TOP CONTROL JCONT=2 BOTTOM CONTROL
C                    JCONT=3 TOP AND BOTTOM CONTROL
C            JM      PRINT CONTROL KEY  (S.I.)
C         KCONT      KCONT=1 OPEN LOOP    KCONT=2 CLOSE LOOP
C            KB      PROCESS TIME DELAY FOR BOTTOM CONTROL  (S.I.)
```

```
C          KT      PROCESS TIME DELAY FOR TOP    CONTROL  (S.I.)
C          KPB     PROPORTIONAL GAIN CONSTANT USED FOR BOTTOM PI CONTROL
C          KPT     PROPORTIONAL GAIN CONSTANT USED FOR TOP    PI CONTROL
C          KPAR    TIME WHEN STR IS EFFECTED  (S.I.)
C          KPIT    PRIOR TO THIS TIME,TOP SYSTEM CONTROL IS ON PI
C                  BEYOND STR IS EFFECTED  (S.I.)
C          KPIB    PRIOR TO THIS TIME,BOTTOM SYSTEM CONTROL IS ON PI
C                  BEYOND STR IS EFFECTED  (S.I.)
C          KSTR    TIME WHEN STR PARAMETERS ESTIMATION IS STOPPED (S.I.)
C          KST     PROGRAM SAMPLING TIME COUNTER
C          IKT     IKT=1 FOR INITIATION, IKT=3 AFTER,
C                  FOR USE IN SUBROUTINE COMP ENTH VFLOW
C          LB      NUMBER OF PARAMETERS IN MANIPULATED VARIABLE
C                  FOR BOTTOM STR  (BETA)
C          LT      NUMBER OF PARAMETERS IN MANIPULATED VARIABLE
C                  FOR TOP STR  (BETA)
C          LF      FEED STAGE
C          LDUMB   DUMMY VECTOR TO STORE PREVIOUS DIFFERENCE IN
C                  MANIPULATED VARIABLE FOR BOTTOM STR
C          LDUMT   DUMMY VECTOR TO STORE PREVIOUS DIFFERENCE IN
C                  MANIPULATED VARIABLE FOR TOP    STR
C          LTS     INITIAL REFLUX FLOW RATE  (G/S)
C          MB      NUMBER OF PARAMETERS IN OUTPUT VARIABLE
C                  FOR BOTTOM STR  (ALPHA)
C          MT      NUMBER OF PARAMETERS IN OUTPUT VARIABLE
C                  FOR TOP    STR  (ALPHA)
C          MU      EXPONENTIAL FORGETTING FACTOR FOR STR ALGORITHM
C          NI      SIMULATION RUN TIME  (MINUTE)
C          NTT     NUMBER OF TRAYS IN THE COLUMN + REBOILER
C          NUMRUN  SIMULATION RUN NUMBER
C          OLT     TOP MANIPULATED VARIABLE REFLUX: OLT=U(T-2) IN STR
C          OSF     BOTTOM MANIPULATED VARIABLE STEAM: OSF=U(T-2) IN STR
C          PB      COVARIANCE MATRIX FOR STR ALGORITHM  BOTTOM CONTROL
C          PT      COVARIANCE MATRIX FOR STR ALGORITHM  TOP    CONTROL
C          QP      HEAT LOAD
C          RB      NOISE COVARIANCE FOR STR ALGORITHM  BOTTOM CONTROL
C          RT      NOISE COVARIANCE FOR STR ALGORITHM  TOP    CONTROL
C          REFB    REFERENCE BOTTOM COMPOSITION USED FOR CALCULATING
C                  NORMALIZED BOTTOM COMPOSITION DEVIATION ERROR
C          REFT    REFERENCE TOP COMPOSITION USED FOT CALCULATING
C                  NORMALIZED TOP COMPOSITION DEVIATION ERROR
C          RESETB  RESET TIME FOR BOTTOM PI CONTROL  (MIN)
C          RESETT  RESET TIME FOR TOP    PI CONTROL  (MIN)
C          SD      STANDARD DEVIATION OF FEED FLOW NOISE  (G/S)
C          SDR     STANDARD DEVIATION OF REFLUX FLOW NOISE  (G/S)
C          SF      INITIAL STEAM FLOW  (G/S)
C          SFS     INITIAL STEADY STATE STEAM FLOW  (G/S)
C          TE      SECOND HALF OF DATE TO BE PRINTED
C          TR      REBOILER TEMPERATURE  (C)
C          TS      STEAM CHEST TEMPERATURE  (C)
C          UA      HEAT TRANSFER COEFFICIENT  (J/S)
C          UMIN    MINIMUM CONTROLLER CONSTRAINT
C          UMAX    MAXIMUM CONTROLLER CONSTRAINT
C          XBS     INITIAL STEADY STATE BOTTOM COMPOSITION  (WT FN CH3OH)
C          XBSP    NEW DESIRED STEADY STATE BOTTOM COMPOSITION. (WT FN CH3OH)
C          XFT     FEED COMPOSITION  (WT FN CH3OH)
C          XTS     INITIAL STEADY STATE TOP COMPOSITION  (WT FN CH3OH)
C          XTSP    NEW DESIRED STEADY STATE TOP COMPOSITION  (WT FN CH3OH)
C          XDUMB   DUMMY VECTOR TO STORE PREVIOUS OUTPUT VARIABLE
C                  IN BOTTOM STR
```

```
C          XDUMT      DUMMY VECTOR TO STORE PREVIOUS OUTPUT VARIABLE
C                     IN TOP    STR
C-----------------------------------------------------------------------
C-----------------------------------------------------------------------
C
C
C
C
      REAL LT(20),LTS,KPT,KPB,MU,LDUMT(10),LDUMB(10),LTN,XDUMB(10)
      DIMENSION ISTRN(3),PT(7,7),PB(7,7),X1(7),XB(7),XDUMT(10)
      DIMENSION D(4),ICHNG(4),RE(3),DF(4)
      COMMON XT(20),YT(20),LT,VT(20),HLT(20),HVT(20),HL(20),E(20),XF
     1(4),KK,IPROD,T,MTT,TKT,I
      COMMON WT(20),SF,HSI,AHLO,BHLO,UA,TR,F(4),QLP(20),DT,LF,HF,QP,TS
      COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
      DATA ISTRN/'RE  ','FEED','FC  '/
      DATA REST,RESB,X1,XB/0.0,0.0,7*0.0,7*0.0/
      DATA PT,PB,LDUMT,LDUMB,XDUMT,XDUMB/49*0.0,49*0.0,10*0.,10*0.0,
     110*0.,10*0.0/
      DATA KTOP,KBOT/1,1/
C
C
C
  901 FORMAT(F5.2,2I5)
  900 FORMAT(10F8.2)
  902 FORMAT(10I5)
  903 FORMAT(10F8.4)
  904 FORMAT(A4)
  905 FORMAT(8F10.5)
  906 FORMAT(F8.2,8F8.4)
  909 FORMAT(10F8.5)
C
C
C
C
C     THE FOLLOWING READ AND WRITE STATEMENTS CORRESPOND TO THE DATA
C     INPUT REQUIRED FOR THE SIMULATION MODEL
C
C
C  INITIAL VALUES
      READ(5,999)NUMRUN
  999 FORMAT(I5)
      WRITE(6,998) NUMRUN
  998 FORMAT('****************** SIMULATION RUN NUMBER=',I5,' **********
     1*****')
      READ(5,997) DA,TE
  997 FORMAT(2A4)
      WRITE(6,996) DA,TE
  996 FORMAT(/,'*',30X,'DATE PERFORMED BY R TONG',7X,'*',//,'*',45X,
     12A4,10X,'*')
C
C     INPUT DATA USED FOR NL COLUMN MODEL
C
      READ(5,904)ITEST
      READ(5,902)ICHNG
      WRITE(6,1121)
 1121 FORMAT('************************************************************
     1********')
      WRITE(6,995)
  995 FORMAT(//////,2X,'INPUT DATA USED FOR COLUMN MODEL'/)
```

```
C
C
      WRITE(6,986)
      WRITE(6,994)
  994 FORMAT(2X,'ITEST')
      WRITE(6,904)ITEST
      WRITE (6,986)
      WRITE(6,993)
  993 FORMAT(2X,'ICHNG')
      WRITE(6,902)ICHNG
      WRITE(6,986)
      READ(5,903)D
      WRITE(6,992)
  992 FORMAT(2X,'D')
      WRITE(6,903)D
      WRITE(6,986)
      READ(5,902)LF,IR
      WRITE(6,991)
  991 FORMAT(2X,'LF',2X,'IR')
      WRITE(6,902)LF,IR
      WRITE(6,986)
      READ(5,901)DT,NI,NTT
      WRITE(6,990)
  990 FORMAT(2X,'DT',4X,'NI',4X,'NTT')
      WRITE(6,901)DT,NI,NTT
      WRITE(6,986)
      MTT=NTT+1
      READ(5,900) LT(MTT),SD,SDR
      WRITE(6,989)
  989 FORMAT(2X,'LT(MTT)    S.D.    S.D.R ')
      WRITE(6,900)LT(MTT),SD,SDR
      WRITE(6,986)
      READ(5,900)FL,XFT,HF
      WRITE(6,988)
  988 FORMAT(2X,'FL         XFT     HF.')
      WRITE(6,900)FL,XFT,HF
      WRITE(6,986)
      READ(5,900)SF,HSI,BHLO,AHLO,UA,TR
      WRITE(6,987)
  987 FORMAT(2X,'SF',5X,'HSI',8X,'AHLO',8X,'BHLO',2X,'UA',7X,'TR')
      WRITE(6,900)SF,HSI,AHLO,BHLO,UA,TR
      WRITE(6,986)
C
C       INPUT DATA USED FOR STR AND PI CONTROL ALGORITHM
C
      READ(5,903) LTS,XTS,XTSP,SFS,XBS,XBSP
      WRITE(6,985)
  985 FORMAT(////,2X,'CONTROL ALGORITHM INPUT DATA')
      WRITE(6,986)
  986 FORMAT('------ ---------------------------------------------------
     1-----------')
      WRITE(6,984)
  984 FORMAT(2X,'LTS',5X,'XTS',4X,'XTSP',5X,'SFS',5X,'XBS',5X,'XBSP')
      WRITE (6,903) LTS,XTS,XTSP,SFS,XBS,XBSP
      WRITE(6,986)
      READ (5,903) UMIN,UMAX
      WRITE(6,982)
  982 FORMAT(2X,'MIN.     MAX.    CONTROLLER CONTRAINT')
      WRITE(6,903) UMIN,UMAX
      UMIN=1.0+UMIN
```

```
      UMAX=1.0+UMAX
      WRITE(6,986)
      READ (5,902) JM,UCONT,KCONT,KPAR,KSTR,MVR,KPIT,KPIB
      WRITE(6,981)
  981 FORMAT(2X,'JM  UCONT  KCONT  KPAR  KSTR  MVR  KPIT  KPIB')
      WRITE (6,902) JM,UCONT,KCONT,KPAR,KSTR,MVR,KPIT,KPIB
      WRITE(6,986)
      READ (5,900) KPT,KPB,RESETT,RESETB
      WRITE(6,980)
  980 FORMAT(2X,'KPT        KPB       RESETT    RESETB')
      WRITE(6,900) KPT,KPB,RESETT,RESETB
      WRITE(6,986)
      READ(5,902) MT,KT,MB,KB,IBD,IV,LTP,LB
      WRITE(6,979)
  979 FORMAT(2X,'MT     KT     MB     KB    IBD     IV    LTP     LB')
      WRITE(6,902) MT,KT,MB,KB,IBD,IV,LTP,LB
      WRITE(6,986)
      READ(5,905) MU,BOT,BOB,RT,RB
      WRITE(6,978)
  978 FORMAT(2X,'        MU          BOT           BOE           RT          RB')
      WRITE(6,905) MU,BOT,BOB,RT,RB
      WRITE(6,986)
      READ (5,900) (PT(I,I),I=1,7),(PB(I,I),I=1,7)
      WRITE(6,977)
  977 FORMAT(2X,'STR XD & XB INITIAL COVARIANCE MATRIX USED')
      WRITE(6,900)(PT(I,I),I=1,7),(PB(I,I),I=1,7)
      WRITE(6,986)
      READ(5,909) (X1(I),I=1,7),(XB(I),I=1,7)
      WRITE(6,976)
  976 FORMAT(2X,'STR XD & XB INITIAL PARAMETERS USED')
      WRITE(6,909) (X1(I),I=1,7),(XB(I) I=1,7)
      WRITE(6,986)
C-------------------------------------------------------------------------------
C
C
C
C-------------------------------------------------------------------------------
C
C
C       USING SUBROUTINE INII :  READS IN INITIAL DISTILLATION COLUMN
C       VARIABLES-XT,YT,LT,VT,WT,QLT,E,HLT,HVT FOR EACH TRAY AVAILABLE
C       FROM DATA FILE LINE 22-35, SUBTOUTINE  PRINT PRINTS THEM
C       USING LUN=6
      F(1)=FL
      XF(1)=XFT
      E(10)=0.0
      CALL INII
      I=1
      T=0.0
      QP=32747.
      TS=107.7
      CALL PRINT
C
C      CONVERTING FROM NUMBER OF SAMPLING INTERVALS INTO NUMBER OF
C      INTEGRATION TIME STEPS   1 SAMPLING INTERVAL=64 SECONDS
      JM=JM*64/DT
      KPIT=KPIT/DT*64
      KPIB=KPIB/DT*64
      KSTR=KSTR/DT*64
```

```
      KPAR=KPAR/DT*64
      NI=NI/DT*60
      KST=64/DT
      ISPC=20/DT*64
      IB=0
      IST=0
      IN=0
      RE(I)=LT(MTT)
      IKT=1
C
C     DETERMINING THE TYPE OF DISTURBANCE
C
      DO 5 I=1,3
      IF(ISTRN(I)-ITEST)5,7,5
    5 CONTINUE
    7 GO TO (6,8,9),I
C
C     SETTING UP DISTURBANCE VECTOR ACCORDING TO TIME VECTOR
C
    6 DO 61 I=1,3
      ICHNG(I)=ICHNG(I)/DT*64
      XF(I)=XF(1)
      DF(I)=F(1)
      F(I)=F(1)
   61 RE(I)=D(I)
      GO TO 11
    8 DO 81 I=1,3
      ICHNG(I)=ICHNG(I)/DT*64
      XF(I)=XF(1)
      RE(I)=RE(1)
      DF(I)=D(I)
   81 F(I)=D(I)
      GO TO 11
    9 DO 91 I=1,3
      ICHNG(I)=ICHNG(I)/DT*64
      RE(I)=RE(1)
      DF(I)=F(1)
      F(I)=F(1)
   91 XF(I)=D(I)
   11 CONTINUE
C
C     AT THE START OF RUN, SUM OF ERROR SQUARES (AIE AIEB) ARE SET TO ZERO
C     REFERENCE TOP AND BOTTOM COMPOSITIONS ARE ALSO DESIGNATED
C
      I=1
      AIE=0.0
      AIEB=0.0
      DXB=XT(1)
      XT1=DXB
      REFT=0.96
      REFB=0.06
C
C     SIMULATION RUN OUTPUT LABEL
C
      WRITE(7,1119)
 1119 FORMAT(////,2X,'********** SIMULATION RESULTS **********',//,'  TI
     1ME   BOTTOM  STEAM    TOP  REFLUX   FEED   BOTTOM  ACCUM.ERROR
     2SQUARE',/,'  (MIN)         MEAS   (G/S)         (G/S)        ACTUA
     1L  T   B')
      WRITE(7,1005)I,XT1,SF,XT(10),LT(10),F(I),XT(1),AIE,AIEB
```

```
      WRITE(8,1122)
 1122 FORMAT(//////'********** PARAMETERS VALUES **********'/)
      WRITE(8,1123)
 1123 FORMAT(8X,'S T R  CONTROL')
      WRITE(8,1124) MT, KT, LTP
 1124 FORMAT(8X,'M=',I2,2X,'K=',I2,2X,'L=',I2/)
      WRITE(8,1125)
 1125 FORMAT(3X,'TIME' ,2X,'ALPHA2',2X,'BETA2',4X,'ALPHA1',2X,'BETA1',
     12X,'BETA3')
C
C
C
C
C
C
C       COLUMN SIMULATION ITERATION LOOP STARTS
C
C
C
      DO 100 KK=1,NI
      T=KK*DT/60.
C
C        O CHECK TIME FOR SET POINT CHANGES
      IF(KK.GE.ISPC) XTS=XTSP
      IF(KK.GE.ISPC) XBS=XBSP
      AIE=AIE+(XTS-XT(10))*(XTS-XT(10))
      AIEB=AIEB+(XBS-XT(1))*(XBS-XT(1))
C
C       TO CHECK TIME FOR LOAD DISTURBANCE
      DO 23 JJ=1,3
      IF(KK-ICHNG(JJ))23,22,23
   22 I=JJ
   23 CONTINUE
C
C       OBTAIN CORRECT FEED AND TOP PRODUCT FLOW RATES,THEN SIMULATE
      FFF=DF(I)
   25 TPROD=VT(MT)-LT(MT)
C
C
C       SUBROUTINE COMP: USING TABLE LOOK UP,WITH LIQUID COMPOSITION
C       (XT) AVAILABLE, FIND EQUILIBRIUM VAPOR COMPOSITION(YST);
C       AND WITH EFFICIENCIES  SSIGNED,ACTUAL VAPOR COMPOSITION(YT)
C       FOR EACH TRAY IS OBTAINED
      CALL COMP
C
C
C       SUBROUTINE ENTH: USING TABLE LOOK UP ,LIQUID ENTHALPHY(HLT)
C       AND VAPOR ENTHALPHY(HVT) CAN BE OBTAINED FROM XT,YT
   10 CALL ENTH
C
C
C       SUBROUTINE VFLOW: FLOWS LT VT ARE CALCULATED USING TRAY MASS
C       AND ENTHALPY BALANCES
      CALL VFLOW
  107 IKT=3
C
C
C       SUBROUTINE INTGR PERFORMS ACTUAL INTEGRATION
C
      CALL INTGR
```

```
C
C        CHECK PRINT TIME
         IN=IN+1
         IF(IN-JM )20,52,20
      52 CALL PRINT
         IN=0
      20 CONTINUE
C
C        CHECK SAMPLING INTERVAL
     104 IST=IST+1
         IF (IST-KST) 100,888,100
     888 IST=0
C        ROUTINE GAUSS IS USED TO GENERATE NORMALLY DISTRIBUTED
C        RANDOM NUMBER WITH STANDARD DEVIATION 0.05 G/S
C        FOR ADDING NOISE TO THE FEED FLOW
         CALL GAUSS(IV,SD,0.0,VF)
C         REFLUX FLOW NOISE CAN BE ADDED FOR OPEN LOOP IF REQUIRED
C         BY DEFINING SDR TO BE NON-ZERO QUANTITY
         CALL GAUSS(IR,SDR,0.0,VR)
         F(I)=FFF+VF
         IF (KCONT-1) 369,369,963
     369 LT(MTT)=RE(I)
         LT(MTT)=LT(MTT)+VR
C        CHECK FOR OPEN LOOP OR CLOSED LOOP SIMULATION
C
     963 GO TO (222,666), KCONT
     666 CONTINUE
C        CHECK FOR PI OR STR CONTROL---
C
         IF(KK-KPIT) 53,53,57
C        CHECK FOR TOP CONTROL, BOTTOMS CONTROL OR SIMULTANEOUS CONTROL
C
      53 GO TO (33,57,33),JCONT
      33 CALL PID(KPT,RESETT,REST,XTS,LTS,XT(10),LT(10),REFT)
         OLT=LT(10)
      57 IF(KK-KPIB) 56,56,777
      56 GO TO (40,44,44), JCONT
      44 IB=IB+1
C        CHECK BOTTOM G.C. MEASUREMENT TIME DELAY.
C
         IF(IB-IBD) 40,41,40
      41 IB=0
         XT1=DXB
         DXB=XT(1)
         CALL PID(KPB,RESETB,RESB,XBS,SFS,XT1,SF,REFB)
         OSF=SF
      40 CONTINUE
     777 CONTINUE
         IF (KK .GE. KPIR .AND. KK  .LE. KPIT) GO TO 62
         IF(KK-KPIT) 63,63,62
      62 GO TO (55,63,55), JCONT
C     COMPUTE THE NORMALIZED DEVIATION VARIABLES
C
      55 IF(KTOP.EQ.1) OLT=LT(10)
         KTOP=2
         XTN=(XT(10)-XTS)/REFT
         LTN=(LT(10)-OLT)/LTS
         OLTT=OLT
         OLT=LT(10)
         CALL STR(MI,KT,LTP,MU,BOT,RT,PT,X1,LDUMT,XDUMT,XTN,LTN,LTS,
```

```
     1LT(10),OLT)
      WRITE (8,903) T,X1(2),X1(4),X1(1),X1(3),X1(5),X1(6),X1(7)
   63 IF(KK-KPIB) 222,222,65
   65 IF(KBOT.EQ.1) OSF=SF
      KBOT=2
      GO TO (42,66,66), JCONT
   66 IB=IB+1
      IF(IB-IBD) 42,43,42
   43 IE=0
      XT1=DXB
      DXB=XT(1)
C  COMPUTE NORMALIZED DEVIATION VARIABLES
C
      XBN=(XT1-XBS)/REFB
      SFN=(SF-OSF)/SFS
      OSFF=OSF
      OSF=SF
      XBN=-XBN
      CALL STR(MB,KB,LB,MU,BOB,RB,PB,XB,LDUMB,XDUMB,XBN,SFN,SFS,SF,OSFF)
      WRITE (9,906)T,XB(2),XB(4),XB(1),XB(3),XB(5),XB(6),XB(7)
   42 CONTINUE
  222 WRITE(7,906) T,XT1,SF,XT(10),LT(10),F(I),XT(1),AIE,AIEB
      WRITE(11,531)XT(10),LT(10)
  531 FORMAT(2F8.4)
  100 CONTINUE
      WRITE(7,905) T,AIE,AIEB
      STOP
      END
C
C
C        SELF-TUNING REGULATOR CONTROL ALGORITHM
C
      SUBROUTINE STR(M,K,L,MU,BO,R2,P,X,LDUM,XDUM,XTN,LTN,USP,UCT,OUCT)
      REAL LT(20),MU,LTS,LTN,LDUM(10),LTGS,XDUM(10),[illegible]K(7)
      DIMENSION P(7,7),PHI(7,7),Y(7),X(7),PY(7),AK(7),AKY(7,7),PPH(7,7)
      COMMON XT(20),YT(20),LT,VT(20),HLT(20),HVT(20),HL(20),E(20),XF
     1(4),KK,TPROD,T,MTT,IKT,II
      COMMON WT(20),SF,HSI,AHLO,BHLO,UA,[illegible]R,F(4),OLP(20),[illegible],LF,HF,QP,TS
      COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
      DATA LUNR,LUNW,ACCUM,PHI/5,6,0.0,49*0.0/
C  M= NO. OF PARAMETERS IN Y,  K= NO. OF TIME DELAYS
C
      ULOW=UMIN*USP
      UHI=UMAX*USP
      N=L+M
      K2=K+4
      K4=K+6
      DO 208  I=1,N
  208 PHI(I,I)=1.0
C  GET THE DATA FROM  PROCESS MODEL
C
      DO 204 I=1,K2
      KIN=K2-I+2
  204 XDUM(KIN)=XDUM(KIN-1)
      DO 205 I=1,K4
      KIN=K4-I+2
  205 LDUM(KIN)=LDUM(KIN-1)
      DO 206 I=1,M
      IK1=I+K+1
  206 Y(I)=XDUM(IK1)
```

```
      DO 207 I=1,L
      IM=I+M
      IK2=I+K+1
207   Y(IM)=LDUM(IK2)
      XDUM(1)=-XTN
      LDUM(1)=LTN*BO
      IF(MVR-1) 50,51,50
50    IF(KK-KST.) 52,51,51
52    CONTINUE
C  COMPUTE THE  GAIN VALUE
C
      DO 22 I=1,N
      PY(I)=0.0
      DO 22 J=1,N
22    PY(I)=PY(I)+P(I,J)*Y(J)
      YPY=0.0
      DO 23 I=1,N
      AK(I)=PHI(I,I)*PY(I)
23    YPY=YPY+Y(I)*PY(I)
      YPY=YPY+R2
      DO 25 I=1,N
25    AK(I)=AK(I)/YPY
C   COMPUTE  COVARIANCE MATRIX
C
      DO 26 I=1,N
      DO 26 J=1,N
26    AKY(I,J)=PHI(I,J)-AK(I)*Y(J)
      DO 27 I=1,N
      DO 27 J=1,N
      PPH(I,J)=0.0
      DO 27 IJ=1,N
   27 PPH(I,J)=PPH(I, )+P(I,I )*PHI(IJ,J)
      DO 24 I=1,N
      DO 24 J 1,N
      P(I,J) 0.0
      DO 28 IJ=1,
28    P(I,J)=P(I,J)+AKY(I,IJ *PPH(IJ,J)
24    P(I,J)=(P(I,J))/MU
C  UPDATE  PARAMETER  VECTOR
C
      YX=0.0
      DO 29 I=1,N
      X(I)=PHI(I,I)*X(I)
29    YX=YX+Y(I) X(I)
C  COMPUTE  ERRC
C  Y(4) IS EQUAL TO BO*U(T-2)
C
      ERR=X N-LDU (K+1)-YX
      DO 201 I=1,
      ERK(I)=AK(I)*ERR
201   X(I)=X(I)+ERK(I)
C   COMPUTE  CONTROL  VECTOR
C
51    CONTINUE
      YX=0.0
      DO 202 I=1,M
202   YX=YX+XD  (I)*X(I)
      YY= .0
      DO 203    1,L
      IM=I+M
```

```
  203   YY=YY+LDUM(I+O)*X(IM)
        LTN=(-YY-YX)/BO
        IF(KK-KPIT) 46,46,53
   53   CONTINUE
C  CONSTRAINTS  ON THE  CONTROLLER
C
        UCT=DUCT+LTN*USP
        IF(UCT-ULOW)44,44,45
   45   IF(UCT-UHI)46,47,47
   47   UCT=UHI
        GO TO 46
   44   UCT=ULOW
   46   CONTINUE
        WRITE(LUNW,37) (Y(I),X(I),P(I,I),AK(I),I=1,N)
   37   FORMAT(4(7X,E10.3))
   40   CONTINUE
        RETURN
        END
C
C
C       PROPORTIONAL INTEGRAL CON  OL ALGORITHM
C
C
        SUBROUTINE PID(KP,RESET,RES,XSP,USP,XTV,UCT,REF)
        REAL KP,LT(20),LTS,LTGS
        COMMON XT(20),YT(20),LT,VT(20),HLT(20),HVT(20),HL(20),E(20),XF
       1(4),KK,TPR  ,T,MTT,IKT,I
        COMMON W (  ),SF,HSI,AHLP,BHLP,UA,TR,F(4),QLP(20),DT,LF,HF,OP,TS
        COMMON K    ,KSTR,MVR,UMIN,UMA
        ULOW=UMI   USP
        UHI=UMA   USP
        ER=(XTV-XSP)/REF
        RES=  S+ER
        SP=K  ER+KP/RESET*RES
        UCT=(1.+SP)*USP
        IF(UCT-UHI) 1,1,2
    2   UCT=UHI
        GO TO 3
    1   IF(UCT-ULOW) 4,4,3
    4   UCT=ULOW
    3   CONTINUE
        RETURN
        END
        SUBROUTINE GAUSS(IX,SA,AM,V)
        AA=0.0
        DO 7 I=1,12
        IY=IX*65539
        IF(IY) 5,6,6
    5   IY=IY+21474836 7+1
    6   YFL  Y
        YFL  FL*0. 656 13E-9
        IX=IY
    7   AA=AA+YFL
        V=(AA-6. ) SA+AM
        RETURN
        END
        SUBROUTINE PRINT
        REAL LT(20)
        COMMON XT(20),YT(20),LT,V (20),HLT(20),HVT(20),HL( ),E(20),XF
       1(4),KK,IPROD,T,MTT,IKT,I
```

```
      COMMON WT(20),SF,HSI,AHLO,BHLO,UA,TR,F(4),OLP(20),DT,LF,HF,OP,TS
      COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
  101 FORMAT(' ',20X,'TIME',F8.2,'(MIN)')
  102 FORMAT(/,2X,'STAGE',2X,'LCOMP',2X,'VCOMP',2X,'EFFNCY',2X,'LFLOW',2
     1X,'VFLOW',2X,'HOLDUP',2X,'LENTH',2X,'VENTH')
  103 FORMAT(/,I6,F7.3,F7.3,F8.3,F8.3,F7.3,F7.2,F9.2,F8.2)
  104 FORMAT(/,5X,'FEED RATE',2X,'FEED COMP',2X,'STEAM FLOW',2X,'HT COEF
     1F',2X,'HEAT LOAD')
  105 FORMAT(/,F11.2,F11.4,F12.2,F12.3,F15.2)
  106 FORMAT(/,10X,'REBOILER TEMP',F9.3,5X,'STEAM CHEST TEMP',F9.3)
      WRITE(6,101)T
      WRITE(6,102)
      DO 1 N=1,MTT
    1 WRITE(6,103)N,XT(N),YT(N),E(N),LT(N),VT(N),WT(N),HLT(N),HVT(N)
      WRITE(6,104)
      WRITE(6,105) F(I),XF(I),SF,UA,QP
      WRITE(6,106)TR,TS
      RETURN
      END
      SUBROUTINE INTGR
      REAL LT(20)
      DIMENSION D(20),FF(20),XX(20),Q(20)
      COMMON XT(20),YT(20),LT,VT(20),HLT(20),HVT(20),HL(20),E(20),XF
     1(4),KK,TPROD,T,MTT,IKT,I
      COMMON WT(20),SF,HSI,AHLO,BHLO,UA,TR,F(4),OLP(20),DT,LF,HF,OP,TS
      COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
      DO 25 N=1,MTT
   25 XX(N)=XT(N)
      NN=2
   26 DO 10 N=1,MTT
      K=N-1
      J=N+1
      IF(N-1)1,4,1
    1 IF(N-LF)2,5,2
    2 IF(N-MTT)3,6,3
    3 D(N)=LT(J)*XT(J)+VT(K)*YT(K)-LT(N)*XT(N)-VT(N)*YT(N)
      GO TO 10
    4 D(N)=LT(J)*XT(J)-LT(N)*XT(N)-VT(N)*YT(N)
      GO TO 10
    5 D(N)=LT(J)*XT(J)+VT(K)*YT(K)-LT(N)*XT(N)-VT(N)*YT(N)+F(I)*XF(I)
      GO TO 10
    6 D(N)=VT(K)*(YT(K)-XT(N))
   10 CONTINUE
      IF(NN-3)11,27,11
   11 DO 15 N=1,MTT
   15 Q(N)=DT/WT(N)
      DO 22 N=1,MTT
      FF(N)=D(N)
      XT(N)=XT(N)+Q(N)*D(N)
   22 CONTINUE
      IF(NN-2)27,28,27
   28 NN=3
      CALL COMP
      CALL ENTH
      CALL [illegible]FLOW
      GO TO 26
   27 CONTINUE
      DO 30 N=1,MTT
   30 XT(N)=XX(N)+(Q(N)/2)*(D(N)+FF(N))
      RETURN
```

```
      END
      SUBROUTINE INII
      REAL LT(20)
      COMMON XT(20),YT(20),LT,VT(. O),HLT(20),HVT(20),HL(20),E(20),XF
     1(4),KK,IPROD,T,MTT,IKT,I
      COMMON WT(20),SF,HSI,AHLO,BHLO,UA,TR,F(4),QLP(20),DT,LF,HF,QP,TS
      COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
      NTT=MTT-1
      READ(5,100)(XT(N),N=1,MTT)
      READ(5,100)(YT(N),N=1,MTT)
      READ(5,100)(LT(N),N=1,NTT)
      READ(5,100)(VT(N),N=1,NTT)
      READ(5,100)(WT(N),N=1,MTT)
      READ(5,100)(QLP(N),N=1,MTT)
      READ(5,100)(F(N),N=1,NTT)
      READ(5,100)(HLT(N),N=1,MTT)
      READ(5,101)(HVT(N),N=1,MTT)
  100 FORMAT(10F8.4)
  101 FORMAT(10F8.3)
      RETURN
      END
      SUBROUTINE VFLOW
      REAL LT(20)
       DIMENSION C(1 ), B(19),FT(20),SB(19),SFT(19),DHL(20),SW(19)
      COMMON XT(20),YT(20),LT,VT(20),HLT(20),HVT(20),HL(20),E(20),XF
     1(4),KK,IPROD,T,MTT,IKT,I
      COMMON WT(20),SF,HSI,AHLO,BHLO,UA,TR,F(4),QLP(20),DT,LF,HF,QP,TS
      COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
      NTT=MTT-1
C     *TEMPERATURE OF STEAM
C C**HEAT ADDED
      IF(IKT 2)60,51,51
60    READ(5,100)IQP
      QP=IQP
      IF(IQP-0)50,51,50
  100 FORMAT(2I7)
 51   TS=(SF*(HSI-AHLO) UA*TR)/(UA+BHLO*SF)
      QR=UA*(TS-TR)
      QP =QR
   50 CONTINUE
      NL=NTT-1
C C**  VAPOUR AND LIQUID FLOWRAETS
      DO 1 N=1,NL
      M=N+1
      C(N)=(HLT(M)-HVT(N))
    1 B(N)=(HLT(N)-HLT(M))
      C(NTT)=-(HVT(NTT)-HLT(NTT))
C C***  DERIVATIVES OF HL
      DO 35 N=1, TT
   35 DHL(N)=(HL  N)-HL(N))/DT
      M=NTT+1
      DO 2 N=1,NTT
    2 SW(N)=0.0
      DO 4 N=  NL
      IF(N-   ),5,20
   20 IF(N-LF)21,6,21
   21 IF(N-LF)22,22,7
   22 FT(N)=W (N)*DHL( )-B(N)*(SW(M)-F(I)-LT(M))+QLP(N)
      GO TO 4
    5 FT(N)=WT(N)*DHL(N)-B(N)*(SW(N)-F(I)-LT(M))+QLP(N)-QP
```

```
      GO TO 4
    6 FT(N)=WT(N)*DHL(N)-B(N)*(SW(N)-LT(M))-F(I)*(HF-HLT(N))+QLP(N)
      GO TO 4
    7 FT(N)=WT(N)*DHL(N)-B(N)*(SW(N)-LT(M))+QLP(N)
    4 CONTINUE
      FT(NTT)=WT(NTT)*DHL(NTT)-LT(M)*(HLT(M)-HLT(NTT))+QLP(NTT)
      ADD=0.0
      DO 10 N=1,NTT
      SFT(N)=ADD+FT(N)
   10 ADD=SFT(N)
      ADD=0.0
      DO 11 N=1,NL
      SB(N)=ADD+B(N)
   11 ADD=SB(N)
      VT(NTT)=SFT(NTT)/(C(NTT)+SB(NL))
      DO 12 N=1,NL
   12 VT(N)=(SFT(N)-SB(N)*VT(NTT))/C(N)
      DO 13 N=1,NTT
      K=NTT+1-N
      J=K-1
      M=K+1
      IF(K-LF)40,14,40
   40 IF(N-NTT)41,15,41
   41 LT(K)=VT(J)+LT(M)-VT(K)
      GO TO 13
   14 LT(K)=VT(J)+LT(M)-VT(K)+F(I)
      GO TO 13
   15 LT(K)=LT(M)-VT(K)
   13 CONTINUE
      RETURN
      END
      SUBROUTINE  NTH
      REAL LT(20)
      DIMENSION XST(19),YST(19),HXS(19),HYS(19)
      COMMON XT(20),YT(20),LT,VT(20),HLT(20),HVT(20),HL(20),E(20),XF
     1(4),KK,TPROD,T,  T,IKT,I
      COMMON WT(20),SF,HSI,AHLO,BHLO,UA,TR,F( ),QLP(20),DT,LF,HF,JP,TS
      COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
  100 FORMAT(10F8.4)
  101 FORMAT(10F8.3)
  102 FORMAT(10F8.2)
      IF(IKT-2)1,4,4
    1 READ(5,100)XST
      READ(5,100)YST
      READ(5,101)HXS
      READ(5,102)HYS
      DO 50 N=1,19
   50 CONTINUE
    4 DO 5 N=1,MTT
    5 HL(N)=HLT(N)
      K=1
      DO 10 N=1,MTT
      DO 20 J=K,19
      IF(XST(J)-XT(N))20,20,2
   20 CONTINUE
    2 K=J-1
      L=J
      HLT(N)=((XT(N)-XST(K))*(HXS(L)-HXS(K))/(XST(L)-XST(K))+HXS(K))
   10 CONTINUE
      K=1
```

```
          DO 30 N=1,MTT
          DO 40 J=K,19
          IF(YST(J)-YT(N))40,40,3
       40 CONTINUE
        3 K=J-1
          L=J
          HVT(N)=((YT(N)-YST(K))*(HYS(L)-HYS(K))/(YST(L)-YST(K))+HYS(K))
       30 CONTINUE
          RETURN
          END
          SUBROUTINE COMP
C C       YST IS THE VAPOUR COMPOSITION WHICH WOULD
C C       BE IN EQUILIBRIUM WITH THE LIQUID ON THE
C C       CORRESPONDING TRAY. THEY ARE CALCULATED
C C       FROM X-Y DATA, AND ARE EXPRESSED AS LINEARIZED
C C       EQUATIONS. PLATE EFFICIENCIES ARE ALSO
C C       CALCULATED HERE.
          REAL LT(20)
          DIMENSION YST(20),YS(26),XST(26)
          COMMON XT(20),YT(20),LT,VT(20),  T(20),HVT(20),HL(20),E(20),XF
         1(4),KK,TPROD,T,MTT,IKT,I
          COMMON WT(20),SF,HSI,AHLO,BHLO,UA,TR,F(4),OLP(20),DT,LF,HE,QP,TS
          COMMON KPAR,KSTR,MVR,UMIN,UMAX,KPIT,KPIB
          NTT=MTT-1
      101 FORMAT(10F8.3)
          IF(IKT-2)1,3,3
        1 READ(5,101)XST
          READ(5,101)YS
        3 J=1
          DO 20 K=1,MTT
          DO 10 N=J,26
          IF(XST(N)-XT(K))10,10,2
       10 CONTINUE
        2 J=N-1
          L=N
          YST(K)=(((XT(K)-XST(J))*(YS(L)-YS(J))/(XST(L)-XST(J)))+YS(J))
       20 CONTINUE
          YT(1)=E(1)*YST(1)
          DO 21 N=1,NTT
          J=N+1
       21 YT(J)=E(J)*(YST(J)-YT(N))+YT(N)
          RETURN
          END
END OF FILE
```

APPENDIX F: BOTTOM PRODUCT CONTROLLER TUNING BY THE ZIEGLER-NICHOLS CONTINUOUS CYCLING METHOD

The distillation column was operated at the following conditions:

Feed Flow	= 17.98	g/s
Reflux Flow	= 13.35	g/s
Steam Flow	= 13.22	g/s
Top Composition	= 96.83	%
Bottom Composition	= 3.[illegible]	%
Feed Composition	= 50.[illegible]	%

Various proportional gains were tested, under closed loop control, for a 2% step increase in set point to obtain a sustained system oscillation. The results are summarized in Table F1.

Table F1 Results of Bottom Product Composition Ziegler-Nichols Tuning Runs

Run	Proportional Gain	System Performance
F1	0.04	offset
F2	0.06	sustained oscillation
F3	0.12	system unstable
F4	0.20	system unstable

The control and manipulated variable responses for Run F2 are plotted in Figures F1 and F2. On the basis of the Ziegler-Nichols recommended settings, the controller constants for proportional plus integral control are:

$$K_P = 0.45 K_{P\,MAX} = 0.027$$

$$T_R = P_u/1.2 = 80.0 \text{ minutes}$$

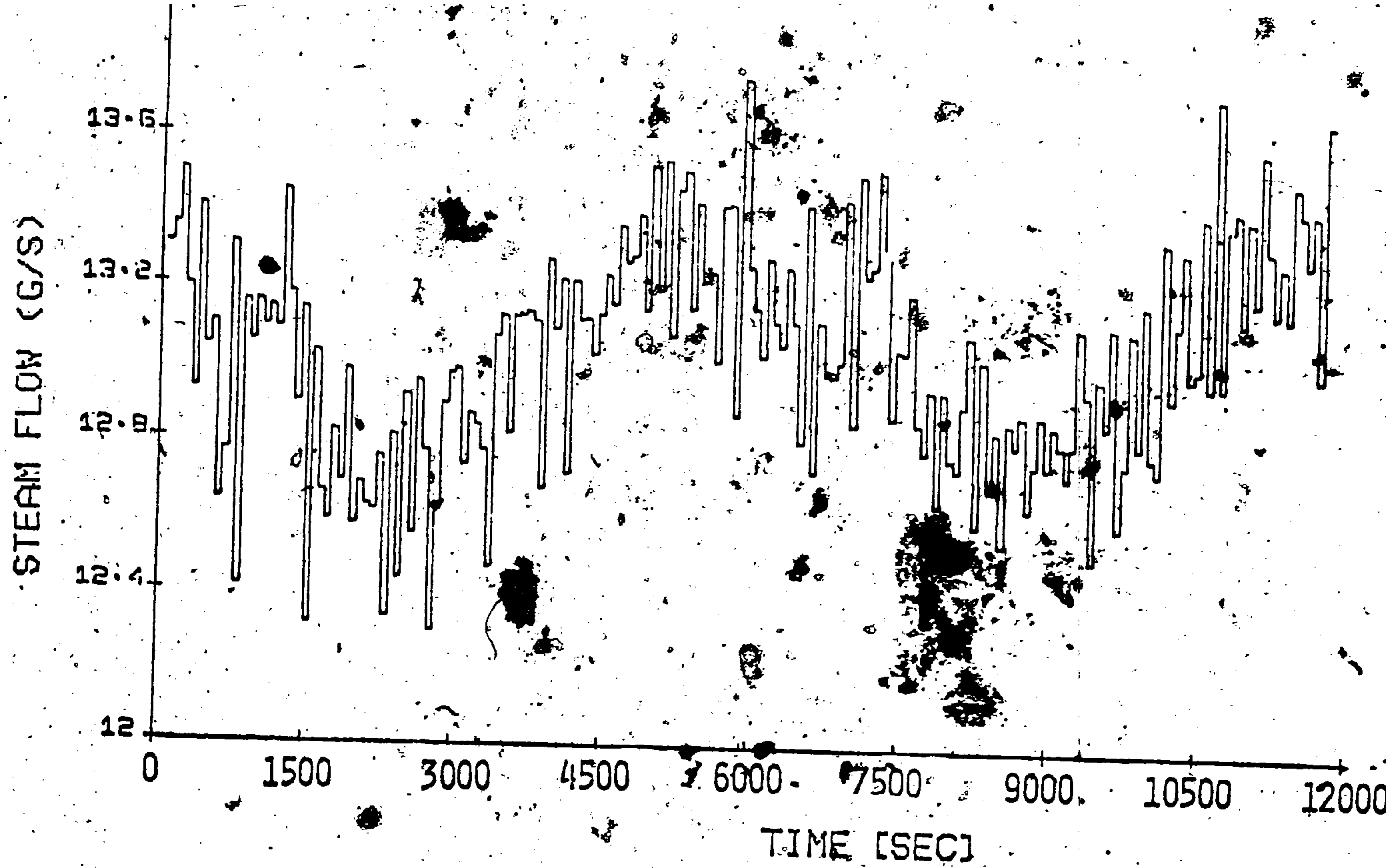


FIGURE F2 STEAM FLOW RATE RESPONSE FOR PROPORTIONAL CONTROLLER WITH $K_P = 0.06$ OF BOTTOM PRODUCT COMPOSITION CONTROL FOR A +2% STEP CHANGE IN SET POINT

APPENDIX G: SUMMARY OF PARAMETER ESTIMATION TESTS FOR STR CONTROL OF BOTTOM PRODUCT COMPOSITION

The closed loop parameter estimation runs given in Table G.1 were undertaken to identify a suitable disturbance flow pattern for obtaining STR parameter estimates for use in tests of the regulator.

Table G.1 Summary Of Parameter Estimation Tests For STR Control Of Bottom Product Composition

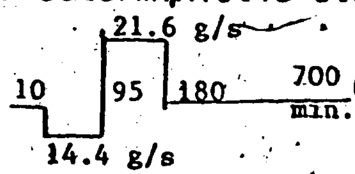
$M = 2 \quad K = 5 \quad L = 6$

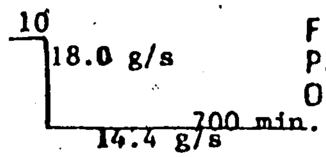
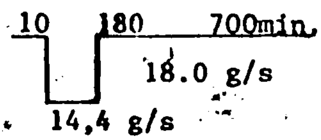
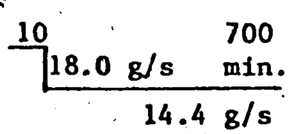
$\beta_0 = 10.0 \quad \underline{P}(0) = 50.0$

For stochastic (Gaussian) changes in feed flow rate:

Run	Standard Deviation	Initial Parameters	Final Parameters α_i	Final Parameters β_i	Parameter Behavior
G1	1.0 g/s	0.0	-0.2754	0.6673	Parameters did not converge
			-0.3796	0.1350	
				-0.1461	
				0.0538	
				-0.0746	
				-0.0972	
G2	3.0 g/s	0.0	-4.2410	-0.4138	Parameters did not converge
			-0.7210	0.6750	
				-0.3219	
				0.5420	
				-0.3189	
				0.1649	

For deterministic step changes in feed flow rate:

Run	Disturbance	Initial Parameters	Final Parameters α_i	Final Parameters β_i	Parameter Behavior
G3		0.0	-1.2274	0.2897	Parameters converged
			0.2885	0.3178	
				0.3419	
				0.2607	
				0.0860	
				0.3481	

G4		Final Parameters Of Run G3	-5.5874 4.6377	-0.1607 -0.0247 0.1027 0.1419 -0.0196 -1.0294	Parameters did not converge
G5		0.0	-0.5038 -0.5259	0.0249 0.2004 0.3102 0.2656 -0.0625 0.6574	Parameters converged
G6		Final Parameters Of Run G5	-5.1508 4.0556	-0.0412 0.0180 0.0731 0.0719 -0.1195 -0.9923	Parameters converged

On the basis of these results, the square pulse disturbance in Run G5 was selected.

APPENDIX H: EFFECT OF SAMPLING INTERVAL ON THE STR CONTROL OF BOTTOM PRODUCT COMPOSITION

The number of parameters to be updated on-line as well as the amount of memory required with a real time process computer are directly affected by the K parameter, which is defined as a nonnegative integer expressing the system time delay as a multiple of the sampling interval. Hence, K can be manipulated by a change in sampling interval. The tests performed using different sampling intervals are summarized in Table H.1.

Table H.1 Summary Of FPC And PI Tests For Bottom Product Composition Control Using Different Sampling Intervals

$M = 2$ $k = 2$ $L = 3$ $\beta_o = 10.$ $\underline{P}(0) = 50.$

Sampling Interval = 512 seconds

Run	ISE	Initial Parameters	Final Parameters α_i	Final Parameters β_i	Disturbance
H1	-	0.0	-1.6044 0.0732	0.0249 0.6328 -0.3466	18.0 g/s; 14.4 g/s (10, 180, 700 min.)
H2	-	Final Parameters Of Run H1	-3.6943 2.0900	0.5255 0.1338 -1.5631	18.0 g/s; 14.4 g/s (10, 700 min.)
H3	0.7628	Final Parameters Of Run H2	FPC		Feed Flow -20%
H4	1.1545	Final Parameters Of Run H2	FPC		Feed Flow +20%
H5	0.2801	Final Parameters Of Run H2	FPC		Set Point + 2%

H6	0.0969	Final Parameters Of Run H2	FPC		Set Point - 2%

PI Control		K_P	T_R	
H7	1.8343	0.08	84.	Feed Flow -20%
H8	2.369	0.08	84.	Feed Flow +20%
H9	0.105	0.08	84.	Set Point + 2%
H10	0.099	0.08	84.	Set Point - 2%

Sampling Interval = 1024 seconds = Time Delay

H11	-	0.00	-1.2148 -0.0478	0.2902 -0.7074	18.0 g/s; 10, 180, 700 min.; 14.4 g/s
H12	-	Final Parameters Of Run H11	-1.9883 1.0378	0.2318 -1.2326	18.0 g/s; 10, 700min.; 14.4 g/s

By increasing the sampling interval from 256 seconds to 512 seconds, K is effectively reduced by half. The FPC performance is superior when compared with the PI performance as shown in Table H.1. When the sampling interval is further increased to 1024 seconds, which is equivalent to the system time delay, the parameters in Run H12 were still drifting at the end of the run.

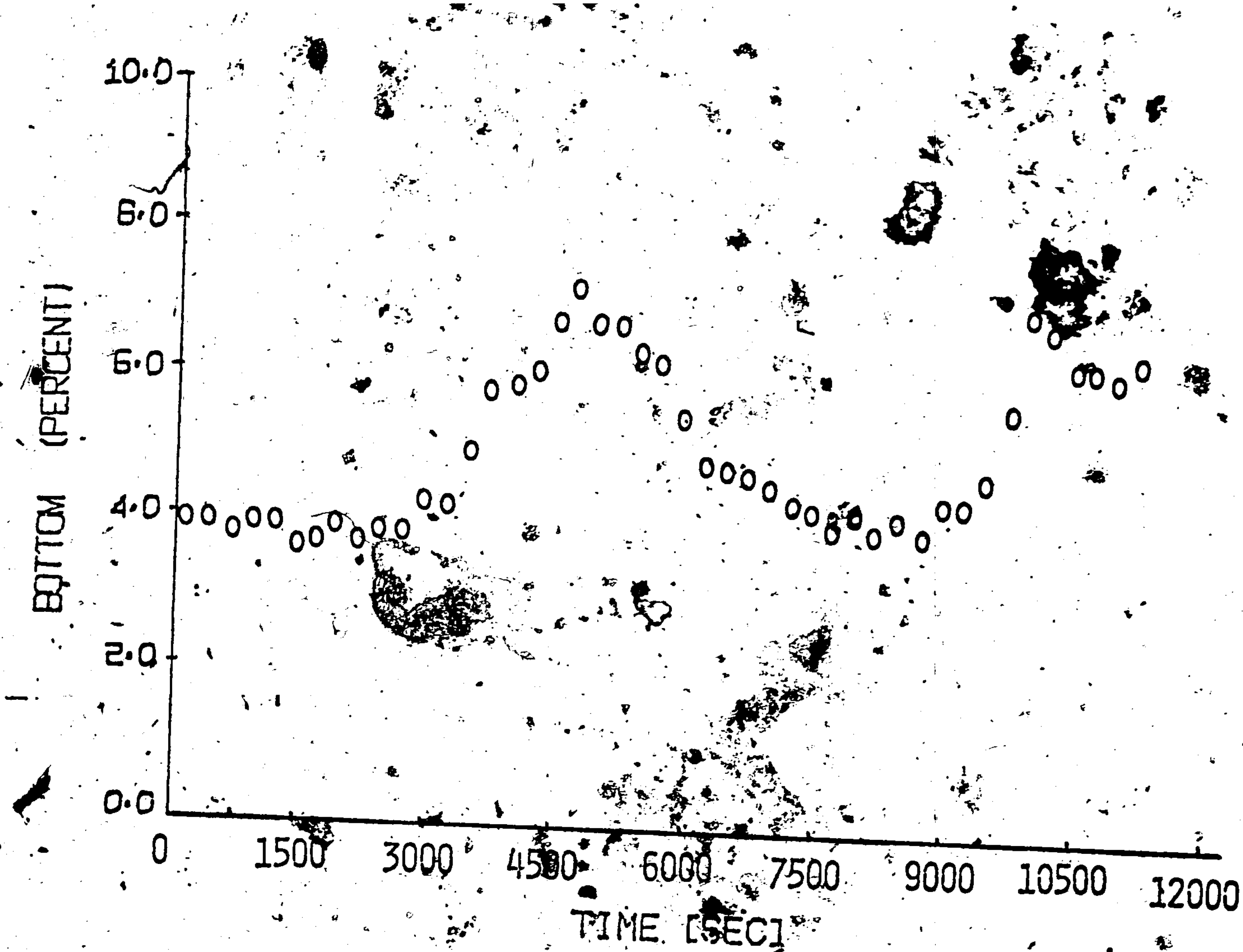


FIGURE F1 BOTTOM PRODUCT COMPOSITION RESPONSE TO A +2% STEP CHANGE IN SET POINT USING A PROPORTIONAL CONTROLLER WITH $K_p = 0.06$