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Full Name of Author — Nom complet de l'auteur
Robert J. Bilec

Date of Birth — Date de naissance
May 6, 1951.

Country of Birth — Lieu de naissance
Canada

Permanent Address — Résidence fixe
9533 95 Avenue,
Edmonton, Alberta.
T6C-2A2.

Title of Thesis — Titre de la thèse
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Dr. R.K. Wood

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THE UNIVERSITY OF ALBERTA

MODELLING AND DUAL CONTROL OF A BINARY DISTILLATION COLUMN

by

ROBERT J. BILEC

A THESIS

SUBMITTED TO THE FACULTY OF GRADUATE STUDIES AND RESEARCH
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THE 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 "MODELLING AND DUAL CONTROL OF A BINARY DISTILLATION COLUMN" submitted by ROBERT J. BILEC in partial fulfilment of the requirements for the degree of Master of Science in Process Control (Chemical Engineering).

Supervisor

CKiparissides

Date APRIL 11, 1980

DEDICATION

To my wife Linda and my daughter Jennifer.

ABSTRACT

When designing and implementing control strategies for the distillation column, a model is often very useful in evaluating those strategies. A nonlinear model which was based on the total and component material and energy balance differential equations was used to predict the dynamic behaviour of the distillation column. Two variations of the model were compared to experimental bottom and top composition responses for step disturbances in steam, reflux, and feed flow rates. One variation of the model assumed constant heat loss and efficiency based on the initial steady state. The other variation of the model used a variable heat loss and efficiency based on the initial and final steady states. Both variations of the nonlinear model assumed constant mass holdup based on the initial steady state. The constant nonlinear model predicted experimental column behaviour for bottom and top composition adequately for the transient and final steady state values. However the variable nonlinear model proved to be superior as it predicted the final steady state exactly for bottom and top compositions.

To aid in the design of the multivariable control strategies for the distillation column, the direct Nyquist array method was used. The direct Nyquist array method is a multivariable frequency domain method that employs a compensator to eliminate interactions between control loops.

A linear model in the form of transfer functions was obtained by pseudo random binary sequence and open loop testing on the distillation column. This linear model was used in the design of the compensators. Static and dynamic compensators were designed using the direct Nyquist array method and later evaluated through simulation studies using the constant nonlinear model for ±20% feed flow disturbances. The proportional plus integral plus derivative controller settings for the individual control loops were established using the Ziegler-Nichols method of tuning. Improved control over multiloop control case was not achieved until the decoupler compensator was used, although the static plus time delay compensator seemed adequate for simplicity of implementation.

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The author would like to congralulate his wife for surviving this rather long time to finish his thesis. The author would also like to thank the DACS centre for assistance in programming, the instrumentation people and workshop personnel for helping keep the column operable during the required period. Finally the author would like to thank Dr. Wood for financial assistance through NRC grants and the numerous hours spent reading the initial drafts.

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Chapter 1

INTRODUCTION

The objective of this work has been the development of a suitable nonlinear distillation column model, then use of the model to study the control behavior of the distillation column. Both linear and nonlinear models were developed to describe the dynamic behaviour of distillation column. The nonlinear model developed in this work was based on a previous model developed by Simonsmeier (53). Simonsmeier's nonlinear model was modified in order to achieve a better representation of the distillation column dynamic behaviour. The modifications were based on experimental tests carried out on a methanol-water pilot plant distillation column. The main assumptions used in the nonlinear model were constant tray efficiency, heat loss, and liquid holdup. Through simulation studies, the model performance was compared to experimental open loop data for steam, reflux, and feed disturbances. These tests provided suitable verification of the model to justify its use in evaluating control strategies for the distillation column.

Conventional proportional plus integral (PI) control works well for controlling the terminal compositions separately. Because of the interaction between the manipulative and control variables, use of the same PI control settings for dual control usually results in degradation in control performance of the two individual loops. The direct Nyquist array method, a frequency domain technique, was used to design compensators to help reduced the interaction that exists in the dual control system. This method was implemented by Kuon (23) as an extension of the work performed by Rosenbrock et al (48). This method uses a plant precompensator to reduce loop interaction and causes the system to behave as two independent loops. Different types of compensators along with conventional proportional plus integral plus derivative control were evaluated through simulation studies. For the design of the compensator, a transfer function model between manipulative and control variables was obtained by dynamic testing of the column using pseudo random binary sequence disturbances. The linear model identification was performed by Mozel (38) using a set of identification programs.

Chapter 2

DEVELOPMENT OF THE DISTILLATION COLUMN MODEL

2.1 Introduction

Most of the research work on distillation column modelling has been carried out on plate type columns for binary systems. The type of models used to describe the dynamic and steady state behaviour of distillation columns can be classed as transfer function models, linearized models and nonlinear models. Numerical solutions to these models, along with some experimental work have been carried out in the frequency and time domain using both analog and digital computers. In most cases, such studies on modelling aided in designing control schemes for distillation columns. No extensive review of the published literature dealing with modelling will be presented here since the topic has been covered adequately elsewhere (57,8,34,41,53,27), with the most comprehensive review of distillation dynamics being given by Rademaker et al (44).

A great deal of research on distillation column modelling has been carried out at the University of Alberta. The dynamic behaviour of the column was first studied by Svrcek (57) with a brief summary of the results published by Svrcek and Ritter (58). Column dynamics were predicted by using the total material balance, component material balance, and energy balance differential equations. Later

McGinnis (34) tried a linear state space model with the states being the liquid enthalpies and flow rates based on the energy balance differential equation. McGinnis offers an excellent review of the literature up to 1968 in which he presents the different models and assumptions used by the various authors. In a later study Liesch (27) evaluated a tenth order enthalpy model and also a concentration model based on the component material balance differential equation.

Besides assuming instantaneous vapour dynamics, constant liquid holdup was generally assumed. Svrcek calculated new liquid mass holdup values based on the relationship between composition changes and density. In a study by Distefano (10) a holdup correlation was determined empirically. This correlation was similiar to the relationship developed by Rademaker et al (44). Simonsmeier (53) used the model proposed by Svrcek and made changes in the numerical methods used to solve the three sets of differential equations. Simonsmeier carried out both experimental and theoretical studies on column dynamics. Also he evaluated the model using three types of holdup equations.

The model used in this study evolved from the nonlinear model used by Simonsmeier (52). Different numerical techniques have been applied to the model in order to improve the model's prediction of the experimental behaviour. It also has been modified in order to compensate for physical changes to the column. In the following sections, a description of the model used in this study will be presented along with a detailed list of changes that were made.

2.2 Mathematical Model

To describe the dynamics of the distillation column, the following set of differential equations were considered for each stage, that is,

(1) total material balance

(2) component material balance

(3) total energy balance

The component mass balance is usually expressed in terms of the more volatile component in the binary mixture. The stages of a column include the number of trays, plus the condenser and reboiler sections.

The major assumptions used to reduce the complexity of the dynamic model are:

(1) The column operates at constant atmospheric pressure which is maintained by manipulating the cooling water flow.

(2) Both liquid and vapour entrainment are negligible.

(3) All liquid and vapour on a stage are well mixed. This assumption means that the temperature, density, and composition throughout the liquid and vapour phases at a

stage will be uniform.

(4) the holdup in the downcomer is negligible compared to the holdup on the trays.

(5) The heat capacity of the metal and glass of each stage is negligible.

(6) The feed and reflux enthalpy are constant.

With these preliminary assumptions, the basic differential equations for a general stage can be stated as

$$d(WT_n + VT_n)/dt = L_{n+1} - L_n + V_{n-1} - V_n \pm S_n \qquad 2.1$$

$$\begin{aligned} d(WT_nX_n + VT_nY_n)/dt = {} & L_{n+1}X_{n+1} - L_nX_n + V_{n-1}Y_{n-1} \\ & - V_nY_n \pm S_nXs_n \end{aligned} \qquad 2.2$$

$$\begin{aligned} d(WT_nh_n + VT_nH_n)/dt = {} & L_{n+1}h_{n+1} - L_nh_n + V_{n-1}H_{n-1} \\ & - V_nH_n - Q_n \pm S_nhs_n \end{aligned} \qquad 2.3$$

where; WT_n - liquid mass holdup (g)
VT_n - vapour mass holdip (g)
L_n - liquid mass flow rate (g/s)
V_n - vapour mass flow rate (g/s)
S_n - side stream mass flow rate (g/s)
X_n - liquid composition (wt. fraction)
Y_n - vapour composition (wt. fraction)

Xs_n - liquid composition of side stream
h_n - liquid enthalpy (J/g)
H_n - vapour enthalpy (J/g)
Q_n - heat loss per stage (W)
hs_n - enthalpy of side stream (J/g)

If the vapour dynamics are considered to occur instantaneously and are small compared to the liquid dynamics, the equations can be further simplified. The following equations are written for a column with no side stream withdrawal and only one feed stream entering the tower.

$$d(WT_n)/dt = L_{n+1} - L_n + V_{n-1} - V_n + F_n \qquad 2.4$$

$$d(WT_nX_n)/dt = L_{n+1}X_{n+1} - L_nX_n + V_{n-1}Y_{n-1} - V_nY_n + F_nXf_n \qquad 2.5$$

$$d(WT_nh_n)/dt = L_{n+1}h_{n+1} - L_nh_n + V_{n-1}H_{n-1} - V_nY_n - Q_n + F_nhf_n \qquad 2.6$$

where; F_n - feed flow rate (g/s)
Xf_n - feed composition (wt. fraction)
hf_n - feed enthalpy (J/g)

For all trays other than the feed tray, the right most term in all of the above equations is deleted. The equations describing the reboiler and condenser sections will be treated separately.

Designating the reboiler as the first stage and assuming the reboiler liquid holdup is constant, the model equations for the process side are:

$$L_2 = L_1 + V_1 \qquad 2.7$$

$$WT_1 d(X_1)/dt = L_2X_2 - L_1X_1 - V_1Y_1 \qquad 2.8$$

$$WT_1 d(h_1)/dt = L_2h_2 - L_1h_1 - V_1H_1 - Q_1 + QR \qquad 2.9$$

where; QR - heat to reboiler (W)

The analysis of the steam side of the reboiler (57) assumes zero heat loss and that the temperature of the reboiler liquid is a function of the bottom liquid composition. With the steam condensate under level control and neglecting the steam dynamics, the following equation can be used to characterize the steam side, that is,

$$St(Hs - (a + bTs)) = UA(Ts - T_1) \qquad 2.10$$

where; St - steam mass flow rate (g/s)

Hs - steam enthalpy (J/g)
a - intercept for heat capacity (J/g)
b - heat capacity for steam (J/g-°C)
Ts - steam temperature (°C)
UA - product of heat transfer times area (W-°C)

The above steady state equation is considered to be applicable because the steam time constant is approximately one thousand times smaller than the liquid time constant on the reboiler process side (57).

With the previous assumptions of perfect mixing, vapour holdup negligible, and liquid level under control, the condenser process equations can be expressed as

$$V_9 = L_{10} + Re \qquad 2.11$$

$$WT_{10}d(X\quad)/dt = V_9Y_{10} - (L_{10} + Re)X_{10} \qquad 2.12$$

where; Re - reflux mass flow rate

The heat transfer dynamics are assumed to be negligible (57), so the energy balance simply reflects the fact that the cooling water used to control pressure removes the latent heat of vapourization.

In order to simulate the dynamic response of the distillation column, the following functional relations are required;

(1) liquid enthalpy, $h_n = f(X_n)$ 2.13

(2) vapour enthalpy, $H_n = f(Y_n)$ 2.14

(3) equilibrium relationship, $Y_n^* = f(X_n)$ 2.15

To predict the actual vapour composition, the efficiency of the trays must be included. The Murphree plate-efficiency (E) expressed in terms of the vapour composition is

$$E_n = (Y_{n-1} - Y_n)/(Y_{n-1} - Y_n^*) \qquad 2.16$$

This equation was used to predict the actual vapour composition for each stage of the column. With the preceding set of equations, the liquid composition for each stage was simulated for reflux, steam, and feed flow step disturbances.

From Simonsmeier's work (53) it was found that, by holding the efficiency and heat loss constant during the simulation, the final simulated state was quite different from the experimental results. Simonsmeier believed the experimental efficiency and heat loss of the final state should have differed from the initial state. Therefore, by simulations, the initial and final efficiencies and heat loss per stage were chosen to match the experimental bottom and top liquid compositions of the initial and final steady states. Simonsmeier used a scheme whereby the heat loss and efficiency were varied in a linear manner between the initial (i) and final (f) steady states, relative to a reference (r) composition, that is,

$$Q_n = Qi_n + (Xr - Xri)/(Xrf - Xri)(Qf_n - Qi_n) \qquad 2.17$$

$$E_n = Ei_n + (Xr - Xri)/(Xrf - Xri)(Ef_n - Ei_n) \qquad 2.18$$

The reference compositions tried were bottom, feed and top composition, with top composition producing the most consistent results. Simonsmeier found that a separate efficiency for the stripping and rectifying section was necessary to match experimental results. Further, a large difference in heat loss was needed to match the initial and final steady states. Based on those results, it seemed necessary to perform experimental tests on the column to verify the model for use in subsequent control studies. From the experimental tests for steam, reflux, and feed flow rate disturbances, it was necessary to modify the model. With the modifications, the constant heat loss and efficiency simulated the column dynamics well. Further, the large difference in heat loss between initial and final steady states was not required. The efficiency varied 2% to 5% between initial and final states.

2.3 Solution Procedure

The original solution procedure used by Simonsmeier was altered to obtain a more stable numerical solution at larger integration intervals (i.e. 64 seconds). As in Simonsmeier's original work, the simulation used in this thesis is based on the integration of the liquid composition only. Once the liquid composition is known, the liquid enthalpy and equilibrium vapour composition are determined. Based on the chosen efficiency, the actual vapour composition can be calculated and subsequently the vapour enthalpy. With this information determined, the liquid and vapour flows are calculated and then the iteration is repeated. The major changes to the model developed by Simonsmeier will now be outlined:

(1) Instead of assuming the reboiler liquid temperature to be constant, it was allowed to be a function of the liquid composition. The program employed a linear

interpolation to determine the temperature based on liquid stage composition.

(2) The use of the first order Euler method of integration to determine the change in the liquid enthalpy per stage, that is,

$$d(h_n)/dt = (h_n - h_{n-1})/\Delta t \qquad 2.19$$

instead of using the total derivative form of

$$d(h_n)/dt = (d(X_n)/dt)(d(h_n)/dX_n) \qquad 2.20$$

This change improved the stability of the solution and allowed the use of a longer integration interval.

(3) The use of a cubic spline (12) function instead of a linear interpolation to determine values between data points. This technique was used for the liquid-vapour equilibrium data, liquid-composition enthalpy data, and vapour- composition enthalpy data (appendix A). Using this technique, the nonlinear regions in the three curves can be described more accurately.

(4) Using the total material and energy balance differential equations (2.4 and 2.5) on each stage, a system of simultaneous equations in the form of $\underline{\underline{A}}\ \underline{X} = \underline{B}$ was obtained. The nonlinear differential equations can be

represented in a linear form because of the constant mass holdup assumption. This system of equations was used to solve for liquid and vapour flow rates by taking the inverse of matrix $\underline{\underline{A}}$. Examination of matrix $\underline{\underline{A}}$ indicates that the determinant is often near to being singular when compared to the working precision of the computer. In recognition of this fact, the method that is used to solve for liquid and vapour flow rates now employs Gaussian elimination with partial pivoting and gives some estimate of the condition (12) of the matrix. On each iteration, the condition (CON($\underline{\underline{A}}$)) of the matrix is determined by using the following method.

$$\mathrm{COND}(\underline{\underline{A}}) = \|\underline{\underline{A}}\| \, \|\underline{\underline{A}}^{-1}\| \qquad 2.21$$

Using this method, if the condition of the matrix is greater than 10^{-6}, the matrix will be considered singular according to the working precision of the computer and the solution will be terminated.

(5) The heat loss was determined experimentally and the heat loss per stage was determined as a percentage of the total experimental heat loss based on the difference between tray liquid temperature and ambient temperature on a weighted basis.

(6) Finally, the vapour enthalpy relationship for a saturated methanol-water system was changed to reflect the

study of Plewes et al (43).

With the above mentioned changes, simulation of the transient and steady-state behaviour of the column was more accurate when compared to the experimental results. In obtaining a suitable model of the column, it was decided to base the model on as much experimental data as possible. The only experimental data not obtainable was the efficiency of each tray. The required data was obtained from the mass and energy balance program (DASS1) on the column.

The liquid composition was the only differential equation which was integrated using a modified Euler method in matrix notation. Using the component material balance differential equations for each stage, the following form was obtained

$$d(\underline{X}_k)/dt = \underline{\underline{H}}_k\underline{X}_k \qquad 2{:}22$$

The matrix $\underline{\underline{H}}$ is a tridiagonal matrix (Appendix B) and the condition of the matrix indicates that it is non-singular. Now the modified Euler method of the above equation is

$$(\underline{X}_{k+1} - \underline{X}_k)/\Delta t = 1/2d(\underline{X}_{k+1})/dt + 1/2d(\underline{X}_k)/dt \qquad 2{:}23$$

or

$$(\underline{X}_{k+1} - \underline{X}_k)/\Delta t = 1/2\underline{\underline{H}}_{k+1}\underline{X}_{k+1} + 1/2\underline{\underline{H}}_k\underline{X}_k \qquad 2{:}24$$

It will be assumed that the coefficients of matrix $\underline{\underline{H}}$ will not change substantially for the integration interval that is, $\underline{\underline{H}}_{k+1} = \underline{\underline{H}}_k$

therefore, collecting like terms yields

$$\underline{X}_{k+1} = (\underline{\underline{I}} - \Delta t/2\underline{\underline{H}}_k)^{-1}(\underline{\underline{I}} + \Delta t/2\underline{\underline{H}}_k)\underline{X}_k \qquad 2.25$$

To solve for the liquid and vapour flow rates the total material and energy balance differential equations were used. Writing the equations for each stage results in a 18x18 matrix of the form shown below.

$$\begin{bmatrix} \underline{\underline{A}}_1 & \underline{\underline{A}}_2 \\ \underline{\underline{A}}_3 & \underline{\underline{A}}_4 \end{bmatrix} \begin{bmatrix} \underline{L}_n \\ \underline{V}_n \end{bmatrix} = \begin{bmatrix} \underline{FT}_n \\ \underline{FT}_m \end{bmatrix} \qquad 2.26$$

Detailed entries of the matrix are given in Appendix B. The liquid and vapour rates are solved for by using Gaussian elimination with partial pivoting to decompose the matrix into an upper triangular matrix, then back substituting to solve for the flow rates.

Chapter 3

EXPERIMENTAL WORK

3.1 Introduction

Originally, simulations using the dynamic model developed by Simonsmeier were to be used in helping to select the controller constants for subsequent experimental evaluation of dual control. Since major modifications to the model were implemented, it became necessary to verify the performance of the model through experimental work. The simulated results were compared to experimental work for step disturbances in feed, reflux and steam flow rates.

Initially, the assumptions of constant heat loss and efficiency based on the initial steady state were used. Later, variable heat loss and efficiency was tried. The first case will be termed as the constant nonlinear model while the second will be the variable nonlinear model. Experimental data was used as input for the theoretical model in order that the comparison have some basis within the theoretical framework.

3.2 Experimental Equipment

The pilot plant distillation column originally built by

Svrcek (57), has been used by various co-workers throughout the years. The column has eight bubble cap trays that are used to separate a 50% methanol-water feed entering the fourth tray. The trays are spaced 30.5 cm apart and contain four bubble caps positioned in a square configuration with centers at 5.40 cm radius from the centre of the tray. Further description of the design can be obtained from Svrcek (57).

The tower has a thermosyphon reboiler and a total condenser, with their respective liquid levels being controlled by manipulating the bottom and top product flow rates. The tower pressure was maintained by adjusting the cooling water flow through the condenser. To maintain constant feed and reflux enthalpies, two steam preheaters under PID analog control were positioned on the feed and reflux streams (Figure 3.1). Both liquid levels and column pressure are controlled by PI analog controllers.

Steam, reflux, and feed flow rates can either be controlled by a standard DDC loop on the IBM 1800 or by analog controllers. Usually the three flow loops were under a supervisory type of control with the output of the DDC loops being used as the setpoint for the analog controllers. A list of DDC and data acquisition loops is presented in Appendix C. The procedure used for the calibration of flow rates and the analog controllers are also found in Appendix C. The top composition was measured by a capacitance probe. This calibration procedure and resulting chart is again

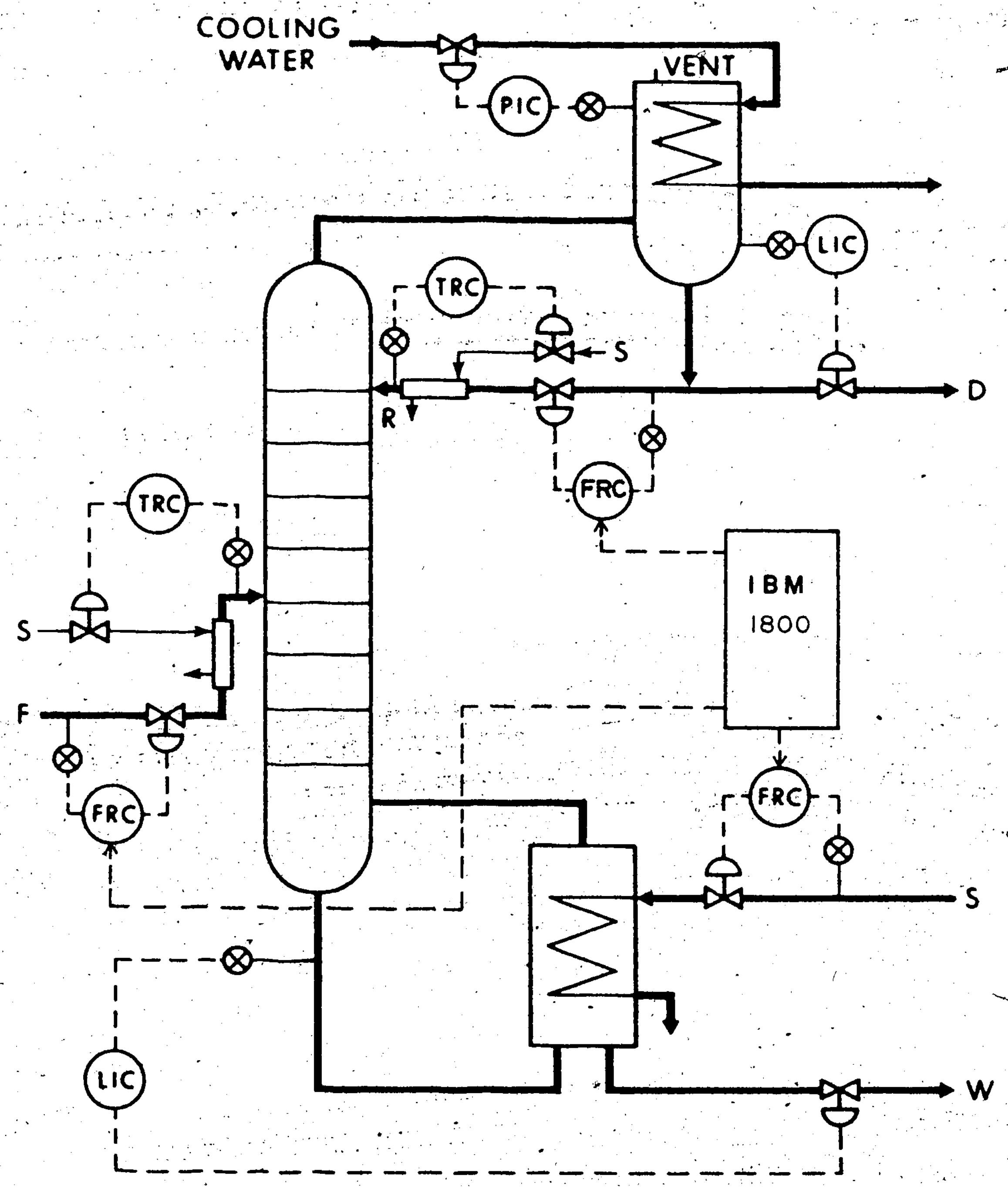


Figure 3.1 Schematic diagram of the distillation column.

given in Appendix C.

The bottom composition was analyzed using a Hewlett-Packard (5702A) gas chromatograph equipped for liquid sampling. Top composition values could be collected at one second intervals if desired, but the composition of the bottom stream could only be obtained at 128 second intervals. The analysis of the bottom composition has historically presented problems when modelling or controlling the distillation column. The previous gas chromatograph sampling system had a 256 second sample time with about 16 minutes dead time due to transportation time. Numerous hours were spent in decreasing the analysis time by one-half. Also, the accuracy was improved by using a different column packing. The sampling system dead time was reduced to approximately 10 seconds. Detailed information regarding the gas chromatograph system is given in Appendix D.

3.3 Experimental Procedure

In performing the open loop experimental work, a procedure was adopted that would ensure consistency in all runs. First, the experimental work was performed at operating conditions that would produce a top composition of about 96% methanol and a bottom composition of about 4%

methanol. These operating conditions are summarized in Table 3.1 and are average values of the initial steady state values for the first six experimental runs.

At the beginning and at the end of each experimental run, the feed composition was analyzed to ensure no change had taken place in the composition. In all runs, the initial and final feed compositions were identical. Once the feed composition had been analyzed, the bottom sample system pump was started and gas chromatograph automatic sampling was initiated. Before any disturbance was introduced, the gas chromatograph was usually operated for about 30 minutes to ensure proper operation and steady state composition. During this time, values of steam, feed, and reflux flows were obtained. This was to insure that all three variables were also at steady state and to obtain an average current value for the variable to be disturbed. The mass and energy balance program 'DASS1', (see Appendix E), was also initiated at the beginning and end of each experimental run. Using the program 'DISTB' and the steady state values, the magnitude of each disturbance was entered and the time of the disturbance was recorded on the teletype. The program 'DWL44' printed the top and bottom composition values at 128 second intervals. The top and bottom composition values were also collected by the data accumulation loops.

Table 3.1 Steady State Operating Conditions

reflux flow	12.1	g/s
steam flow	14.6	g/s
feed flow	18.0	g/s
feed composition	50.6	g/s
bottom composition	4.3	% methanol
top composition	95.8	% methanol
bottom product flow	8.9	g/s
top product flow	8.9	g/s
steam temperature	106.4	° C
reboiler temperature	93.4	° C
tray temperature 1	83.1	° C
tray temperature 2	77.1	° C
tray temperature 3	73.3	° C
tray temperature 4	71.3	° C
tray temperature 5	69.3	° C
tray temperature 6	64.9	° C
tray temperature 7	63.5	° C
tray temperature 8	61.6	° C
condenser temperature	59.6	° C
feed enthalpy	229.3	J/g
steam enthalpy	2683.5	J/g
reflux enthalpy	152.1	J/g
heat transfer	2436.9	W-°C
heat loss	8234.8	W

3.4 Simulation Procedure

In simulating the experimental column dynamic behavior, the experimental steady state data was used for each run. Values for flow rates and enthalpies for steam, feed, and reflux, and also methanol percentages for feed, bottom, and top composition were obtained from the output of the mass and energy balance program. The product of the heat transfer coefficient (U) and area (A) was calculated by using equation 2.10. It was assumed that the reboiler would have an efficiency of 100.0% and the condenser would have an efficiency of zero.

On the basis of the bottom sample loop flow rate and the resulting temperature drop, the heat loss of the reboiler was estimated to be approximately equal to 300 W. An extra 200 W was added in order to allow for the possibility of extra heat loss in the reboiler section. The remaining experimental heat loss was distributed among the trays as a function of the difference between the tray and room temperature (25 °C) on a weighted basis. The experimental liquid mass holdup was measured at the steady state operating conditions using the procedure described in Appendix F. The efficiencies for all the trays were assumed to be constant and at the same value for both the stripping and rectifying sections. The efficiency was the only variable for which no experimental value was available. The efficiency was chosen in such a way as to ensure that the simulated bottom and top compositions matched the experimental compositions.

3.5 Comparison Between Experimental and Simulated Results

The crucial test for any model is comparing the simulated results to experimental data. Step disturbances in steam, reflux, and feed flow rates were introduced into the pilot plant column. The behaviour of top and bottom compositions was monitored. In all the graphs to be presented, only the top and bottom compositions will be shown for the step disturbances. The integration interval for all of the simulations was 64 seconds, unless otherwise stated. The experimental values for top and bottom composition were obtained every 128 seconds (gas chromatograph cycle). The steam, reflux and feed flow rates were disturbed in such a way as to first cause an increase in bottom and top composition. Next, the variable was returned to its original steady state value. The differential equation model does not predict the time delay in either the top or bottom compositions, therefore the simulation results were shifted by an amount equal to the time delay observed in the experimental responses (Table G.1). The time shift was thus equal to the difference between the time of disturbance(arrow on figures) and when the simulated response changes from its initial value. The top composition time delay was at about 36 seconds for all disturbances. The dead time in the bottom composition was the measurement time delay (128 seconds), plus the process time delay. The process time delay varied depending upon the variable being disturbed.

The model used was a constant liquid holdup model with the option of either a constant or variable heat loss and efficiency. Both the heat loss and efficiency had to be adjusted in order to match the simulated top and bottom compositions to experimental values. Figure 3.2 compares the simulated transient response for bottom and top compositions to the experimental response for a 7.5% decrease in steam flow rate. The simulated results are shown for both the constant and the variable heat loss and efficiency models. For this graph the bottom simulated results were shifted 5.40 minutes from the time of disturbance while the top simulated values were shifted 0.60 minutes. The bottom composition was slightly under the final experimental steady state value, while the top composition was above the final experimental steady state value for the constant nonlinear model. Using the variable nonlinear model, the final steady states are identical to the experimental values. The simulated response of the bottom composition behaves like a first order system while the experimental response definitely has the characteristics of a second order system. The transient response difference

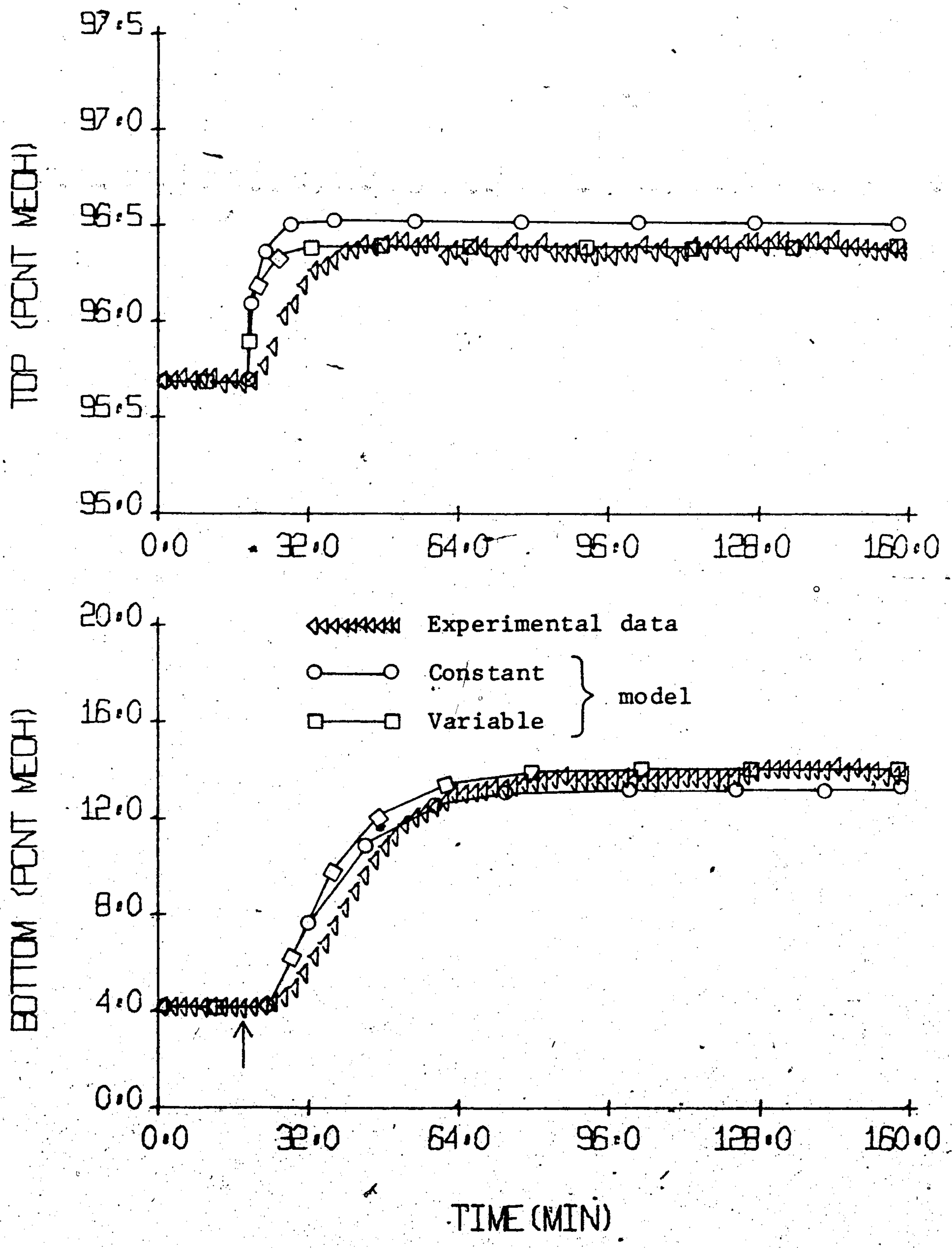


Figure 3.2 Comparison of simulated constant and variable model responses to experimental data for a -7.5% step disturbance in steam flow rate.

between simulated and experimental top composition was greater than that of bottom composition.

Figure 3.3 shows how the integration interval affects the transient behaviour of the top composition. The integration interval was decreased to two seconds. This brought the simulated results closer to the experimental results. This figure also shows that the top composition is more sensitive to the integration interval than the bottom composition. The bottom composition did not differ from that of the 64 second integration interval case. The integration interval was halved five times from 64 seconds to 2 seconds with improvement being noticed in the transient behaviour of the top composition until the 2 second interval was reached. The computer cost for using an interval of 2 seconds was approximately $42.00 compared to $2.50 using a 64 second interval. Therefore to conserve on computing funds, the 64 second interval was chosen.

Appendix G contains the computer mass and energy balance summaries for the data presented in this chapter. In the case of a decrease in steam rate, as can be seen from Figure G.3, the steam temperature does vary during the disturbance. Consequently the steam temperature (Ts) was varied in the models according to equation 2.10. Similarly the changes in reboiler temperature during the disturbance were related to the change in liquid composition. Both condenser and reboiler level control loops were tuned to allow about 10% overshoot without excess oscillations of the

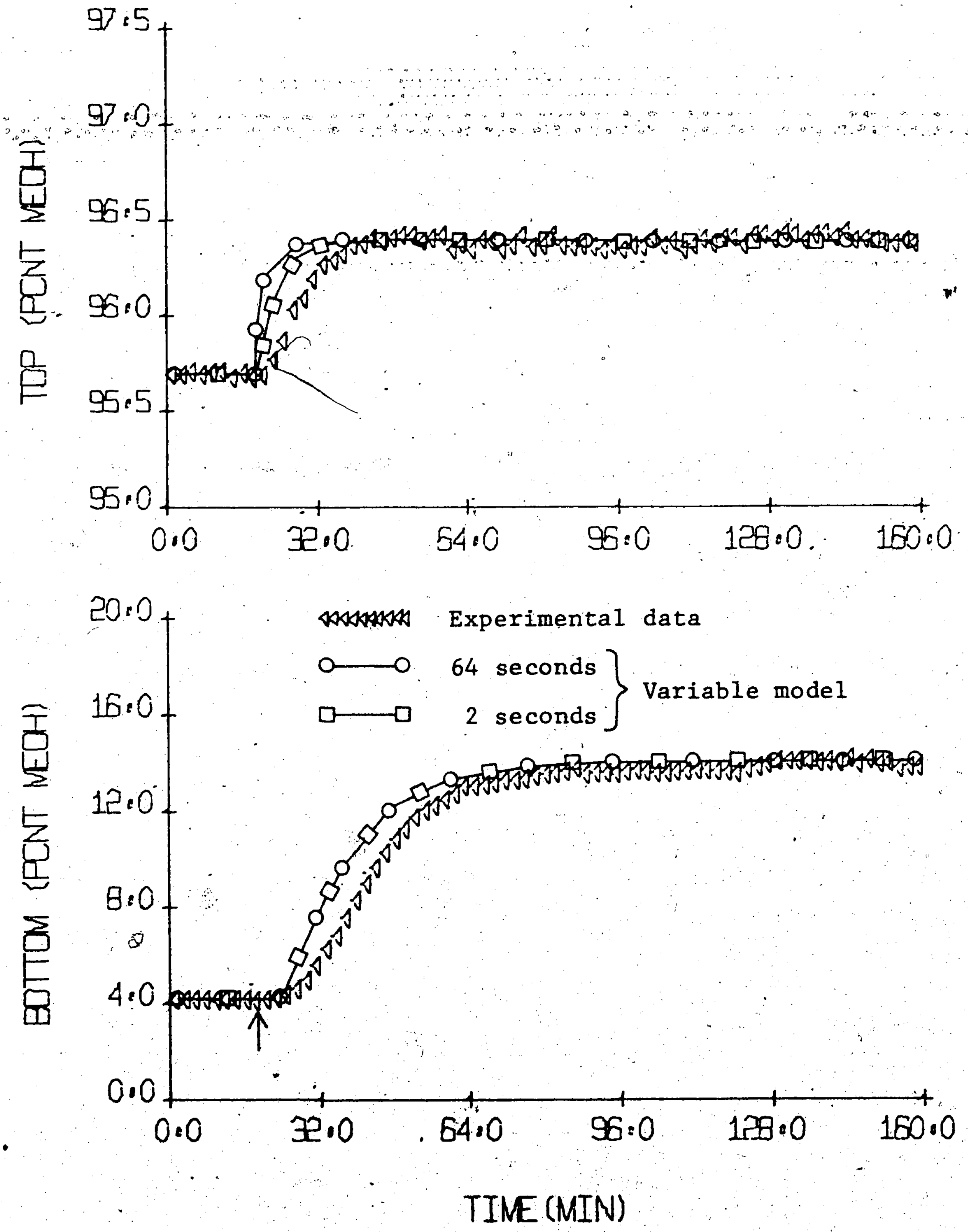


Figure 3.3 Comparison of simulated variable model responses using different integration intervals to experimental data for a -7.5% step disturbance in steam flow rate.

manipulated variable (see Figure G.4). Figure 3.4 shows the top and bottom composition behaviour when the steam flow rate was returned to its original value. The experimental data was compared to both the constant and variable nonlinear models and again, the constant nonlinear model is very close to the experimental data. Located in Appendix G is the computer printout for the final steady state values for the increase in the steam flow case.

Figure 3.5 compares the constant and variable nonlinear model responses to experimental results for a 15% increase in reflux flow. The constant case shows the simulated top composition going above the final experimental steady state value while the opposite was true for the simulated bottom composition. By varying heat loss and efficiency, the simulated values were brought nearer to the final experimental steady state values. Decreasing the integration interval to two seconds (Figure 3.6) improved the top composition transient response. Again plots of the steam flow rate, tower pressure, feed, and reflux flow rates during the run are located in the Appendix G. Figure 3.7 compares the simulated transient response to the experimental response when the system's reflux flow rate was returned to its normal steady state value.

Figures 3.8 and 3.9 show the responses of an increase and a subsequent decrease for a 20% disturbance in feed flow. This was probably the most critical test for the model, in that the experimental results showed that the top

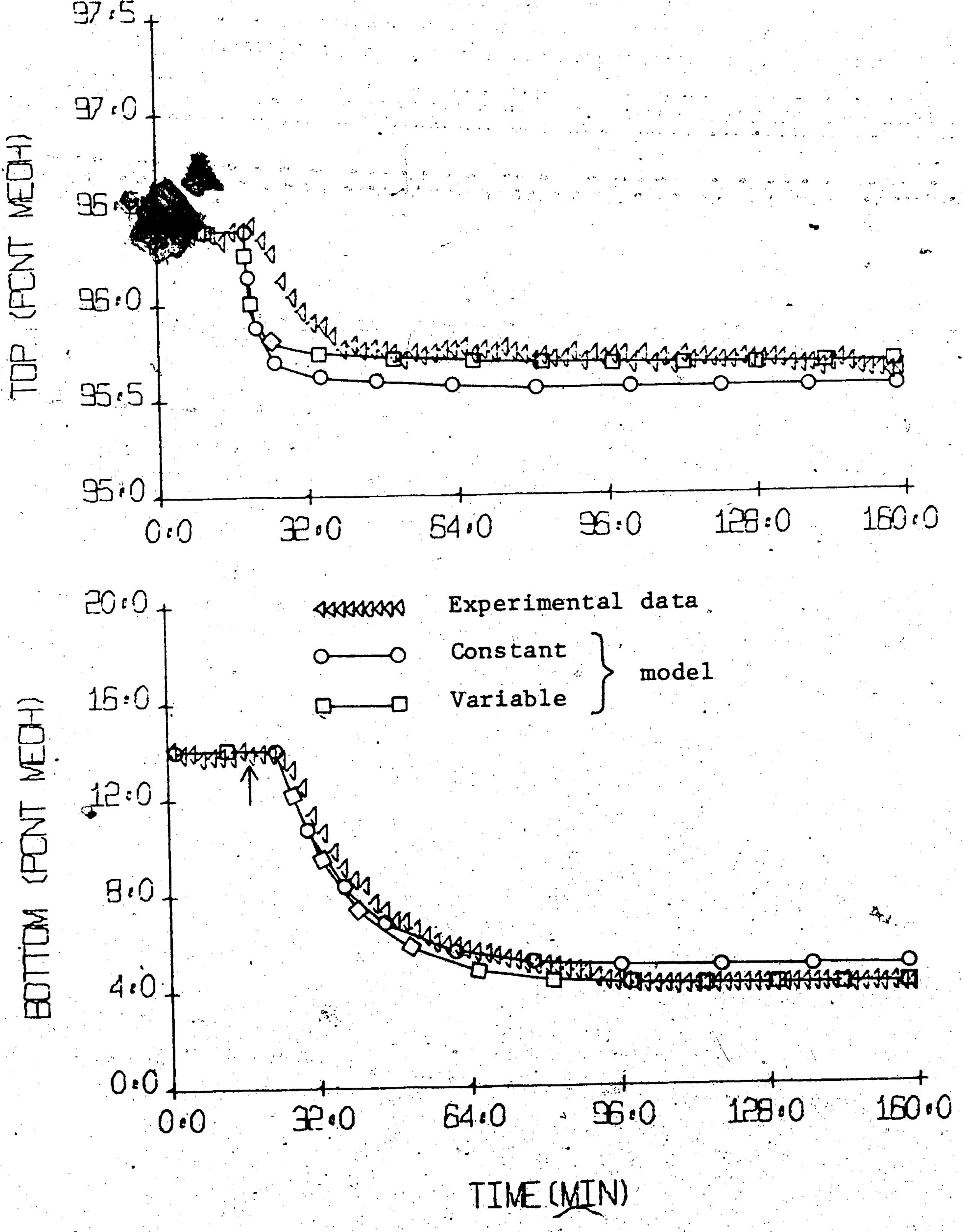


Figure 3.4 Comparison of simulated constant and variable model responses to experimental data for a +7.5% step disturbance in steam flow rate.

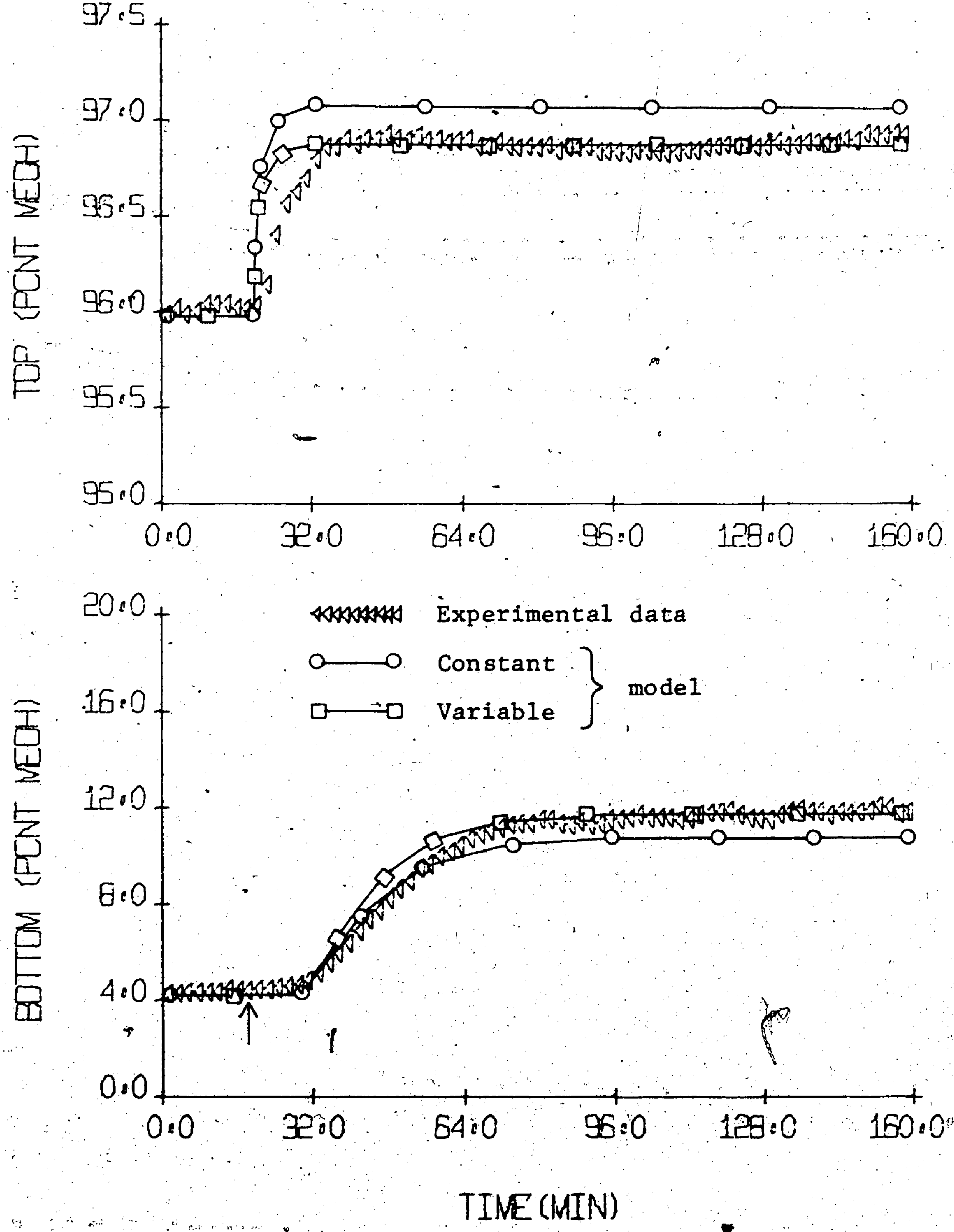


Figure 3.5 Comparison of simulated constant and variable model responses to experimental data for a +15% step disturbance in reflux flow rate.

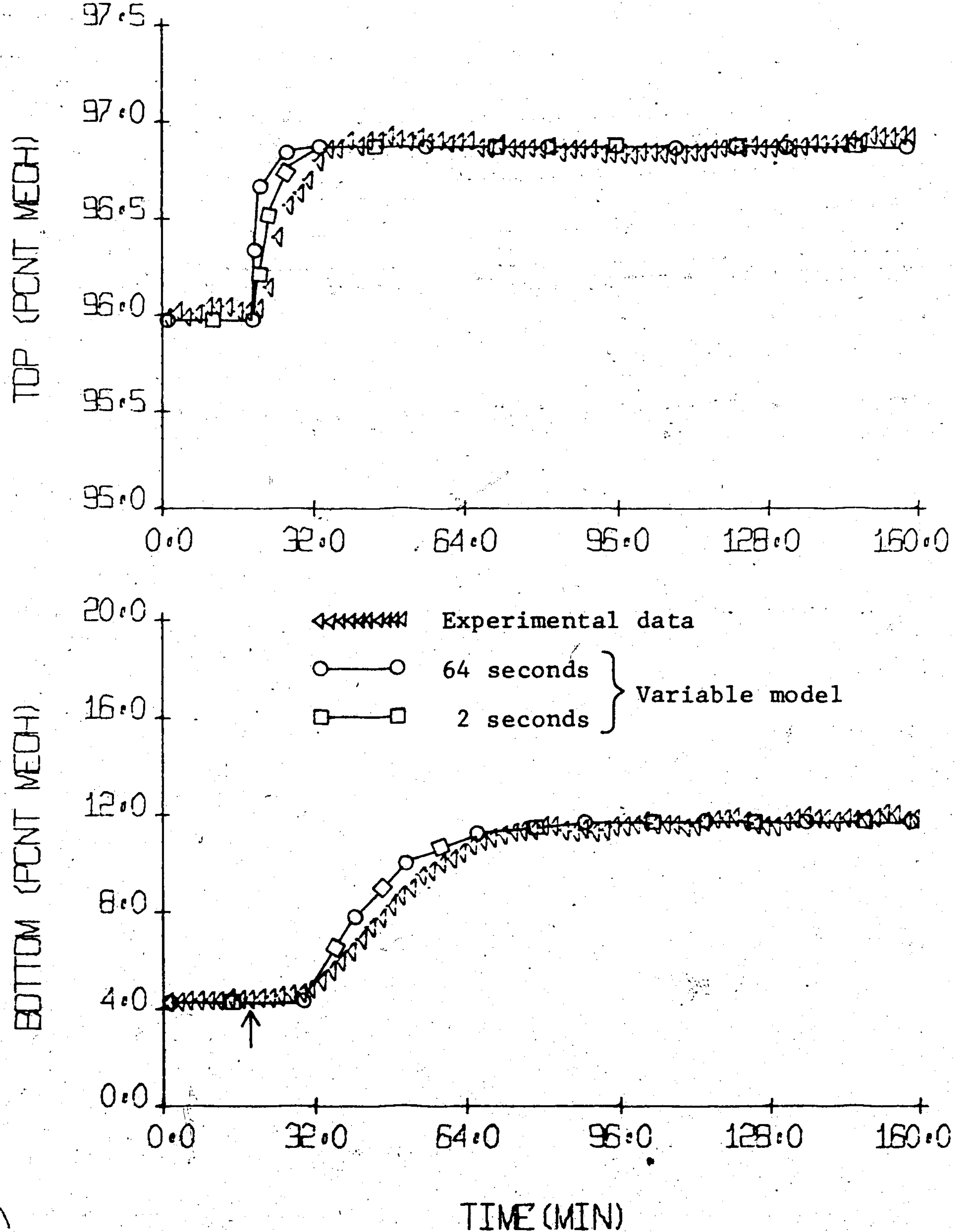


Figure 3.6 Comparison of simulated variable model responses using different integration intervals to experimental data for a +15% step disturbance in reflux flow rate.

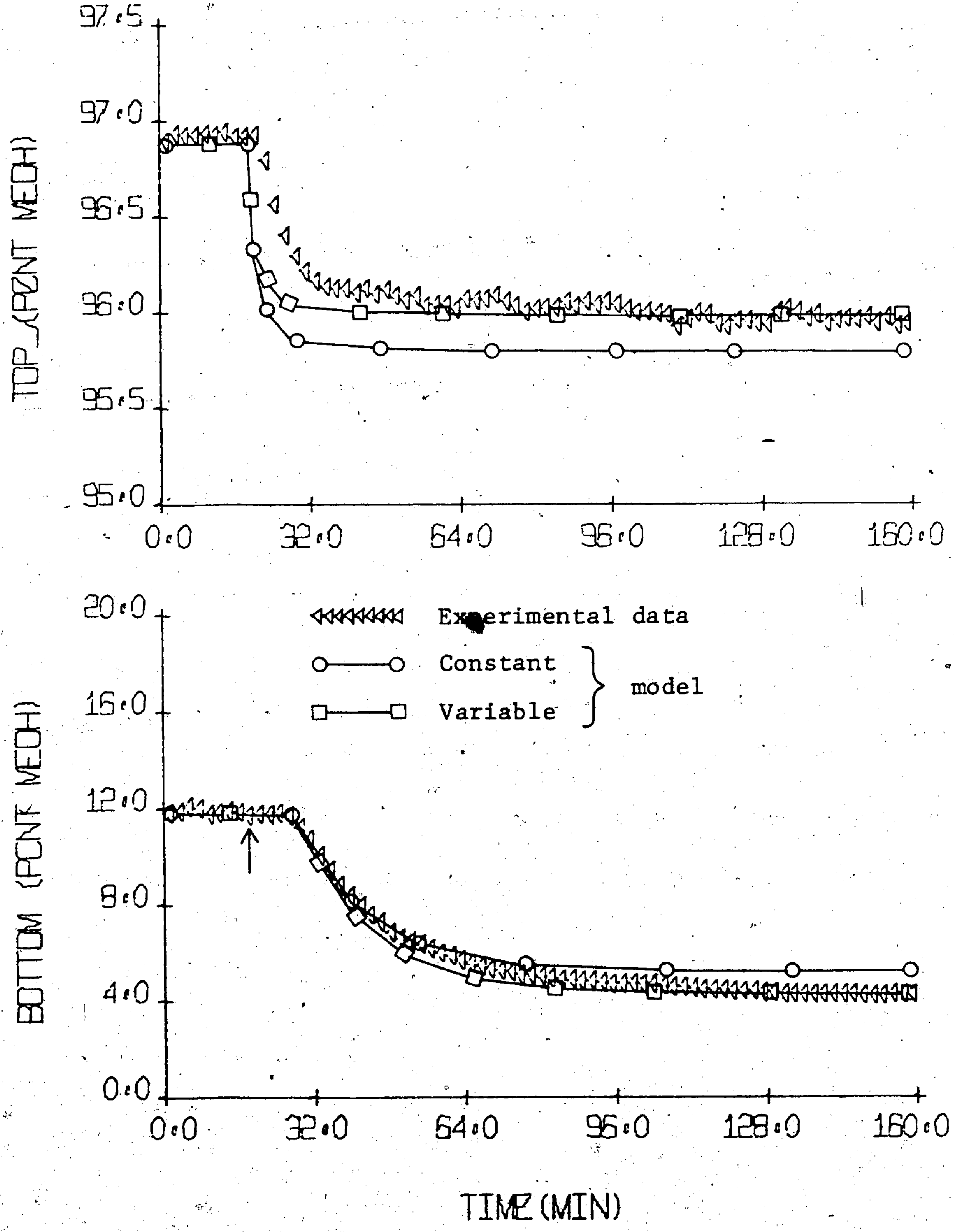


Figure 3.7 Comparison of simulated constant and variable model responses to experimental data for a -15% step disturbance in reflux flow rate.

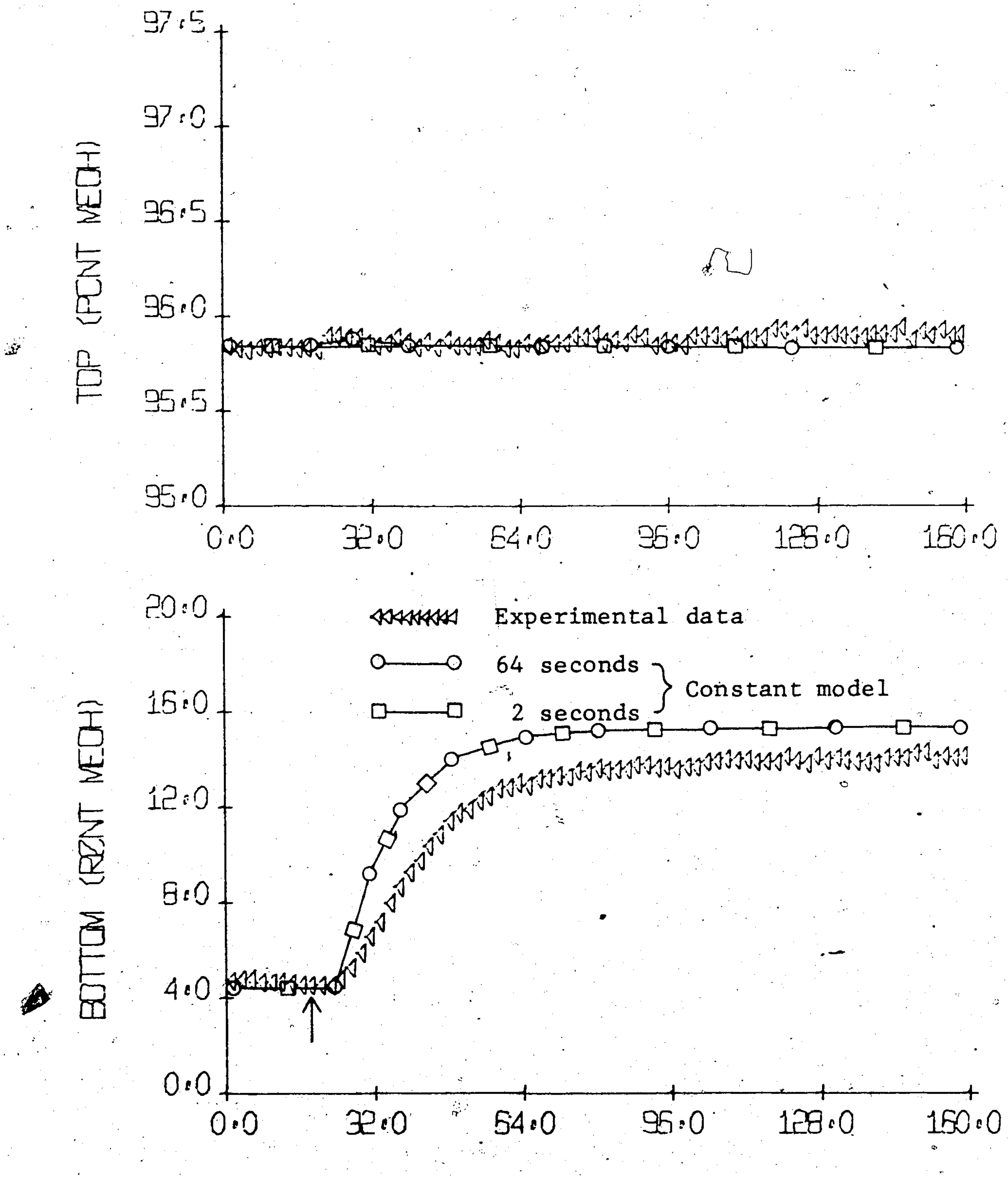


Figure 3.8 Comparison of simulated constant model responses using different integration intervals to experimental data for a +20% step disturbance in feed flow rate.

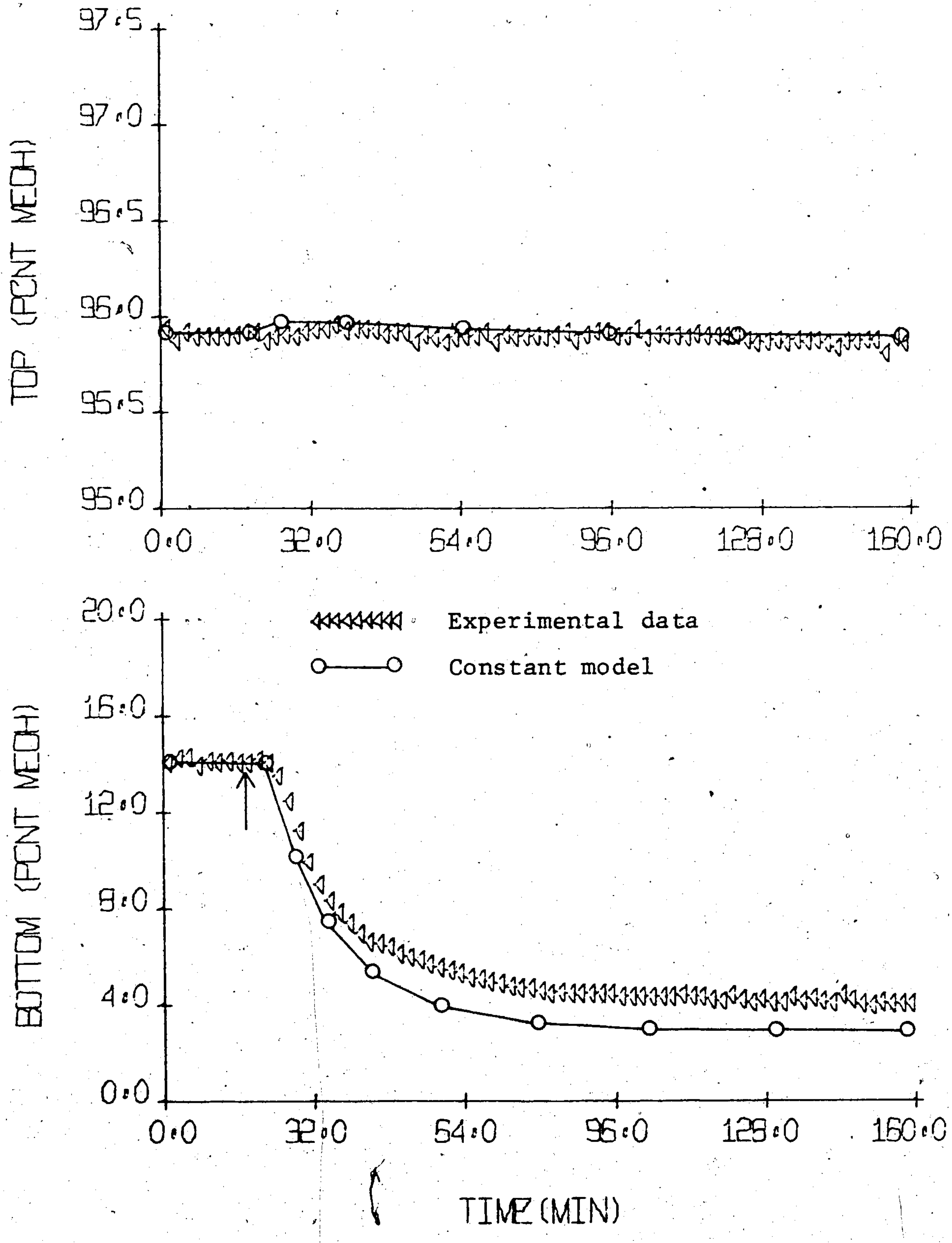


Figure 3.9 Comparison of simulated constant model response to experimental data for a -20% step disturbance in feed flow rate.

composition remained virtually unchanged despite the change in feed flow rate. The simulated top composition response for the constant nonlinear model, at both the 2 and 64 integration intervals compares well with the experimental responses. Since the top composition does not change, the variable nonlinear model would produce the same results because it is based on the top composition. For the feed disturbance studies, the reference composition could have been changed to the bottom composition so that the variable nonlinear model would have an effect on the bottom composition. In the case of bottom composition, as can be seen from Figures 3.8 and 3.9 there was some difference between the experimental and simulated responses in the final steady state for the increase in feed rate. This difference was smaller for the decrease case. Plots of steam, tower pressure, reflux flow and feed flow rates are located in Appendix G for this feed flow disturbance.

Figures 3.10 and 3.11, 3.12 show duplicate results for decreasing steam, and increasing reflux and feed flows. Both the constant and the variable nonlinear models are compared to experimental data for steam and reflux disturbances. Only the constant heat loss and efficiency model was used for the feed flow disturbance. The top and bottom composition match closely the experimental data for the constant nonlinear model. The duplicate runs show results similar to the initial experimental runs.

Figures 3.13 and 3.14 show the transient behaviour when

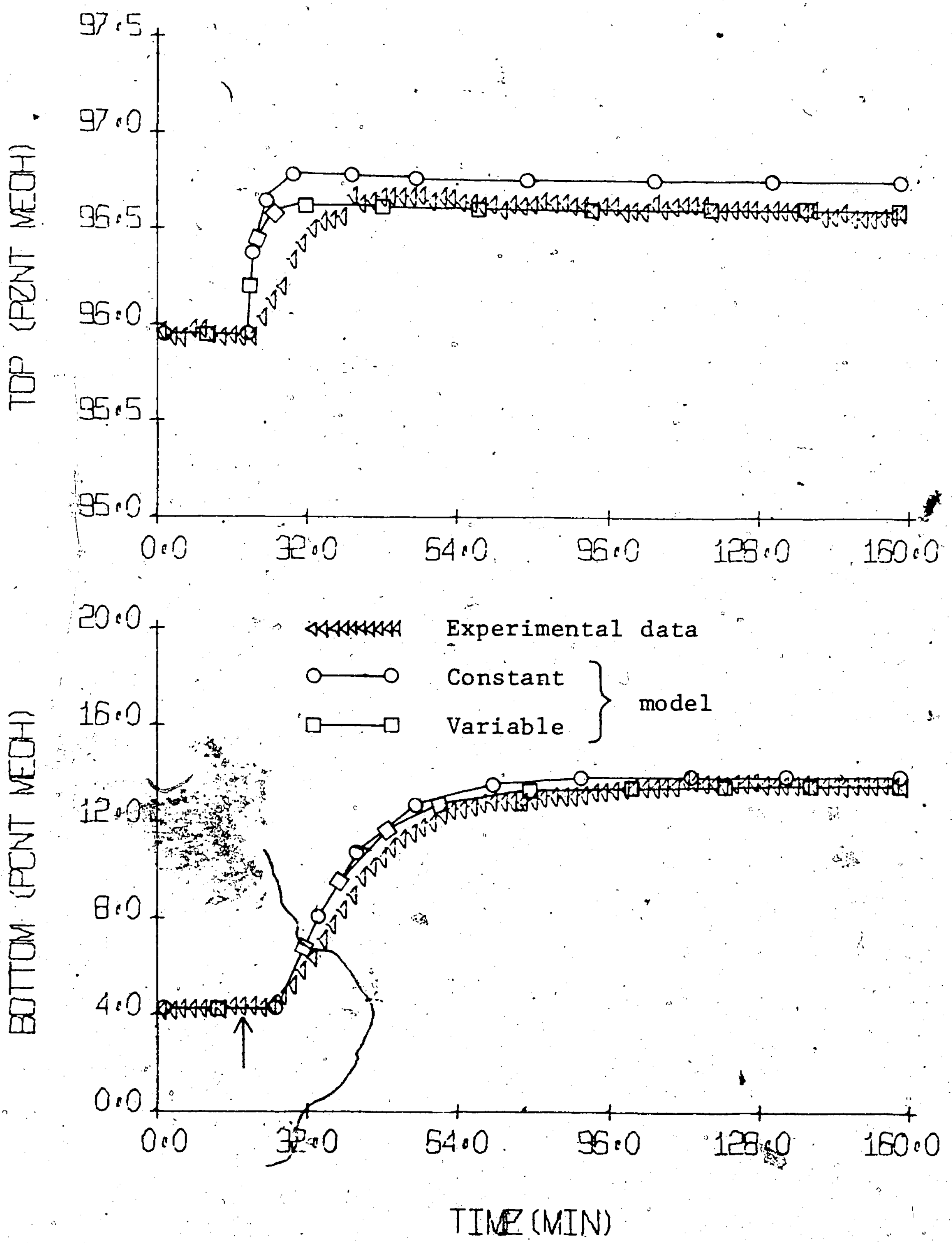


Figure 3.10 Comparison of simulated constant and variable model responses to experimental data for a -7.5% step disturbance in steam flow rate.

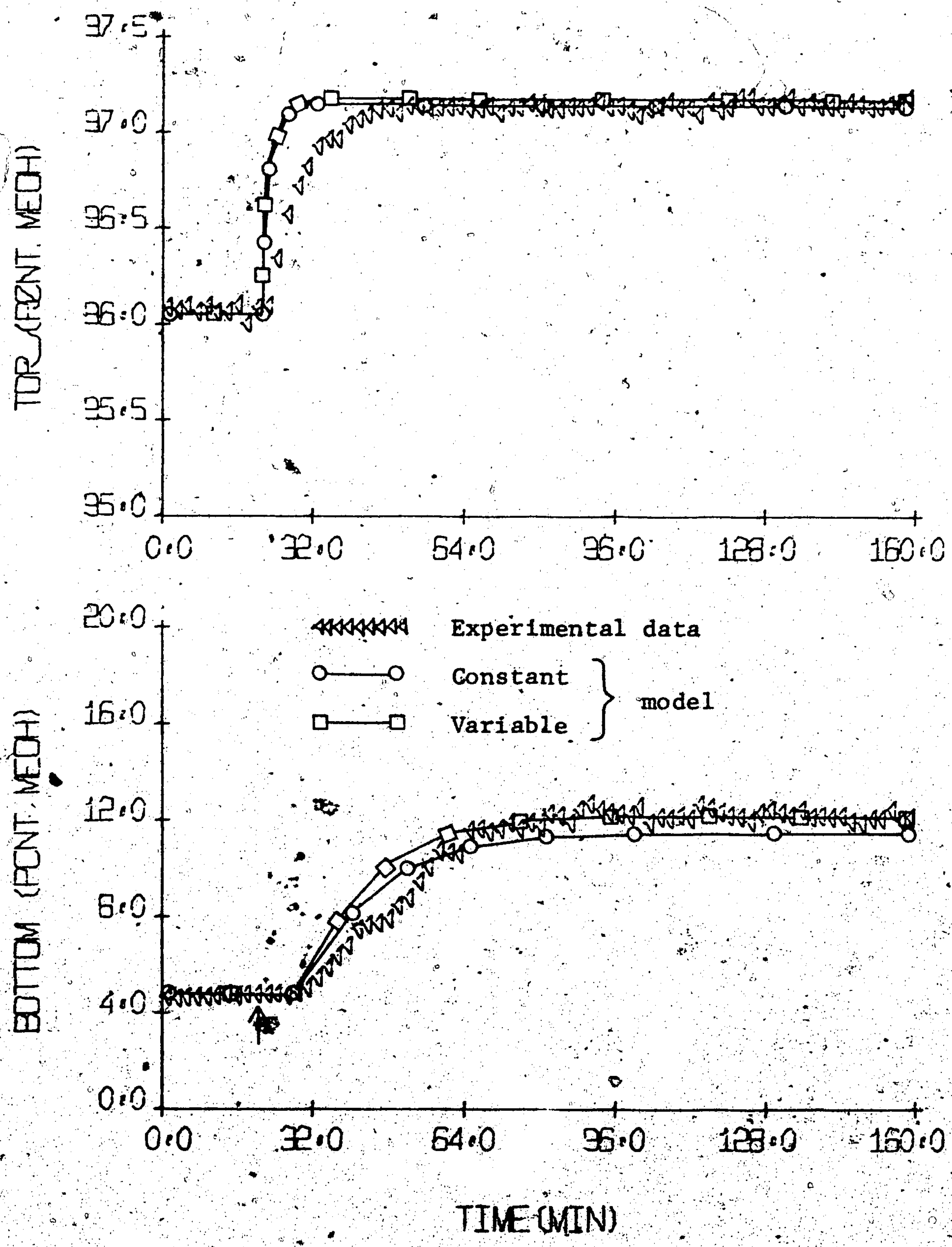


Figure 3.11 Comparison of simulated constant and variable model responses to experimental data for a +15% step disturbance in reflux flow rate.

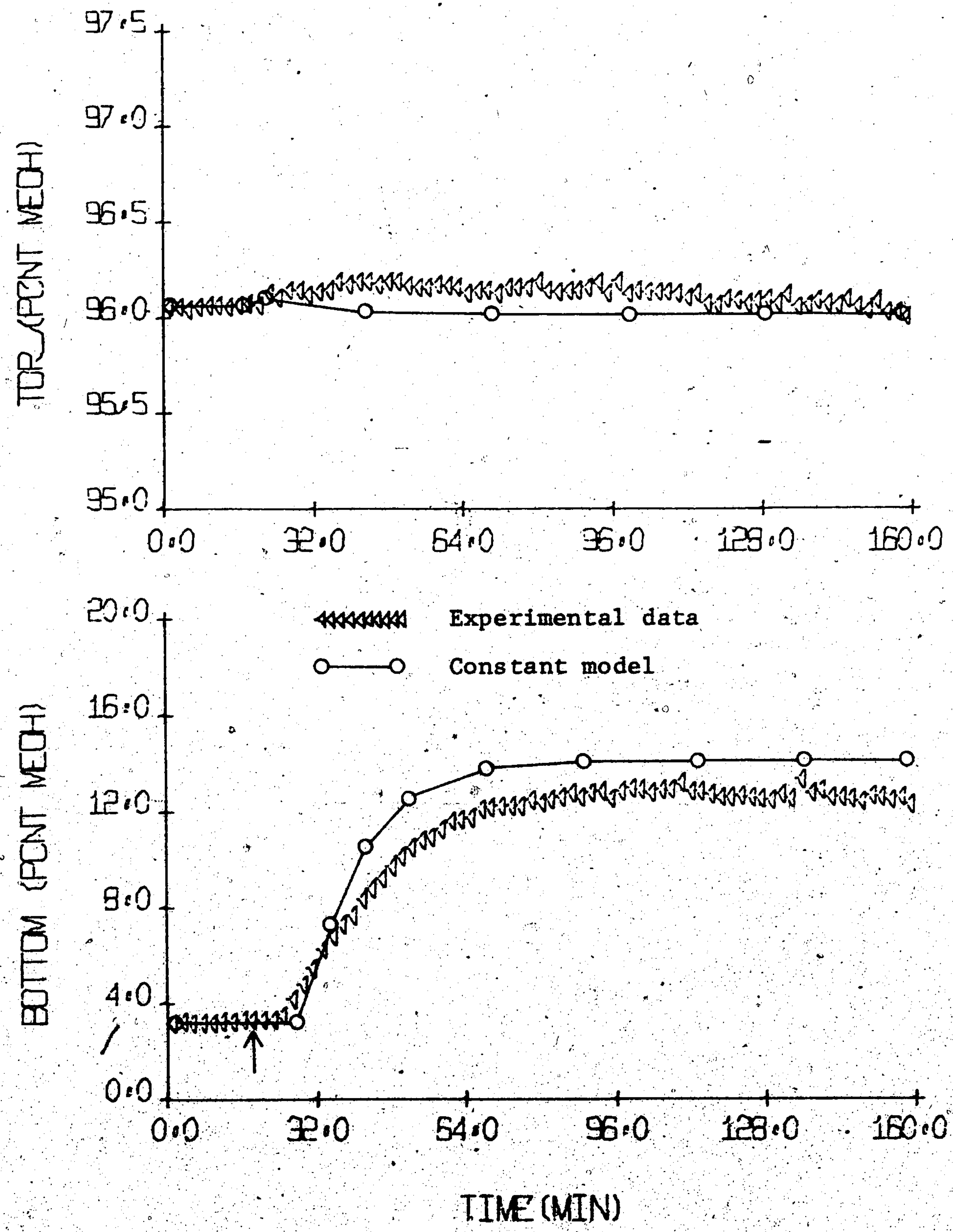


Figure 3.12 Comparison of simulated constant model response to experimental data for a +20% step disturbance in feed flow rate.

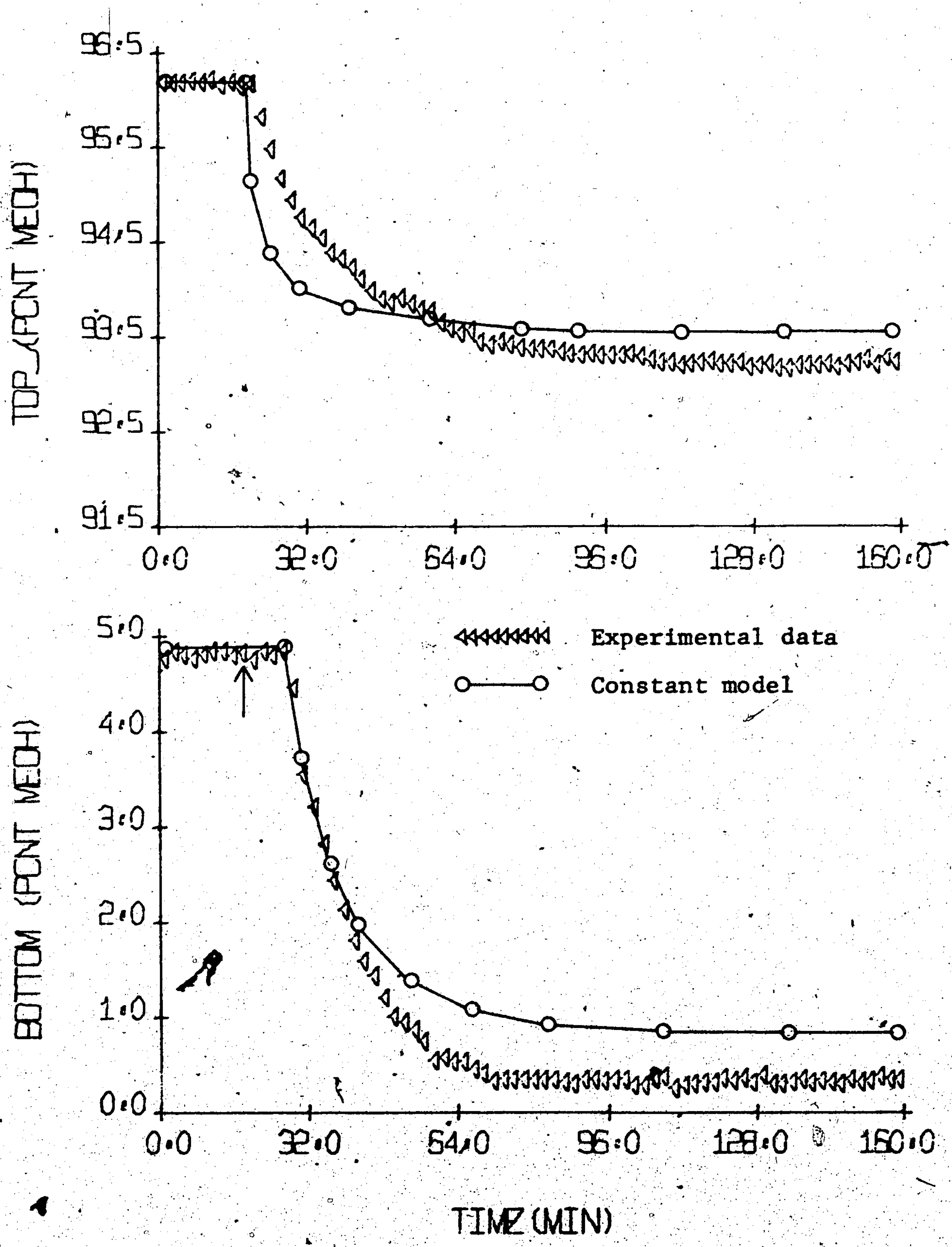


Figure 3.13 Comparison of constant model response to experimental data for a 15% decrease in reflux flow below its normal operating value.

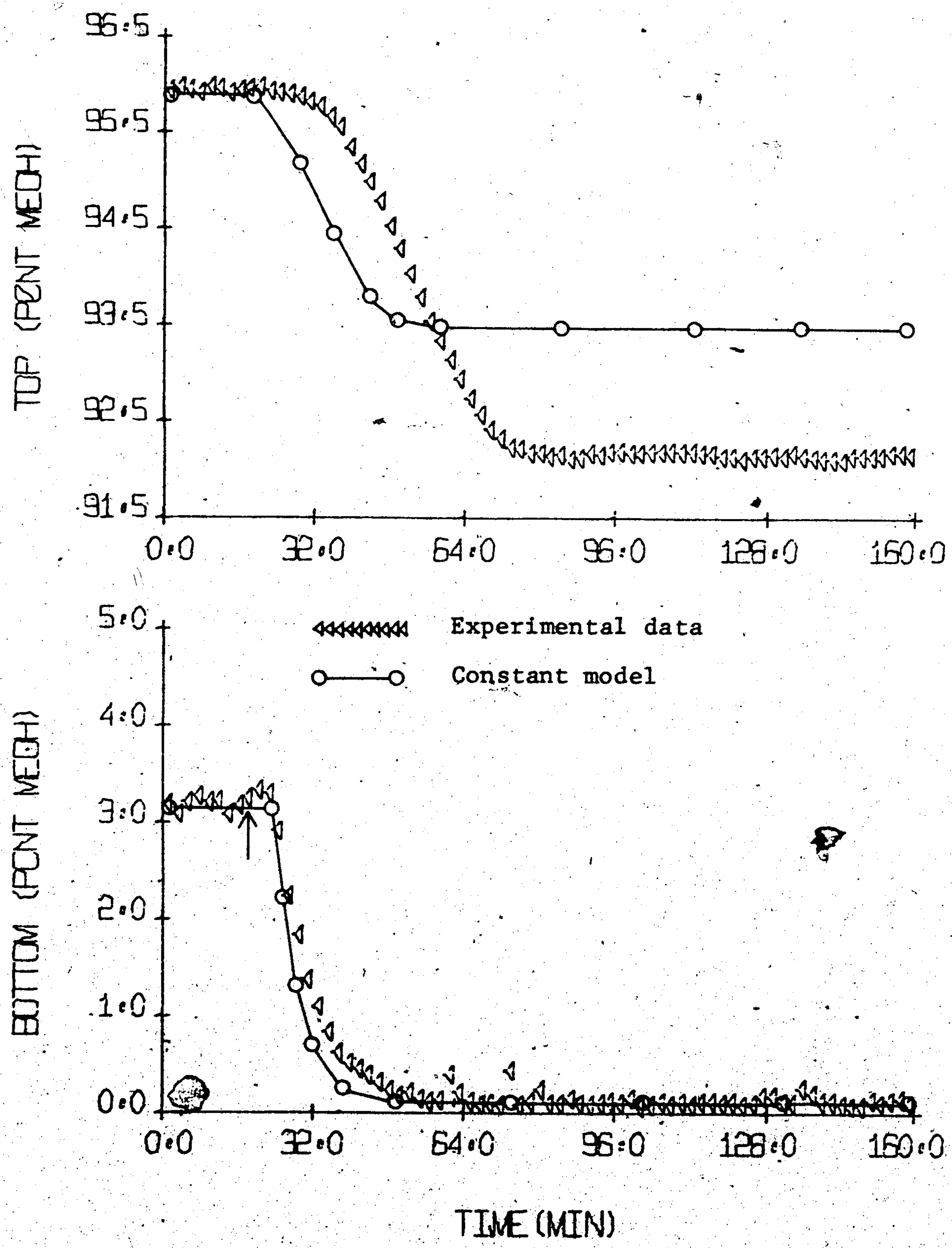


Figure 3.14 Comparison of constant model response to experimental data for a 20% decrease in feed flow below its normal operating value.

the reflux and feed flow rates were decreased below their normal operating values. The constant heat loss and efficiency model predicted the bottom experimental composition well for the feed disturbance only. However, there were large deviations between the predicted response and the experimental response for top composition. The most drastic differences in predicted versus experimental responses occurred for changes in the flow rate of steam and feed for top composition. A 7.5% increase in steam flow rate above normal operating values is not shown because the top composition response was out of the range of the analyzer. When in this operating region, the column has reached a physical constraint, that is, with the bottom solution being virtually methanol free, the additional heat input to the reboiler continues to vapourize the solution. This results in more water being driven up the column which continues to lower the top composition.

3.6 Discussion of Results

The major shortcoming of the constant nonlinear model is its inability to predict the final top composition steady state values for steam and reflux disturbances, and bottom composition in the case of an increase in feed flow rate. The variable heat loss and efficiency model did well for reflux and steam disturbances but was not completely satisfactory for feed disturbances because of the reference composition chosen. Since the top composition was insensitive to a feed disturbance, that is, it remained the same, varying heat loss and efficiency with top composition would have no effect. The constant nonlinear model predicted the bottom composition well for decreases in steam and reflux flow rates. In the case of the duplicate open loop runs, simulations using the constant nonlinear model predicted small deviations when compared to results for experimental top and bottom composition (Figure 3.11). However, the model did not satisfactorily predict the top and bottom composition for increases in steam (Figure 3.10) and feed (Figure 3.12) disturbances respectively.

Experimental tests were conducted in which the reflux, feed and steam flows were changed to cause a decrease in the top and bottom compositions below their normal operating values. The constant nonlinear model predicted the final steady state quite well for the reflux (Figure 3.13) disturbance, however values for both top and bottom experimental responses were slightly below the simulated final values. In the case of a feed disturbance(Figure 3.14) to 20% below its normal value, the constant nonlinear model satisfactorily predicted the bottom composition but the predicted top composition response exhibited a large final steady state deviation. The same situation occurred with the steam disturbance, however the deviation was even more pronounced. This phenomenon can be explained by the fact that for a given heat input to the reboiler for a decrease in feed rate all the methanol will be vapourized. Thus, when the bottom solution becomes depleted in methanol as indicated by Figure 3.14, the remaining energy input goes toward evaporating more water, thus lowering the concentration of methanol in the top product.

The same analysis can be applied to the steam disturbance with more drastic results. Attempts were made to use the variable nonlinear model for these cases, but unrealistic heat losses and efficiencies were necessary in order to match experimental values for top and bottom composition. As long as the methanol composition remained above 0.2%, the constant heat loss and efficiency model predicted the final steady state quite well. In all simulated results, the heat loss had to be reduced to between 65% and 75% of the experimental values. The procedure used was to take about 60% of the total experimental loss value and divide it on a weighted basis of the difference between the tray and room temperatures starting with stage 2 to stage 9 inclusive. The resultant heat loss value for stage 9 was then multipled by 2.25. The temperatures for the different trays can be obtained from the mass and energy balances for each type of disturbance which is located in Appendix G. The heat loss for stage 1 and 10 was 500.0 W and 0.0 W respectively. Table 3.2 shows the initial steady state heat loss values for stages 1

Table 3.2 Distribution of Heat Loss for Disturbances

Stage	Steam W	Reflux W	Feed W
1	500.0	500.0	500.0
2	694.7	760.5	714.5
3	622.8	683.9	638.9
4	577.4	633.3	591.9
5	551.1	607.4	564.8
6	528.4	582.7	543.6
7	474.6	525.6	485.4
8	456.7	506.2	473.0
9	1025.2	1112.9	1138.9
Total	5430.8	5912.5	5151.0
Experimental	8268.1	8315.1	8121.1

through 9 for steam, reflux and feed disturbances. The values for each disturbance are not the same because each type of disturbance was treated separately.

Table 3.3 shows the changes in heat loss and efficiency required to use the variable heat loss and efficiency model. In all cases one efficiency was used for all the trays and only the top tray heat loss required adjustment in order to match the final steady state values to the experimental top and bottom compositions. The largest heat loss difference was for the feed disturbance, while the largest change in efficiency between initial and final states occurred for the reflux disturbance. Table 3.4 shows the model's sensitivity in terms of the product compositions to small deviations in the model parameters and process inputs for top and bottom compositions of 95.985% and 4.267% methanol respectively. The greatest change occurred for ±2% changes in the steam flow rate, while the feed and reflux changes of ±2% showed the next most significant effect. The sensitivity of the model clearly shows the importance of accurate experimental work.

Table 3.3 Tray 9 Heat Loss and Efficiency Values for the Variable Nonlinear Model

	initial	final
steam disturbance		
heat loss(W)	1025.2	1070.2
efficiency	89.1	86.0
reflux disturbance		
heat loss(W)	1112.9	1132.9
efficiency	92.4	88.1
feed disturbance		
heat loss(W)	1138.9	838.9
efficiency	90.5	94.0

Table 3.4 Model Sensitivity to Changes in Process Inputs and Parameter Values

		% Methanol	
	% Change	Bottom (4.267)	Top (95.985)
F (18.055 g/s)	+2	5.465	96.009
	-2	3.234	95.927
Re (12.087 g/s)	+2	5.061	96.168
	-2	3.630	95.771
St (14.577 g/s)	+2	2.520	95.662
	-2	6.651	96.228
Xf (50.012% meth)	+2	5.962	96.086
	-2	3.124	95.787
Hfe (231.963 J/g)	+5	3.672	95.868
	-5	4.981	96.084
Hre (154.129 J/g)	+5	4.069	95.923
	-5	4.516	96.047
UA (2468.039 W-°C)	+5	4.141	95.967
	-5	4.435	96.007

( ) - the value in parentheses is the steady state value

Chapter 4

TRANSFER FUNCTON MODEL

4.1 Introduction

Interaction exists between the top and bottom composition control loops which presents problems in a multiloop control scheme. Three different compensators were evaluated in order to determine which was most effective in reducing this interaction. The direct Nyquist method was used to design only the plant precompensator. Before the plant precompensator could be designed it was necessary to obtain a transfer function model of the column.

4.2 Transfer Punction Model

Pseudo random binary sequence (PRBS) tests were carried out on the distillation column to obtain a transfer function model. PRBS tests were conducted on top and bottom compositions by manipulating reflux and steam flow rates one at a time. The input-output data collected was then analyzed by Mozel (38) (flow rate versus composition) using a set of identification programs. The three different methods used to obtain the parameters of the pulse transfer function, were the generalized least squares, instrumental variables, and maximum likelihood techniques. For a detailed description of the three identification techniques used refer to Mozel (38). The pulse transfer function model obtained has the following form, that is,

$$\frac{W(z^{-1})}{V(z^{-1})} = \frac{z^{-k}Bz^{-1}}{(1 + Az^{-1})} \qquad 4.1$$

The parameters of the pulse transfer function were determined using the following sequences, that is, a double 31, 63 and 127 with a sampling interval of 128 seconds. The sampling interval of 128 seconds was based on the minimum sampling interval of the two composition analyzers. Also, the sampling interval of 128 seconds was chosen in order to identify the model at the same sampling interval that would be used in the control studies.

4.3 Experimental Procedure

The different binary sequences were stored on disk before the start of the experimental run. The program "PRBS1" (Appendix H) was queued and the necessary information entered using a teletype attached to the IBM 1800. The steam and reflux flow rates were adjusted according to a predetermined percentage of the steady state values. The flows were adjusted by changing the output of the DDC loop which in turn changed the setpoint of the analog flow controller. This involved operating the DDC loop on manual while the analog flow controller was on automatic control. Before the change in reflux or steam flow was introduced, the current values of top and bottom composition, and either reflux or steam flow were recorded. This procedure was continued until the binary sequence was finished. One sampling interval, that is, the measurement time delay (128 seconds) was removed before the data was analyzed. The elimination of one sampling interval was done in order that the parameters of the pulse transfer function would represent only the process dynamics.

The operating conditions of the distillation column were chosen as follows:

(1) Steam and reflux flow rates were operated at 14.75 g/s and 11.85 g/s respectively.

(2) Top and bottom steady state compositions were 95.89% methanol and 4.23% methanol respectively.

(3) The magnitude of disturbance for steam and reflux flow rates was ±7.5% and ±15% respectively.

4.4 Results

Figures 4.1 to 4.4 show the behaviour of the bottom and

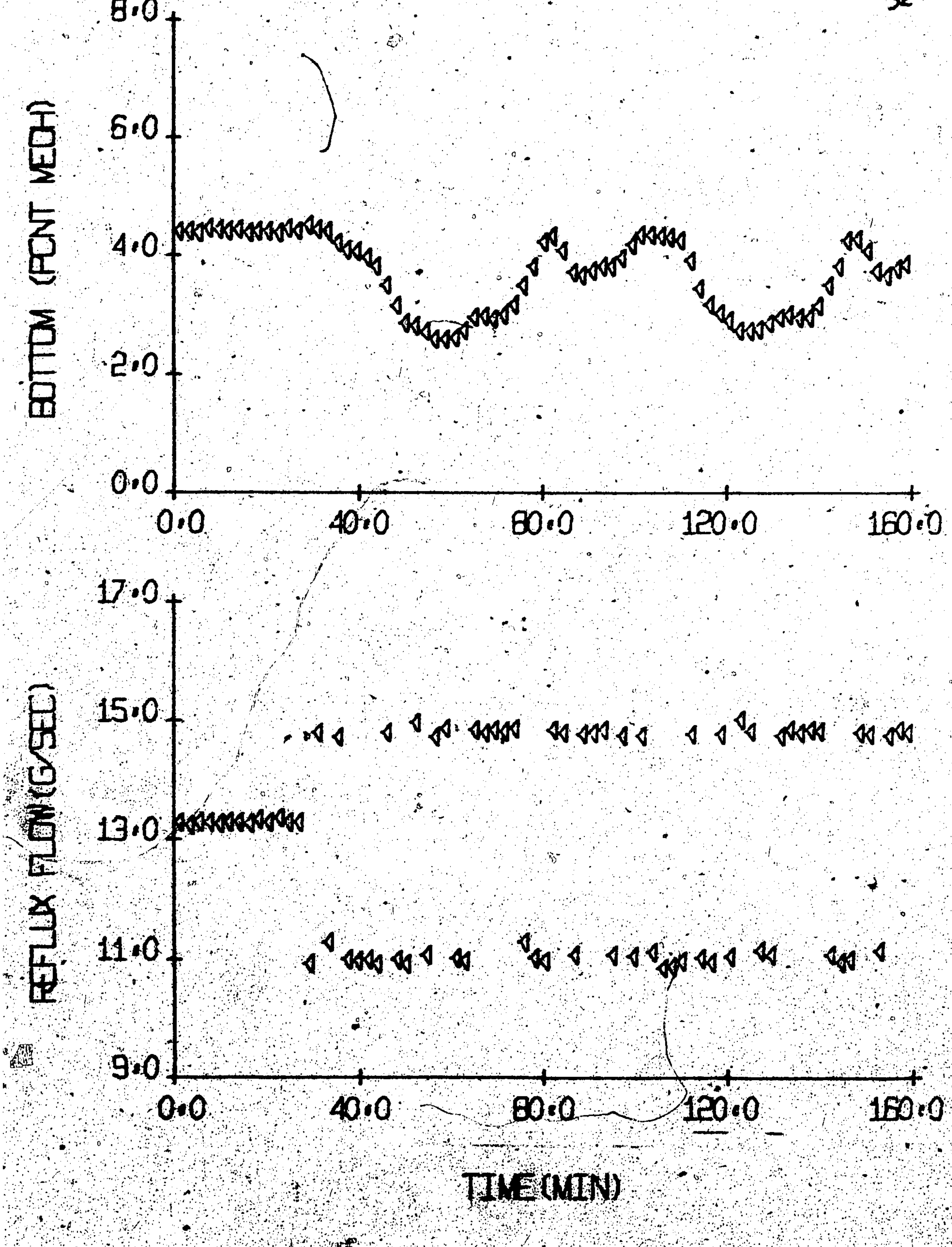


Figure 4.1 Response of bottom composition to a double 31 PRBS on reflux flow.

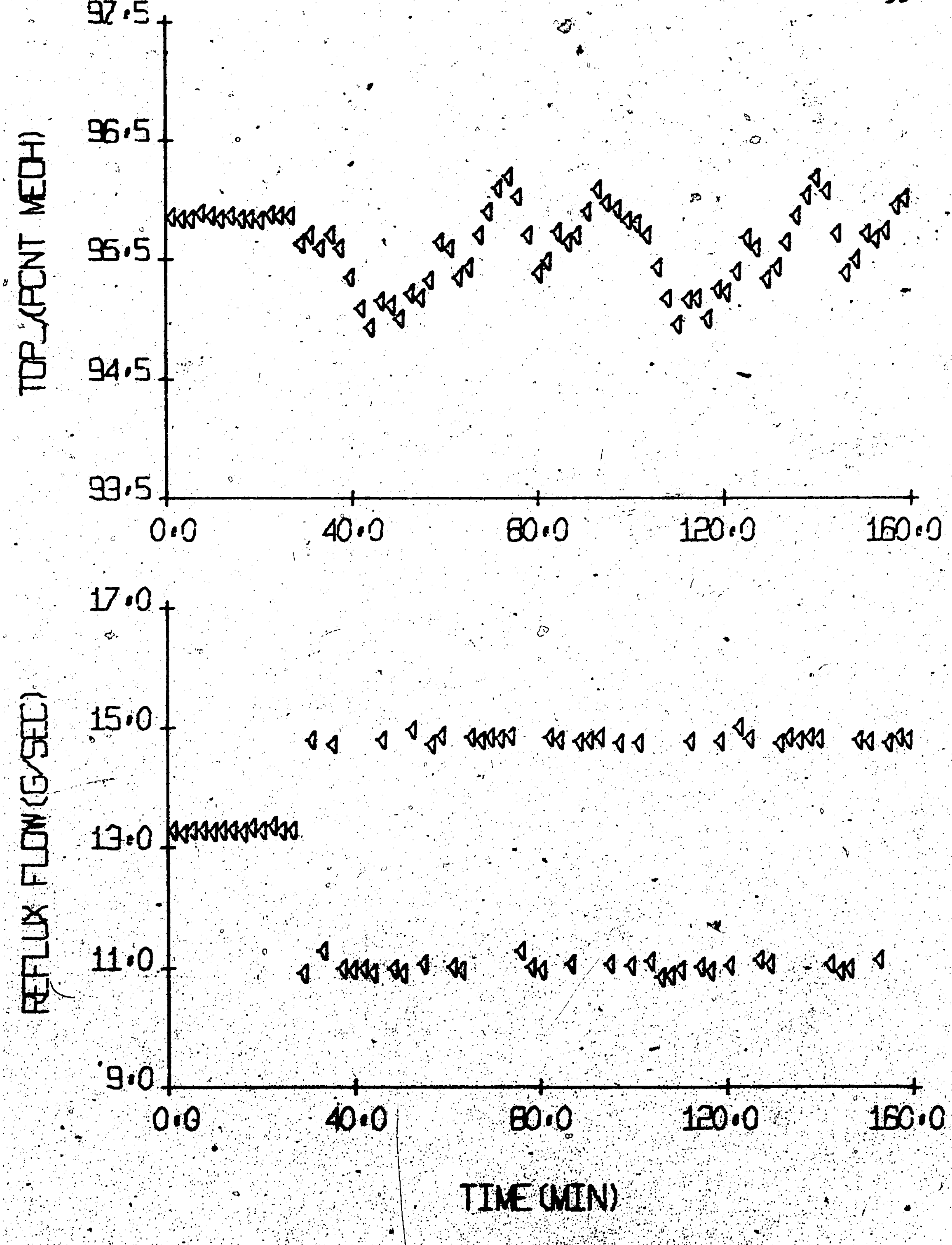


Figure 4.2 Response of top composition to a double 31 PRBS on reflux flow.

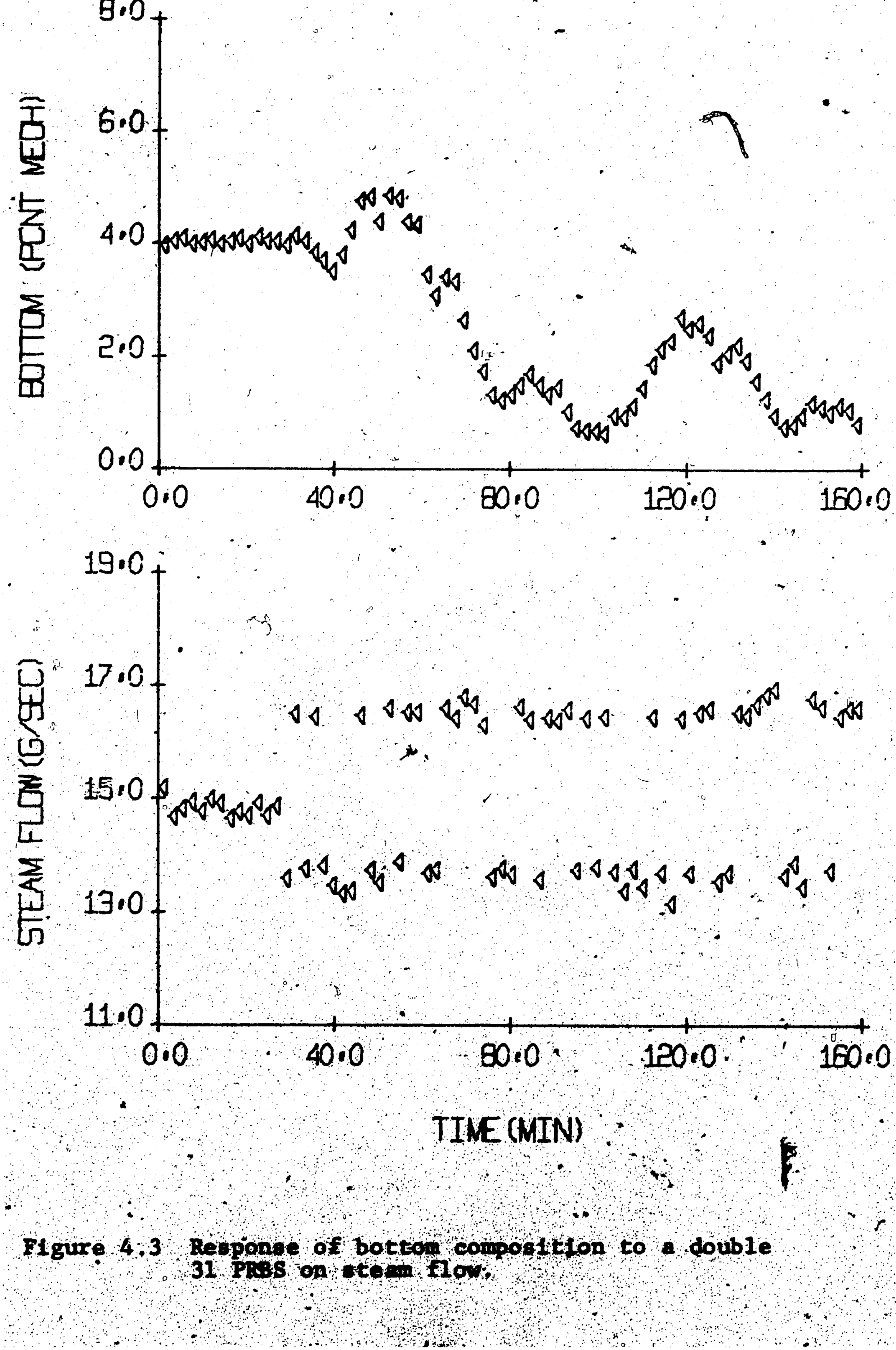


Figure 4.3 Response of bottom composition to a double 31 PRBS on steam flow.

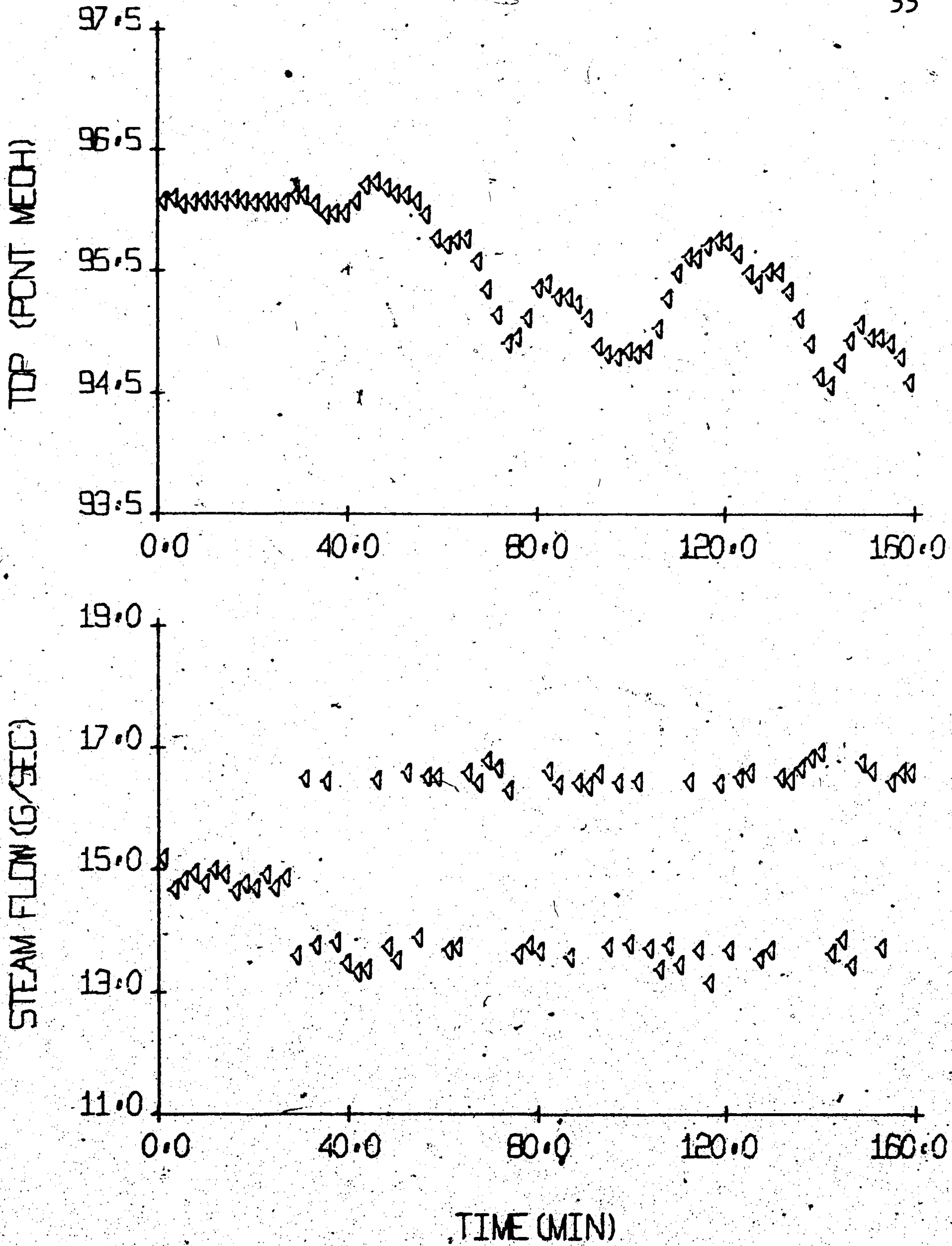


Figure 4.4 Response of top composition to a double 31 PRBS on steam flow.

top composition for a double 31 PRBS for reflux and steam respectively. Figure 4.5 shows the resulting pulse transfer functons for the distillation column from the analysis of the double 31 PRBS. The parameters of the pulse transfer function are the average values estimated using the instrumental variable and maximum likelihood techniques. The results from analysis of the data using the generalized least squares technique differed significantly from the other two methods and therefore were not used. On the same figure, the corresponding transfer function model in the s domain is shown (See Appendix I to convert from z to s). When performing control studies, the minimum sampling interval would be 128 seconds, therefore the model of the column was obtained for those circumstances. Because of the large sampling interval used, the analysis of the data showed no delay in the transfer functions between top composition and reflux and steam flow rates respectively. In reality, there is a delay of approximately 32 seconds for both transfer functions.

Figures 4.6 to 4.9 shows the response of bottom and top composition to the use of a 63 length PRBS for reflux and steam flow respectively. Figure 4.10 shows the pulse transfer functions obtained from analysis of this data. The parameters of the pulse transfer function were obtained from averaging the results obtained using the instrumental variable and maximum likelihood techniques (38).

Figures 4.11 to 4.14 display the bottom and top

$$\begin{bmatrix} Xb(z^{-1}) \\ Xd(z^{-1}) \end{bmatrix} = \begin{bmatrix} \dfrac{-0.1194z^{-2}}{1.0 - 0.801z^{-1}} & \dfrac{0.0754z^{-4}}{1.0 - 0.898z^{-1}} \\ \\ \dfrac{-0.0685z^{-1}}{1.0 - 0.860z^{-1}} & \dfrac{0.0549z^{-1}}{1.0 - 0.737z^{-1}} \end{bmatrix} \begin{bmatrix} St(z^{-1}) \\ Re(z^{-1}) \end{bmatrix}$$

$$\begin{bmatrix} Xb(s) \\ Xd(s) \end{bmatrix} = \begin{bmatrix} \dfrac{-0.628e^{-128s}}{576.9s + 1.0} & \dfrac{0.736e^{-384s}}{1184.9s + 1.0} \\ \\ \dfrac{-0.488}{846.1s + 1.0} & \dfrac{0.208}{418.5s + 1.0} \end{bmatrix} \begin{bmatrix} St(s) \\ Re(s) \end{bmatrix}$$

Figure 4.5 The pulse transfer functions from fitting the double 31 PRBS data and the corresponding continuous form approximations.

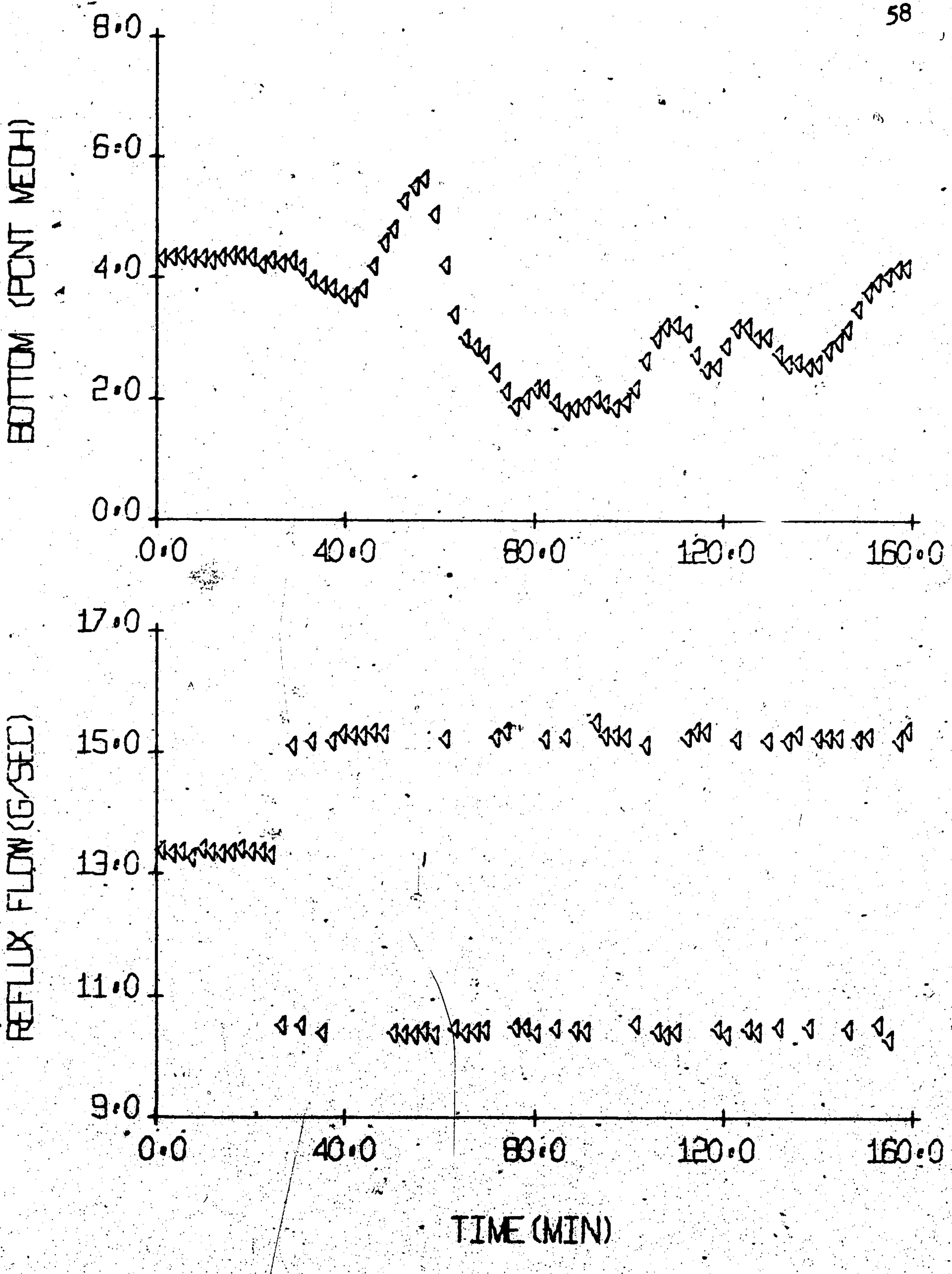


Figure 4.6 Response of bottom composition to a 63 PRBS on reflux flow.

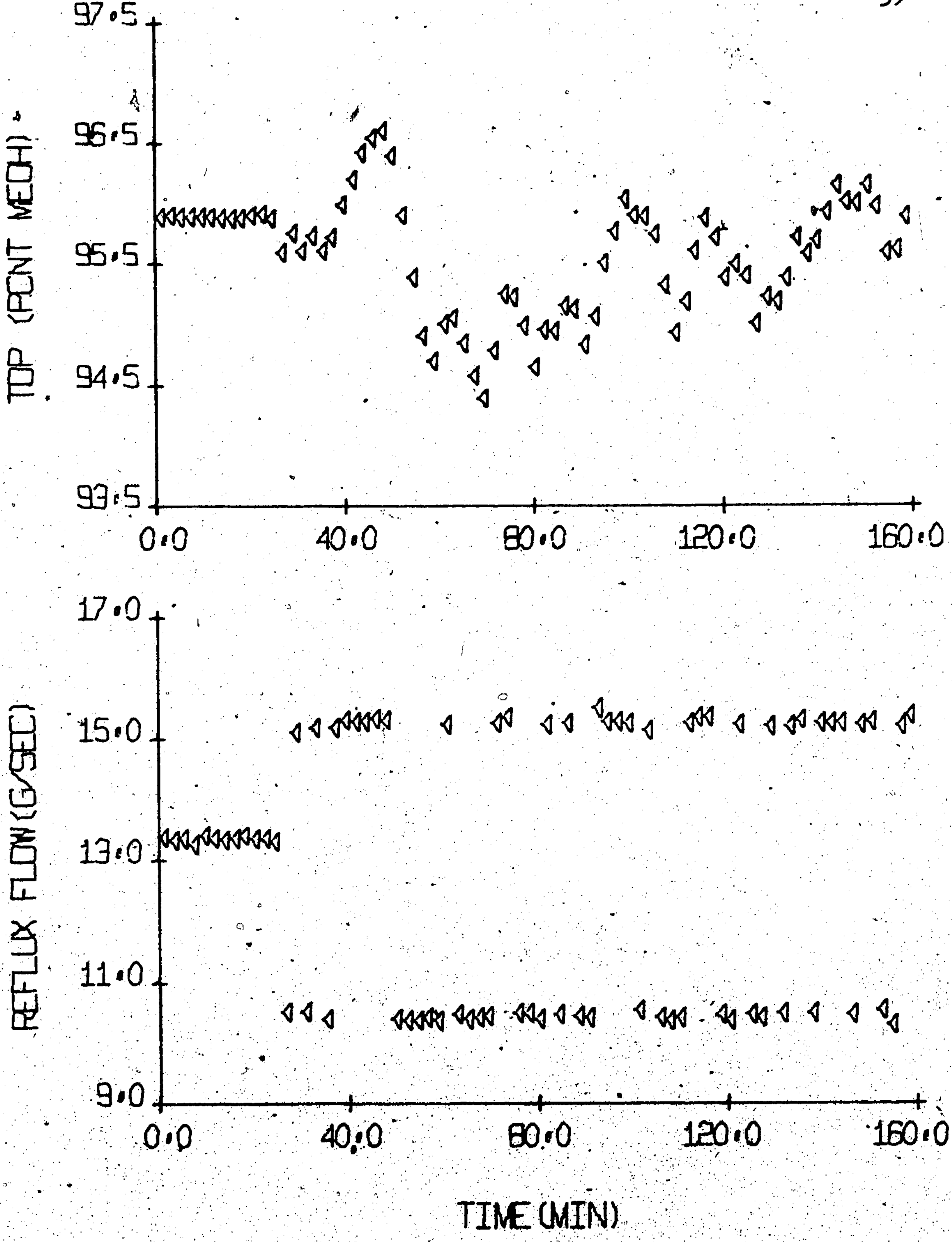


Figure 4.7 Response of top composition to a 63 PRBS on reflux flow.

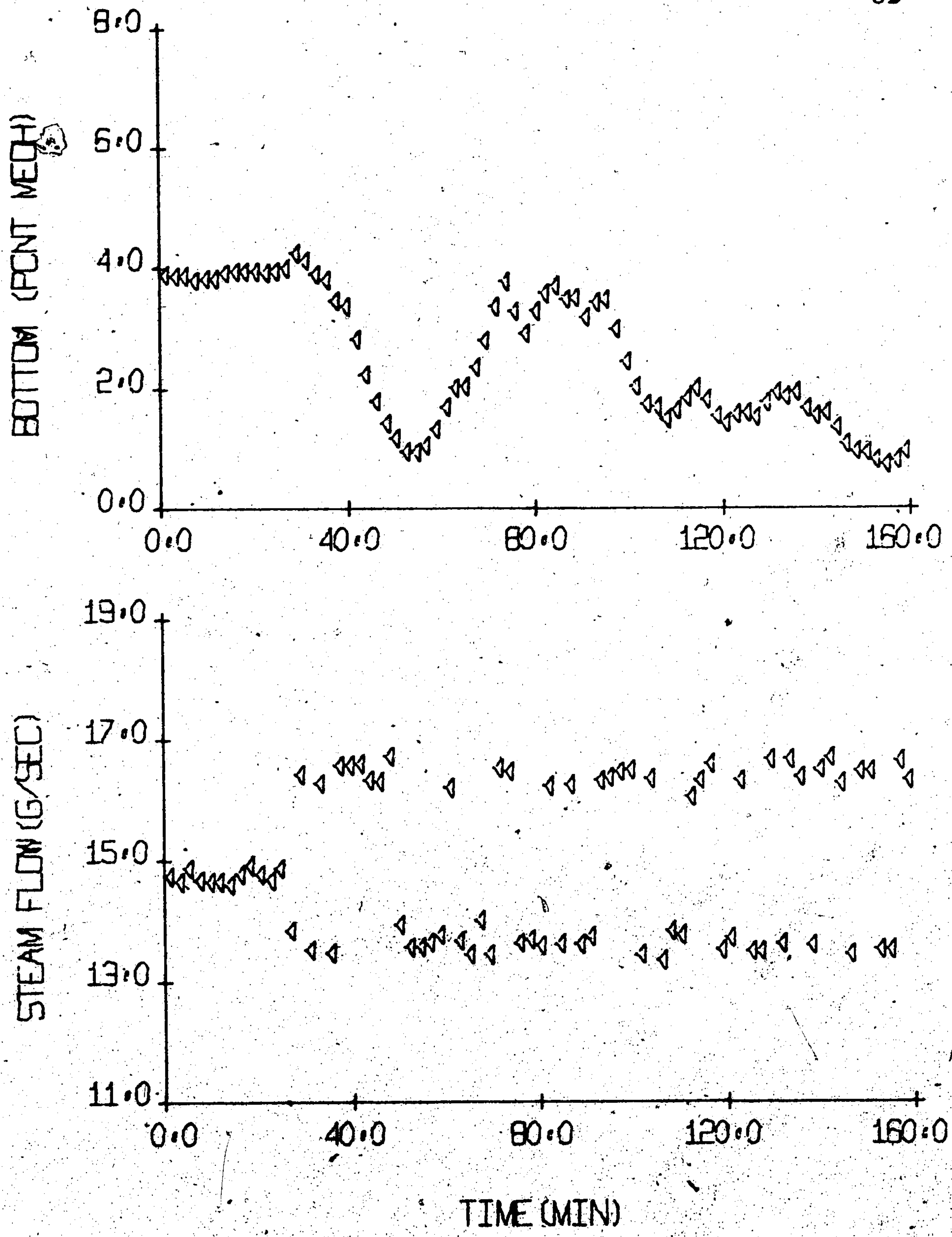


Figure 4.8 Response of bottom composition to a 63 PRBS on steam flow.

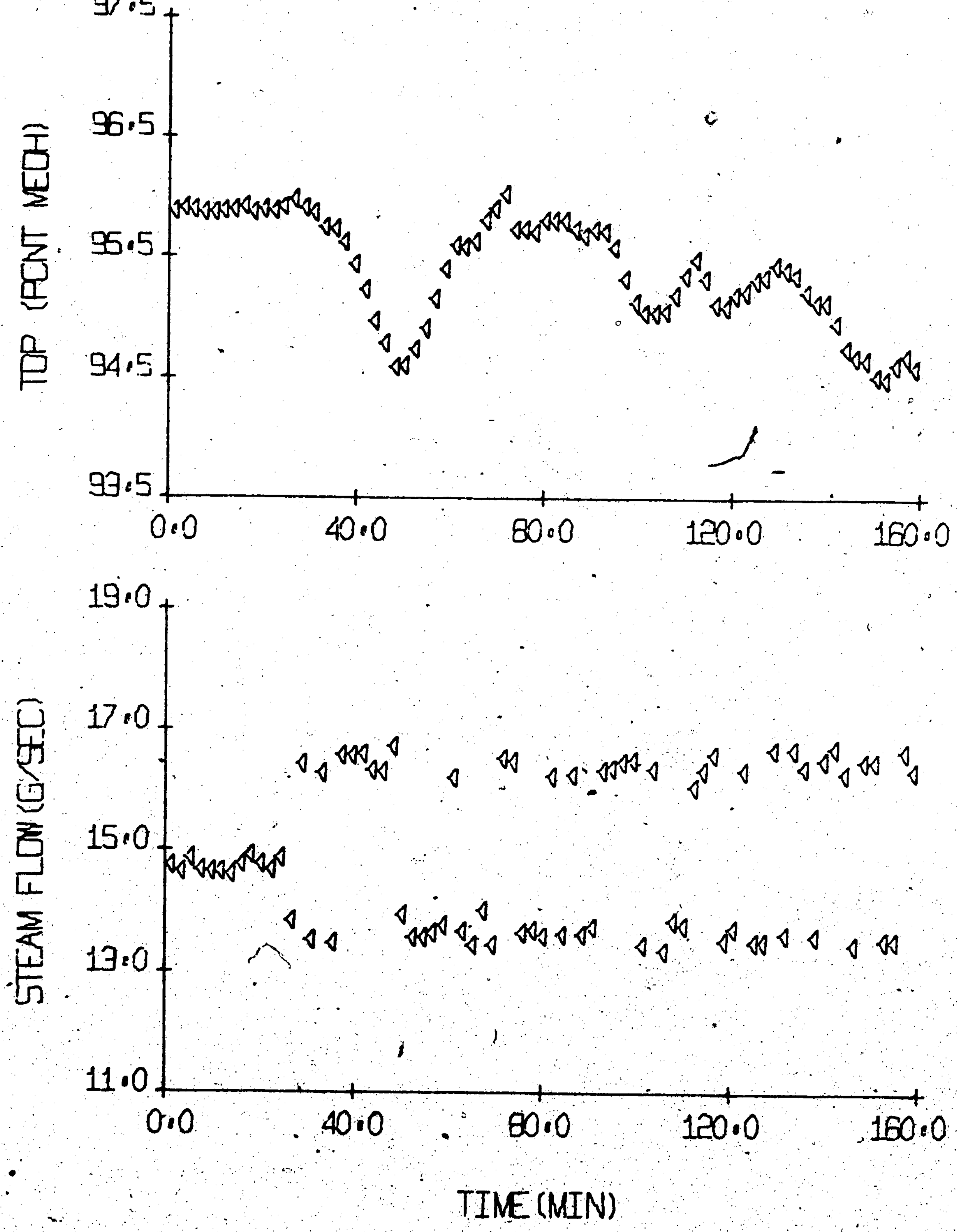


Figure 4.9 Response of top composition to a 63 PRBS on steam flow.

$$\begin{bmatrix} Xb(z^{-1}) \\ Xd(z^{-1}) \end{bmatrix} = \begin{bmatrix} \dfrac{-0.1496z^{-2}}{1.0 - 0.918z^{-1}} & \dfrac{0.0578z^{-4}}{1.0 - 0.886z^{-1}} \\ \dfrac{-0.0618z^{-1}}{1.0 - 0.864z^{-1}} & \dfrac{0.0596z^{-1}}{1.0 - 0.671z^{-1}} \end{bmatrix} \begin{bmatrix} St(z^{-1}) \\ Re(z^{-1}) \end{bmatrix}$$

$$\begin{bmatrix} Xb(s) \\ Xd(s) \end{bmatrix} = \begin{bmatrix} \dfrac{-1.827e^{-\text{[illegible]}8s}}{1497.9s + 1.0} & \dfrac{0.505e^{-384s}}{1053.6s + 1.0} \\ \dfrac{-0.453}{872.9s + 1.0} & \dfrac{0.181}{320.7s + 1.0} \end{bmatrix} \begin{bmatrix} St(s) \\ Re(s) \end{bmatrix}$$

Figure 4.10 The pulse transfer functions from fitting the 63 PRBS data and the corresponding continuous form approximations.

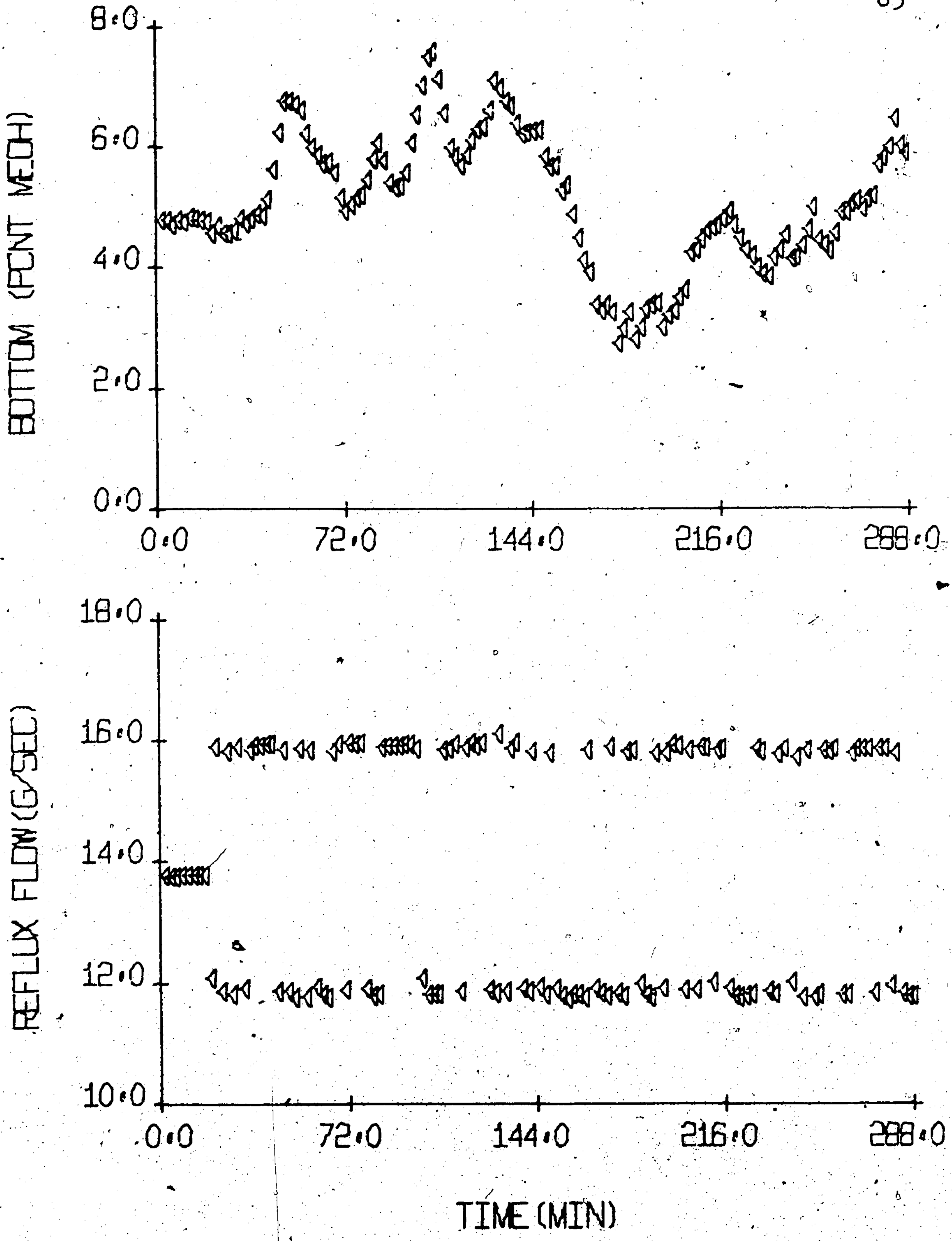


Figure 4.11 Response of bottom composition to a 127 PRBS on reflux flow.

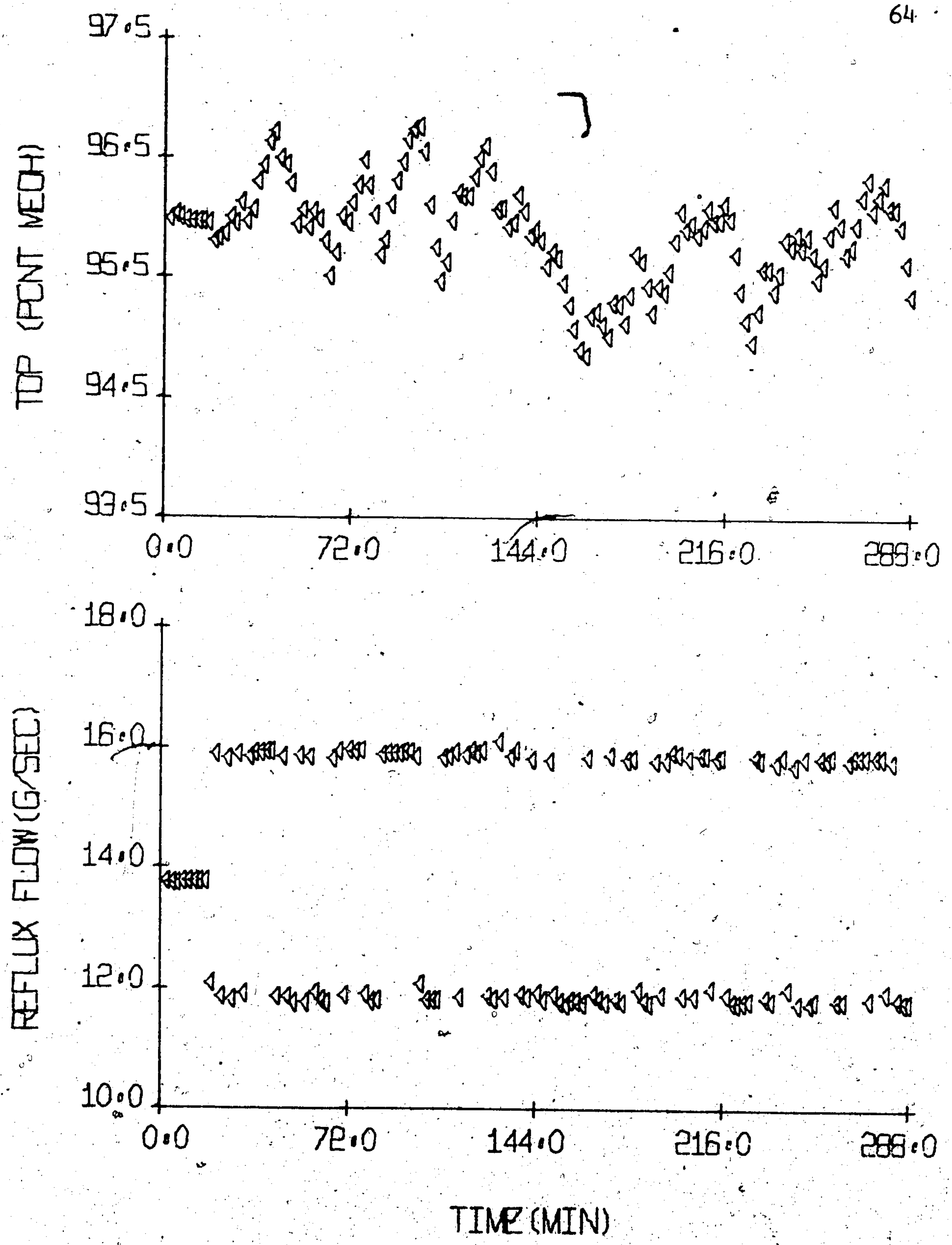


Figure 4.12 Response of top composition to a 127 PRBS on reflux flow.

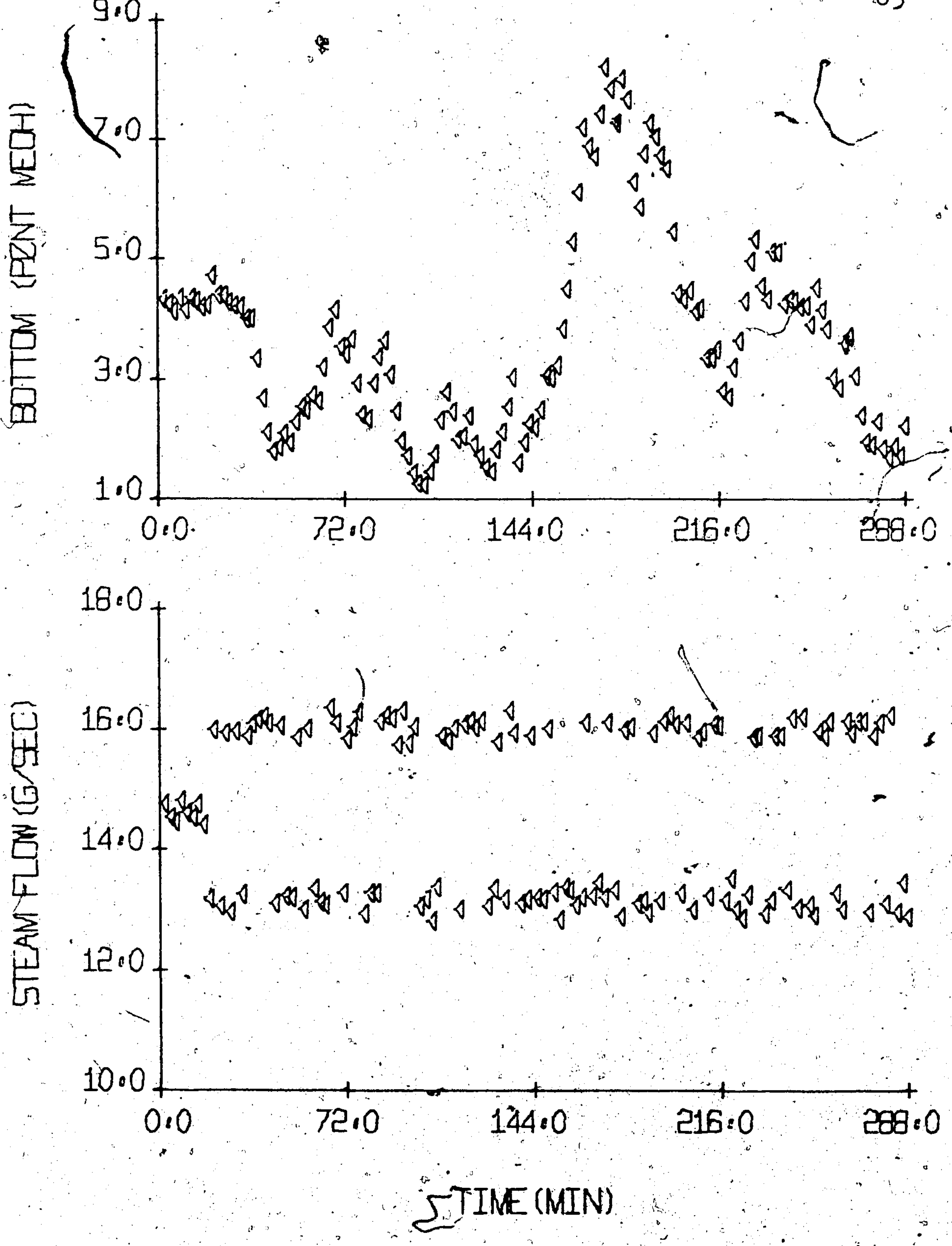
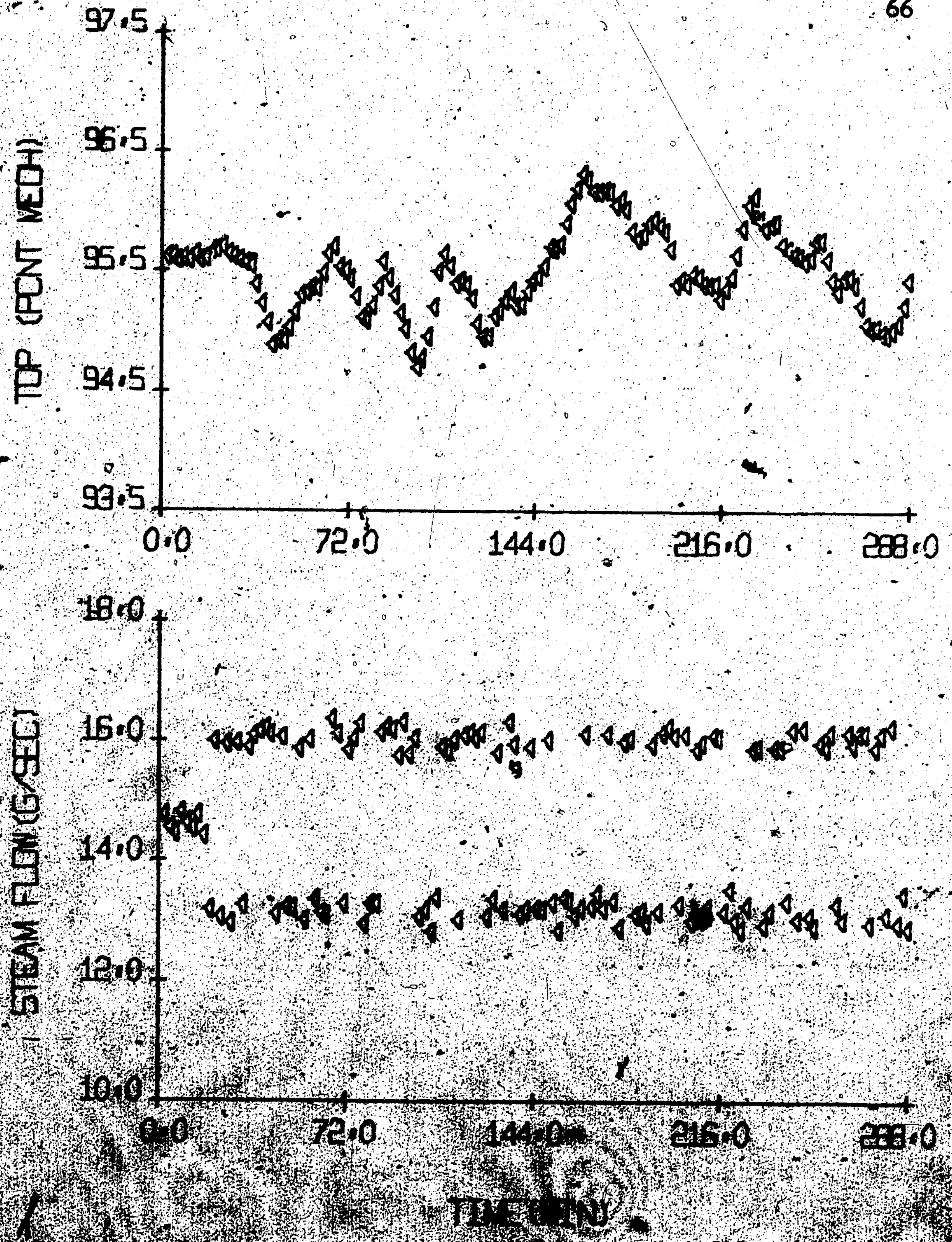


Figure 4.13 Response of bottom composition to a 127 PRBS on steam flow.


97.5
96.5
95.5
94.5
93.5
TOP (PCNT MEOH)
0.0
72.0
144.0
216.0
288.0
18.0
16.0
14.0
12.0
10.0
STEAM FLOW (G/SEC)
0.0
72.0
144.0
216.0
288.0
TIME (MIN)

[illegible]

compositon response to changes in reflux and steam flow rates. The pulse transfer functions shown in Figure 4.15 are based on analysis of the 127 PRBS, but in this case, the parameters are average values from the generalized least squares, instrumental variable, and maximum likelihood techniques (38).

4.5 Discussion of Results

In determining the transfer function model of the column by PRBS studies, only one magnitude of disturbance was used. The sampling interval was not varied because a time shorter than 128 seconds was not obtainable due to physical constraints in analyzing the bottom composition. Figure 4.16 compares the time constants and gains of the open loop responses found in Chapter 3 to those established by approximating the pulse transfer functions obtained from the PRBS testing with continuous form (s domain) transfer functions. The approximations from the longest sequence shows values for the time constant which are nearly double those from the open loop experimental runs (with one exception, top composition to reflux flow rate). As the sequence increased in length, the gains began to more closely match the gains of the experimental open loop responses. The only transfer function that showed consistent results [illegible]

$$\begin{bmatrix} Xb(z^{-1}) \\ \\ Xd(z^{-1}) \end{bmatrix} = \begin{bmatrix} \dfrac{-0.2582z^{-2}}{1.0 - 0.941z^{-1}} & \dfrac{0.0749z^{-4}}{1.0 - 0.956z^{-1}} \\ \\ \dfrac{-0.0561z^{-1}}{1.0 - 0.908z^{-1}} & \dfrac{0.0465z^{-1}}{1.0 - 0.798z^{-1}} \end{bmatrix} \begin{bmatrix} St(z^{-1}) \\ \\ Re(z^{-1}) \end{bmatrix}$$

$$\begin{bmatrix} Xb(s) \\ \\ Xd(s) \end{bmatrix} = \begin{bmatrix} \dfrac{-4.339e^{-128s}}{2086.6s + 1.0} & \dfrac{1.710e^{-384s}}{2857.9s + 1.0} \\ \\ \dfrac{-0.609}{\text{132[illegible]3}s + 1.0} & \dfrac{0.230}{567.3s + 1.0} \end{bmatrix} \begin{bmatrix} St(s) \\ \\ Re(s) \end{bmatrix}$$

(A) Time Constants

Test	Xb/St	Xd/St	Xb/Re	Xd/Re
Open Loop	1142.8	568.9	1274.9	431.7
PRBS-31D	576.9	846.1	1184.9	418.5
PRBS-63	1497.9	872.9	1053.6	320.7
PRBS-127	2086.6	1326.3	2857.9	567.3

(B) Gains

Test	Xb/St	Xd/St	Xb/Re	Xd/Re
Open Loop	-8.841	-0.631	4.218	0.542
PRBS-31D	-0.628	-0.488	0.736	0.208
PRBS-63	-1.827	-0.453	0.505	0.181
PRBS-127	-4.339	-0.609	1.710	0.230

Figure 4.16 Comparison of time constants and gains between open loop tests and [illegible] to [illegible] [illegible] [illegible] PRBS testing

transfer function between top composition and reflux flow rate. The longest sequence, 127 was approximately equal to five hours of operation to complete the sequence. Reliable data was not obtained in using longer sequences because of operational problems with the liquid sampling system for the gas chromatograph, that is, the filters became plugged and require cleaning.

The inability of the identification programs to determine the transfer functions more accurately can be attributed to the short sequence length, large sampling interval, and the possible nonminimum phase response behaviour of the column. The time delay was the only parameter determined accurately by the identification programs when compared to the experimental values obtained from open loop testing.

The transfer function model that was used consisted of the gains from the open loop tests and the time constants used were average values from the continuous approximations of the pulse transfer functions obtained from the double 31 and 63 PRBS studies. The time delays used were those determined by the PRBS testing. Figure 4.17 shows the final form of the transfer function model chosen. The transfer functions were then compared to the open loop tests reported in Chapter 3 for steam and reflux disturbances. Figures 4.18 to 4.21 show the response of the linear models to the open loop tests using the same magnitude of disturbances for both steam and reflux flow rates.

$$\begin{bmatrix} Xb(s) \\ \\ Xd(s) \end{bmatrix} = \begin{bmatrix} \dfrac{-8.841e^{-126s}}{1037.4s + 1.0} & \dfrac{4.218e^{-384s}}{1119.3s + 1.0} \\ \\ \dfrac{-0.631}{859.5s + 1.0} & \dfrac{0.542}{369.6s + 1.0} \end{bmatrix} \begin{bmatrix} St(s) \\ \\ Re(s) \end{bmatrix}$$

Figure 4.17 The transfer function model of the column used for designing the compensators.

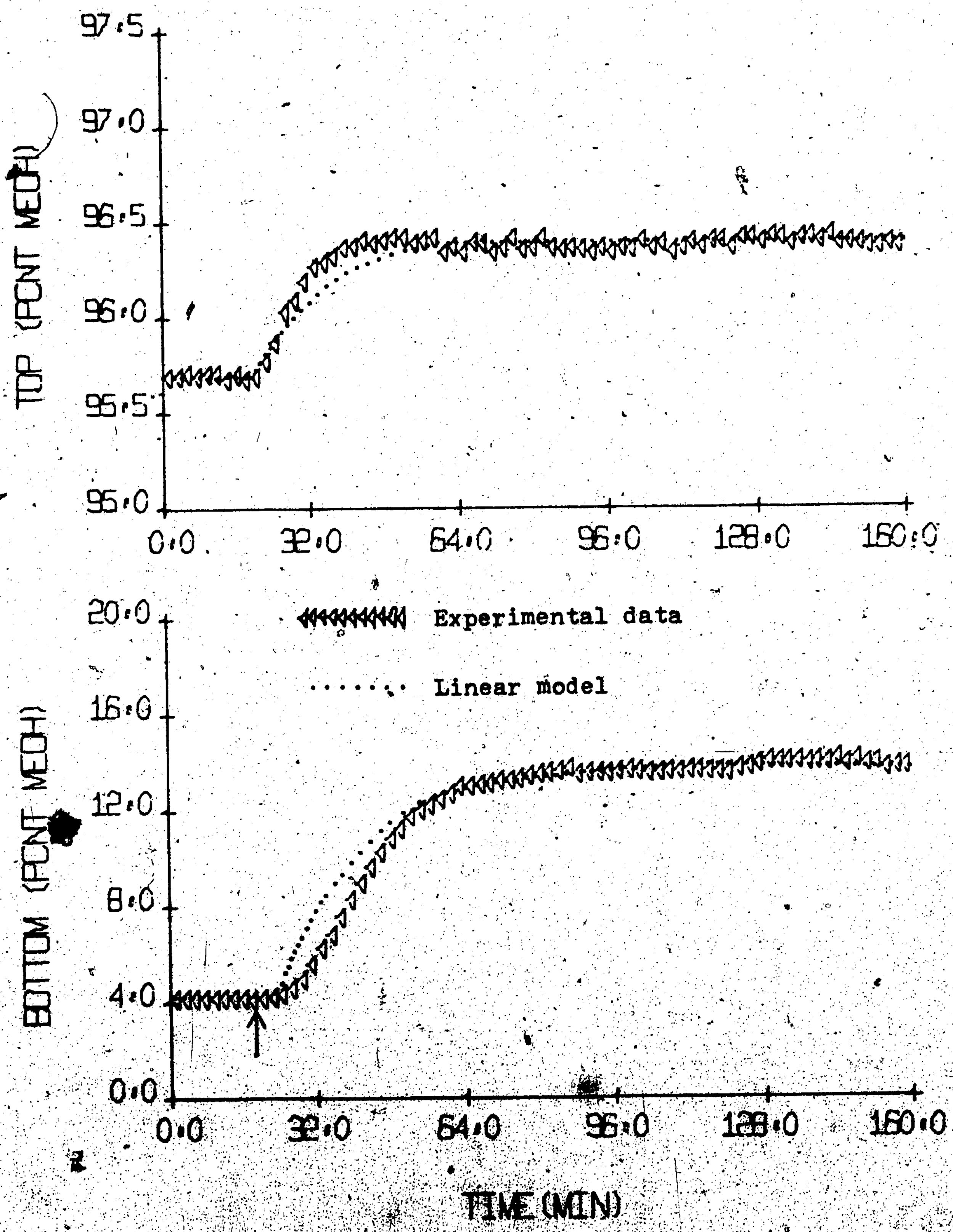


Figure 4.18 Comparison of linear models to experimental top and bottom compositions for a -1.11 g/s step change in steam flow rate.

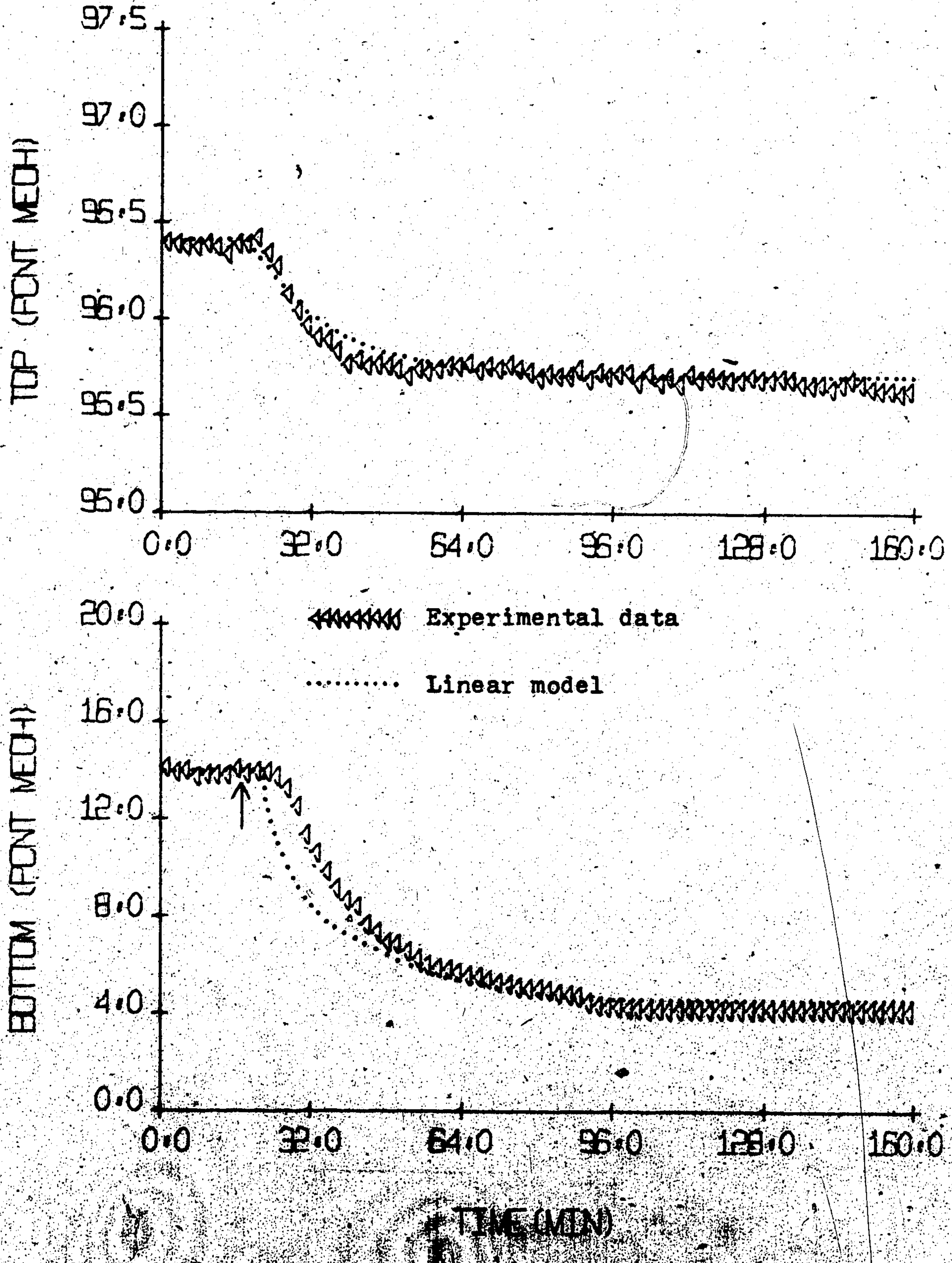


Figure 4.19 Comparison of linear models to experimental top and bottom compositions for a [illegible] g/s step change in [illegible] flow rate

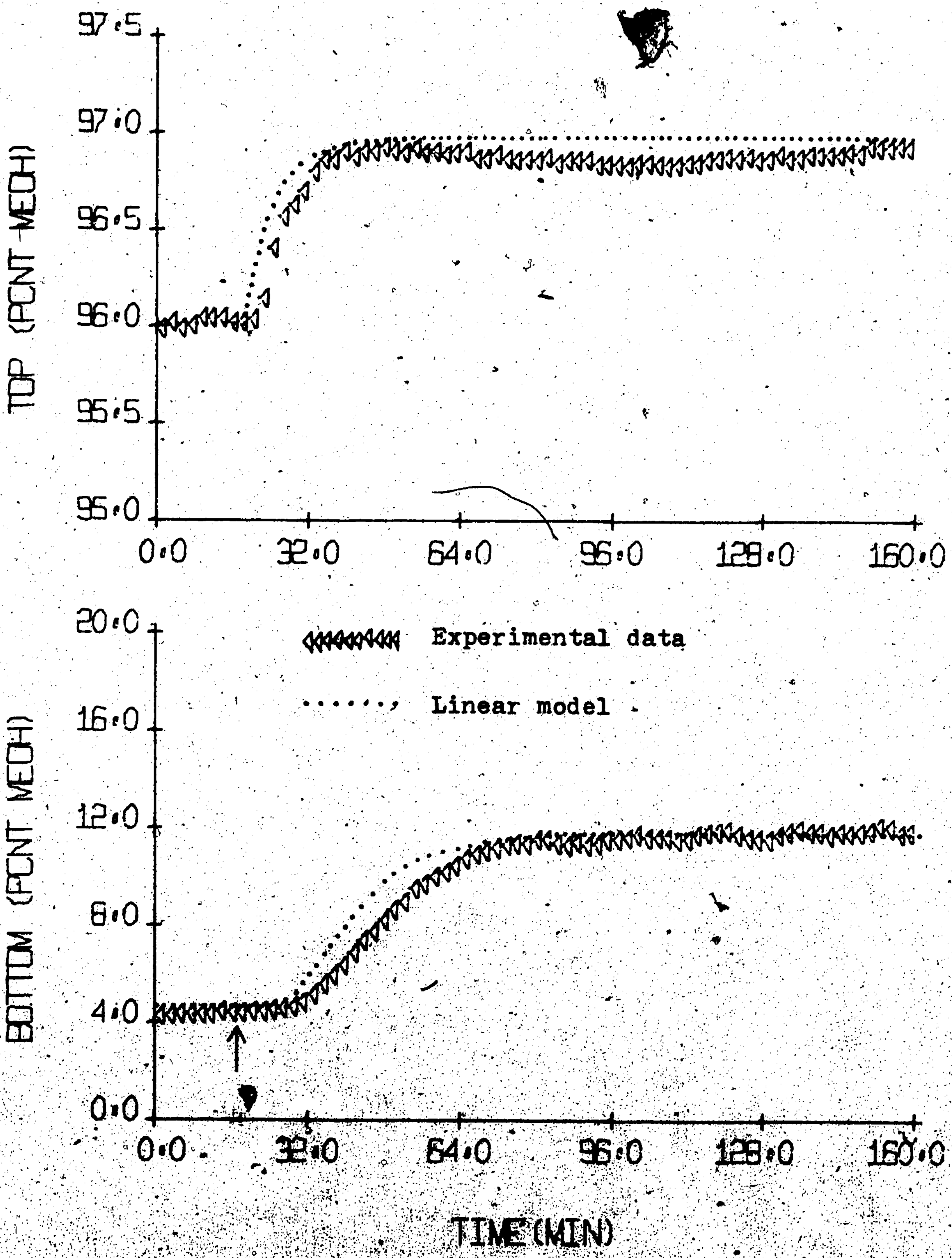


Figure 4.20 Comparison of linear models to experimental top and bottom compositions for a +1.79 g/s step change in reflux flowrate.

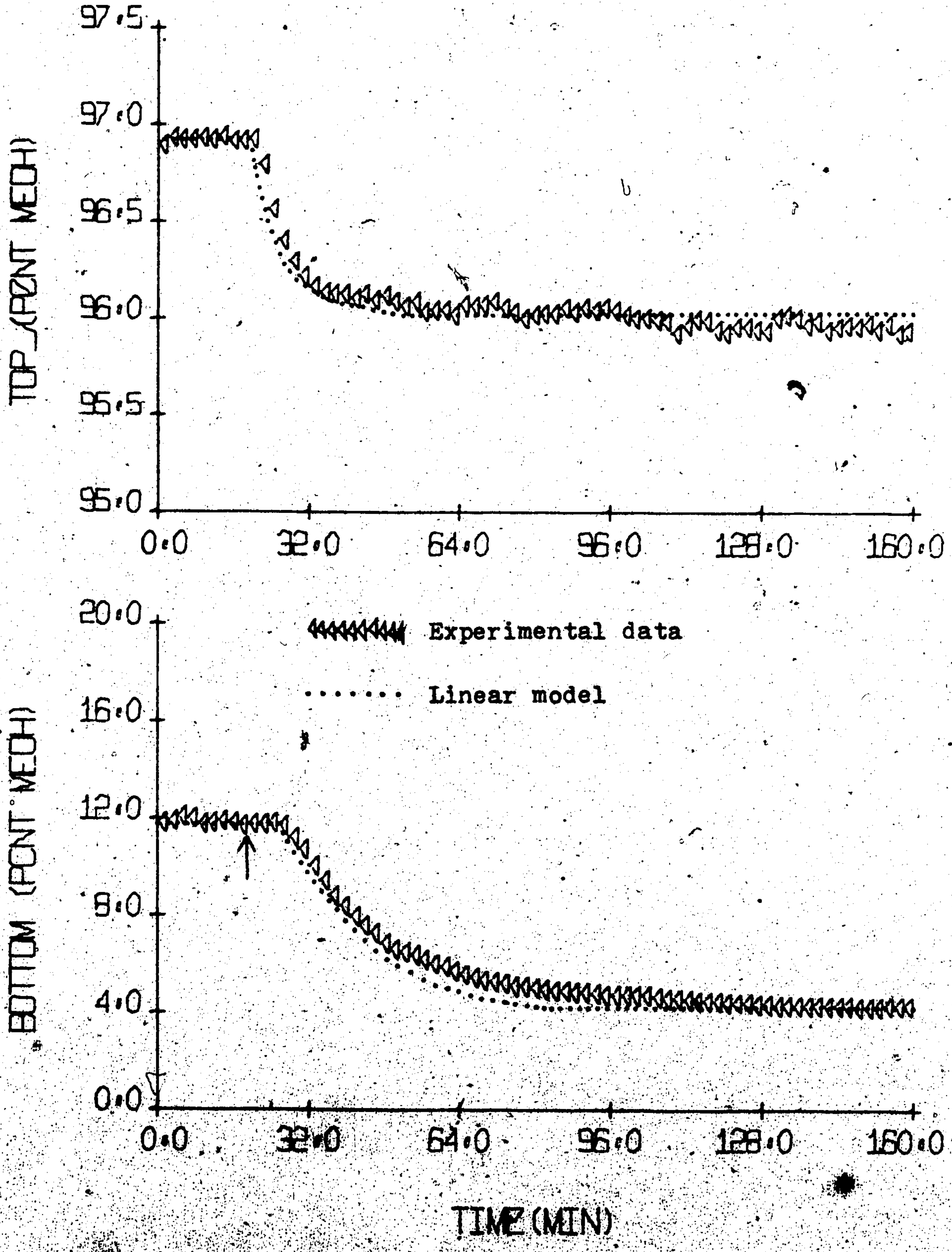


Figure 4.21 Comparison of linear models to experimental top and bottom compositions for a -1.79 g/s step change in reflux flow rate.

Chapter 5

CONTROLLING THE DISTILLATION COLUMN

5.1 Introduction

The interaction between the top and bottom composition control loops presents problems in a multiloop control system without compensation. Three types of compensators were evaluated in order to determine which was most effective in reducing interaction. The direct Nyquist array method was used to design a plant precompensator. The individual loops were controlled using the conventional PID controller.

Most control strategies for distillation columns are designed to hold the product compositions at some desired setpoints. In most cases, the manipulative variables chosen for controlling the top and bottom compositions cause undesirable interactions between the two loops. Conventional PID control may work fine when the controllers are operated individually but when operated together the control performance is decreased when compared to the individual control loops. Therefore the control strategy used should have some means of compensating or eliminating the interactions between the two loops.

Buckley (6) suggested the idea of inserting two compensators to cancel out the effects each manipulated variable has on the other control variable. This idea was tried by Luyben (29) where he compared ideal decoupling to simplifed decoupling for a binary distillation column. He used a linear model in the frequency domain to obtain the decoupling elements. Other techniques were used in attempts to reduce interaction between the control loops. Rosenbrock used the sum of and the difference between the values for the two terminal compositions to manipulate the reflux and vapour boil-up rates respectively. Rijnsdorp proposed a ratio control scheme between reflux and top vapour flow. This type of control scheme was tried by Liesch (27) on a binary distillation column. Berry (2) designed a decoupled system based on a linear transfer function model. Another method tried by Pacey (41), was to use feedforward control on the disturbance variable.

5.2 Theory for Design of Compensators

Due to the work of Rosenbrock et al (49), the design of controllers for multivariable systems using frequency domain techniques has increased substantially in the past few years. Single variable design techniques have been extended to handle multivariable systems. The most important single feature of frequency domain techniques is their ability to handle irrational functions, that is, time delays. Another advantage is that the method does not require a mathematical model; that is, the frequency response data can be used directly.

The design techniques that can be used for controller design are the direct Nyquist array (DNA) method, the inverse Nyquist array (INA) method or the Characteristic Locus (CL) method. To aid in the design, a transfer function model was obtained (see Chapter 4). Although the time delays contained in the distillation column transfer function matrix can be handled using the INA method, the type of diagrams that result make the procedure somewhat awkward so the INA method was not used. Faced with the choice of the DNA method and the CL method, the DNA was chosen because of its ease of use and interpretation for design decisions.

For the schematic representation of the multivariable system shown in Figure 5.1, the return ratio ($\underline{\underline{T}}(s)$) and difference ($\underline{\underline{F}}(s)$) matrices are defined as

$$\underline{\underline{T}}(s) = \underline{\underline{G}}(s)\underline{\underline{K}}(s)\underline{\underline{H}}(s) \qquad 5.1$$

and

$$\underline{\underline{F}}(s) = (\underline{\underline{I}} + \underline{\underline{G}}(s)\underline{\underline{K}}(s)\underline{\underline{H}}(s))^{-1}\underline{\underline{G}}(s)\underline{\underline{K}}(s)\underline{\underline{H}}(s) \qquad 5.2$$

For the special case where $\underline{\underline{H}}(s) = \underline{\underline{I}}$, the return ratio and

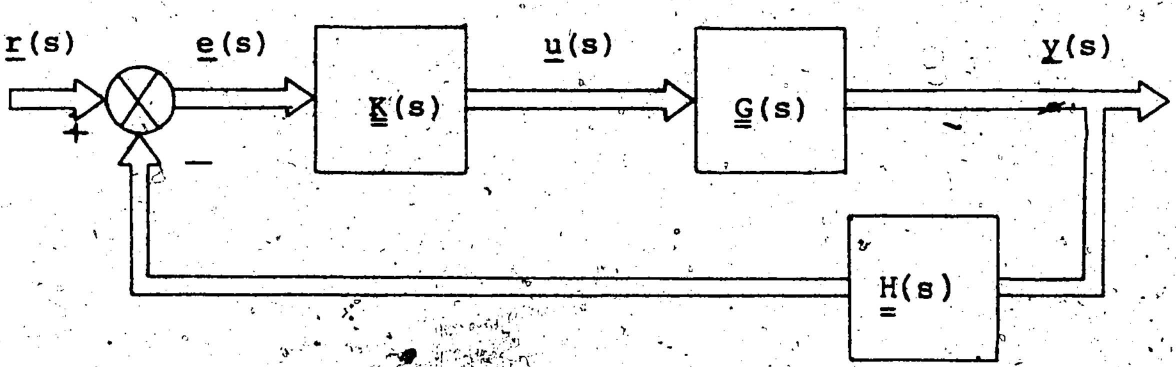


Figure 5.1 (a) Block diagram of a multivariable feedback control system.

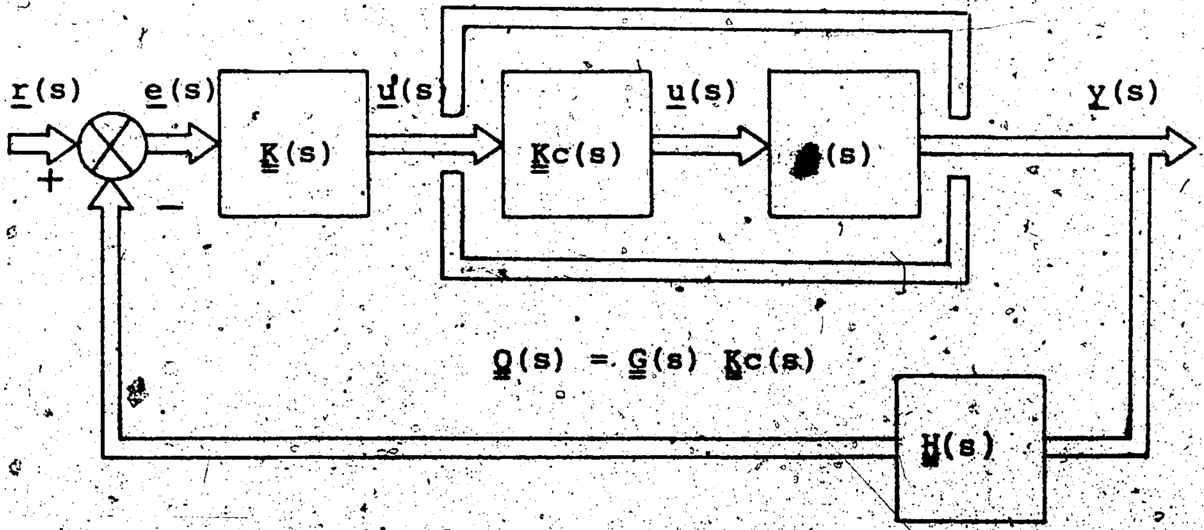


Figure 5.1 (b) Block diagram of a multivariable feedback control system showing precompensator system.

difference matrices will become the open loop ($\underline{\underline{Q}}(s)$) and closed loop ($\underline{\underline{R}}(s)$) transfer function matrices respectively. Before the stability criteria of single variable frequency domain techniques can be applied to multivariable systems, the return difference matrix must be diagonally dominant. Using a plant precompensator, the open loop transfer function matrix ($\underline{\underline{Q}}(s)$) can be made diagonally dominant and inherently $\underline{\underline{F}}(s)$ will also be diagonally dominant. This open loop transfer function matrix is as follows

$$\underline{\underline{Q}}(s) = \underline{\underline{G}}(s)\underline{\underline{K}}(s) \qquad 5.3$$

where $\underline{\underline{G}}(s)$ is the plant transfer function matrix and $\underline{\underline{K}}(s)$ is the controller matrix. To determine diagonal dominance for the open loop transfer function matrix, Gershgorin circles were used. The Gershgorin circles have the advantage of being independent of the controller gains and therefore can be used in the design procedure. The centre of the circles lay on the Nyquist locus of the diagonal element at specified frequencies. The magnitude of the radius for the Gershgorin circles is the summation of all the off diagonal elements. The return difference matrix is diagonally dominant if the Gershgorin circles do not include the point (-1,0). For a more detailed discusson of Gershgorin circles and their respective equations see Kuon Ph. D thesis (23).

When $\underline{\underline{F}}(s)$ is diagonally dominant, the closed loop system is stable if and only if

$$\sum_{i=1}^{m} nt_i = -Po \qquad 5.4$$

where nt_i is the net number of clockwise encirclements that the Nyquist locus of the diagonal element t_i of matrix $\underline{\underline{T}}(s)$ makes of the critical point (-1,0). The symbol Po denotes the number of roots of the open loop characteristic polynomial that are in the right half of the complex plane. The open loop characteristic polynomial is the least common denominator of all minors of the open loop transfer function matrix.

5.3 Evaluation of Control Systems

Originally it was planned that only proportional plus integral (PI) control action would be used for controlling both the top and bottom compositions. However, it was found necessary to add derivative action because of the poor PI response on the bottom composition due to the large time delay and time constant. The following control law was used in this study, that is,

$$\underline{u}(s) = \underline{\underline{K}}(s)(\underline{r}(s) - \underline{y}(s)) \qquad 5.5$$

where

$$\underline{\underline{K}}(s) = \underline{\underline{K}}b(s)\underline{\underline{K}}c(s) \qquad 5.6$$

where $\underline{\underline{K}}b(s)$ is initially the identity matrix until diagonal dominance has been achieved and $\underline{\underline{K}}c(s)$ is the plant precompensator and is designed by column operations. Now the open loop transfer function matrix becomes

$$\underline{\underline{Q}}(s) = \underline{\underline{G}}(s)\underline{\underline{K}}c(s) \qquad 5.7$$

If $\underline{\underline{Q}}(s)$ can be made diagonally dominant, then $\underline{\underline{K}}b(s)$ can be designed using single frequency domain techniques.

The precompensator was designed using the direct Nyquist array (DNA) method as described in the Gemscope User's manual (62) at the University of Alberta. The Nyquist array method was also used to choose the pairing of manipulative and control variables. Another method for the pairing of variables is Bristol's relative gain approach. Both methods were used to determine the pairing as a check of each other.

The first step in the design procedure using the DNA method was the pairing of variables (25). Figure 5.2 shows the Nyquist plots for each element of the open loop transfer function matrix shown on Figure 4.17. The Nyquist plots are

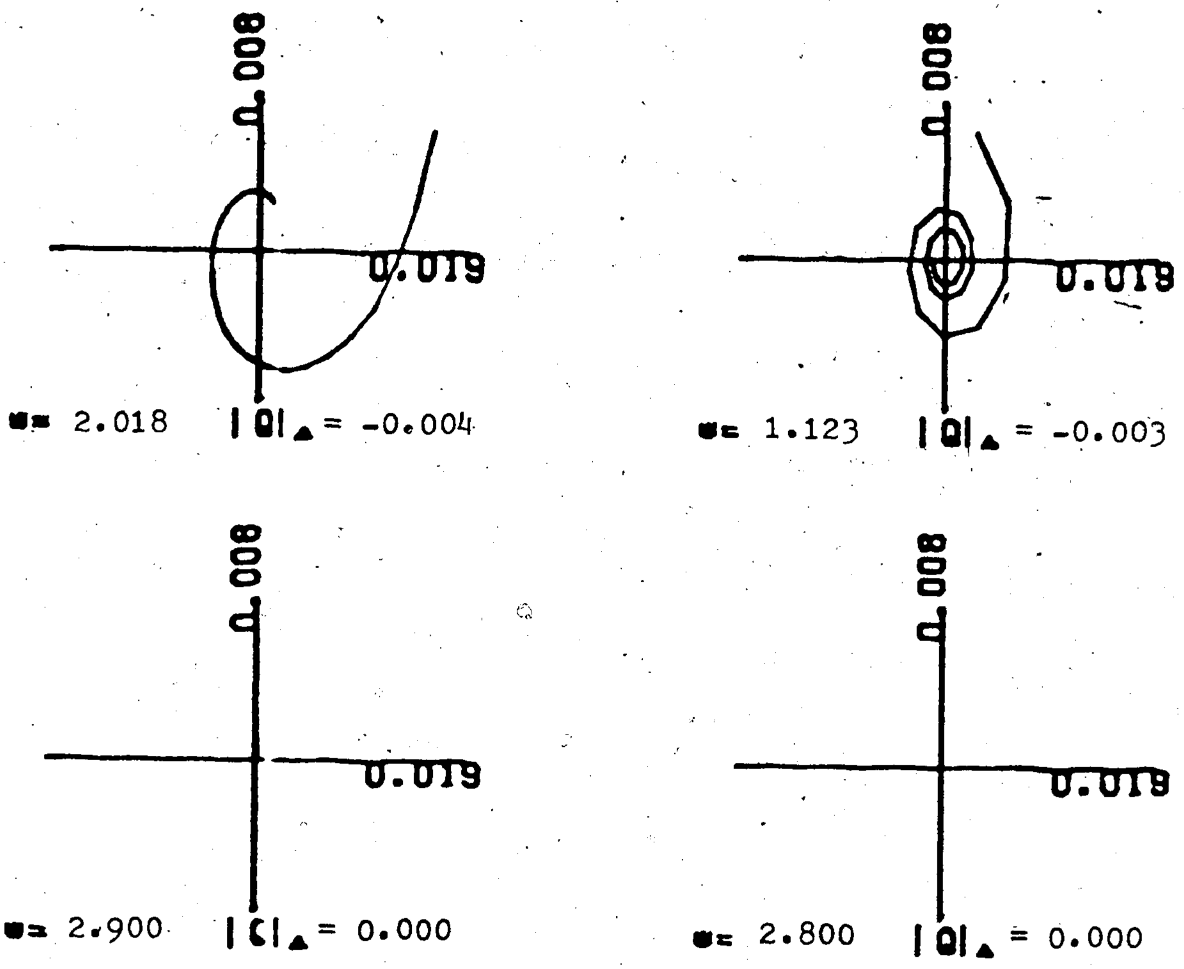


Figure 5.2 Nyquist plots for elements of the plant transfer function matrix $\underline{\underline{G}}(s)$.

representing Equation 5.7 and in the case of Figure 5.2 the precompensator matrix ($\underline{\underline{K}}c(s)$) in the identity matrix initially. The frequency used for all Nyquist plots is from 0.5 to 3.0 radians/second. For each element, the cross over frequency and the magnitude at that frequency is determined for each transfer function. The magnitude of each axis is indicated on the axis. The lower Nyquist plots on Figure 5.2 are both first order systems with no time delay. The control variables for the distillation column are the top and bottom compositions, while the possible manipulative variables are reflux and steam flow rates. Figure 5.3 shows the pairing of steam flow with bottom composition and reflux flow with top composition as the manipulative and control variables in each pairing. This figure shows the Nyquist locus of the diagonal elements along with the Gershgorin circles which are used as a measure of interaction between the chosen pairing of variables. Figure 5.4 is a similar figure but for the reverse pairing; that is, controlling top composition with steam and bottom composition using reflux flow. Comparing the above two figures clearly indicates the former pairing of variables is a better choice as indicated by the smaller Gershgorin circles overall. In control terms, this would mean that there is more interaction if the latter pairing is chosen and therefore would be more difficult to control.

As mentioned earlier, Bristol's relative gain matrix approach (4) was also used to verify the pairing of control

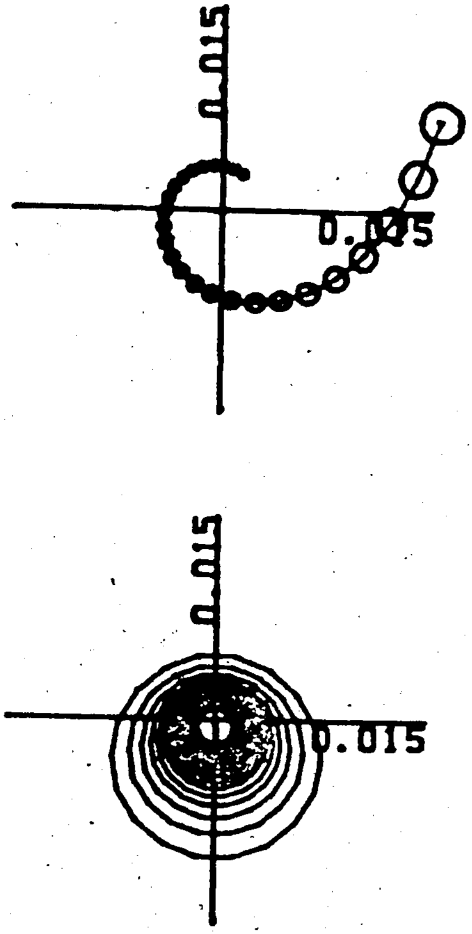


Figure 5.3 Nyquist plots of diagonal elements for $\underline{\underline{G}}(s)$ plus Gershgorin circles for Xb-St and $X\bar{d}$-Re pairing of variables

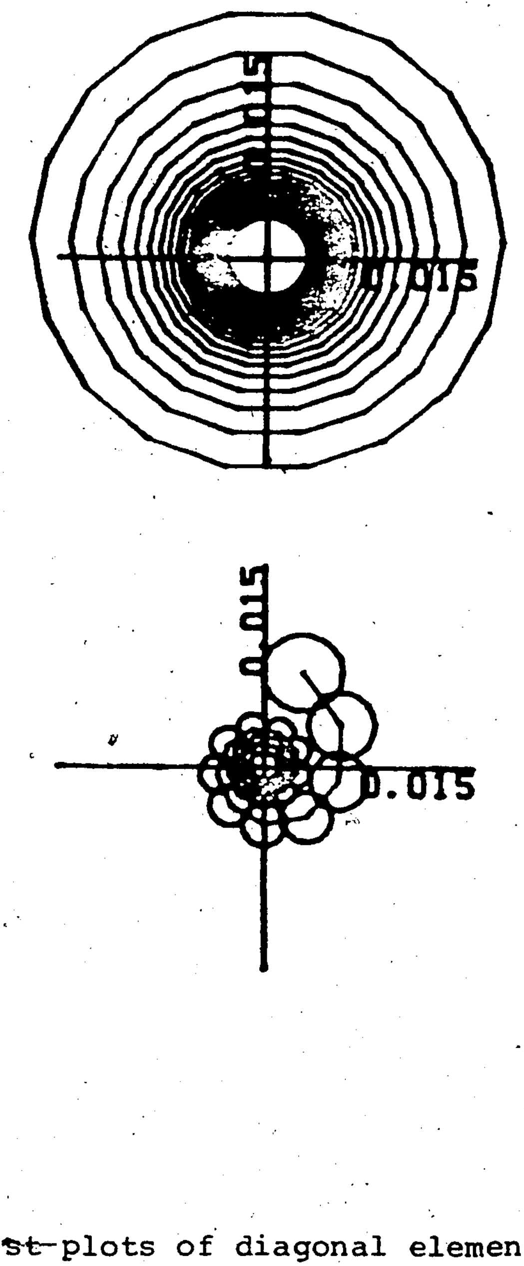


Figure 5.4 Nyqui~~st~~-plots of diagonal elements for $\underline{\underline{G}}(s)$ plus Gershgorin circles for Xb-Re and X$\bar{d}$-St pairing of variables.

and manipulated variables. The method is based on the steady state gains; that is, when s=0. The steady state gains (wt.%/g/s) for the column are

$$\underline{\underline{G}}(0) = \begin{bmatrix} -8.841 & 4.218 \\ -0.631 & 0.542 \end{bmatrix} \qquad 5.8$$

If $\underline{\underline{B}}$ is defined as follows

$$\underline{\underline{B}} = (\underline{\underline{G}}(0)^{-1}) \qquad 5.9$$

therefore,

$$\underline{\underline{B}} = \begin{bmatrix} -0.254 & -1.980 \\ 0.296 & 4.150 \end{bmatrix} \qquad 5.10$$

with the relative gain matrix being defined as
$\underline{\underline{C}}_{ij} = \underline{\underline{G}}(0)_{ij}\underline{\underline{B}}_{ij}$ or

$$\underline{\underline{C}}_{ij} = \begin{bmatrix} 2.25 & -1.25 \\ -1.25 & 2.25 \end{bmatrix} \qquad 5.11$$

Based on the relative gain matrix, top and bottom compositions should be controlled by reflux and steam flow rates respectively. Both methods agree on the pairing of variables. Both are based on a measurement of steady-state interactions only.

The next step was to design a plant precompensator to reduce the interaction between the two control loops. The magnitude of the Gershgorin circles is an indication of the degree of interaction. The key to the selection of the compensator is to minimize the magnitude of the Gershgorin circles. The smaller the radius of the Gershgorin circles indicates less interaction. Three different compensators were designed, that is,

(1) static only

$$\underline{\underline{K}}c_{\bar{1}} = \begin{bmatrix} 1.000 & 0.477 \\ 1.164 & 1.000 \end{bmatrix} \qquad 5.12$$

(2) static plus time delay

$$\underline{\underline{K}}c_2(s) = \begin{bmatrix} 1.000 & 0.477e^{-256s} \\ 1.164 & 1.000 \end{bmatrix} \qquad 5.13$$

(3) decoupler.

$$\underline{\underline{K}}c_3(s) = \begin{bmatrix} 1.000 & \dfrac{0.477(1037s + 1)e^{-256s}}{(1119s + 1)} \\ \dfrac{1.164(369s + 1)}{(860s + 1)} & 1.000 \end{bmatrix} \qquad 5.14$$

For the static compensator (Equation 5.12) which is shown in Figure 5.5 the interaction was reduced in the steam loop but increased in the reflux loop. The above conclusion was arrived at by comparing Figure 5.5 with Figure 5.3. This occurred due to the fact that the original system (Figure 4.17) is an upper triangular matrix. The controller matrix can be designed without a precompensator because the system is inherently decoupled. For this reason the static precompensator did not substantially improve the system in comparison with no precompensator as shown in Figure 5.3. It is not until the time delay (Equation 5.13) is added to the precompensator that there is a significant reduction in

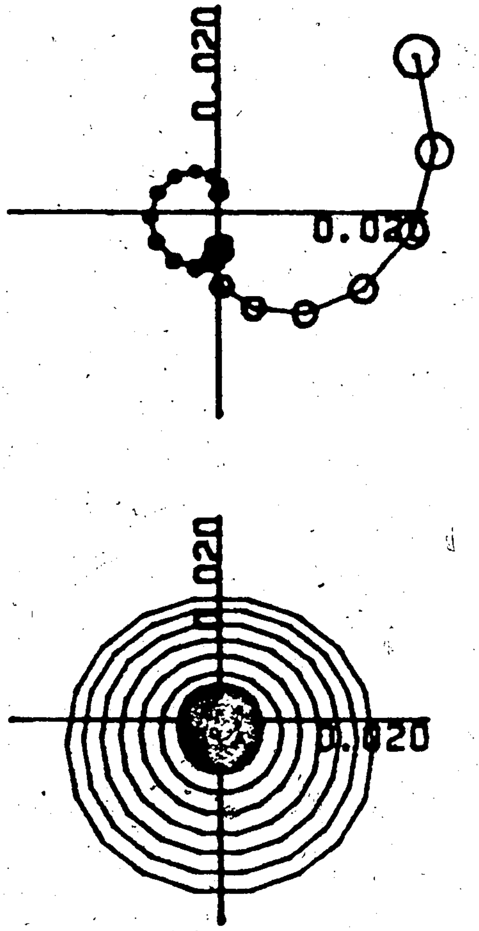


Figure 5.5 Nyquist plots of diagonal elements for $\underline{Q}(s)\underline{\underline{K}}o_1(s)$ system plus Gershgorin circles showing column dominance.

interaction between the two loops as shown in Figure 5.6. Finally, adding the dynamics to the precompensator (Figure 5.7) completely reduces the interaction or decouples the loops. The absence of the Gershgorin circles is the indication that the two loops are decoupled.

5.4 Implementation of Control Law

The different compensators were evaluated through simulation studies using the nonlinear distillation column model for ±20% feed disturbances. The Z-transform of the compensator was taken and the resulting form implemented in the computer. For each of the following equations u_1 and u_2 denotes the change in the steam and reflux flows respectively, while the prime symbol (′) associated with u_1 and u_2 is the change determined by the PID algorithm for the manipulated variables. For the static compensator the following two equations were used,

$$u_1(z) = u_1'(z) + 0.477u_2'(z) \quad 5.15$$

$$u_2(z) = 1.164u_1'(z) + u_2'(z) \quad 5.16$$

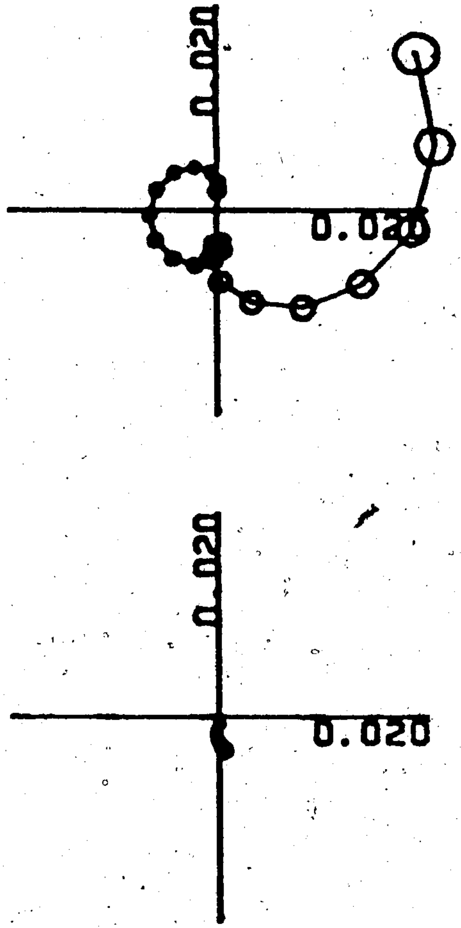


Figure 5.6. Nyquist plots of diagonal elements for $\underline{Q}(s)\underline{K}c_2(s)$ system plus Gershgorin circles showing column dominance.

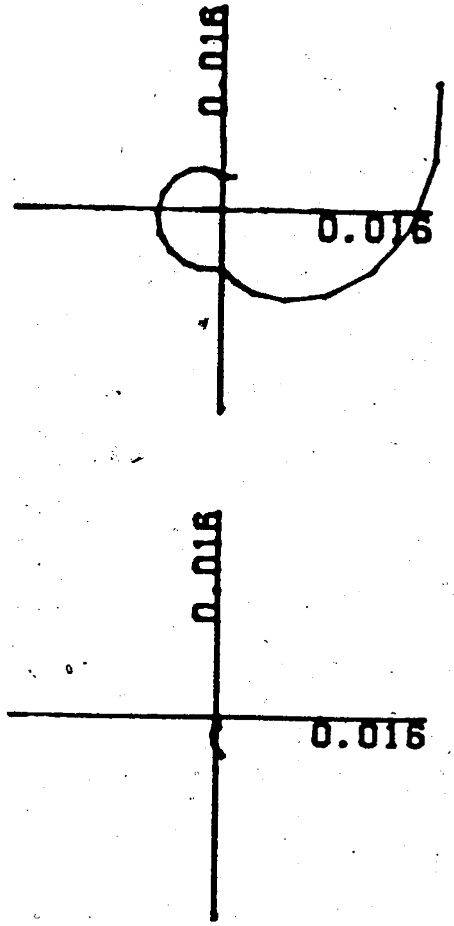


Figure 5.7 Nyquist plots of diagonal elements for $\underline{Q}(s)\underline{K}c3(s)$ system plus Gershgorin circles showing column dominance.

The static plus time delay compensator used the following equations

$$u_1(z) = u'_1(z) + 0.477u'_2(z)z^{-2} \qquad 5.17$$

$$u_2(z) = 1.164u'_1(z) + u'_2(z) \qquad 5.18$$

And finally, the form of the decoupler (Equation 5.13) used is as follows

$$u_1(z) = u'_1(z) + 0.892(u_1(z)z^{-1} - u'_1(z)z^{-1}) + 0.442u'_2(z)z^{-2} - 0.391u'_2(z)z^{-3} \qquad 5.19$$

$$u_2(z) = u'_2(z) + 0.928(u_2(z)z^{-1} - u'_2(z)z^{-1}) + 0.500u'_1(z) - 0.417u'_2(z)z^{-3} \qquad 5.20$$

In all the above expressions u is the delta change to be added to the manipulated variable while u' is the output of the PID algorithm as defined by Equation 5.21.

$$u_n = K_p(e_n + Pt/T_i \sum_{k=1}^{n} e_k + (T_d/Pt)(e_n - e_{n-1}) \qquad 5.21$$

The above equation is the PID algorithm based on the time domain form.

Before simulation studies were carried out on the different compensators for dual control, studies on the individual loops were conducted. Proportional plus integral (PI) control was less than successful in controlling bottom composition because of time delay and the large time constant. PI was successfully used on the top composition because of the smaller time constant and the absence of a time delay. The criteria used in determining the best controller constants was to minimize the Integral of Absolute Error (IAE) value. Therefore, in this study when control performance is said to increase or decrease, it is in relationship to the IAE value. Thus, a lower IAE value would mean better control performance for the control loop. However, the IAE criterion does not penalize control behaviour for oscillatory effects and for that reason it may be less than optimal as a criterion to use. Tables 5.1 and 5.2 summarize the controller settings used to control the bottom and top compositions with the control behaviour shown in Figures 5.8 to 5.16 inclusive. Table 5.1 shows controller settings for single variable control that is, top and bottom composition control.

Table 5.1 PID Constants Used For Figures 5.8 - 5.13

Figure	K_p	T_i	T_d	Pt
5.8 (1)-Xb	27.0	658.2	164.5	128
(2)-Xb	27.0	8000.0	164.5	128
(3)-Xb	27.0	8000.0	164.5	64
5.9 (1)-Xb	5.1	1497.6	374.4	128
(2)-Xb	5.1	2995.2	374.4	128
(3)-Xb	6.6	2995.2	374.4	128
(4)-Xb	6.6	2995.2	374.4	64
5.10 (1)-Xb	3.1	2074.1	518.5	128
(2)-Xb	5.1	1497.6	374.4	128
(3)-Xb	27.0	658.2	164.5	128
5.11 (1)-Xb	3.1	2074.1	518.5	128
(2)-Xb	5.2	2074.1	518.5	128
(3)-Xb	6.0	2074.1	700.5	128
(4)-Xb	5.2	2074.1	518.5	64
5.12 (1)-Xd	24.8	1000.0	-----	128
(2)-Xd	24.8	1000.0	-----	64
(3)-Xd	24.8	1000.0	-----	32
5.13 (1)-Xd	24.8	1000.0		128
(2)-Xd	33.0	600.0	150.0	128
(3)-Xd	95.0	3000.0	150.0	32

Table 5.2 PID Constants Used For Dual Control

(A) -20% Feed Disturbance

	Xb	Xd
K_p	6.0	33.0
T_i	2074.7	600.0
T_d	700.5	150.0
Pt	128.0	128.0

(B) +20% Feed Disturbance

	Xb	Xd
K_p	10.5	33.0
T_i	1573.2	600.0
T_d	393.3	150.0
Pt	128.0	128.0

5.5 Discussion of Results

The individual loops were tuned using the Ziegler Nichols (Z-N) method. Figure 5.8 shows the PID Z-N controller constants if the bottom composition control loop had no time delay. If the integral action was reduced or the sampling interval decreased by one half, the control behaviour improved as indicated by the lower IAE values. It should be remembered that the Z-N method for tuning controllers is for continuous systems, but as indicated by Figure 5.8, the method does perform quite well as an initial guess for sampled data systems. Figure 5.9 displays bottom composition control for a time delay of one sampling interval (128 seconds) using PID control action. The controller settings were determined again by using the Z-N method as the initial guess. In this case, the Z-N controller settings had to be modified more than in the previous case, that is, the integral action reduced and the proportional action increased before the IAE value was reduced. Again, decreasing sampling interval by one half reduces the IAE value considerably. Figure 5.10 indicates the decreasing control performance of a PID controller (Z-N tuned) as the time delay is increased. The largest decrease in control performance (or increase in the IAE value) occurred when the first time delay of one sampling interval was added. The addition of the second sampling interval to the time delay (256 seconds) was not as significant in

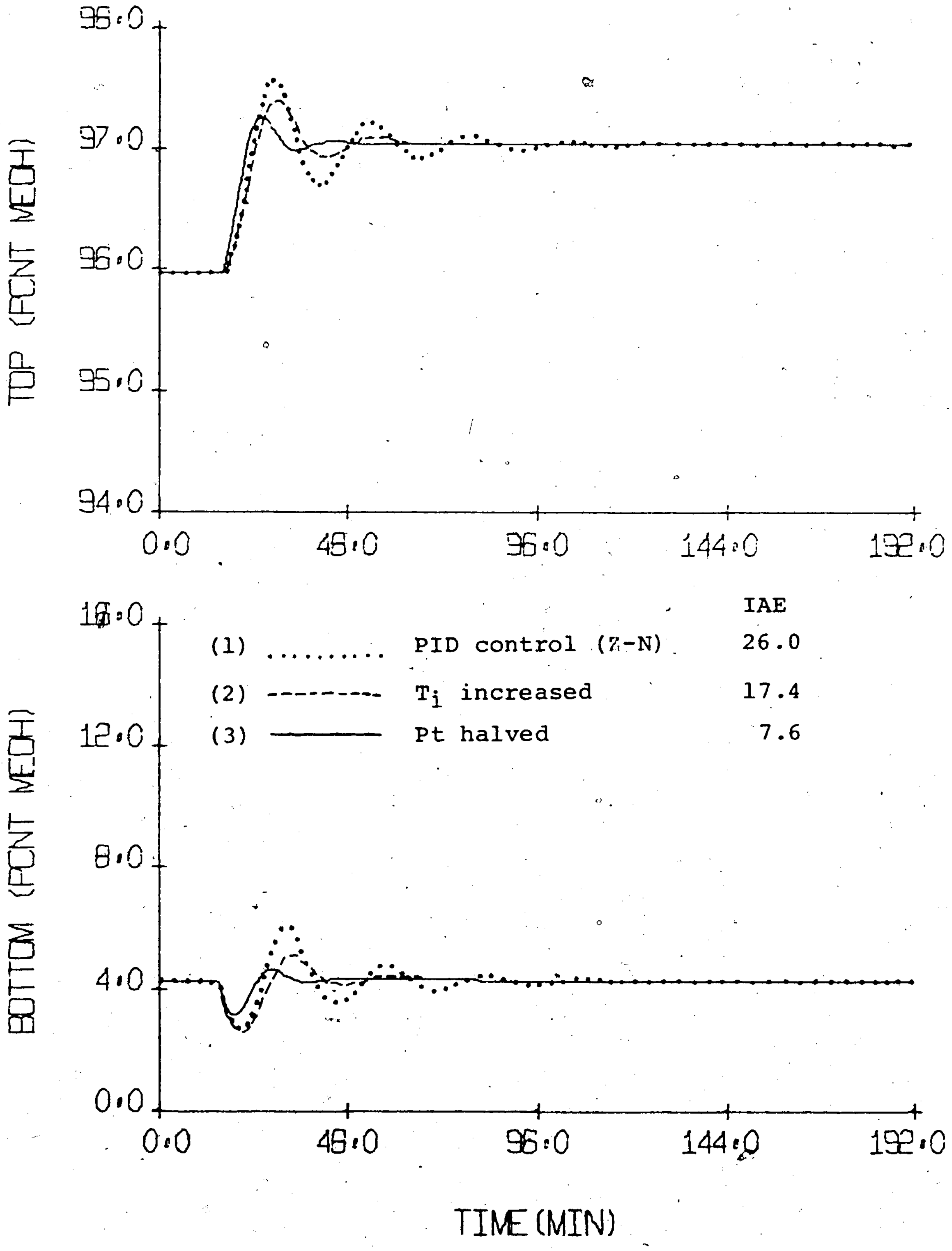


Figure 5.8 Comparison of bottom composition control (no time delay) for different controller settings and sampling interval halved.

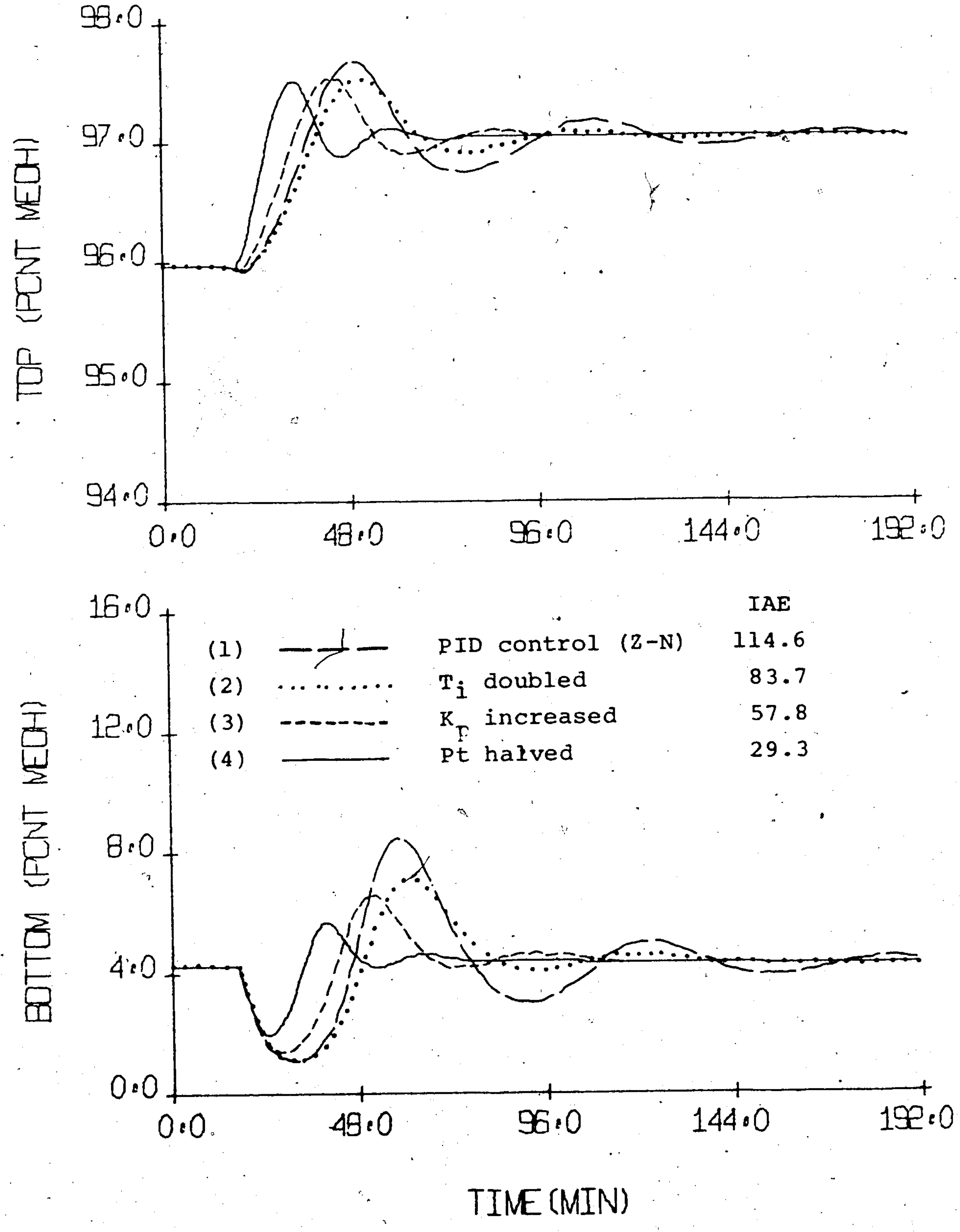


Figure 5.9 Comparison of bottom composition control (time delay = 128 seconds) for different controller settings and sampling interval halved.

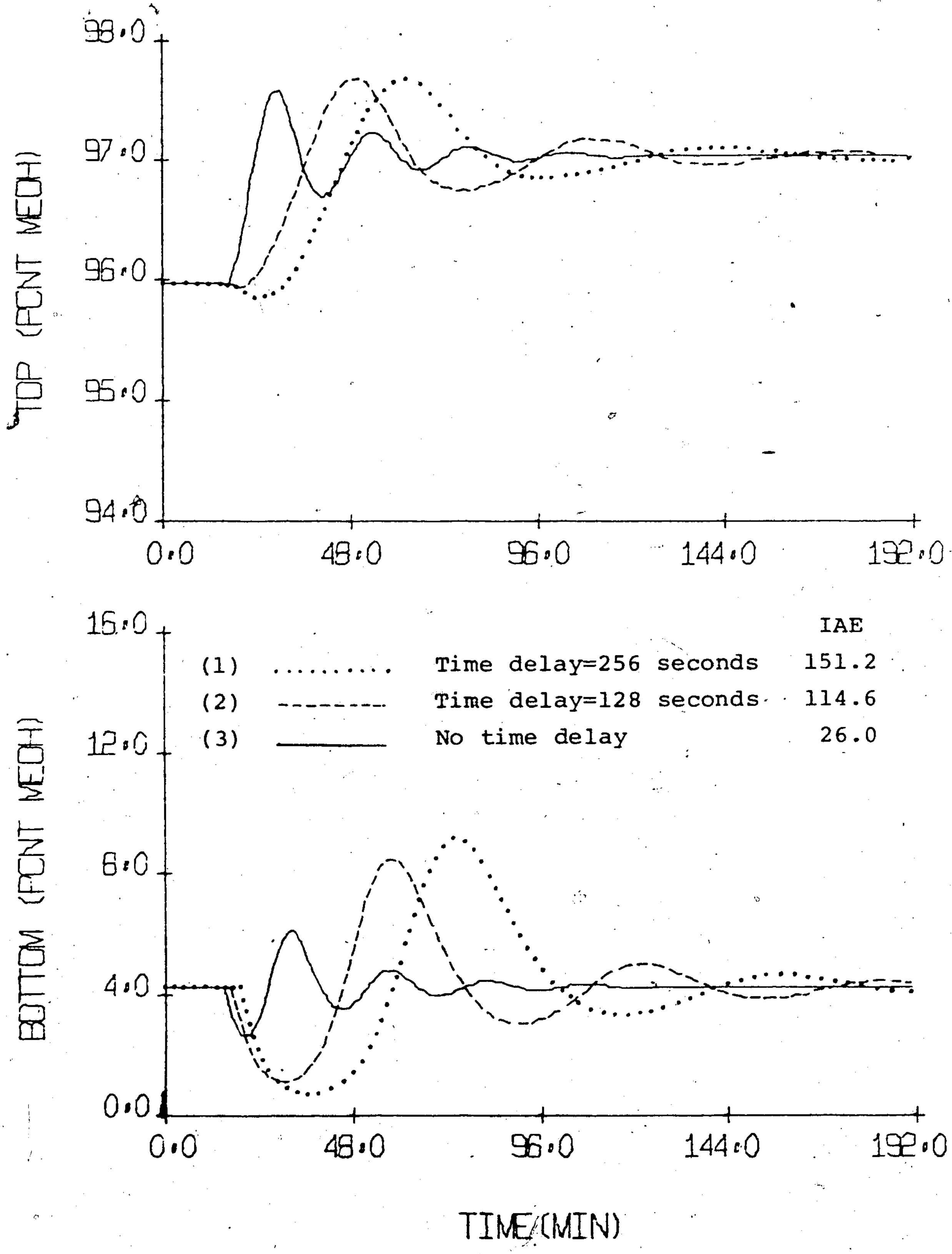


Figure 5.10 Comparison of bottom composition under PID control (Z-N tuned) for different time delays with same sampling interval.

decreasing the control behaviour. This is true until the point is reached at which the system becomes unstable. The basis of comparison for all three curves is the use of the Z-N method of tuning. Figure 5.10 illustrates that for systems with large time delays, PID control performance decreases when compared to systems with no time delay. Figure 5.11 shows an attempt to improve the control performance, that is by reducing the IAE value. For a 256 second time delay it was necessary to use the ultimate gain as determined by Z-N method, a decreased value for integral time and also increase the derivative action. This result provides a good example of how derivative action can improve control performance of systems with time delays. It should be noted that reducing the sampling interval in the above case has a lesser effect (IAE being reduced) than when the sampling period was reduced for the cases of 128 second time delay and no time delay.

Next, the top composition control loop was tuned. Again, the loop was Z-N tuned using a sampling interval of 128 seconds. Experimentally, the sampling interval of one second was possible for the top loop because a capacitance probe was used to measure top composition. Figure 5.12 shows that PI control can effectively be used for the top composition. The sampling interval was halved two times from the original 128 seconds and the performance improved substantially each time. This shows clearly that the Z-N method is definitely more accurate when used for continuous

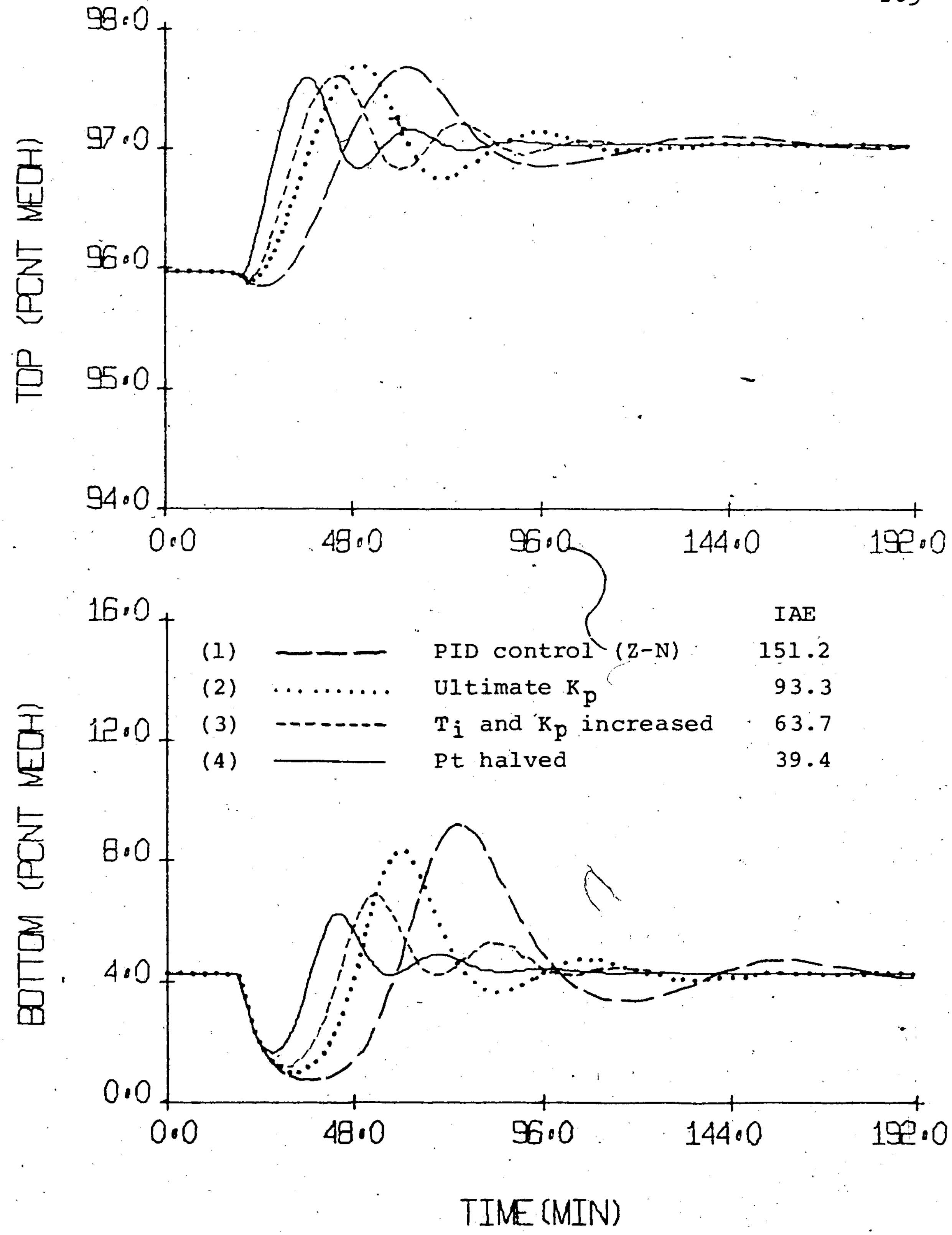


Figure 5.11 Comparison of bottom composition control (time delay = 256 seconds) for different controller settings and sampling interval halved.

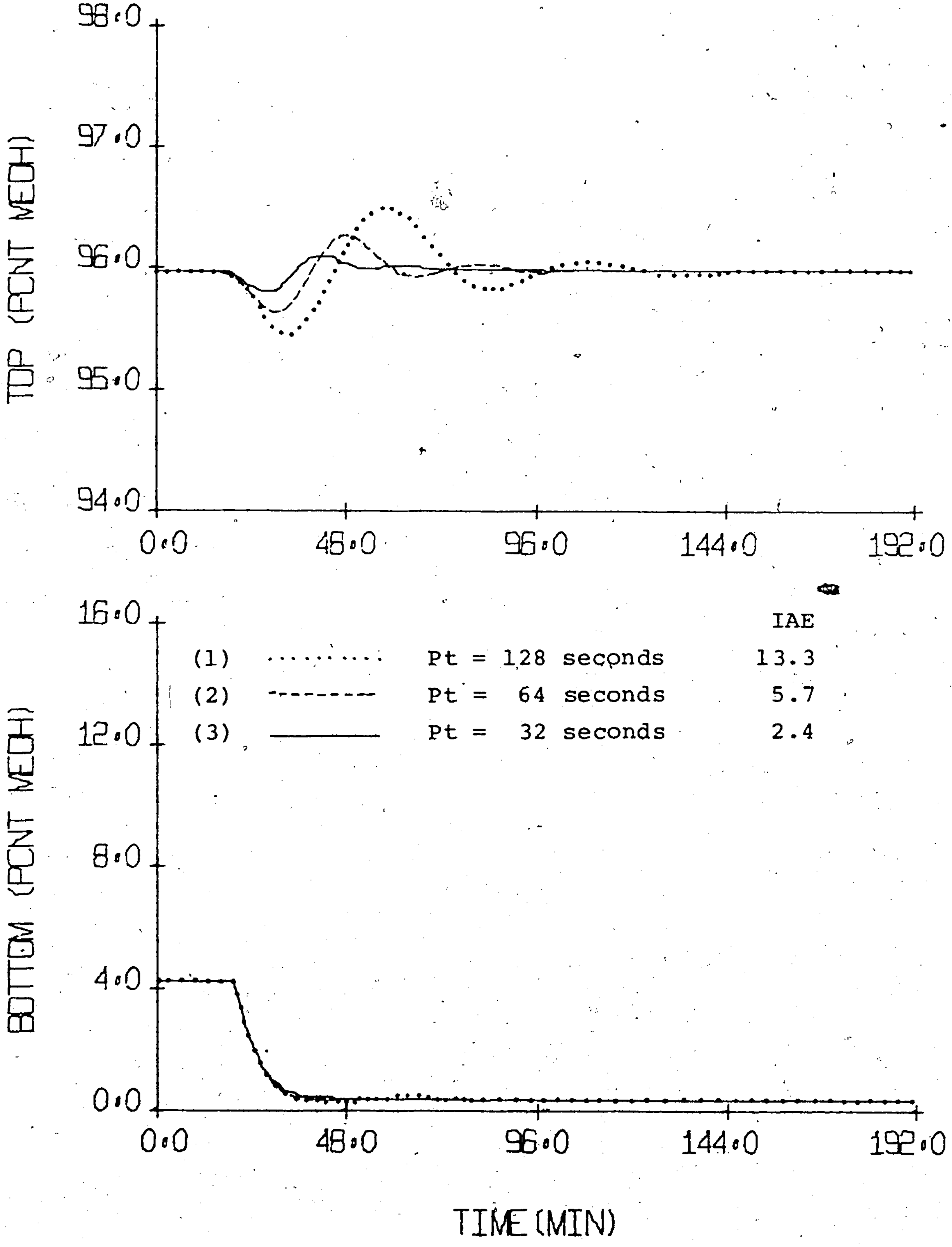


Figure 5.12 Comparison of top composition control only using PI control action for different sampling intervals.

systems, that is, as the sampling interval becomes continuous, control performances improves. It is interesting to note that the bottom composition behaved as if only a feed disturbance was experienced. This indicates that the reflux has little influence on the bottom composition during a -20% feed disturbance. Figure 5.13 shows the differences in control behaviour between PI and PID Z-N tuned controller settings using a sampling interval of 128 seconds. With the addition of derivative action, the IAE value was almost halved indicating better control. The final curve shown on Figure 5.13 shows the control behaviour of a tightly tuned PID loop using a sampling interval of 32 seconds. And finally, tight tuning the controller gave no deviation from the setpoint.

For multivariable control systems, if it is a multiloop case (that is no compensation for loop interactions) then different sampling times for each control loop are acceptable. As soon as the loops are connected in an interacting control system (use of a plant precompensator) then the sampling interval must be the same for all loops, otherwise the system becomes unstable. The above behaviour was encountered while performing control simulations. In this study, the smallest sampling interval is the maximum sampling interval of the two loops of which the gas chromatograph is capable (128 seconds).

The following three figures show results for the cases when the two control loops are compensated for the

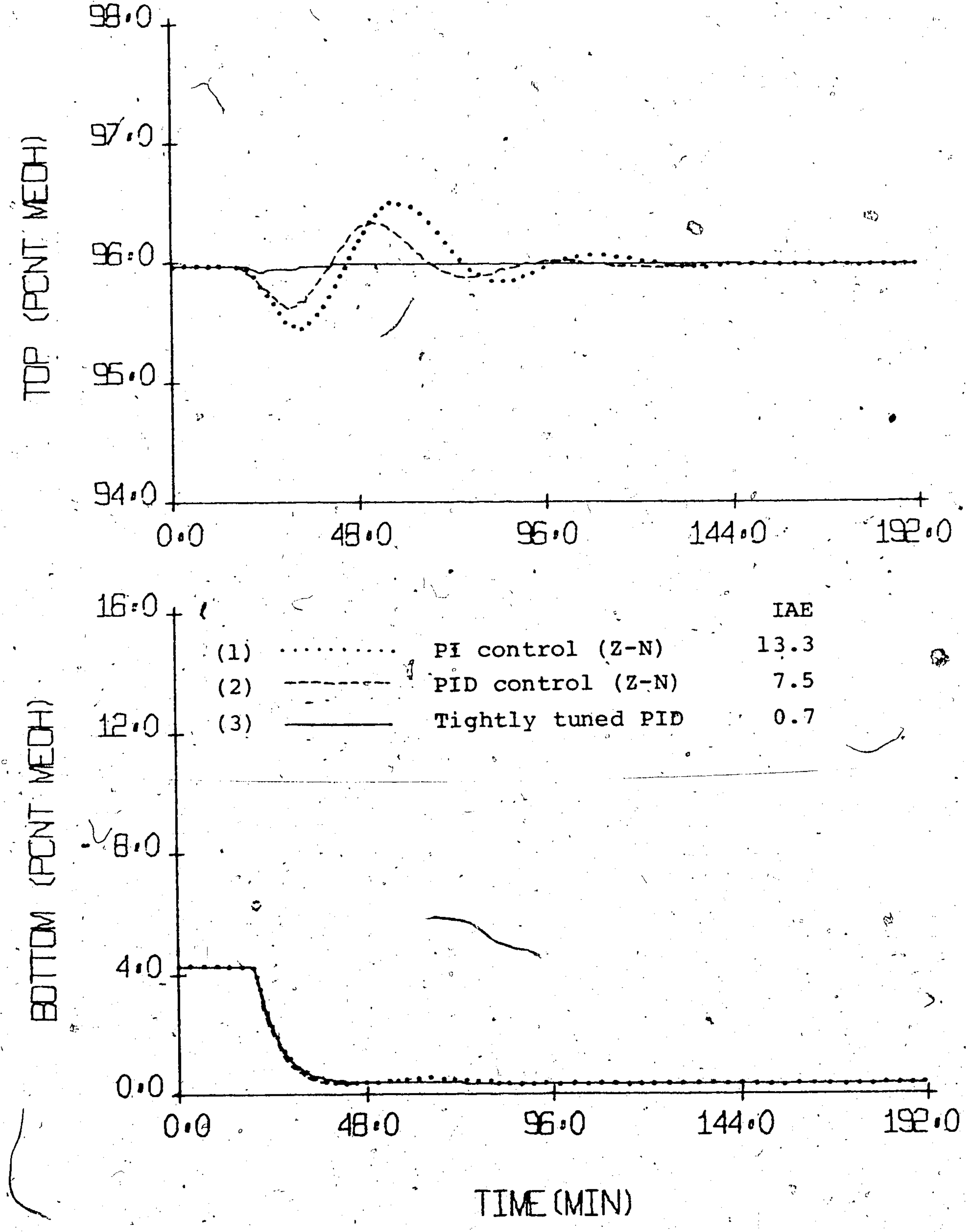


Figure 5.13 Comparison of top composition control for different control action and controller settings.

respective effects on each other. Figure 5.14 makes a comparison between the multiloop case and the three types of compensators designed by the DNA method for dual control. For the multiloop case, the two control loops deteriorated in control performance when compared to their individual loop performance ,that is, the top control loop IAE value increased from 7.5 to 17.3 and while the bottom control loop IAE value increased from 63.7 to 105.0. The top performance improved for all three types of compensators tried. The decoupler was the only precompensator to achieve an IAE value less than the IAE value for bottom composition in comparison to the mutiloop case. The top control loop IAE value increased from 7.6 to 15.4 but remained less than the mutiloop case. Attempts to retune the control loops for the multivariable control system lead to better control behaviour, but the IAE values for single loop control could not be reached. The reason for the less than optimal results for the compensator, that is, static and static plus time delay, is the fact that the system is a dominant system. The steam flow determines the performance of the top control loop and has a greater effect on the column operation than the reflux flow. The system is an upper triangular matrix and the control strategy for the distillation column can be designed using PID control without compensation.

Figure 5.15 present results that are similar to Figure 5.14, except that the sampling interval has been reduced by

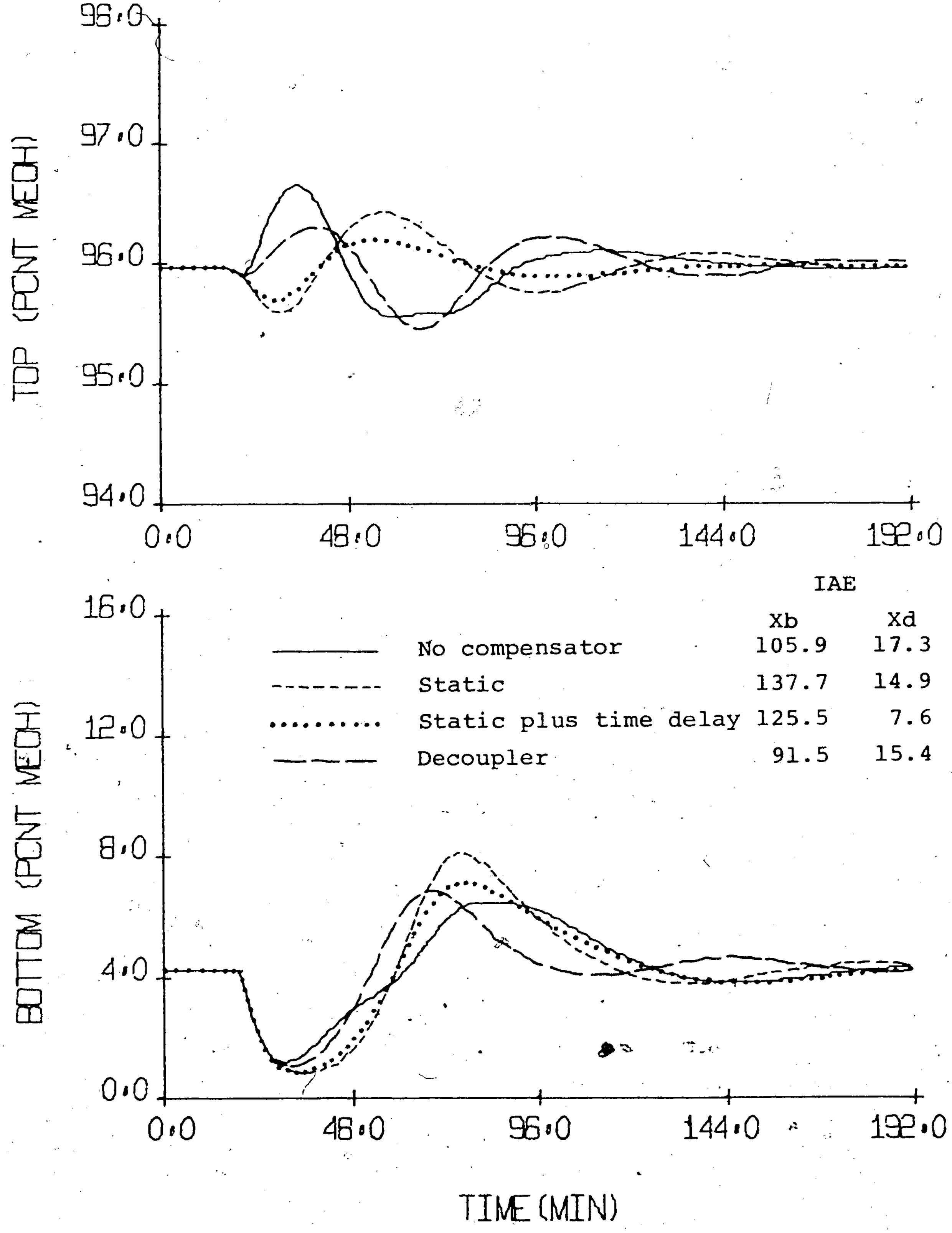
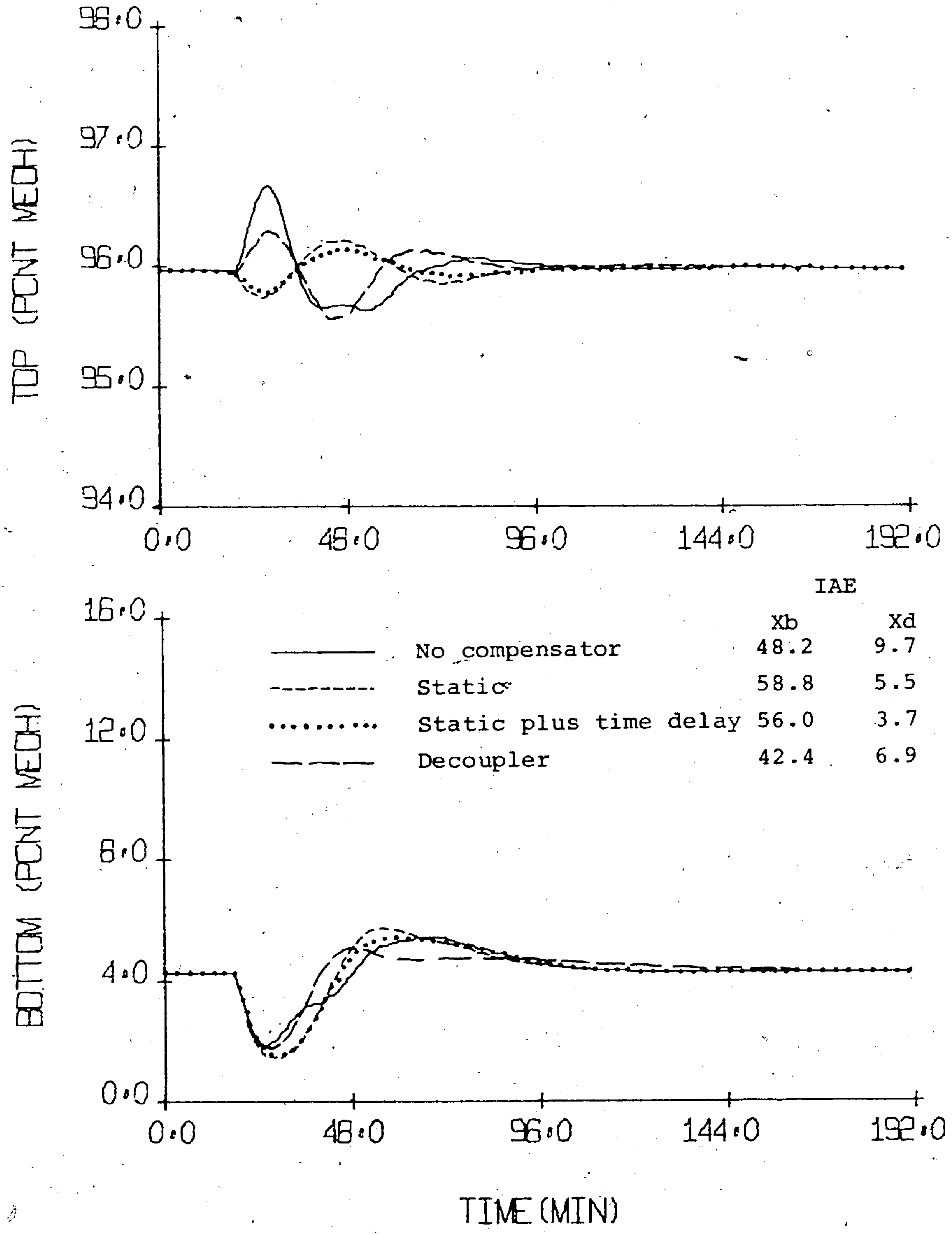


Figure 5.14 Comparison of multiloop and multivariable control systems for dual control with sampling interval of 128 seconds (-20% feed disturbance).



		IAE	
		Xb	Xd
————	No compensator	48.2	9.7
--------	Static	58.8	5.5
·········	Static plus time delay	56.0	3.7
— — —	Decoupler	42.4	6.9

Figure 5.15 Comparison of multiloop and multivariable control systems for dual control with sampling interval of 64 seconds (-20% feed disturbance).

one half for both control loops and consequently the performance has improved. The results shown for the PID controller of 128 seconds were similar except the overshoot of bottom composition was reduced significantly while the top composition performance stayed about the same.

Table 5.3 summarizes static gain values used along with the resulting IAE values when attempts were made to improve the overall control performance. In Table 5.3 the different compensators are represented in the following manner, that is, "1" for the static, "2" for the static plus time delay, and "3" for the decoupler. The effects of changes in the compensator gain values were investigated since it was felt that this would have the greatest effect on the performance of the interacting control system. Therefore the static gains K_1 and K_2 were varied by ±10%. Decreasing the static gains individually and simultaneously did improve IAE values when compared to the first two cases. Using the controller PID constants for the tight top composition control the IAE values decreased for bottom and top loops with a greater percentage for the top.

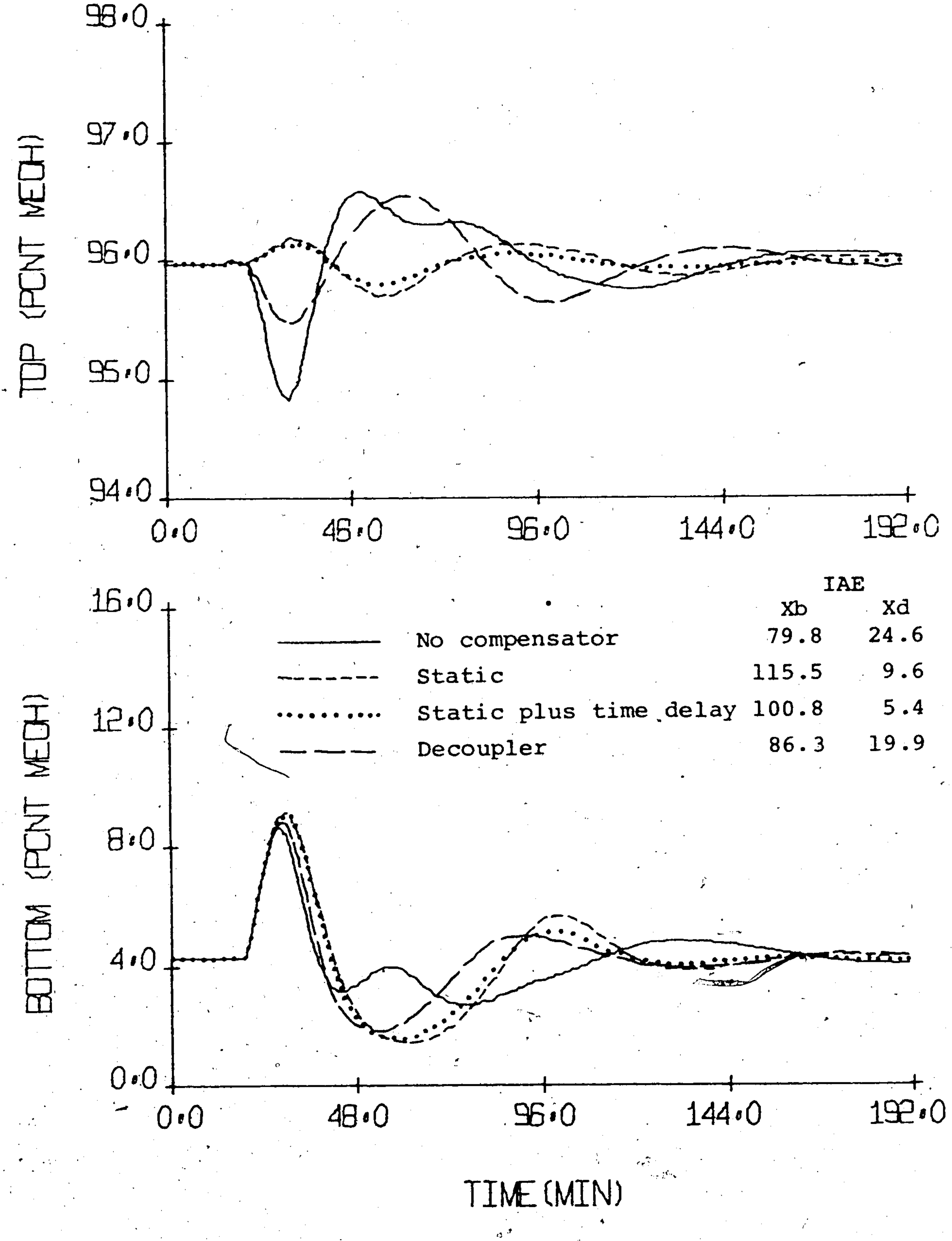
Figure 5.16 shows the control performance of the mutiloop and three compensators for a +20% feed disturbance. Since an increase in feed does not affect the top loop (Figure 3.8) the same controller settings as were used in the -20% disturbance multiloop case were again chosen. The bottom control loop PID constants had to be retuned for the +20% feed disturbance to the values shown in Table 5.2. For

Table 5.3 Effect of Compensator Gains K_1 and K_2 on Control Performance

		Compensator Type					
		1		2		3	
K_1	K_2	IAE1	IAE2	IAE1	IAE2	IAE1	IAE2
0.477	1.164	137.7	14.9	125.5	7.6	91.5	15.4*
0.477	1.164	131.0	4.7	124.6	3.1	87.6	5.1**
0.525	1.164	159.4	17.2	131.6	7.9	96.1	16.3*
0.429	1.164	123.4	12.9	120.1	7.3	89.4	14.6*
0.477	1.280	162.0	19.0	135.3	9.6	90.4	13.5*
0.477	1.048	117.8	13.0	118.0	7.1	93.2	17.5*
0.429	1.048	110.1	11.5	113.4	6.8	89.8	16.5*

* -- PID constants used for -20% feed disturbance

** -- PID constants using tight top control with bottom loop constants unchanged.



	IAE Xb	 Xd
———— No compensator	79.8	24.6
- - - - - - Static	115.5	9.6
········· Static plus time delay	100.8	5.4
— — — Decoupler	86.3	19.9

Figure 5.16 Comparison between multiloop and multivariable control systems for dual control of column using sampling interval of 128 seconds (+20% feed disturbance).

the increase case, it is interesting to note that the IAE value for the bottom loop was not improved over the multiloop case, although the top loop performed better for all three cases. Performance again would improve if the sampling interval were decreased by one half.

Chapter 6

CONCLUSIONS

Conclusions can be drawn from the three phases of this work. The first phase was to develop an accurate nonlinear model to describe the dynamic behaviour of the distillation column. The second phase of the thesis work was to develop a linear model in the form of transfer functions relating the chosen controlled variables to manipulated variables. The linear model was used to help design compensators to reduce the interaction between controlled and manipulated variables. The final stage was to design the compensators using the direct Nyquist array method and then to compare the different compensators in dual control studies using the nonlinear model. The following conclusions were made;

(1) Experimental data for the bottom and top compositions closely matched the simulated results for the nonlinear distillation column model for steam, reflux, and feed flow disturbances. The nonlinear model used in this case was a constant holdup, efficiency, and heat loss based on the initial steady state operating conditions.

(2) The nonlinear model using variable efficiency and heat loss produces values for the final steady state bottom and top compositions which match identically to the experimental results. The variable model worked well for both steam and reflux disturbance but was not useful for the

feed disturbance because the top composition which was the reference composition used was insensitive to feed flow increases. As expected, the heat loss and efficiency between the initial and final steady states did not vary significantly.

(3) Significant improvement was accomplished in the analysis of the bottom composition using the gas chromatograph. The sampling interval was reduced by one halve to 128 seconds. The long transportation delay which was experienced by previous co-workers was completely eliminated. Finally, the accuracy of measuring low concentrations (less than 1.0% methanol) was improved by the use of a different column packing for the gas chromatograph.

(4) A procedure for determining the liquid mass holdup on the trays more accurately was developed.

(5) The PRBS identification technique used was very accurate in determining the time delay of the transfer functions considered. The identification technique gave some consistent results for the parameters between the PRBS sequences tried, but the gains were not close to the experimental results obtained from the open loop responses.

(6) The results between the open loop responses and the PRBS studies differed due to the use of the large sampling interval and only one magnitude of disturbance.

(7) A non-interactive control strategy for the distillaton column does slightly improve dual control when compared to the multi-loop case for <u>dominant</u> multivariable

systems.

(8) A non-interacting control system only reduces interactions in control loops for the mulitivariable system and does not improve loop performance when compared to single loop control performance.

(9) Significant improvement of the multivariable control performance was achieved by reducing the sampling interval.

(10) A non-interacting control structure has a greater effect on systems that have similar effects on each other's control loops than for a dominant multivariable control systems.

(11) If the dynamic elements of each control loop are similar in the precompensator, then improved control performance can be obtained by using only the static gains.

NOMENCLATURE

A - constant of pulse tranfer function

$\underline{\underline{A}}_1$-$\underline{\underline{A}}_2$ - material balance equation matrix

$\underline{\underline{A}}_3$-$\underline{\underline{A}}_4$ - energy balance equation matrix

a - intercept for heat capacity at 0°C (J/g)

B - constant of pulse tranfer function

b - constant in heat capacity (=4.184 J/g-°C)

$\underline{\underline{C}}_{ij}$ - relative gain matrix

E - tray efficiency

Ef - final efficiency

Ei - initial efficiency

e - error = (setpoint - measurement)

F - feed flow rate (g/s)

$\underline{\underline{F}}(s)$ - m x m return difference matrix

$\underline{FT}$ - right hand side of matrix $\underline{\underline{A}}$

$\underline{\underline{G}}(s)$ - m x t plant transfer function matrix

H - vapour enthalpy (J/g)

$\underline{\underline{H}}$ - tridiagonal composition matrix

Hfe - feed flow enthalpy (J/g)

Hre - reflux flow enthalpy (J/g)

Hs - steam enthalpy (J/g)

$\underline{\underline{H}}(s)$ - m x m measurement transfer function matrix

h - liquid enthalpy (J/g)

hf - feed enthalpy (J/g)

hs - liquid enthalpy of side stream (J/g)

$\underline{\underline{I}}$ - identity matrix

IAE1 and IAE2 - integral of absolute error for two loops

$\underline{\underline{K}}b(s)$ - controller matrix

$\underline{\underline{K}}c(s)$ - precompensator matrix

K_p - proportional action of PID algorithm

$\underline{\underline{K}}(s)$ - t x m overall precompensator matrix

K_1 and K_2 - gains of compensator matrix

L - liquid mass flow rate (g/s)

Po - roots of open loop characteristic polynomial

$\underline{\underline{P}}(s)$ - zero order hold

Pt - sampling interval

Q - heat loss per stage (W)

Qf - final heat loss (W)

Qi - initial heat loss (W)

QR - heat to reboiler (W)

$\underline{\underline{Q}}(s)$ - m x m open loop transfer function matrix

Re - reflux mass flow rate (g/s)

$\underline{\underline{R}}(s)$ - m x m closed loop transfer function matrix

$\underline{r}(s)$ - m x 1 vector of setpoints

S - side stream mass flow rate (g/s)

St - steam mass flow rate (g/s)

s - Laplace operator

Tc - time constant for continuous transfer function form

T_d - derivative action of PID algorithm

T_i - integral action of PID algorithm

Ts - steam temperature (°C)

$\underline{\underline{T}}(s)$ - m x m return ratio matrix

T_1 - reboiler or first stage temperature (°C)

t_i - number of encirclements of the origin for $\underline{\underline{I}}(s)$
UA - product of heat transfer times area (W-°C)
$\underline{u}(s)$ - m x 1 vector of manipulative variables
u_1 - overall change in steam flow rate
u'_1 - change in steam flow as a result of PID control
u_2 - overall change in reflux flow rate
u'_2 - change in reflux flow as a result of PID control
V - vapour mass flow rate (g/s)
VT - vapour mass holdup of a stage (g)
WT - liquid mass holdup of a stage (g)
X - liquid composition in wt. fraction
Xb - bottom composition % methanol
Xd - top compositon % methanol
Xf - feed composition in wt. fraction (methanol)
Xr - reference liquid composition wt. fraction methanol
Xrf - final reference composition
Xri - initial reference composition
Xs - liquid composition of side stream in wt. fraction
$\underline{y}(s)$ - m x 1 vector plants output
Y - vapour composition in wt. fraction
Y^* - equilbrium vapour composition
z - pulse transfer operator

Subscripts
k - integration interval for model
n - stag number

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APPENDIX A

A.1 Physical Properties for Methanol-Water

The following physical relationships are taken directly from the M.Sc. thesis of Pacey (41) but are presented here in SI units. These functional relationships are used when performing a mass and energy balance on the distillation column. The designations of the functions are as used in computer programs.

(a) density of liquid water

$$\mathrm{DENSW(T)} = 1.0050-0.2142\mathrm{E}-03*\mathrm{T}-0.2508\mathrm{E}-05*\mathrm{T}^2 \qquad \text{A-1}$$

Range: $40 < T < 120$

Units: DENSW(T) - g/cm^3

T - °C

(b) density of methanol-water mixture

$$\begin{aligned}\mathrm{DENS(T,X)} = {} & 1.0201-0.5123\mathrm{E}-03*\mathrm{T}-0.1512\mathrm{E}-05*\mathrm{T}^2\\ & -0.1519\mathrm{E}-02*100.0*\mathrm{X}\\ & -0.8106\mathrm{E}-05*(100.0*\mathrm{X})^2 \qquad \text{A-2}\end{aligned}$$

Range: $40 < T < 120$

$0 < X < 1.0$

Units: DENS(T,X) - g/cm^3

T - °C

X - wt. fraction methanol

(c) heat capacity of liquid methanol

$$CPML(T) = 2.3305+7.8241E-03*T+37.6978E-06*T^2 \qquad A-3$$

Range: $0 < T < 120$

Units: CPML(T) - J/g-°C

T - °C

(d) heat capacity of liquid water

$$CPWL(T) = 4.1882-5.6902E-04*T+8.4935E-06*T^2 \qquad A-4$$

Range: $35 < T < 100$

Units: CPWL(T) - J/g-°C

T - °C

(e) heat of vapourization of water

$$HEATW(T) = 2510.35-9.27E-01*T-13.06E-04*T^2 \qquad A-5$$

Range: $150 < T < 300$

Units: HEATW - J/g

T - °C

(f) enthalpy of saturated methanol-water vapour

$$ENTHV = 2657.4-1424.8*X \qquad A-6$$

Range: $0 < X < 1.0$

Units: ENTHV - J/g
X - wt. fraction methonal

The equilibrium data shown in Table A.1 was taken directly from the Ph.D. thesis of Svrcek (57) except for one value. The one value(*) was replaced by an estimated value after the curve was hand drawn. Figure A.1 is a plot of the values as generated by use of the spline function (12) for increments of 0.001 in liquid composition. Similarly, Table A.2 contains the liquid enthalpy data and Figure A.2 shows the representation using values generated by the spline function for liquid composition versus enthalpy while Figure A.3 shows the generated values using the spline function corresponding to data values from Equation A-6.

Table A.1 Liquid Vapour Equilibrium Data

X COMP wt. fraction methanol	Y COMP	TEMP ° C
0.0000	0.0000	100.00
0.0350	0.2158	96.39
0.0689	0.3469	93.61
0.1020	0.4372	91.39
0.1339	0.5055	89.05
0.1650	0.5609	87.50
0.2026	0.6143	85.40
0.2389	0.6556	84.28
0.2739	0.6870	82.70
0.3078	0.7100	81.67
0.3405	0.7331	80.25
0.3722	0.7500	79.61
0.4028	0.7652	78.78
0.4325	0.7793	78.00
0.4612	0.7927	77.17
0.4892	0.8043	76.60
0.5162	0.8154	75.80
0.5665*	0.8300	74.80
0.5926	0.8449	74.06
0.6400	0.8624	73.22
0.7273	0.8934	71.33
0.8058	0.9224	69.44
0.8767	0.9504	67.94
0.9412	0.9759	66.11
0.9713	0.9881	65.44
1.0000	1.0000	64.70

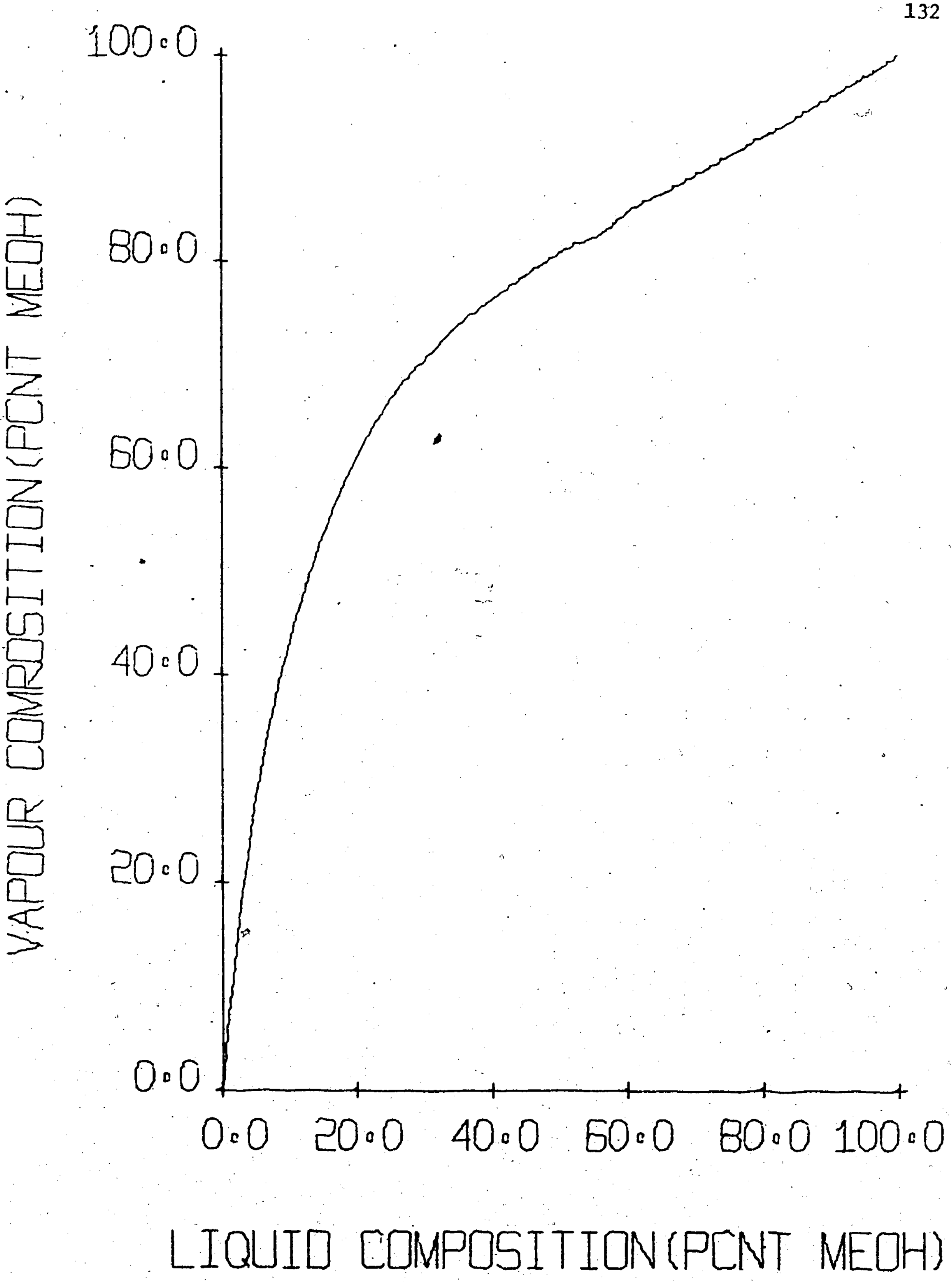

Figure A.1 Liquid vapour equilibrium curve.

Table A.2 Liquid Enthalpy Data

X COMP wt. fraction methonal	ENTHALPY J/g
0.00	418.39
0.05	399.56
0.10	378.64
0.15	355.63
0.20	331.46
0.25	305.43
0.30	280.32
0.35	256.38
0.40	235.46
0.45	221.75
0.50	203.85
0.55	191.53
0.60	182.93
0.65	177.82
0.70	173.63
0.75	170.61
0.80	168.52
0.90	165.50
1.00	163.17

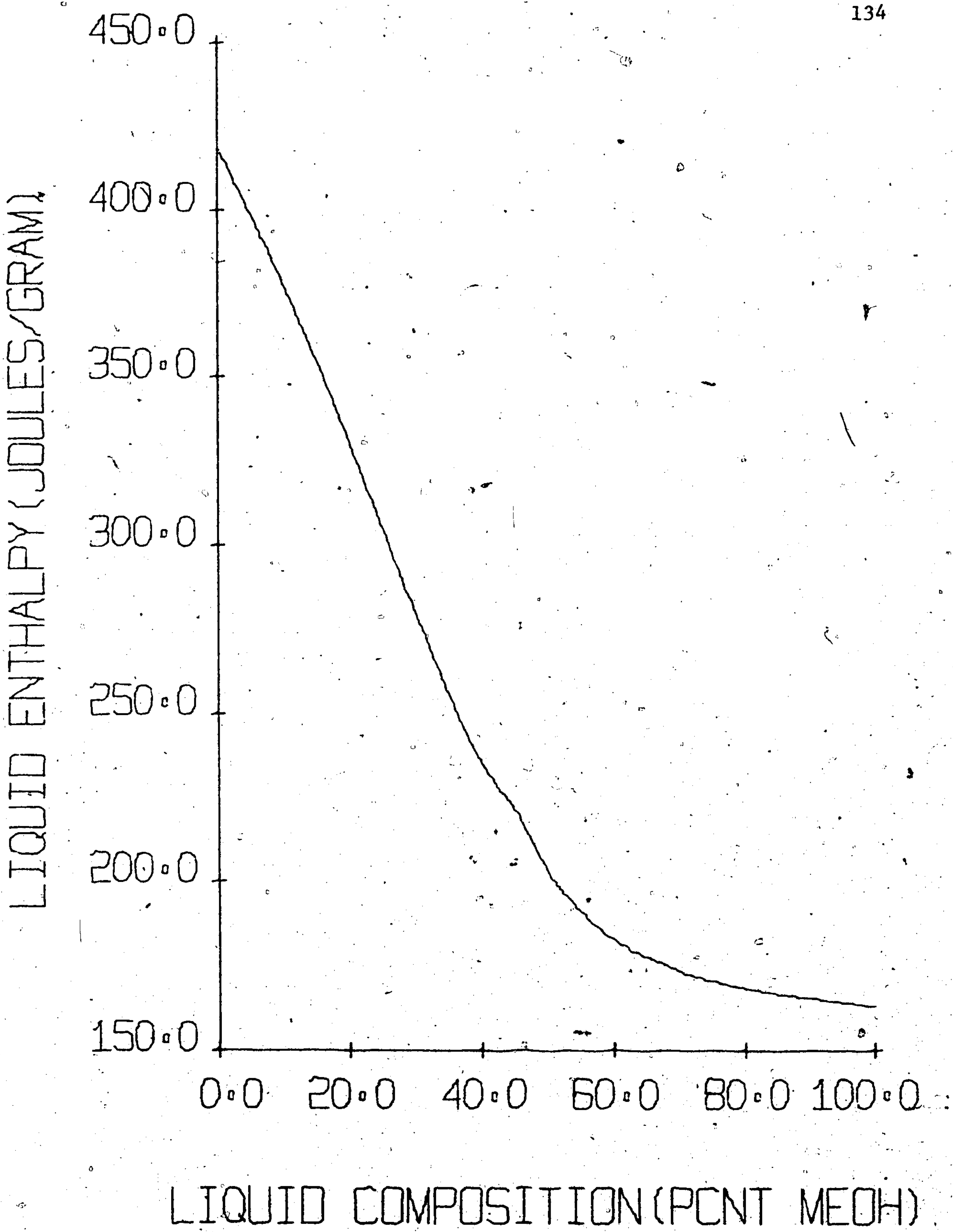

Figure A.2 Liquid enthalpy curve.

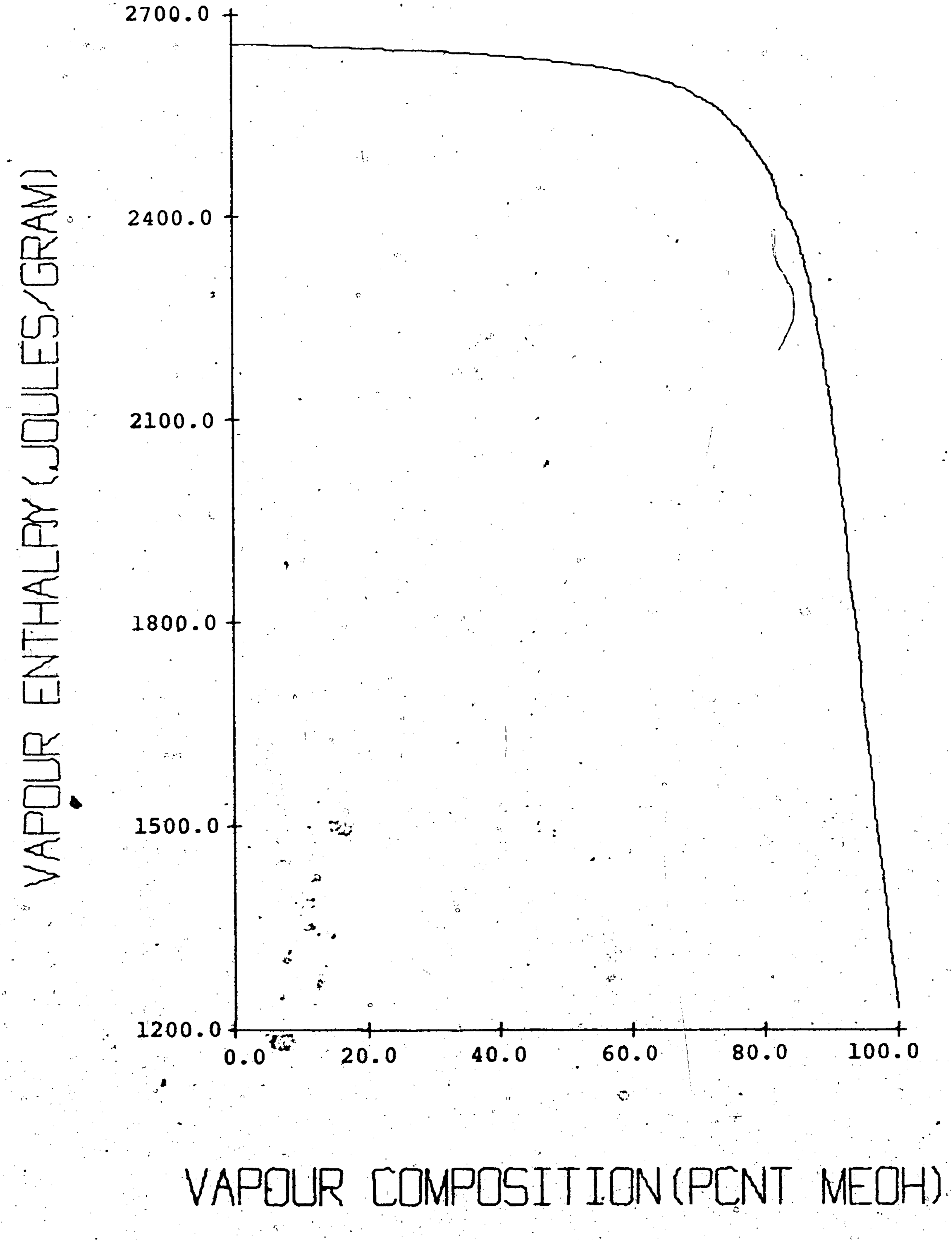

Figure A.3 Vapour enthalpy curve.

APPENDIX B

B.1 Differential Equations

This section outlines the development of the entries for the $\underline{\underline{H}}$ matrix as defined in Equation 2.22. The entries for the $\underline{\underline{H}}$ matrix are established in the following manner. Substituting Equation 2.4 into 2.5 yields

$$WT_n(dX_n/dt) = L_{n+1}X_{n+1} - L_nX_n + V_{n-1}Y_{n-1} - V_nY_n - X_n(L_{n+1} - L_n + V_{n-1} - V_n + F_n) + F_nXf_n \qquad \text{B-1}$$

The equations presented will be for a general tray including the feed stream. The feed stream will be deleted when not applicable. To eliminate the vapour composition from Equation B-1 an equilbrium constant (K) is defined as follows

$$K_n = Y_n/X_n \qquad \text{B-2}$$

which is performed for each stage. The feed composition is eliminated from the feed stage by using the following ratio

$$RR = Xf_n/X_n \qquad \text{B-3}$$

Using the above two equations to eliminate Y_n and Xf_n from Equation B-1 and collecting like terms yields

$$dX_n/dt = X \quad (L_{n+1}/WT_n) + X_n(V_n(1-K_n) - L_{n+1} - V_{n-1}$$
$$+ F_nRR - F_n)/WT_n + X_{n-1}(V_{n-1}K_{n-1})/WT_n \qquad \text{B-4}$$

Writing the equations for each stage results in a 10x10 tridiagonal matrix $\underline{\underline{H}}$ shown in Figure B.1. Each element in the matrix is multiplied by $\Delta t/2$ with blank entries being zero. For the special stages the following equations will apply that is,

$$a_1 = (V_1(1.0 - K_1) - L_2)/WT_1 \qquad \text{B-5}$$

$$a_5 = a_5 + (RRF_5 - F_5)/WT_5 \qquad \text{B-6}$$

$$a_9 = (V_9(1.0 - K_9) - Re - V_8)/WT_9 \qquad \text{B-7}$$

$$a_{10} = -V_9/WT_{10} \qquad \text{B-8}$$

$$c_9 = Re/WT_9 \qquad \text{B-9}$$

$$\begin{bmatrix}
a_1 & c_1 & & & & & & & & \\
b_2 & a_2 & c_2 & & & & & & & \\
 & b_3 & a_3 & c_3 & & & & & & \\
 & & b_4 & a_4 & c_4 & & & & & \\
 & & & b_5 & a_5 & c_5 & & & & \\
 & & & & b_6 & a_6 & c_6 & & & \\
 & & & & & b_7 & a_7 & c_7 & & \\
 & & & & & & b_8 & a_8 & c_8 & \\
 & & & & & & & b_9 & a_9 & c_9 \\
 & & & & & & & & b_{10} & a_{10}
\end{bmatrix}$$

where; $c_n = L_{n+1}/WT_n, n=1,8$

$b_n = V_{n-1}K_{n-1}/WT_n, n=2,10$

$a_n = (V_n(1.0 - K_n) - L_{n+1} - V_{n-1})/WT \ , n=2,8$

Figure B.1 The tridiagonal matrix $\underline{\underline{H}}$ used to solve for liquid composition.

The following analysis was used to solve for the liquid and vapour flow rates as defined by Equation 2.26. The following equation can be obtained by rearranging Equation 2.4.

$$L_{n+1} - L_n + V_{n-1} - V_n = d(WT_n)/dt - F_n \tag{B-10}$$

Using the above equation for the first 9 stages, the upper two partitioned matrices $\underline{\underline{A}}_1$ and $\underline{\underline{A}}_2$ are formed. At this point the effects of changing holdup can be studied by having a method to evaluate the change of holdup. In this study, since constant liquid mass holdup was assumed, the term $d(Wt_n)/dt$ is not retained. Substituting Equation 2.4 into Equation 2.6 and rearranging yields

$$\begin{aligned} &L_{n+1}(h_{n+1}-h_n) + V_n(h_n-H_n) + V_{n-1}(H_{n-1} - h_n) \\ &= Q_n + WT_n d(h_n)/dt - F_n hf_n + F_n h_n \end{aligned} \tag{B-11}$$

Using the above equation for each stage, excluding the condenser, the lower two partitioned matrices $\underline{\underline{A}}_3$ and $\underline{\underline{A}}_4$ can be generated. The complete matrix showing all the entries is shown in Figures B.2 and B.3. The resultant matrix is a 18x18 matrix that is used to solve for the liquid and vapour flow rates on each iteration. Not all of the entries on the right hand side are non zero. They are as follows

$FT_n = 0.0 \qquad n=1,9$ B-12

except

$FT_5 = -F$ B-13

$FT_9 = -Re$ B-14

and

$FT_m = WT_n d(h_n)/dt + Q_n \qquad n=2,9,\ m=n+9,18$ B-15

$FT_{10} = WT_1 d(h_1)/dt + Q_1 - QR$ B-16

$FT_{14} = FT_{14} - F(hf_5 - h_5)$ B-17

$FT_{18} = FT_{18} - Re(hre - h_9)$ B-18

$$\begin{bmatrix}
-1.0 & +1.0 & & & & & & & \\
 & -1.0 & +1.0 & & & & & & \\
 & & -1.0 & +1.0 & & & & & \\
 & & & -1.0 & +1.0 & & & & \\
 & & & & -1.0 & +1.0 & & & \\
 & & & & & -1.0 & +1.0 & & \\
 & & & & & & -1.0 & +1.0 & \\
 & & & & & & & -1.0 & +1.0 \\
 & & & & & & & & -1.0
\end{bmatrix}$$

$$\begin{bmatrix}
-1.0 & & & & & & & & \\
+1.0 & -1.0 & & & & & & & \\
 & +1.0 & -1.0 & & & & & & \\
 & & +1.0 & -1.0 & & & & & \\
 & & & +1.0 & -1.0 & & & & \\
 & & & & +1.0 & -1.0 & & & \\
 & & & & & +1.0 & -1.0 & & \\
 & & & & & & +1.0 & -1.0 & \\
 & & & & & & & +1.0 & -1.0
\end{bmatrix}$$

Figure B.2 Entries for the upper two partitioned matrices $\underline{\underline{A}}_1$ and $\underline{\underline{A}}_2$.

$$\begin{bmatrix}
 & h_2-h_1 & & & & & & & \\
 & & h_3-h_2 & & & & & & \\
 & & & h_4-h_3 & & & & & \\
 & & & & h_5-h_4 & & & & \\
 & & & & & h_6-h_5 & & & \\
 & & & & & & h_7-h_6 & & \\
 & & & & & & & h_8-h_7 & \\
 & & & & & & & & h_9-h_8
\end{bmatrix}$$

$$\begin{bmatrix}
h_1-H_2 & & & & & & & & \\
H_1-h_2 & h_2-H_2 & & & & & & & \\
 & H_2-h_3 & h_3-H_3 & & & & & & \\
 & & H_3-h_4 & h_4-H_4 & & & & & \\
 & & & H_4-h_5 & h_5-H_5 & & & & \\
 & & & & H_5-h_6 & h_6-H_6 & & & \\
 & & & & & H_6-h_7 & h_7-H_7 & & \\
 & & & & & & H_7-h_8 & h_8-H_8 & \\
 & & & & & & & H_8-h_9 & h_9-H_9
\end{bmatrix}$$

Figure B.3 Entries for the lower two partitioned matrices $\underline{\underline{A}}_3$ and $\underline{\underline{A}}_4$.

B.2 Programs Used for Simulation of Model

The simulation of the model is performed by using a set of subroutines designed to carry out specific tasks. Figure B.4 shows the order of executing each subroutine. Along with these subroutines are a few system subroutines plus programs from Forsythe (12). A short description of each subroutine will be given here;

(a) INII -sets up the initial conditions

(b) PRINT -prints desired variables at time equal zero

(c) -next the program checks if it is time for the open or close loop disturbances

(d) COMP - calculates the liquid temperature and the equilibrium vapour composition based on the liquid composition, first using SPLINE (12) and then SEVAL (12). The actual vapour composition is calculated using the Murphree vapour efficiency.

(e) ENTH -calculates the liquid and vapour enthalpy based on the liquid and vapour composition.

(f) VFLOW -calculates the liquid and vapour flow rates using first DECOMP (12) and then SOLVE (12). DECOMP decomposes the matrix into an upper triangle matrix and SOLVE back substitutes to determine the individual flows.

(g) PRIN2 -prints top and bottom composition at specified intervals.

(h) PRINT -prints out desired variables at a specified print interval

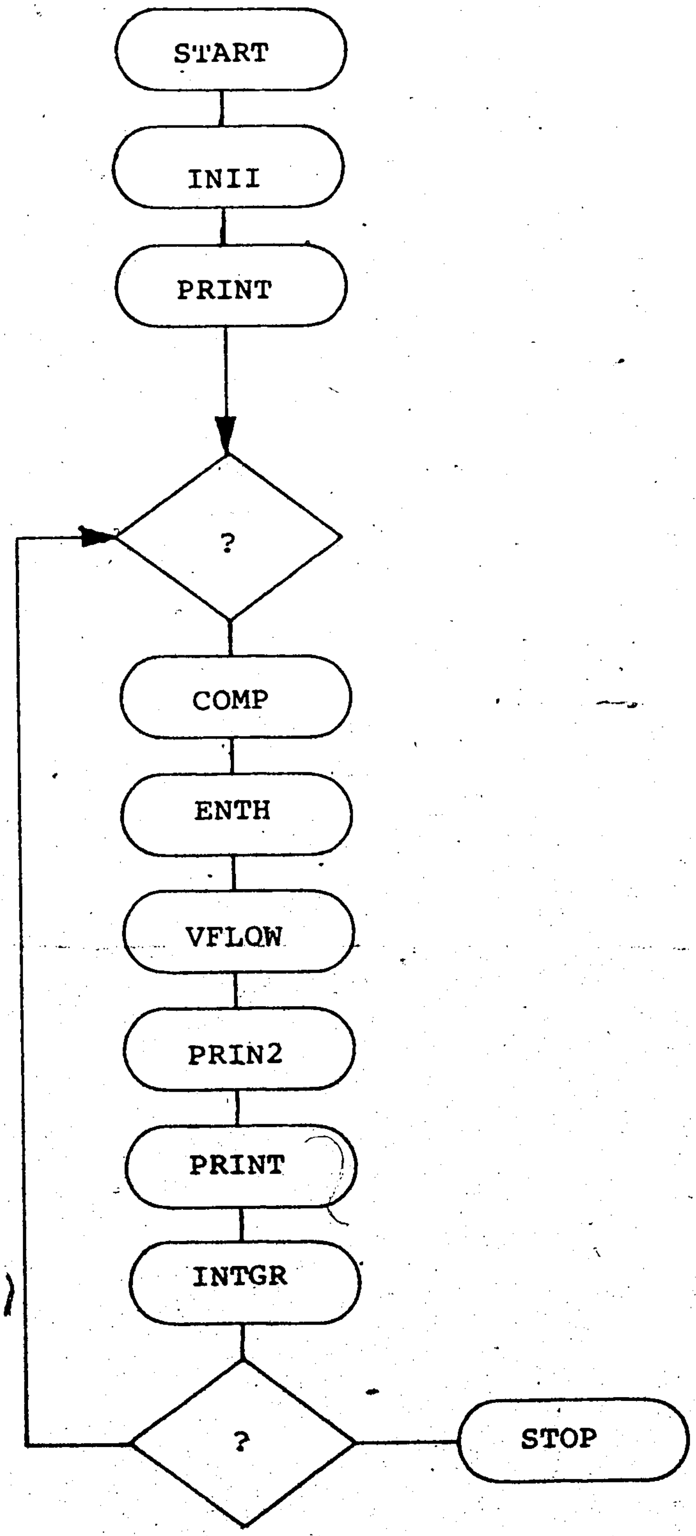

Figure B.1 Computer flow diagram of nonlinear model.

```
C THIS PROGRAM USES MASS,ENERGY AND COMPONENT BALANCES TO
C SOLVE FOR COMPOSITION TRANSIENT RESPONSES FOR A 10 STAGE
C DISTILLITION COLUMN.THE ASSUMPTIONS ARE:
C    NEGLIGBLE VAPOUR HOLDUP
C    PERFECT MIXING
C    CONSTANT LIQUID HOLDUP
C    VARIABLE EFFICIENCIES AND HEAT LOSS PER TRAY BASED ON TOP COMPOSTION
C
      REAL LT(10),DIST(3),CHAN(2),CHAN1(4),CHAN2(2)
      INTEGER ISTRN(3),ICHNG(3),IACT(2)
      COMMON T,MTT,XT(10),YT(10),LT,EF(10),VT(10),WT(10),HLT(10)
     1,HVT(10),DWT(10),TL(10),QSP(10),DXT(10),HLTO(10),DHL(10)
      COMMON E(10),EE(10),QLP(10),QRP(10),XF(3),FE(3),ST(3),RE(3)
      COMMON XS(26),YS(26),TLS(26),XST(19),YST(19),HXS(19),HYS(19)
      COMMON HSI,UA,TR,DT,LF,HF,HR,QR,KK,XIN,XIF,IDT,TS,IEQ
      COMMON B1(26),C1(26),D1(26),B2(19),C2(19),D2(19),B3(19),C3(19)
     1,D3(19)
      DATA ISTRN/'RE','FE','ST'/
      READ(5,100) ITEST
      READ(5,101) ICHNG
      READ(5,103) DIST
      READ(5,101) ITYPE,ICON,LF,MTT,NI
      READ(5,102) XIN,XIF
      READ(5,103) FEI,REI,STI
      READ(5,104) HF,HR,HSI,UA
      READ(5,103) XFI,DT
      READ(5,101) IEQ,JM,JN
      JM=JM/DT
      JN=JN/DT
      IN=0
      II=0
      NI=60*NI/DT
      DO 10 ID=1,3
      IF(ISTRN(ID)-ITEST)10,1,10
   10 CONTINUE
    1 GO TO (2,3,4),ID
    2 DO 20 I=1,3
      ICHNG(I)=60*ICHNG(I)/DT
      XF(I)=XFI/100.0
      FE(I)=FEI
      ST(I)=STI
      RE(I)=DIST(I)
   20 CONTINUE
      GO TO 5
    3 DO 30 I=1,3
      ICHNG(I)=60*ICHNG(I)/DT
      XF(I)=XFI/100.0
      ST(I)=STI
      RE(I)=REI
      FE(I)=DIST(I)
   30 CONTINUE
      GO TO 5
    4 DO 40 I=1,3
      ICHNG(I)=ICHNG(I)*60/DT
      RE(I)=REI
      FE(I)=FEI
      XF(I)=XFI/100.0
      ST(I)=DIST(I)
   40 CONTINUE
    5 CONTINUE
```

```
      IDT=1
      CALL INII
      T=0.0
      CALL PRINT
      DO 80 KK=1,NI
      T=KK*DT/60
      DO 50 JJ=1,3
      IF(KK-ICHNG(JJ))50,6,50
    6 IDT=JJ
   50 CONTINUE
      GO TO (16,12), ITYPE
C
C       ITYPE=TYPE OF SIMULATION IE.,
C           1 -- OPEN LOOP
C           2 -- CLOSE LOOP
C
   12 CONTINUE
      CALL PI(XT,DT,KK,NI,ICHNG,ICON,IACT,CHAN1,XB)
      CHAN(1)=CHAN1(1)
      CHAN(2)=CHAN1(2)
      GO TO (15,15,15,13), ICON
C
C       ICON=TYPE OF CONTROL IE.,
C           1 -- BOTTOM CONTROL ONLY
C           2 -- TOP CONTROL ONLY
C           3 -- BOTTOM AND TOP CONTROL WITHOUT COMPENSATION
C           4 -- BOTTOM AND TOP CONTROL WITH COMPENSATION
C
   13 CONTINUE
      IF(KK-ICHNG(2)) 15,14,14
   14 CALL NICMP(CHAN1,KK,ICHNG,IACT,CHAN2)
      CHAN(1)=CHAN2(1)
      CHAN(2)=CHAN2(2)
   15 CONTINUE
      ST(IDT)=ST(IDT)+CHAN(1)
      RE(IDT)=RE(IDT)+CHAN(2)
      WRITE(14,201) KK,RE(IDT),ST(IDT),CHAN(1),CHAN(2)
  201 FORMAT(1X,I4,2(2X,F6.3),2(2X,F12.6))
   16 CONTINUE
      CALL COMP
      DO 60 N=1,MTT
      HLTO(N)=HLT(N)
   60 CONTINUE
      CALL ENTH
      DO 70 N=1,MTT
      DHL(N)=(HLT(N)-HLTO(N))/DT
   70 CONTINUE
      CALL VFLOW
      II=II+1
      IN=IN+1
      IF(II-JM)7,8,7
    8 II=0
      CALL PRIN2(JM,KK,MTT,XT,LT,VT,ST(IDT),RE(IDT),ITYPE,XB)
    7 IF(IN-JN)9,11,9
   11 IN=0
      CALL PRINT
    9 CONTINUE
      CALL INTGR
   80 CONTINUE
  100 FORMAT(A2)
```

```
  101 FORMAT(5I5)
  102 FORMAT(2(1X,F8.6))
  103 FORMAT(3(1X,F6.3))
  104 FORMAT(4(1X,F8.3))
      STOP
      END
C
      SUBROUTINE INII
C
      REAL LT(10)
      COMMON T,MTT,XT(10),YT(10),LT,EF(10),VT(10),WT(10),HLT(10)
     1,HVT(10),DWT(10),TL(10),QSP(10),DXT(10),HLTO(10),DHL(10)
      COMMON E(10),EE(10),QLP(10),QRP(10),XF(3),FE(3),ST(3),RE(3)
      COMMON XS(26),YS(26),TLS(26),XST(19),YST(19),HXS(19),HYS(19)
      COMMON HSI,UA,TR,DT,LF,HF,HR,QR,KK,XIN,XIF,IDT,TS,IEQ
      COMMON B1(26),C1(26),D1(26),B2(19),C2(19),D2(19),B3(19),C3(19)
     1,D3(19)
      READ(5,100)(XT(N),N=1,MTT)
      READ(5,101)(LT(N),N=1,MTT)
      READ(5,101)(VT(N),N=1,MTT)
      READ(5,102)(WT(N),N=1,MTT)
      READ(5,103)(QLP(N),N=1,MTT)
      READ(5,103)(QRP(N),N=1,MTT)
      READ(5,104)(E(N),N=1,MTT)
      READ(5,104)(EE(N),N=1,MTT)
      READ(5,104)(XS(N),N=1,26)
      READ(5,104)(YS(N),N=1,26)
      READ(5,105)(TLS(N),N=1,26)
      WRITE(13,200)
  200 FORMAT(//,11X,'********EQUILIBRIUM DATA********')
      WRITE(13,201)
  201 FORMAT(11X,'LCOMP',8X,'VCOMP',8X,'LTEMP',//)
      DO 10 N=1,26
      WRITE(13,202)XS(N),YS(N),TLS(N)
   10 CONTINUE
  202 FORMAT(11X,F6.4,7X,F6.4,7X,F6.2)
      CALL SPLINE(26,XS,YS,B1,C1,D1)
      WRITE(13,203)
  203 FORMAT(//,2X,'INTERVAL',10X,'**YS**',12X,'**B1**'
     1,15X,'**C1**',15X,'**D1**',//)
      DO 20 I=1,26
      WRITE(13,204) I,YS(I),B1(I),C1(I),D1(I)
   20 CONTINUE
  204 FORMAT(5X,I2,13X,F6.4,9X,E12.5,9X,E12.5,9X,E12.5)
      CALL COMP
      READ(5,104)(XST(N),N=1,19)
      READ(5,105)(HXS(N),N=1,19)
      DO 30 I=1,19
      YST(I)=SEVAL(26,XST(I),XS,YS,B1,C1,D1)
      HYS(I)=2657.4-1424.8*XST(I)
   30 CONTINUE
      WRITE(13,205)
  205 FORMAT(//,10X,'****ENTHALPY COMPOSITION DATA****')
      WRITE(13,206)
  206 FORMAT(10X,'LCOMP',4X,'LENTH',4X,'VCOMP',4X,'VENTH',//)
      DO 40 N=1,19
      WRITE(13,207) XST(N),HXS(N),YST(N),HYS(N)
   40 CONTINUE
  207 FORMAT(10X,F6.4,3X,F6.2,3X,F6.4,3X,F6.1)
      EF(MTT)=0.0
```

```
      YT(MTT)=0.0
      HVT(MTT)=0.0
      CALL SPLINE(19,XST,HXS,B2,C2,D2)
      WRITE(13,208)
  208 FORMAT(//,2X,'INTERVAL',10X,'**HXS**',12X,'**B2**'
     1,15X,'**C2**',15X,'**D2**',//)
      DO 50 I=1,19
      WRITE(13,209) I,HXS(I),B2(I),C2(I),D2(I)
   50 CONTINUE
  209 FORMAT(5X,I2,14X,F6.2,9X,E12.5,9X,E12.5,9X,E12.5)
      CALL SPLINE(19,YST,HYS,B3,C3,D3)
      WRITE(13,210)
  210 FORMAT(//,2X,'INTERVAL',10X,'**HYS**',12X,'**B3**'
     1,15X,'**C3**',15X,'**D3**',//)
      DO 60 I=1,19
      WRITE(13,211) I,HYS(I),B3(I),C3(I),D3(I)
   60 CONTINUE
  211 FORMAT(5X,I2,14X,F6.1,9X,E12.5,9X,E12.5,9X,E12.5)
      CALL ENTH
      DO 70 N=1,MTT
      HLTO(N)=HLT(N)
      QSP(N)=QLP(N)
      DHL(N)=(HLT(N)-HLTO(N))/DT
   70 CONTINUE
      AHLO=0.0
      BHLO=4.18
      TR=TL(1)
      TS=(ST(IDT)*(HSI-AHLO)+UA*TR)/(UA+BHLO*ST(IDT))
      QR=UA*(TS-TR)
      NTT=MTT-1
      NL=MTT-2
      DO 80 N=2,NL
      DWT(N)=LT(N+1)+VT(N-1)-LT(N)-VT(N)
   80 CONTINUE
      DWT(1)=LT(2)-LT(1)-VT(1)
      DWT(LF)=DWT(LF)+FE(IDT)
      DWT(NTT)=RE(IDT)+VT(NL)-LT(NTT)-VT(NTT)
      DWT(MTT)=VT(NTT)-LT(MTT)-RE(IDT)
  100 FORMAT(10(1X,F8.6))
  101 FORMAT(10(1X,F8.5))
  102 FORMAT(10(1X,F7.2))
  103 FORMAT(10(1X,F6.1))
  104 FORMAT(10(1X,F6.4))
  105 FORMAT(10(1X,F6.2))
      RETURN
      END
C
      SUBROUTINE PI(XT,DT,KK,NI,ICHNG,ICON,IACT,CHAN1,XB)
C
      REAL ER(2),RES(2),PK(2),TI(2),DELTA(2),CHAN1(4),XT(10),PT(2)
     $,SIAE(2),XNT(128),ERR(2),DER(2)
      INTEGER KK,ICHNG(3),ICON,KCNT(2),ITIM(2),IACT(2)
      XB=XT(1)
      IF(KK-ICHNG(2)) 14,1,2
    1 READ(4,101) XBSP,XDSP
  101 FORMAT(2(1X,F8.6))
      READ(4,102)IMODE,PK(1),PK(2),TI(1),TI(2),DER(1),DER(2),PT(1),PT(2)
  102 FORMAT(I3,2(1X,F6.2),2(1X,F8.2),2(1X,F7.2),2(1X,F5.1))
      READ(4,103) ITIM(1),ITIM(2)
  103 FORMAT(2(1X,I3))
```

```
      DO 10 I=1,2
      SIAE(I)=0.0
      RES(I)=0.0
      ER(I)=0.0
      IACT(I)=0
      KCNT(I)=0
      ITIM(I)=ITIM(I)/DT
      CHAN1(I+2)=0.0
   10 CONTINUE
      IDEL=(ITIM(1)*2)
      NUM=IDEL+1
      DO 20 I=1,NUM
      XNT(I)=XT(1)
   20 CONTINUE
      GO TO (18,19,21), IMODE
C
C       IMODE=CONTROL MODE
C          1 -- PROPORTIONAL
C          2 -- PROPORTIONAL + INTEGRAL
C          3 -- PROPORTIONAL + INTEGRAL + DERIVATIVE
C
   18 CONTINUE
      TI(1)=0.0
      TI(2)=0.0
      DER(1)=0.0
      DER(2)=0.0
      GO TO 22
   19 CONTINUE
      TI(1)=PK(1)/TI(1)
      TI(2)=PK(2)/TI(2)
      DER(1)=0.0
      DER(2)=0.0
      GO TO 22
   21 CONTINUE
      TI(1)=PK(1)/TI(1)
      TI(2)=PK(2)/TI(2)
   22 CONTINUE
      GO TO 14
C
C       DETERMINE WHEN TO TAKE CONTROL ACTION FOR BOTTOM
C       (IE. EVERY 128 SECONDS) AND TOP(IE. EVERY 32 OR 64
C       SECONDS).  THE ONLY RESTRICTION IS THAT THE INTEGRATION
C       INTERVAL BE EQUAL OR SMALLER THAN THE SMALLEST CONTROL
C       ACTION.  THE INTEGRATION INTERVAL MUST BE A MULTIPLE OF
C       2 NOT GREATER THAN 64 OR LESS THAN 2.
C
    2 CONTINUE
      DO 30 I=1,IDEL
      XNT(I)=XNT(I+1)
   30 CONTINUE
      XB=XNT(1)
      XNT(NUM)=XT(1)
      KCNT(1)=KCNT(1)+1
      KCNT(2)=KCNT(2)+1
      IACT(1)=KCNT(1)-ITIM(1)
      IF(IACT(1)) 4,3,3
    3 IACT(2)=KCNT(2)-ITIM(2)
      KCNT(1)=0
      KCNT(2)=0
      GO TO 7
```

```
    4 DELTA(1)=0.0
      IACT(2)=KCNT(2)-ITIM(2)
      IF(IACT(2)) 5,6,6
    5 DELTA(2)=0.0
      GO TO 9
    6 KCNT(2)=0
      GO TO 8
    7 CONTINUE
C
C       BOTTOM CONTROL PART
C
      ERR(1)=ER(1)
      ER(1)=XBSP-XNT(1)
      RES(1)=RES(1)+ER(1)*PT(1)
      SIAE(1)=SIAE(1)+ABS(ER(1)*PT(1))
      DELTA(1)=PK(1)*ER(1) + TI(1)*RES(1) + (PK(1)*DER(1)/PT(1))
     $*(ER(1)-ERR(1))
    8 CONTINUE
C
C       TOP CONTROL PART
C
      ERR(2)=ER(2)
      ER(2)=XDSP-XT(10)
      RES(2)=RES(2)+ER(2)*PT(2)
      SIAE(2)=SIAE(2)+ABS(ER(2)*PT(2))
      DELTA(2)=PK(2)*ER(2) + TI(2)*RES(2) + (PK(2)*DER(2)/PT(2))
     $*(ER(2)-ERR(2))
    9 CONTINUE
      GO TO (11,12,13,13), ICON
   11 CONTINUE
C
C       BOTTOM CONTROL ONLY
C
      CHAN1(1)=-DELTA(1)
      CHAN1(2)=0.0
      GO TO 15
   12 CONTINUE
C
C       TOP CONTROL ONLY
C
      CHAN1(1)=0.0
      CHAN1(2)=DELTA(2)
      GO TO 15
   13 CONTINUE
C
C       BOTTOM AND TOP CONTROL WITH OR WITHOUT COMPENSATION
C
      CHAN1(1)=-DELTA(1)
      CHAN1(2)=DELTA(2)
      GO TO 15
   14 CONTINUE
      CHAN1(1)=0.0
      CHAN1(2)=0.0
   15 CONTINUE
      IF(KK-NI) 16,17,17
   17 WRITE(6,201) ICON, SIAE(1),SIAE(2)
  201 FORMAT('TYPE OF CONTROL= ',I2,' IAE(1)=  ',F8.3,' IAE(2)=  ',F8.3)
      WRITE(6,202) PK(1),TI(1),DER(1),PK(2),TI(2),DER(2)
  202 FORMAT(/,'BOTTOM CONTROL LOOP      PK(1)=  ',F6.2,'  TI(1)=  ',F7.5
     $,'  DER(1)=  ',F7.2
```

```
     $./'TOP CONTROL LOOP        PK(2)=  ',F6.2,'  TI(2)=  ',F7.5
     $,'  DER(2)=  ',F7.2)
   16 CONTINUE
      RETURN
      END
C
      SUBROUTINE NICMP(CHAN1,KK,ICHNG,IACT,CHAN2)
C
      REAL CHAN1(4),CHAN2(2),COMP(2),DALE(4),CHIP(2)
      INTEGER KK,IACT(2),ICHNG(3),ICOMP
      IF(KK-ICHNG(2)) 1,1,2
    1 READ(4,101) ICOMP,COMP(1),COMP(2)
  101 FORMAT(I3,2(1X,F6.3))
      DO 10 I=1,2
      CHAN2(I)=0.0
      DALE(I)=0.0
      CHIP(I)=0.0
   10 CONTINUE
      DALE(3)=0.0
      DALE(4)=0.0
      GO TO 8
    2 CONTINUE
      IF(IACT(1)) 4,3,3
    3 CONTINUE
      GO TO 6
    4 CONTINUE
      CHAN2(1)=0.0
      IF(IACT(2)) 5,7,7
    5 CHAN2(2)=0.0
      GO TO 8
    6 CONTINUE
C
C      COMPENSATION PART IS ADDED TO THE MANIPULATIVE VARIABLES
C
C      ICOMP=TYPE OF CONPENSATOR IE.
C          1 -- STATIC
C          2 -- STATIC PLUS TIME DELAY
C          3 -- DECOUPLER
C
      GO TO (9,11,12), ICOMP
    9 CONTINUE
      CHAN2(1)=CHAN1(1)+COMP(1)*CHAN1(2)
      CHAN2(2)=CHAN1(2)+COMP(2)*CHAN1(1)
      GO TO 8
   11 CONTINUE
      DO 20 I=1,2
      DALE(I)=DALE(I+1)
   20 CONTINUE
      DALE(3)=CHAN1(2)
      CHAN2(1)=CHAN1(1)+COMP(1)*DALE(1)
      CHAN2(2)=CHAN1(2)+COMP(2)*CHAN1(1)
      GO TO 8
   12 CONTINUE
      DO 30 I=1,3
      DALE(I)=DALE(I+1)
   30 CONTINUE
      DALE(4)=CHAN1(2)
      CHIP(1)=CHIP(2)
      CHIP(2)=CHAN1(1)
      CHAN2(1)=CHAN1(1)+0.8919*(CHAN1(3)-CHIP(1))
```

```
     *+0.9268*COMP(1)*DALE(2)-0.8187*COMP(1)*DALE(1)
      CHAN1(3)=CHAN2(1)
      CHAN2(2)=CHAN1(2)+0.9282*(CHAN1(4)-DALE(3))
     *+0.4300*COMP(2)*CHAN1(1)-0.3582*COMP(2)*CHIP(1)
      CHAN1(4)=CHAN2(2)
      GO TO 8
    7 CONTINUE
      GO TO (13,13,14), ICOMP
   13 CONTINUE
      CHAN2(2)=CHAN1(2)+COMP(2)*CHAN1(1)
      GO TO 8
   14 CONTINUE
      DO 40 I=1,3
      DALE(I)=DALE(I+1)
   40 CONTINUE
      DALE(4)=CHAN1(2)
      CHAN2(2)=CHAN1(2)+0.9282*(CHAN1(4)-DALE(3))
      CHAN1(4)=CHAN2(2)
    8 CONTINUE
      WRITE(14,201) KK,IACT(1),IACT(2),CHAN1(1),CHAN1(2),CHAN1(3)
     $,CHAN1(4),CHAN2(1),CHAN2(2)
  201 FORMAT(1X,I4,2(1X,I3),6(1X,F12.6))
      RETURN
      END
C
      SUBROUTINE COMP
C
      REAL LT(10)
      DIMENSION YSE(10)
      COMMON T,MTT,XT(10),YT(10),LT,EF(10),VT(10),WT(10),HLT(10)
     1,HVT(10),DWT(10),TL(10),QSP(10),DXT(10),HLTO(10),DHL(10)
      COMMON E(10),EE(10),QLP(10),QRP(10),XF(3),FE(3),ST(3),RE(3)
      COMMON XS(26),YS(26),TLS(26),XST(19),YST(19),HXS(19),HYS(19)
      COMMON HSI,UA,TR,DT,LF,HF,HR,QR,KK,XIN,XIF,IDT,TS,IEQ
      COMMON B1(26),C1(26),D1(26),B2(19),C2(19),D2(19),B3(19),C3(19)
     1,D3(19)
      NL=MTT-2
      DO 20 N=1,MTT
      DO 10 J=1,26
      IF(XS(J)-XT(N))10,10,1
   10 CONTINUE
    1 K=J-1
      L=J
      TL(N)=((XT(N)-XS(K))*((TLS(L)-TLS(K))/(XS(L)-XS(K))))+TLS(K)
   20 CONTINUE
      EF(1)=E(1)
      YSE(1)=SEVAL(26,XT(1),XS,YS,B1,C1,D1)
      YT(1)=EF(1)*YSE(1)
      DO 30 N=1,NL
      J=N+1
C
C       IF 'IEQ' EQUALS
C           1 -- CONSTANT EFFICIENCY PER STAG
C           2 -- VARIABLE EFFICIENCY BASED ON TOP COMPOSITION
C
      GO TO (2,3), IEQ
    2 EF(J)=E(J)
      GO TO 4
    3 EF(J)=E(J)+(EE(J)-E(J))*(XT(MTT)-XIN)/(XIF-XIN)
    4 CONTINUE
```

```
      YSE(J)=SEVAL(26,XT(J),XS,YS,B1,C1,D1)
      YT(J)=EF(J)*(YSE(J)-YT(N))+YT(N)
   30 CONTINUE
      RETURN
      END
C
      SUBROUTINE ENTH
C
      REAL LT(10)
      COMMON T,MTT,XT(10),YT(10),LT,EF(10),VT(10),WT(10),HLT(10)
     1,HVT(10),DWT(10),TL(10),QSP(10),DXT(10),HLTO(10),DHL(10)
      COMMON E(10),EE(10),QLP(10),QRP(10),XF(3),FE(3),ST(3),RE(3)
      COMMON XS(26),YS(26),TLS(26),XST(19),YST(19),HXS(19),HYS(19)
      COMMON HSI,UA,TR,DT,LF,HF,HR,QR,KK,XIN,XIF,IDT,TS,IEQ
      COMMON B1(26),C1(26),D1(26),B2(19),C2(19),D2(19),B3(19),C3(19)
     1,D3(19)
      NTT=MTT-1
      DO 10 N=1,MTT
      HLT(N)=SEVAL(19,XT(N),XST,HXS,B2,C2,D2)
   10 CONTINUE
      DO 20 N=1,NTT
      HVT(N)=SEVAL(19,YT(N),YST,HYS,B3,C3,D3)
   20 CONTINUE
      RETURN
      END
C
      SUBROUTINE INTGR
C
      REAL LT(10)
      DIMENSION FY(10),TSU(10,10),TA(10,10)
      DIMENSION TT(10,10),FX(10),FT(10)
      COMMON T,MTT,XT(10),YT(10),LT,EF(10),VT(10),WT(10),HLT(10)
     1,HVT(10),DWT(10),TL(10),QSP(10),DXT(10),HLTO(10),DHL(10)
      COMMON E(10),EE(10),QLP(10),QRP(10),XF(3),FE(3),ST(3),RE(3)
      COMMON XS(26),YS(26),TLS(26),XST(19),YST(19),HXS(19),HYS(19)
      COMMON HSI,UA,TR,DT,LF,HF,HR,QR,KK,XIN,XIF,IDT,TS,IEQ
      COMMON B1(26),C1(26),D1(26),B2(19),C2(19),D2(19),B3(19),C3(19)
     1,D3(19)
      NTT=MTT-1
      NL=MTT-2
      DO 20 N=1,MTT
      DO 10 M=1,MTT
      TT(N,M)=0.0
      TSU(N,M)=0.0
   10 TA(N,M)=0.0
   20 CONTINUE
      DO 30 N=1,MTT
      FT(N)=YT(N)/XT(N)
   30 CONTINUE
      RR=XF(IDT)/XT(LF)
      DO 40 N=2,NL
      M=N-1
      J=N+1
      TT(N,N)=(VT(N)*(1.0-FT(N))-LT(J)-VT(M))*DT/(WT(N)*2.0)
      TT(N,M)=VT(M)*FT(M)*DT/(WT(N)*2.0)
      TT(N,J)=LT(J)*DT/(WT(N)*2.0)
   40 CONTINUE
      TT(1,1)=(VT(1)*(1.0-FT(1))-LT(2))*DT/(WT(1)*2.0)
      TT(1,2)=LT(2)*DT/(WT(1)*2.0)
      TT(LF,LF)=TT(LF,LF)+(RR*FE(IDT)-FE(IDT))*DT/(WT(LF)*2.0)
```

```
      TT(NTT,NTT)=(VT(NTT)*(1.0-FT(NTT))-RE(IDT)-VT(NL))*DT/(WT(NTT)
     1*2.0)
      TT(NTT,NL)=VT(NL)*FT(NL)*DT/(WT(NTT)*2.0)
      TT(NTT,MTT)=RE(IDT)*DT/(WT(NTT)*2.0)
      TT(MTT,NTT)=VT(NTT)*FT(NTT)*DT/(WT(MTT)*2.0)
      TT(MTT,MTT)=-VT(NTT)*DT/(WT(MTT)*2.0)
      CALL GMPRD(TT,XT,DXT,MTT,MTT,1)
      DO 50 N=1,MTT
      TA(N,N)=1.0+TT(N,N)
      TSU(N,N)=1.0-TT(N,N)
   50 CONTINUE
      DO 60 N=1,NTT
      TA(N,N+1)=TT(N,N+1)
      TA(N+1,N)=TT(N+1,N)
      TSU(N,N+1)=-TT(N,N+1)
      TSU(N+1,N)=-TT(N+1,N)
   60 CONTINUE
      CALL MINV(TSU,MTT,D,FX,FT)
      CALL GMPRD(TSU,TA,TT,MTT,MTT,MTT)
      CALL GMPRD(TT,XT,FT,MTT,MTT,1)
      DO 70 N=1,MTT
      XT(N)=FT(N)
   70 CONTINUE
      RETURN
      END
C
      SUBROUTINE PRIN2(JM,KK,MTT,XT,LT,VT,YM1,YM2,ITYPE,XB)
C
      REAL LT(10)
      DIMENSION XXT(10),XBT(10),XT(10),VT(10),YST(10),YRE(10)
      NTT=MTT-1
      IF(KK.GT.JM) GO TO 1
      DO 10 N=1,MTT
      XBT(N)=0.0
      XXT(N)=0.0
      YST(N)=0.0
      YRE(N)=0.0
   10 CONTINUE
    1 IF(KK-JM)2,2,3
    2 NN=0
    3 CONTINUE
      WRITE(7,200)(XT(N),N=1,MTT)
  200 FORMAT(10(1X,F8.6))
      WRITE(10,201)(LT(N),N=1,MTT)
  201 FORMAT(10(1X,F8.5))
      WRITE(10,202)(VT(N),N=1,MTT)
  202 FORMAT(10(1X,F8.5))
      DO 20 N=1,NTT
      XBT(N)=XBT(N+1)
      XXT(N)=XXT(N+1)
      YST(N)=YST(N+1)
      YRE(N)=YRE(N+1)
   20 CONTINUE
      GO TO (6,7),ITYPE
    6 XBT(10)=XT(1)*100.0
      GO TO 8
    7 XBT(10)=XB*100.0
    8 CONTINUE
      XXT(10)=XT(MTT)*100.0
      YST(10)=YM1
```

```
      YRE(10)=YM2
      NN=NN+1
      IF (NN-10)5,4,4
    4 WRITE(8,203)(XXT(N),N=1,10)
      WRITE(9,203)(XBT(N),N=1,10)
      WRITE(15,203)(YST(N),N=1,10)
      WRITE(16,203)(YRE(N),N=1,10)
      NN=0
    5 CONTINUE
  203 FORMAT(10F8.3)
      RETURN
      END
C
      SUBROUTINE PRINT
C
      REAL LT(10)
      COMMON T,MTT,XT(10),YT(10),LT,EF(10),VT(10),WT(10),HLT(10)
     1,HVT(10),DWT(10),TL(10),QSP(10),DXT(10),HLTO(10),DHL(10)
      COMMON E(10),EE(10),QLP(10),QRP(10),XF(3),FE(3),ST(3),RE(3)
      COMMON XS(26),YS(26),TLS(26),XST(19),YST(19),HXS(19),HYS(19)
      COMMON HSI,UA,TR,DT,LF,HF,HR,QR,KK,XIN,XIF,IDT,TS,IEQ
      COMMON B1(26),C1(26),D1(26),B2(19),C2(19),D2(19),B3(19),C3(19)
     1,D3(19)
      WRITE(12,200)T
  200 FORMAT(38X,'TIME=',F8.2,'  (MIN)')
      WRITE(12,201)
  201 FORMAT(/,2X,'STAGE',2X,'LCOMP',3X,'VCOMP';3X,'EFFNCY',3X,'LFLOW'
     1,3X,'VFLOW',2X,'HOLDUP',2X,'LENTH',3X,'VENTH',5X,'DWT',5X,'LTEMP'
     2,3X,'HLOSS',/)
      DO 10 N=1,MTT
      WRITE(12,202)N,XT(N),YT(N),EF(N),LT(N),VT(N),WT(N),HLT(N),HVT(N)
     1,DWT(N),TL(N),QSP(N)
   10 CONTINUE
  202 FORMAT(I5,3X,F7.4,1X;F7.4,1X,F7.4,F8.2,F8.2,F8.1,F8.2,F8.1,F9.4
     1,F8.2,F8.1)
      WRITE(12,203) FE(IDT)
  203 FORMAT(/,15X,'FEED FLOW=   ',F5.2,' G/SEC')
      WRITE(12,204) RE(IDT)
  204 FORMAT(/,15X,'REFLUX FLOW= ',F5.2,' G/SEC')
      WRITE(12,205) ST(IDT)
  205 FORMAT(/,15X,'STEAM FLOW=  ',F5.2,' G/SEC')
      WRITE(12,206) TS,QR
  206 FORMAT(/,15X,'STEAM TEMP= ',F5.1,' DEC-C',5X,'HEAT DUTY= ',
     1F7.1,' J/SEC',//)
      RETURN
      END
C
      SUBROUTINE VFLOW
C
      REAL LT(10)
      REAL TT(18,18),FT(18),WORK(18),COND,CONDP1
      INTEGER IPVT(18),MTT,NTT,NL,NNT
      COMMON T,MTT,XT(10),YT(10),LT,EF(10),VT(10),WT(10),HLT(10)
     1,HVT(10),DWT(10),TL(10),QSP(10),DXT(10),HLTO(10),DHL(10)
      COMMON E(10),EE(10),QLP(10),QRP(10),XF(3),FE(3),ST(3),RE(3)
      COMMON XS(26),YS(26),TLS(26),XST(19),YST(19),HXS(19),HYS(19)
      COMMON HSI,UA,TR,DT,LF,HF,HR,QR,KK,XIN,XIF,IDT,TS,IEQ
      COMMON B1(26),C1(26),D1(26),B2(19),C2(19),D2(19),B3(19),C3(19)
     1,D3(19)
      NTT=MTT-1
```

```
      NL=MTT-2
      NNT=NTT*2
      DO 20 J=1,NNT
      DO 10 K=1,NNT
   10 TT(K,J)=0.0
   20 CONTINUE
      AHLO=0.0
      BHLO=4.18
      TR=TL(1)
      TS=(ST(IDT)*(HSI-AHLO)+UA*TR)/(UA+BHLO*ST(IDT))
      QR=UA*(TS-TR)
      DO 30 N=1,NTT
      FT(N)=0.0
C
C       IF 'IEQ' EQUALS
C           1 -- CONSTANT HEAT LOSS PER STAG
C           2 -- VARIABLE HEAT LOSS BASED ON TOP COMPOSITION
C
      GO TO (1,2), IEQ
    1 QSP(N)=QLP(N)
      GO TO 3
    2 QSP(N)=QLP(N)+(QRP(N)-QLP(N))*(XT(MTT)-XIN)/(XIF-XIN)
    3 CONTINUE
   30 CONTINUE
      FT(LF)=-FE(IDT)
      FT(NTT)=-RE(IDT)
      DO 40 N=2,NTT
      NN=N+NTT
      FT(NN)=WT(N)*DHL(N)+QSP(N)
   40 CONTINUE
      FT(MTT)=WT(1)*DHL(1)+QSP(1)-QR
      N=NTT+LF
      FT(N)=FT(N)-FE(IDT)*(HF-HLT(LF))
      FT(NNT)=FT(NNT)-RE(IDT)*(HR-HLT(NTT))
      TT(1,1)=-1.0
      TT(1,2)=1.0
      TT(1,NTT+1)=-1.0
      DO 50 N=2,NTT
      TT(N,N+NTT)=-1.0
      TT(N,N)=-1.0
      TT(N-1,N)=1.0
      TT(N,NL+N)=1.0
   50 CONTINUE
      DO 60 N=2,NTT
      M=N+NL
      TT(M,N)=HLT(N)-HLT(N-1)
   60 CONTINUE
      DO 70 N=2,NTT
      M=N+NTT
      TT(M,M)=HLT(N)-HVT(N)
      TT(M,M-1)=HVT(N-1)-HLT(N)
   70 CONTINUE
      TT(MTT,MTT)=HLT(1)-HVT(1)
      CALL DECOMP(NNT,NNT,TT,COND,IPVT,WORK)
      WRITE(11,200) COND
  200 FORMAT(/,'CONDITION OF "TT" MATRIX IS ',E15.5)
      CONDP1=COND+1
      IF(CONDP1.EQ.COND) WRITE(6,201)
  201 FORMAT(1X,'MATRIX "TT" IS SINGULAR TO WORKING PRECISION')
      IF(CONDP1.EQ.COND) STOP
```

```
      CALL SOLVE(NNT,NNT,TT,FT,IPVT)
      DO 80 N=1,NTT
      LT(N)=FT(N)
      M=NTT+N
      VT(N)=FT(M)
   80 CONTINUE
      LT(MTT)=VT(NTT)-RE(IDT)
      DO 90 N=2,NTT
      DWT(N)=LT(N+1)+VT(N-1)-LT(N)-VT(N)
   90 CONTINUE
      DWT(1)=LT(2)-LT(1)-VT(1)
      DWT(LF)=DWT(LF)+FE(IDT)
      DWT(NTT)=RE(IDT)+VT(NL)-LT(NTT)-VT(NTT)
      DWT(MTT)=VT(NTT)-LT(MTT)-RE(IDT)
      RETURN
      END
```

(i) INTGR -intergrates the component material balance differential equation using a modified Euler method.

(j) finally the program either stop or continues depending on the run time.

APPENDIX C

C.1 Calibration Procedures

Before experimental data was collected, all the flow rates and analyzers were calibrated and the analog controllers were retuned. The calibration procedures followed were close to that of Pacey (41). The design of the sampling system used in this study and calibration of the gas chromatograph used in analyzing the bottom composition will be presented in Appendix D.

C.2 Flow Rates

The flows were measured using orifice plates connected to standard Foxboro differential pressure (d/p) cells. All liquid flow rates were calibrated using water at 15 °C and the following equation.

$$\text{Flow}/(sg) = k(\Delta P)^{\frac{1}{2}} \tag{C-1}$$

During the calibration procedure, it was assumed that the specific gravity (sg) of water was unity. The flow rate was measured using a scale and a stop watch, and the pressure drop ($\triangle$P) was measured in percent of the chart recorders. Before the flow rates were calibrated, the d/p cells were checked for air locks and then zeroed. The methanol-water flow rates were determined using equation C-1 and the density equation A-2 for methanol- water mixtures. The steam flow was calibrated by measuring the condensate flow at various chart readings. The condensate discharge was under level control and the line was surrounded by an ice bath to insure no vapour remained. Points for all calibrations were chosen to cover the operating range and then a least squares regression method was used to fit a straight line through the points. Figures C.1 through C.6 are the calibration curves for feed, reflux, bottom product, top product, steam, and cooling water flow rates respectively. Table C.1 summarizes the measurement device parameters or the type and size of orifice plate used for each flow calibration.

C.3 Top Composition

The procedure used to calibrate the Dynalog capacitance analyzer is outlined in the Foxboro manual for the instrument. In order to determine the relationship between capacitance and methanol concentration, samples at various chart readings were taken. The methanol concentration of

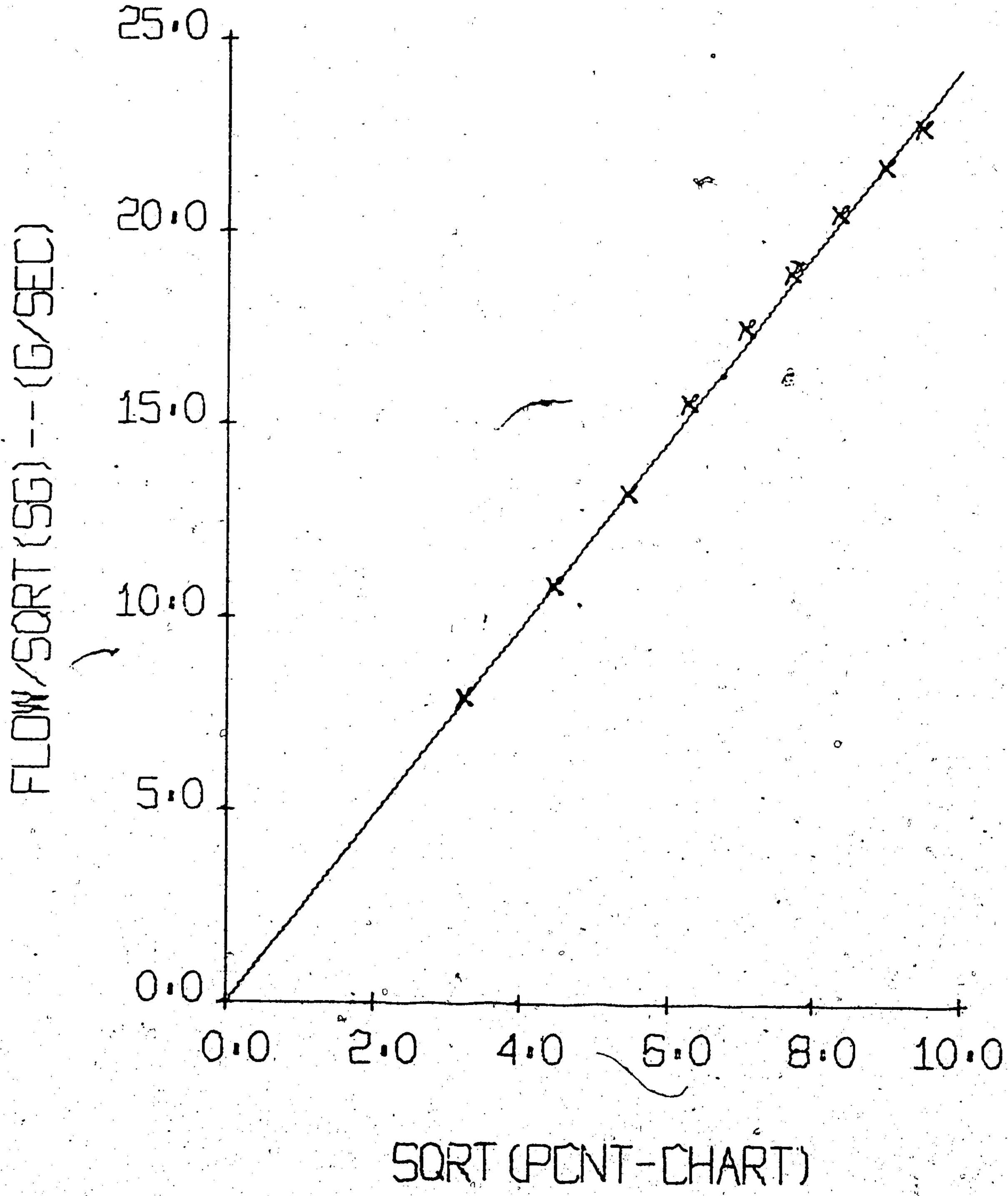

Figure C.1 Feed flow rate recorder calibration curve.

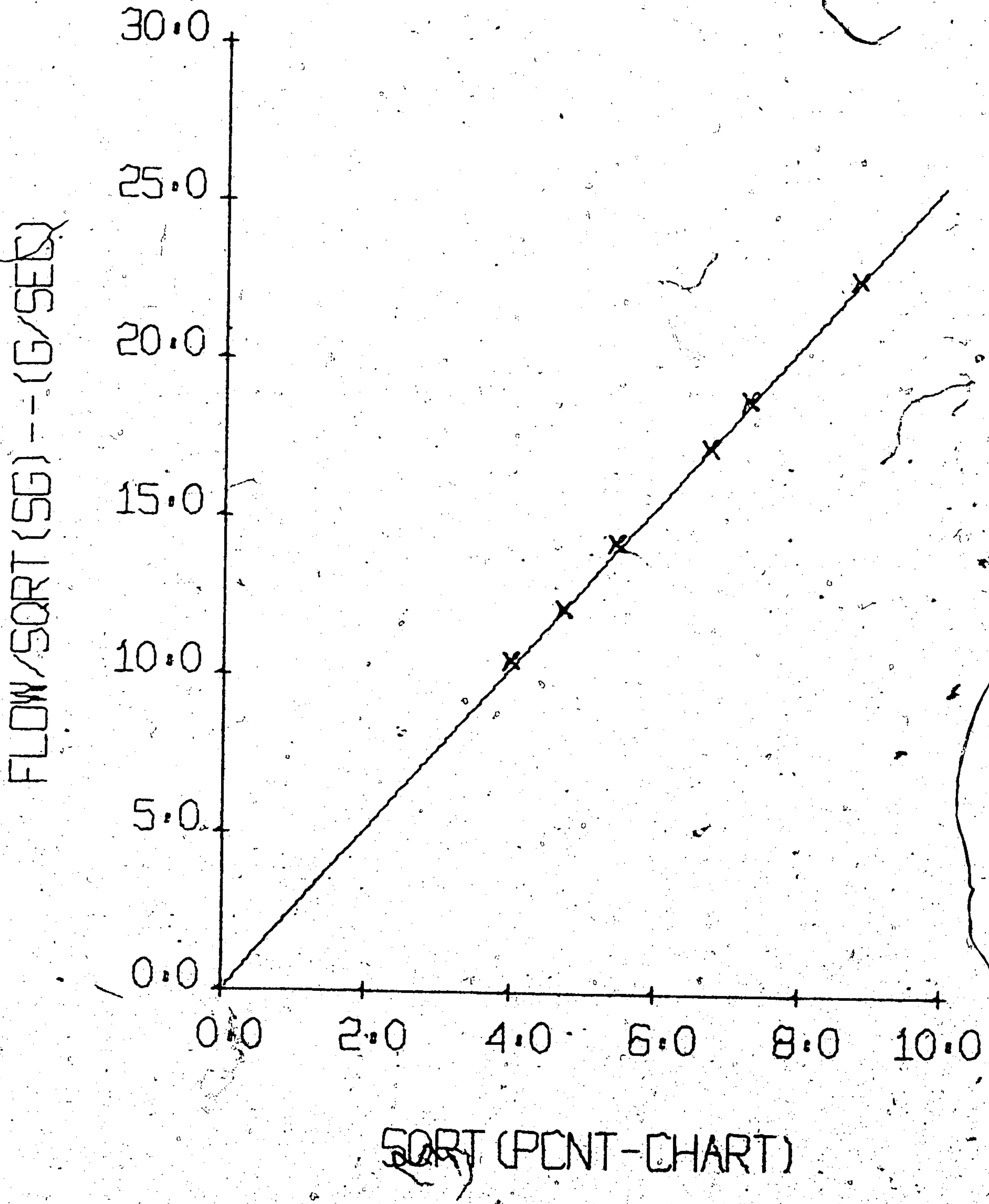

Figure C.3 Reflux flow rate recorder calibration curve.

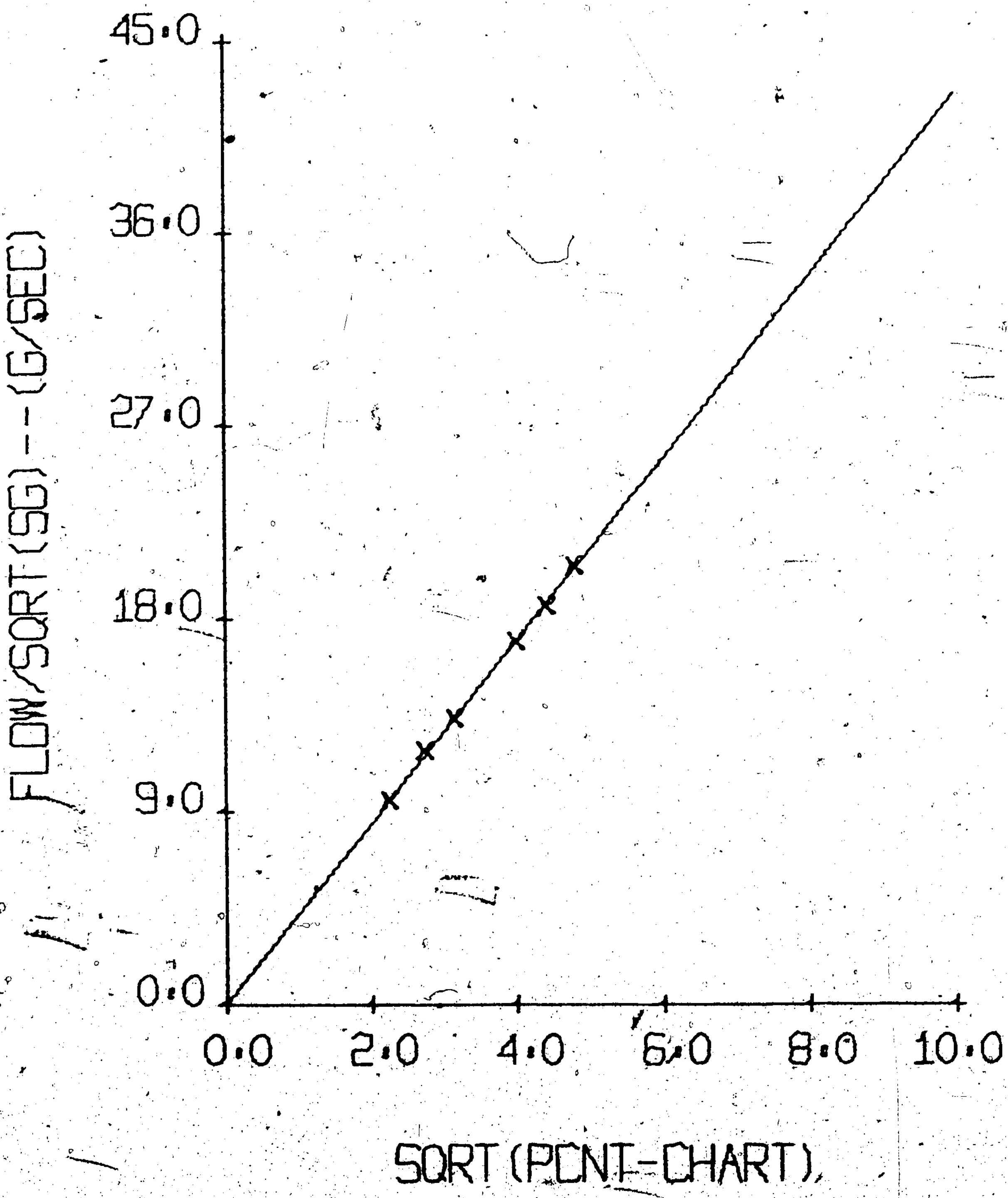

Figure C.3 Steam flow rate recorder calibration curve.

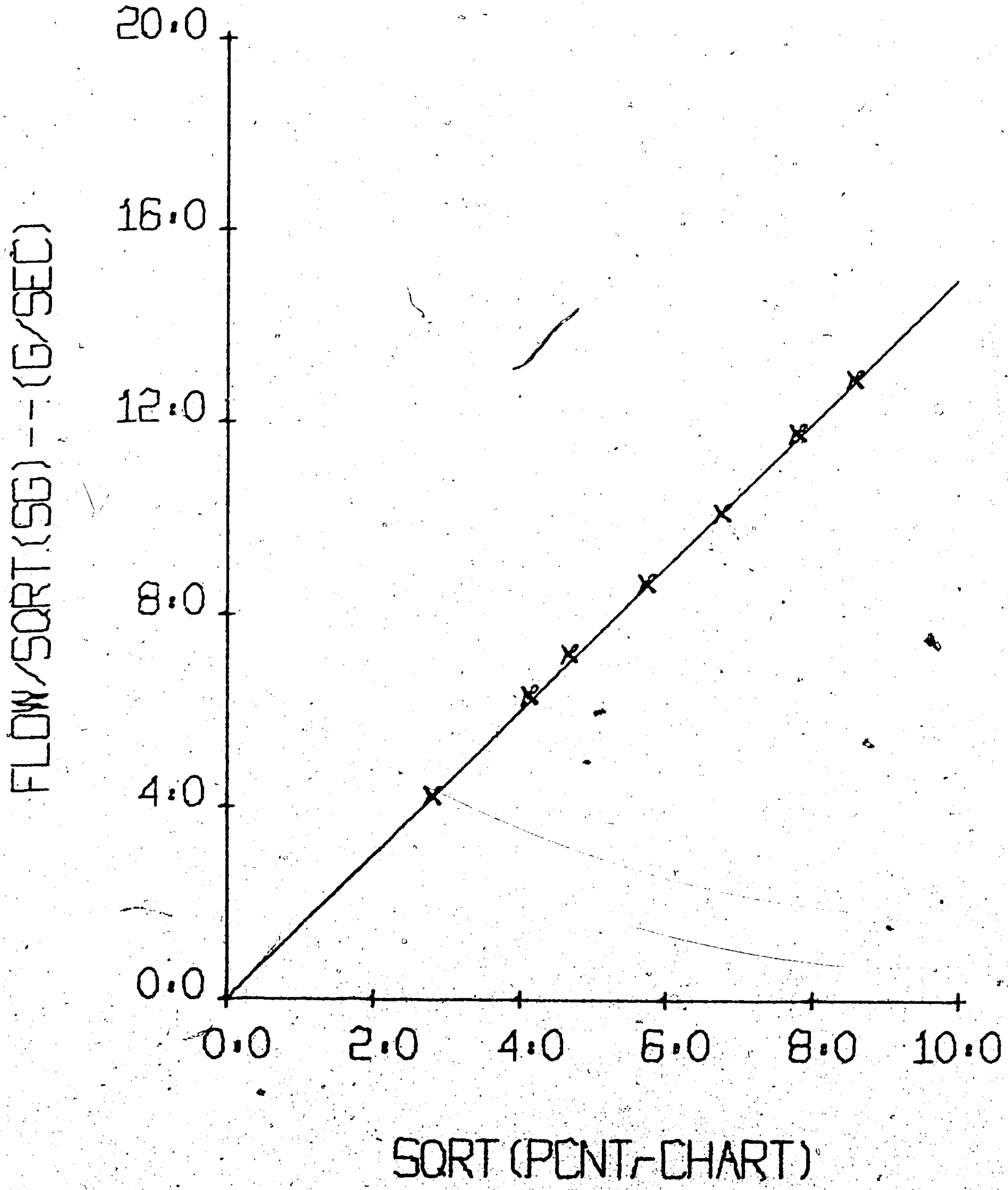

Figure C.4 Bottom flow rate recorder calibration curve.

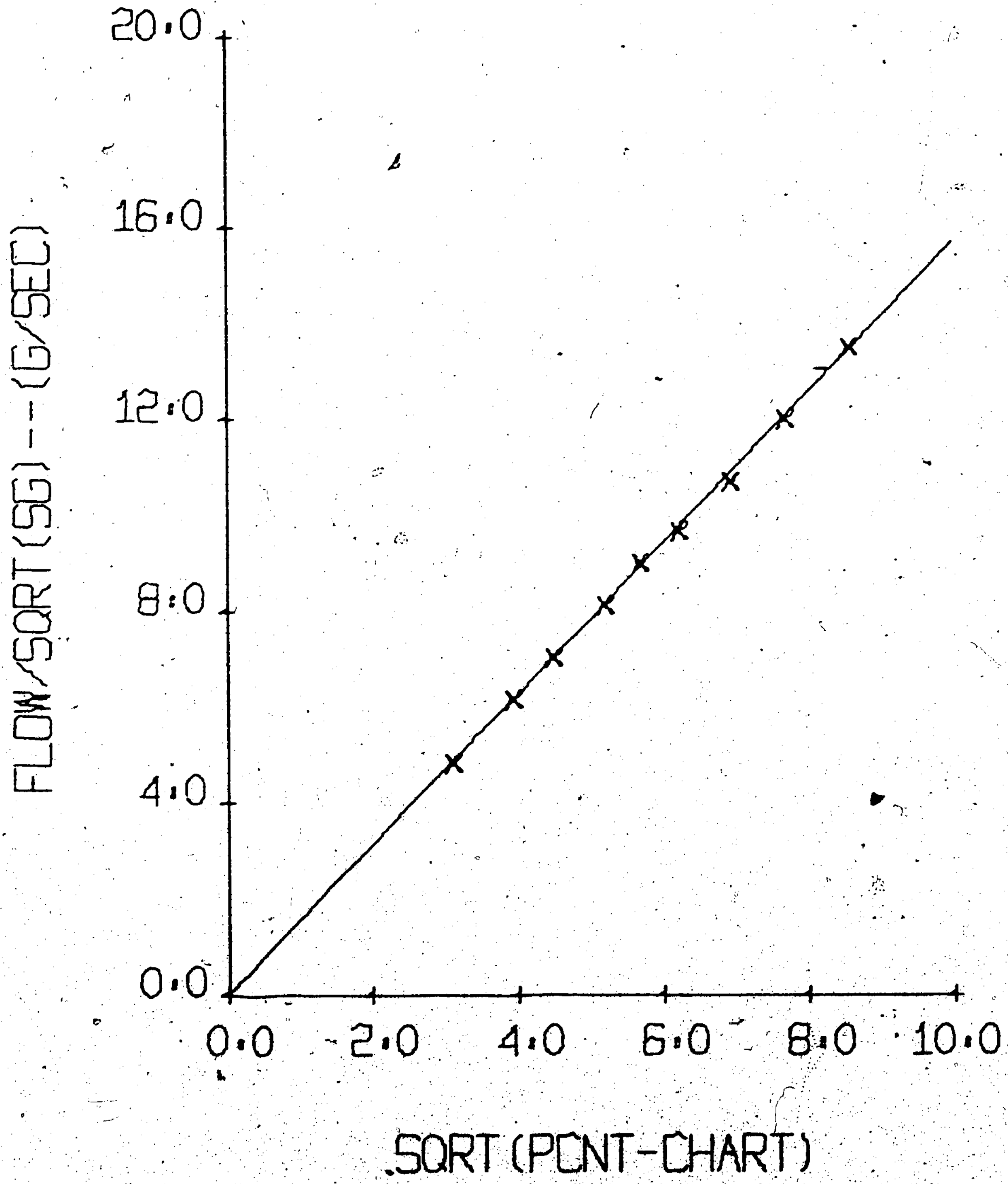

Figure C.5 Top flow rate recorder calibration curve.

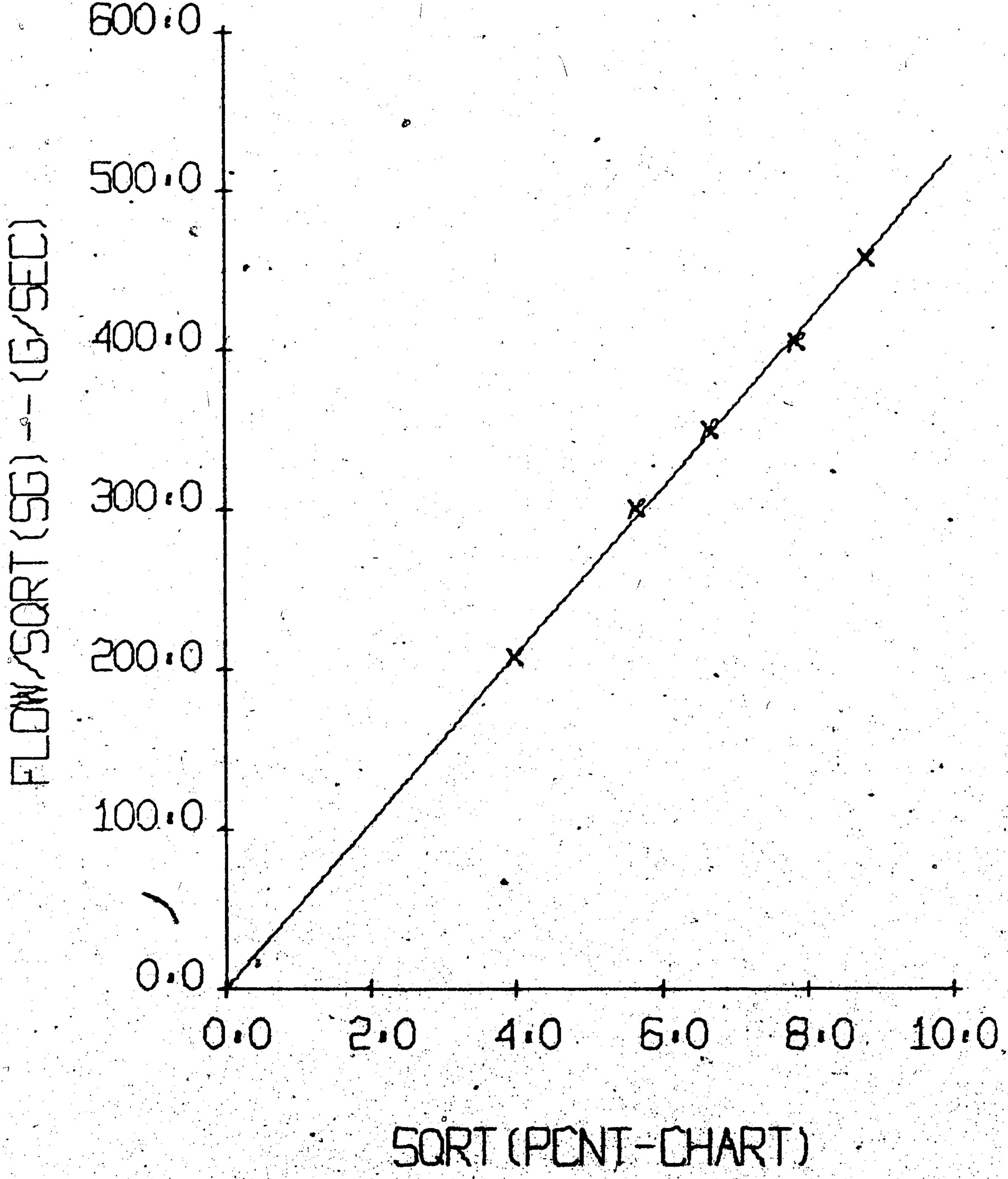

Figure C.6 Cooling water flow rate recorder calibration curve.

Table C.1 Summary of Measurement Device Parameters

Process Variable	Orifice Size (cm)	Orifice Type	A	B	Eq'n Type	Units
feed flow	0.254	quadrant	2.42	0.0	1	g/s
reflux flow	0.229	quadrant	2.57	0.0	1	g/s
steam flow	0.762	sharp	4.25	0.0	1	g/s
top flow	0.254	quadrant	1.57	0.0	1	g/s
bottom flow	0.254	quadrant	1.50	0.0	1	g/s
cooling flow	0.889	sharp	52.16	0.0	1	g/s
top composition			13.26	88.85	2	%methanol
tower pressure			4.981	2.491	2	KPa
condenser level			82.74	20.68	2	KPa
reboiler level			82.74	20.68	2	KPa

Equation types 1 $Y = AX^{\frac{1}{2}} + B$

2 $Y = AX^{\frac{1}{2}} + B$

Y - measurement

X - % chart

the samples was determined by measuring the specific gravity using a Christian Becker specific gravity balance (Model SG1). Figure C.7 shows the relationship between chart percent and the methanol concentration in the top product.

C.4 Analog Controllers

The feed, reflux, and steam flow analog controllers were tuned to give the minimum overshoot and oscillations for step disturbances in set point. The reboiler and condenser level controllers and the tower pressure controller were tuned to have virtually no overshoot. This was necessary to comply with the assumptions of constant tower pressure and constant reboiler and condenser levels. The controller settings for the feed and reflux temperature control were chosen in such a manner as to maintain a relatively constant temperature for all disturbances. The feed and reflux enthalpies were controlled by maintaining slightly lower temperatures than the flow temperatures on those trays to ensure that no flashing occurred. Table C.2 summarizes the controller settings used.

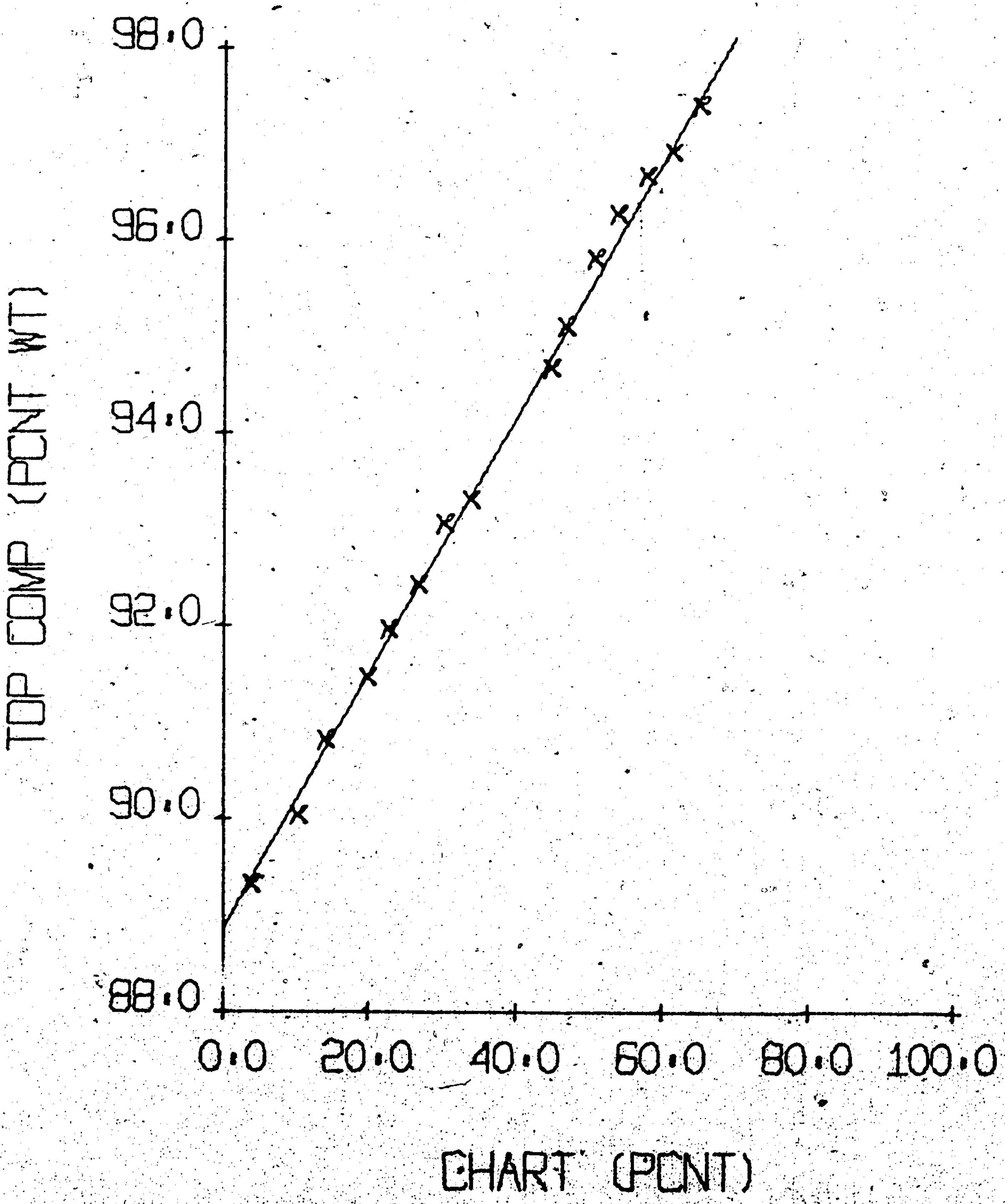

Figure C.7 Top composition (% methanol) recorder calibration curve.

Table C.2 Analog Controller Settings

Variable	Pb(%)	Reset(min)	Deriv(min)
Feed Flow	110	0.2	0.01
Reflux Flow	120	0.2	0.01
Steam Flow	180	0.2	
Reboiler level	120	6.0	
Condenser Level	150	3.0	
Tower Pressure	150	0.1	0.22
Feed Temp	50	0.6	
Reflux Temp	30	6.5	

C.5 DDC and Data Accumulation Loops

Table C.3 summarizes the DDC loops and their corresponding variable names used in the experimental work. Table C.4 shows the hexadecimal words used to specify each of the loops listed in Table C.3. The meaning of each word is described in the DDC manual located in the DACS Centre of the Department of Chemical Engineering.

Table C.3 DDC and Data Accumlation Loops

Loop ID		DDC Loops Units
0601	feed flow	g/s
0602	reflux flow	g/s
0603	steam flow	g/s
0604	top composition	% methanol
0605	top product flow	g/s
0606	bottom product flow	g/s
0607	bottom filtered composition	% methanol
0608	bottom raw composition	% methanol
0609	steam temperature	°C
0610	reboiler vapour temperature	°C
0611	tray 1 temperature	°C
0612	tray 2 temperature	°C
0613	tray 3 temperature	°C
0614	tray 4 temperature	°C
0615	tray 5 temperature	°C
0616	tray 6 temperature	°C
0617	tray 7 temperature	°C
0618	tray 8 temperature	°C
0619	condenser temperature	°C
0620	reflux inlet temperature	°C
0621	feed inlet temperature	°C
0622	cooling water inlet temperature	°C
0623	cooling water outlet temperature	°C
0624	tower pressure	KPa
0625	differential pressure	KPa
0626	reboiler level	KPa
0627	cooling water flow	g/s
0628	reboiler liquid temperature	°C
	Data Accumulation Loops	
0629	reboiler level	KPa
0630	feed flow	g/s
0631	reflux flow	g/s
0632	steam flow	g/s
0633	top composition	% methanol
0634	bottom raw composition	% methanol
0635	bottom filtered composition	% methanol
0636	reboiler liquid temperature	°C
0637	steam temperature	°C
0638	bottom product flow	g/s
0639	top product flow	g/s
0683	differential pressure	KPa

Table C.4 IBM 1800 DDC Loop Records (Hexidecimal)

```
          LOOP RECORD 0601

0601  E21B  105C  1006  3800  9E40  12E8  0000  0000  6615
7FFF  7FFF  2DDC  7FFF  0000  2DD0  7FFF  0000  2DD0  106A
528A  0080  0000  71A1  C840  00CD  FFC1

          LOOP RECORD 0602

0602  E21B  105C  1005  1B00  9E40  1414  0000  0000  433C
7FFF  7FFF  000C  7FFF  0000  0029  7FFF  0000  2110  116A
2663  0380  0000  223A  C000  0400  FFF0

          LOOP RECORD 0603

0603  E21B  105C  1004  3800  9E40  2134  0000  0000  2D2D
7FFF  7FFF  200C  7FFF  0FA4  2DD0  7FFF  0000  2000  126A
0C28  0080  0000  7FFF  3B40  0066  FFDB

          LOOP RECORD 0604

0604  E21B  114F  100A  0000  9100  0A5D  22B5  E729  2A1A
7FFF  7FFF  255C  35E8  0000  0000  7FFF  0000  0000  0602
9470  2020  0000  810F  D5C0  001F  FFDB

          LOOP RECORD 0605

0605  E21B  005E  1007  3340  9E40  0C44  0000  0000  588F
7FFF  7FFF  2DDC  7FFF  0000  2DD0  7FFF  0000  0000  1000
0000  0000  0000  0000  0000  0000  0000

          LOOP RECORD 0606

0606  E010  205F  1008  1000  9EE0  0B88  0000  0000  70E7
7FFF  7FFF  211C  7FFF  0000  A11C

          LOOP RECORD 0607

0607  A010  407F  0608  3A00  9180  4E20  0000  04CC  0676
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0608

0608  8010  1007  0000  0000  9180  4E20  0000  04D6  0E56
7FFF  7FFF  2110  7FFF  0000  0000
```

```
          LOOP RECORD 0609

0609   F010   21BF   E08C   3800   A0E0   2B66   FF4C   09DF   0000
0000   0000   C000   0000   0000   C00C

          LOOP RECORD 0610

0610   F010   21BF   E091   3800   A0C0   2B66   FF4C   09F2   0000
7FFF   7FFF   2110   7FFF   0000   2110

          LOOP RECORD 0611

0611   F010   21BF   E083   3800   A0C0   2B66   FF4C   09FE   0000
7FFF   7FFF   2110   7FFF   0000   2110

          LOOP RECORD 0612

0612   F010   21BF   E084   3800   A0C0   2B66   FF4C   09F4   0000
7FFF   7FFF   2110   7FFF   0000   2110

          LOOP RECORD 0613

0613   F010   21BF   E085   3800   A0C0   2B66   FF4C   0A0E   0000
7FFF   7FFF   2110   7FFF   0000   2110

          LOOP RECORD 0614

0614   F010   21BF   E086   3800   A0C0   2B66   FF4C   0A1A   0000
7FFF   7FFF   2110   7FFF   0000   2110

          LOOP RECORD 0615

0615   F010   21BF   E087   3800   A0C0   2B66   FF4C   09EA   0000
7FFF   7FFF   2110   7FFF   0000   2110

          LOOP RECORD 0616

0616   F010   21BF   E088   3800   A0C0   0000   FF4C   097E   0000
7FFF   7FFF   2110   7FFF   0000   2110
```

```
          LOOP RECORD 0617

0617  F010  21BF  E089  3800  A0C0  2B66  FF4C  0976  0000
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0618

0618  F010  21BF  E08A  3800  A0C0  2B66  FF4C  0974  0000
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0619

0619  F010  21BF  E08B  3800  A0C0  2B66  FF4C  0940  0000
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0620

0620  F010  21BF  E093  3800  A0C0  2B66  FF4C  094F  0000
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0621

0621  F010  21BF  E092  3800  A0C0  2B66  FF4C  0B73  0540
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0622

0622  F010  21BF  E095  3800  A0C0  2B66  FF4C  0945  0000
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0623

0623  F010  21BF  E096  3800  A0C0  2B66  FF4C  093C  0000
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0624

0624  E010  204F  100D  0000  9540  03E4  EF07  E72E  0000
0000  7FFF  300C  7FFF  0000  3000
```

```
          LOOP RECORD 0625

0625  E010  204F  100E  0000  A540  0677  00CF  E695  0000
7FFF  7FFF  300C  7FFF  0000  3000

          LOOP RECORD 0626

0626  E010  204F  100C  0000  A540  F989  040A  E9BF  0000
7FFF  7FFF  300C  7FFF  0000  3000

          LOOP RECORD 0627

0627  E010  205F  1009  3340  AEE0  28BE  0000  0000  0000
7FFF  7FFF  211C  7FFF  0000  A11C

          LOOP RECORD 0628

0628  F010  21BF  E082  3800  A0C0  2B66  FF4C  09D0  0000
7FFF  7FFF  2110  7FFF  0000  2110

          LOOP RECORD 0629

0629  86[illegible]  6F04  0626  001A  08AF  E112  EAFD  EA0D  EA32
E9C4  EAB9  EABE  EAE1  E9AE  E9B0  EAFD  EB2A  E9CE  E9BA
E9D3  EA12  EAF5  EB0E  EA35  E9B0  E9C4  E9C4  E9CE  EAAF
EAD5  E9AB

          LOOP RECORD 0630

0630  8620  6000  0601  001D  08A[illegible]  FF03  0000  0000  0000
0000  0000  0000  0000  0000  0000  0000  0000  0000  0000
0000  0000  0000  0000  0000  0000  0000  0000  0000  0000
0000  0000

          LOOP RECORD 0631

0631  8620  6F04  0602  1008  08AF  E112  0000  0000  0000
0000  0000  0000  0000  0000  0000  0000  0000  0000  0000
0000  0000  0000  0000  0000  0000  0000  0000  0000  0000
0000  0000
```

```
          LOOP RECORD 0632

0632  8620  6F04  0603  0000  08AF  E112  0000  0000  0000
0000  0000  0000  0000  0000  0000  0000  0000  0000  0000
0000  0000  0000  0000  0000  0000  0000  0000  0000  0000
0000  0000

          LOOP RECORD 0633

0633  8620  6F04  0604  000E  08AF  E112  E715  E715  E717
E72E  E726  E712  E710  E712  E70B  E712  E735  E703  E70B
E710  E710  E6FE  E712  E701  E72B  E71A  E708  E712  E717
E70B  E712

          LOOP RECORD 0634

0634  8620  6F04  0608  0014  08AF  FF12  04D6  04D6  04D6
04D6  04D6  04D6  04D6  04D6  04D6  04D6  04D6  04D6  04D6
04D6  04D6  04D6  04D6  04D6  04D6  04D6  04D6  04D6  04D6
04D6  04D6

          LOOP RECORD 0635

0635  8620  7F04  0607  0009  08AE  FF0E  04CC  04CC  04CC
04CC  04CC  04CC  04CC  04CC  04CC  04CC  04CC  04CC  04CC
04CC  04CC  04CC  04CC  04CC  04CC  04CC  04CC  04CC  04CC
04CC  04CC

          LOOP RECORD 0636

0636  8620  6000  0628  000A  08B0  E107  09CB  09D4  09CD
09C8  09E2  09DA  09DF  09D4  09D6  09DC  09CE  09E2  09CD
09D3  09CF  09E1  09D0  09D6  09E5  09D8  09CC  09CB  09CB
09CB  09CB

          LOOP RECORD 0637

0637  8620  6F04  0609  001F  08AF  E112  09DA  09EB  09E7
09E9  09E2  09E4  09E9  09DF  09E9  09DA  09DA  09EF  09DF
09E8  09F1  09EC  09EF  09DF  09E3  09DC  09DC  09DC  09E4
09EE  09E6
```

```
          LOOP RECORD 0638

0638   8620   6F04   0606   0000   08B0   E116   0000   0000   0000
0000   0000   0000   0000   0000   0000   0000   0000   0000   0000
0000   0000   0000   0000   0000   0000   0000   0000   0000   0000
0000   0000

          LOOP RECORD 0639

0639   8620   6F04   0605   001B   08B0   E116   0000   0000   0000
0000   0000   0000   0000   0000   0000   0000   0000   0000   0000
0000   0000   0000   0000   0000   0000   0000   0000   0000   0000
0000   0000

          LOOP RECORD 0683

0683   8620   6000   0624   101A   08B0   E107   E72E   E71C   E735
E721   E71F   F726   E72E   E71A   E72B   E744   E729   E721   E721
E724   E729   E726   E721   E71F   E71F   E729   E733   E721   E738
E726   E729
```

APPENDIX D

D.1 Bottom and Feed Composition Analysis

A Hewlett Packard (HP-5720A) gas chromatograph connected to the IBM 1800 was used to analyze the bottom product and feed compositions. The feed composition was determined by manual sample injections whereas the bottom composition had an automatic liquid sampling valve. The previous gas chromatograph sample system had to be improved for the following reasons;

(1) analysis time was too long ie. 4.0 minutes

(2) transportation time was too long ie. 16 minutes

(3) difficulty in detecting the methanol peak at low concentrations, and

(4) sampling valve was often plugged

If the gas chromatograph was to be used for control studies, one can appreciate the problems that are involved if all of the above four problems exist simultaneously.

D.2 Design of Sampling System

When redesigning the sampling system, the objective was to reduce the long transportation and dead time delays. These delays were caused by the large pressure drop across the liquid sample valve which reduced the flow rate through it to about 26.0 cubic centimetres per minute. The sample line contained two filters since it was necessary to remove particles which tended to plug the sample valve. (A detailed description and operation of the sample valve is given by Simonsmeier (53).) The filters capacities were 200 cm^3 and 5 cm^3 respectively, therefore based on the flow rate, a dead time of approximately 8 minutes resulted.

The design of the system is shown in Figure D.1. Since the liquid only needs to flow through the sample valve at injection time, it was decided to structure the system so that the flow by-passed the small filter and the sample valve. A solenoid valve was place in the line and was in the open state under normal operating conditions. It was closed shortly before each new sample was injected into the gas chromatograph for analysis. The flow was directed through the sample valve long enough so that the line was flushed with a new sample. After some experimentation, it was found that about 10 seconds was adequate to flush the line. With this method the performance of the sample valve improved. To improve the flow through the system, it was also necessary to clean the filters in the line from time to time. Upon implementation of the new design of the sampling system and based on a new flow rate of 30 cm^3/s, the transportation delay was reduced to 6 seconds. This represents a drastic reduction when compared to the 16 minute time delay which existed previously.

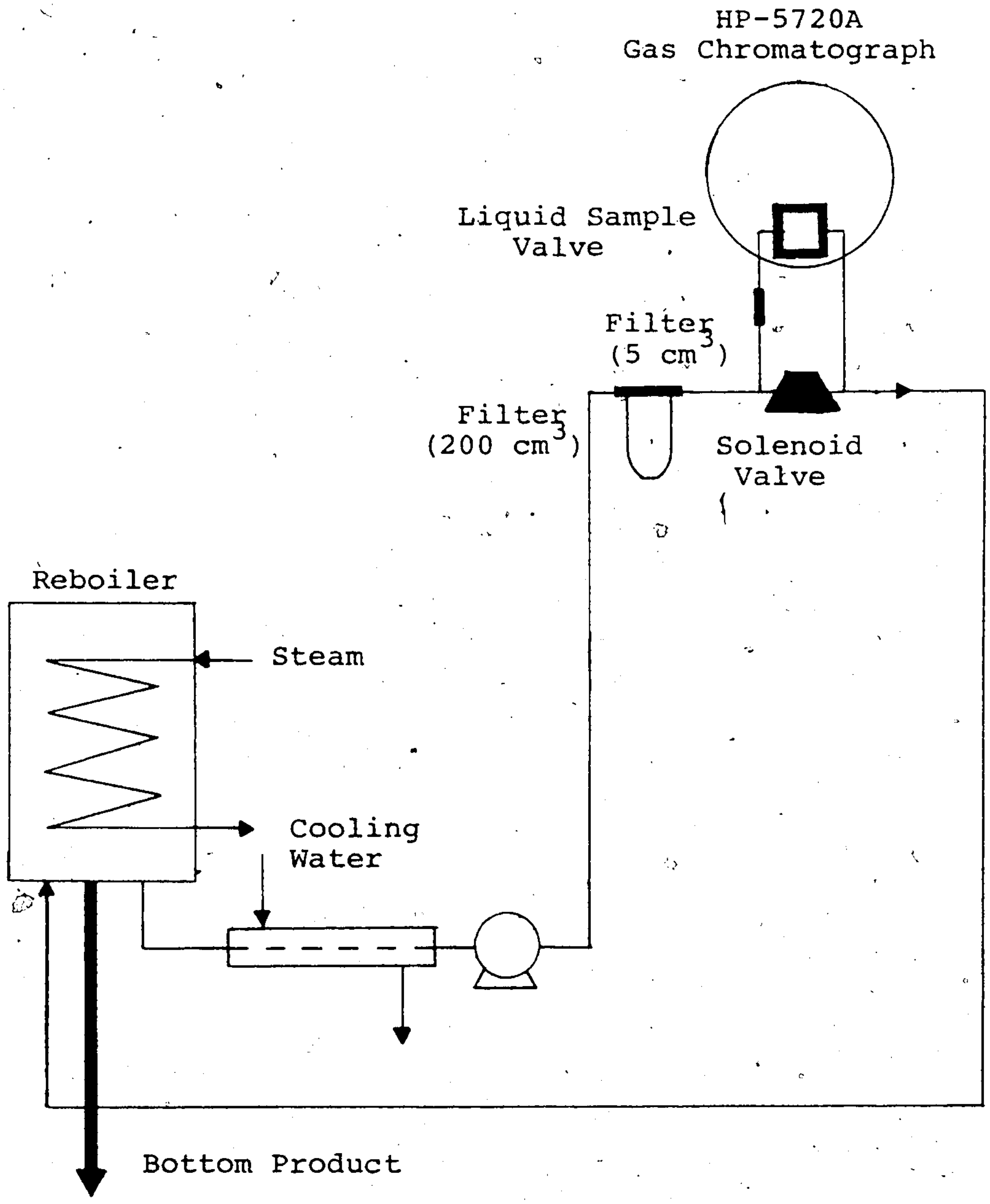

Figure D.1 Schematic diagram of gas chromatograph sampling system.

The heat exchanger in the line was used to cool the liquid down before it entered the pump. This was done to prevent flashing of the liquid and heating up of the pump. Under normal operations it was found that the heat removed from the system was quite large at maximum cooling water flow. This affected the selection of desired operating conditions, therefore flow of cooling water was kept to a minimum during experimental runs.

D.3 Gas Chromatograph Operation

With the previous length of column and type of packing, the analysis time was 256 seconds. The packing used resulted in water being the first component to be analysed and methanol the second. At low concentrations (less than 1.5 percent), it was difficult for the gas chromatograph package program to accurately determine the small methanol peak on the back side of the large water peak.

With the assistance and advice of Jerry Moser (Chemical Technologist, in the Department of Chemical Engineering, University of Alberta) a different packing was developed. The packing material was developed to elute methanol before water and consisted of 3 wt. % Apiezon L(obtained from Mandel Scientific Co. Ltd., Montreal, Quebec) on 80/100 mesh Chromosorb W. A/W(from Chromatogrph Specialties, Brockville, Ontario). The packing was prepared by using oil-free, dry air in a fluidized bed to remove the chloroform from the chloroform-Apiezon L-Chromosorb W. A/W slurry. The resulting almost chloroform free packing was then heated under nitrogen to 150 °C for 16 hours to remove all traces of chloroform. The packing was sieved in a Tyler sieve arrangement to yield the 80/100 mesh fraction. The use of this new packing material gave good separation of water and methanol in one half the previous analysis time. Secondly, the order of elution was reversed, that is, the methanol peak now was analysed first. This method facilitates more accurate and consistent analysis by the gas chromatograph. Finally, the gas chromatograph was later found to be accurate for both the bottom and feed composition analysis at similar operating conditions. The logic of peak detection by the gas chromatograph package is based on the first and second derivatives. Figures D.2 and D.3 show the behaviour of the first and second derivative for a methanol composition of 5.0% and 50.0%. It can be seen that the gas chromatograph package on the IBM 1800 was easily able to analyze the behaviour.

D.4 Feed Composition

Since the feed composition remained constant, it was

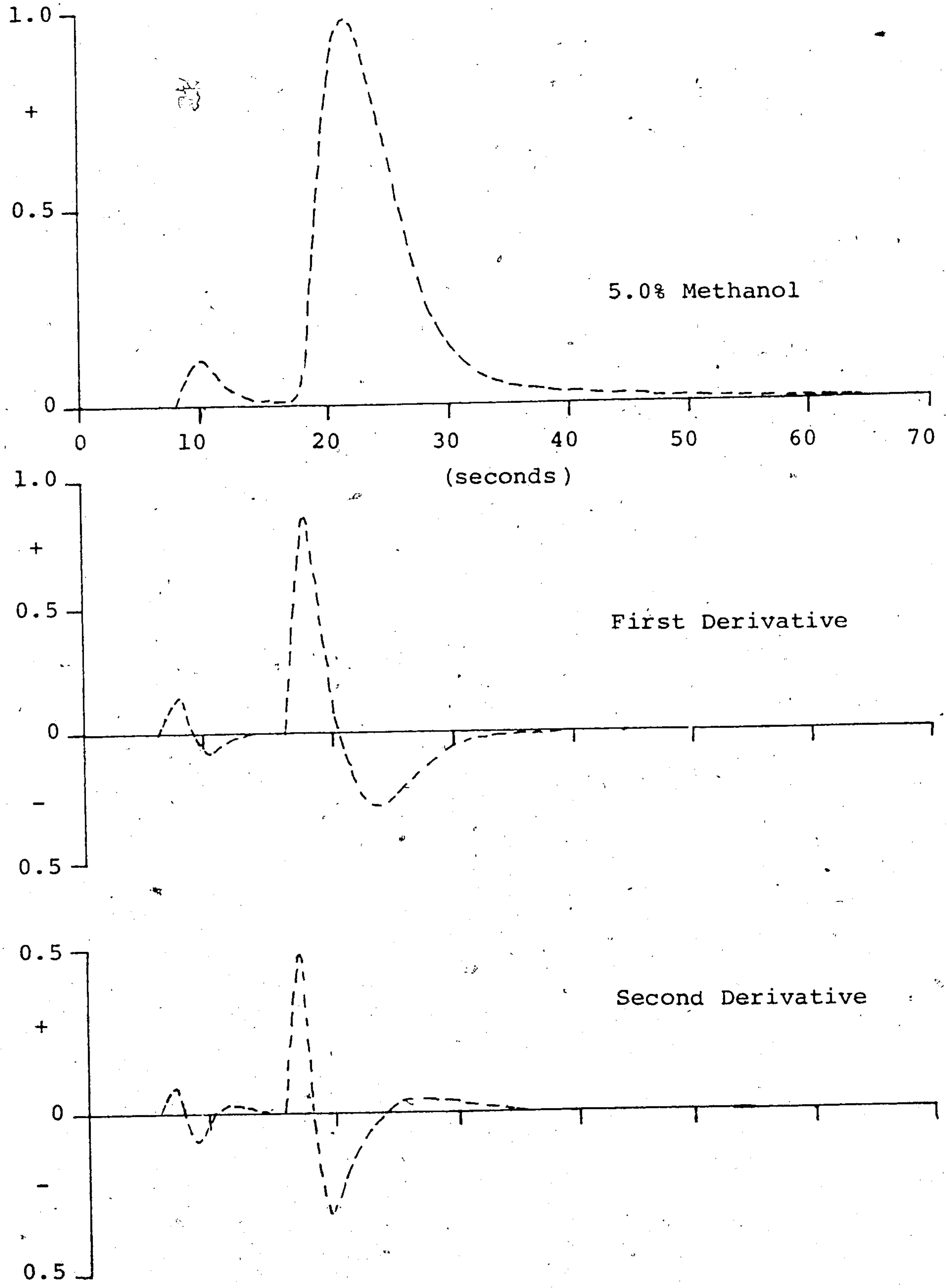

Figure D.2 Response of gas chromatograph to a 5.0% methanol sample.

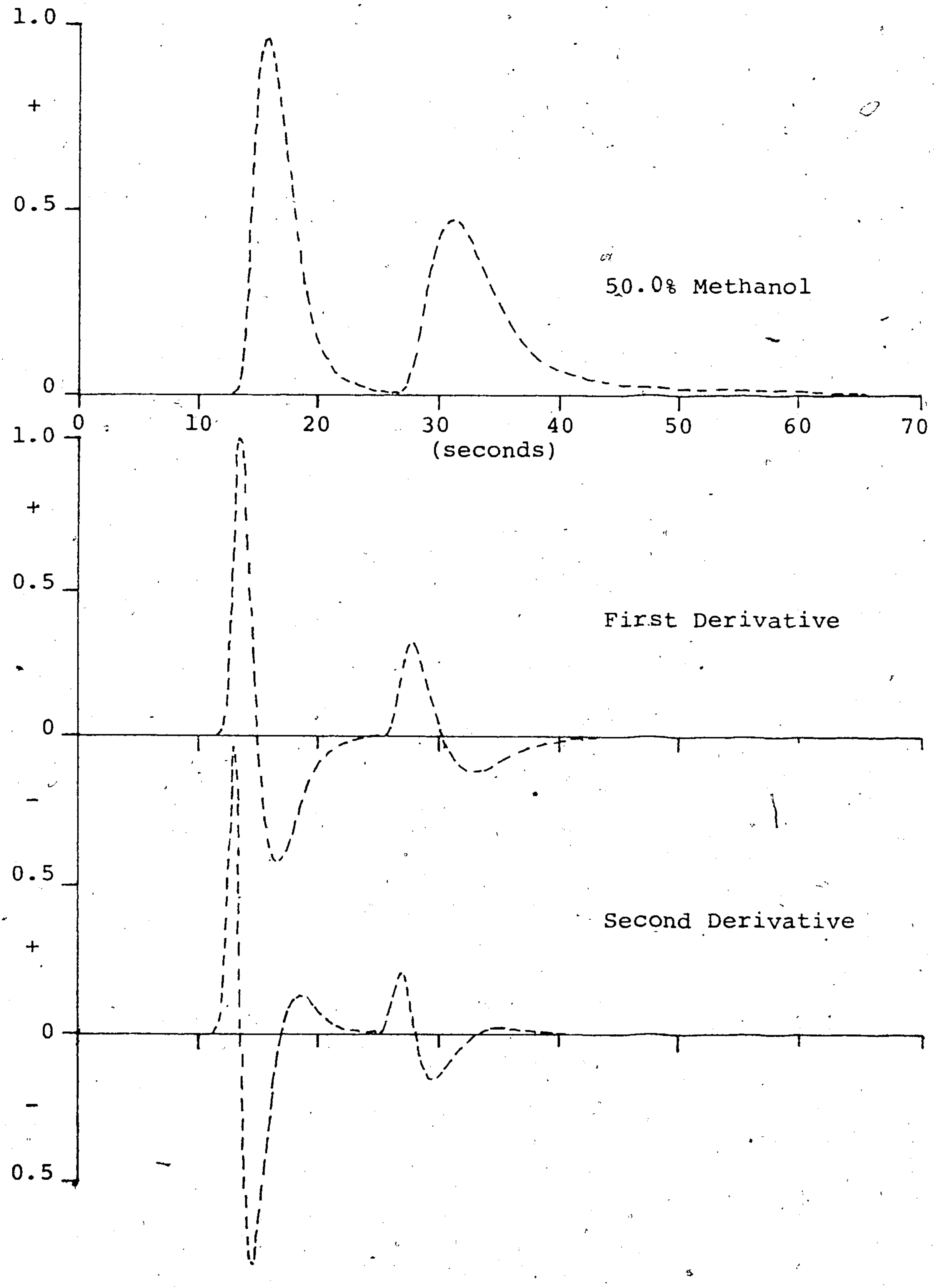

Figure D.3 Response of gas chromatograph to a 50.0% methanol sample.

not necessary to have an automatic sampling system. The feed composition was analysed by manual sample injections. The response factor seemed to change from day to day, therefore it was determined before each feed composition was analyzed. Gas chromatograph job #72 was used for the analysis of the feed composition. The operating conditions of the gas chromatograph for the bottom composition are listed in Table D.1 and are the same for the feed composition.

Table D.1 Gas Chromatograph Operating Conditions

sample flow	1.286 cm^3/s
oven temperature	98 °C
detector temperture	200 °C
sample valve temperature	150 °C
reference flow	0.366 cm^3/s
detector current	150 mA
sample size	1 µl

D.5 Bottom Composition

The response factor for the bottom composition, unlike that for feed composition, was more consistent on a day to day basis. The response factor for methanol is defined as follows;

RFmeth=wt. fraction*AH2O/Ameth*(1.0-wt. fraction) D-1

where RF - is the response factor for methanol
AH2O - area of water peak
Ameth - area of methanol peak
wt. fraction - wt. fraction of the methanol sample

Gas chromatograph job #73 was used to calculate the response factor of methanol. Table D.2 summarizes the response factor of methanol for compositions ranging from 0.5% to 15.0%. Three injections were analysed for each composition. The average reponse factor was 1.0245 which was close to the theoretical value of 1.035 (9). The elution time for methanol was about 15.5 seconds and for water it was about 40.0 seconds.

Gas chromatograph job #74 was used as the inline sampling program. This job was used to control the sample valve and solenoid valve action. On the first second of the automatic sampling, the solenoid valve was closed to allow liquid to flow through the sample valve. Eleven seconds later the sample valve injected a 1.0 microliter sample into the gas chromatograph. This process took about 2 seconds and one second later the solenoid valve reopened. Gas chromatograph job #74 was used only to calculate the peak areas and the elution time. Program DWL44 was used to calculate the composition on a percent methanol basis

Table D.2 Response Factor for Methanol

Wt. Fraction	Response Factor
0.1401	0.9938
	0.9983
	0.9993
0.1109	1.0002
	1.0024
	1.0023
0.0830	1.0061
	1.0169
	1.0054
0.0465	1.0119
	1.0076
	1.0131
0.0273	1.0075
	1.0192
	1.0133
0.0185	1.0297
	1.0356
	1.0252
0.0100	1.0519
	1.0606
	1.0646
0.0054	1.0786
	1.0101
	1.0856

and put the result into DDC loop 0608. The listing for each of the Gas Chromatograph jobs is given in Tables D.3 to D.5.

Table D.3 Gas Chromatograph Job Number 72

```
                                                                   28 JUL 78
                                                                   22/42 HRS
                        ** GAS CHROMATOGRAPH JOB LISTING  **
                           ******************************

JOB NUMBER =  72           CALCULATION OPTION =   6
GC NUMBER  =   7           TOTAL PEAKS        =   6

FINISH TIME            =      100
NORMALIZATION CONST. =        100.

                             * REFERENCE PEAK DATA  *
                               -------------------

LOW TIME OF REF PEAK                        10.
HIGH TIME OF REF PEAK                       28.
CONCENTRATION OF REF PEAK                    0.
RESPONCE FACTOR OF REF PEAK                  0.

                                * TIME BAND DATA *
                                  --------------

LOW TIME  HIGH TIME  CONC FOR STND  FACTOR  JTYPE  COMPONENT
   10.       28.          0.        1.000     1     METHANOL
   28.      100.          0.        1.000     1     WATER

   COMPONENT       SEQNO  IPLUS  INREF  IFCGO  IREST  UNITS
METHANOL             1      0      9      0      0     WT. %
WATER                2      0      0      0      0     WT. %

                      * PARAMETER AND CONTROL ACTION DATA  *
                        ---------------------------------

PARAMETER ACTIONS

 ACTION TIME  KNDTM   ISTS   IRATE  IHIGH   ILOW  IHARD  ISOFT  IEXP1  IEXP2
      0        AB      11    8PPS    750   -750     8      8      1      2
     10        AB       1   16PPS    100    -50     6      4      1      2
    100        AB      20    8PPS    500   -500     6      4      1      2
    100        RL      19    8PPS    500   -500     6      4      1      2

ECO ACTIONS

   NO ECO ACTIONS SPECIFIED

JOB COMPLETE
```

Table D.4 Gas Chromatograph Job Number 73

```
                                                                        28 JUL 78
                                                                        22/43 HRS
                    ** GAS CHROMATOGRAPH JOB LISTING  **
                       ********************************

   JOB NUMBER =  74            CALCULATION OPTION =   8
   GC NUMBER  =   7            TOTAL PEAKS        =   6

   FINISH TIME            =        90
   NORMALIZATION CONST. =         100.

                           * REFERENCE PEAK DATA   *
                             -------------------

   LOW TIME OF REF PEAK                        22.
   HIGH TIME OF REF PEAK                       37.
   CONCENTRATION OF REF PEAK                    0.
   RESPONCE FACTOR OF REF PEAK                  0.

                              * TIME BAND DATA *
                                --------------

   LOW TIME   HIGH TIME   CONC FOR STND   FACTOR   JTYPE   COMPONENT
       22.         37.           0.        1.000      1      METHANOL
       37.        105.           0.        1.000      1      WATER

       COMPONENT        SEQNO   IPLUS   INREF   IFCGO   IREST   UNITS
   METHANOL                 1      0       9       0       0     WT. %
   WATER                    2      0       0       0       0     WT. %

                    * PARAMETER AND CONTROL ACTION DATA   *
                      ---------------------------------

   PARAMETER ACTIONS

    ACTION TIME   KNDTM    ISTS   IRATE   IHIGH    ILOW   IHARD   ISOFT   IEXP1   IEXP2
            0       AB      11     8PPS    500    -500      8       8       1       2
           22       AB       1    16PPS    150     -75      8       6       1       2
          110       AB      20     8PPS    750    -750      6       4       1       2
           90       RL      19     8PPS    750    -750      6       4       1       2

   ECO ACTIONS

    ACTION TIME   KNDTM   GROUP   BIT      ON/OFF
            1       AB      16     6       ON
           11       AB      16     5       ON
           13       AB      16     5       OFF
           14       AB      16     6       OFF
```

Table D.5 Gas Chromatograph Job Number 74

```
                                                                         28 JUL 78
                                                                         22/42 HRS
                     ** GAS CHROMATOGRAPH JOB LISTING   **
                        ******************************

JOB NUMBER =  73            CALCULATION OPTION =   7
GC NUMBER  =   7            TOTAL PEAKS        =   6

FINISH TIME            =       90
NORMALIZATION CONST. =        100.

                          * REFERENCE PEAK DATA   *
                            -------------------

LOW TIME OF REF PEAK                          11.
HIGH TIME OF REF PEAK                         26.
CONCENTRATION OF REF PEAK                      0.
RESPONCE FACTOR OF REF PEAK                    0.

                            * TIME BAND DATA *
                              --------------

LOW TIME   HIGH TIME   CONC FOR STND   FACTOR   JTYPE   COMPONENT
    11.        26.            0.        1.000     1       METHANOL
    26.        95.            0.        1.000     1       WATER

    COMPONENT          SEQNO   IPLUS   INREF   IFCGO   IREST   UNITS
METHANOL                 1       0       9       0       0      WT. %
WATER                    2       0       0       0       0      WT. %

                   * PARAMETER AND CONTROL ACTION DATA   *
                     ---------------------------------

PARAMETER ACTIONS

 ACTION TIME   KNDTM   ISTS   IRATE   IHIGH   ILOW   IHARD   ISOFT   IEXP1   IEXP2
       0         AB     11     8PPS    500    -500     8       8       1       2
      11         AB      1    16PPS    150     -75     8       6       1       2
      95         AB     20     8PPS    750    -750     6       4       1       2
      90         RL     19     8PPS    750    -750     6       4       1       2

ECO ACTIONS

   NO ECO ACTIONS SPECIFIED

JOB COMPLETE
```

APPENDIX E

E.1 Computer Program Listings

The following computer programs were created for use on the IBM 1800 computer. No description of the programs will be given because the programs are described by comment cards. The program names are as follows DASS1, DATA1, BALN1, INFF, DWL44, DISTB.

```
C
C                    PROGRAM  DASS1
C                    ---------------
C
C
C     PURPOSE
C        START DATA ACQUISITION FOR STEADY-STATE DATA AND
C        PERFORM A MATERIAL AND ENERGY BALANCE ON THE
C        DISTILLATION COLUMN
C
C     DESCRIPTION OF PARAMETERS
C        MAX       NUMBER OF DATA POINTS
C        TIME      SAMPLE TIME(SECONDS)
C        RUN       RUN NUMBER
C        DATE      DATE OF THE EXPERIMENT
C        BTCMP     BOTTOM PRODUCT COMPOSITION(%MEOH)
C        FDCMP     FEED COMPOSITION(%MEOH)
C        FDPLT     FEED PLATE NUMBER
C
C     USAGE
C        QUEUE FROM ANY TELETYPE AND ENTER DATA AS REQUIRED
C
C        EXAMPLE
C         QDASS1
C                ENTER OPTION 1=GO, 2=STOP
C
C         1
C
C                ENTER--MAX, TIME, RUN(2A4), DATE(2A4)
C
C         10,1,RBIL-001,03/07/78
C                ENTER BOTTOM COMPOSITION,FEED COMPOSITION
C                AND FEED PLATE
C
C         5.0,50.0,4
C                DATA1 QUEUED
C
C     SUBROUTINES AND PROGRAMS REQUIRED
C        DATA1,INFF
C
C     SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C        GETTY,FFINP,QLEVEL,CANCL
C
C

      INTEGER FDPLT
      EXTERNAL DATA1
      DIMENSION RUN(2), DATE(2), TITLE(10,3)
      DIMENSION ZERO(32)
```

...(CONT'D)

```
      DATA ZERO/32*0.0/
      DEFINE FILE 10(60,80,U,NEXT)
      KOUNT=0
      CALL GETTY(LUW)
      WRITE (LUW,200)
  200 FORMAT (T5, 'ENTER OPTION 1=GO, 2=STOP')
    1 CALL FFINP(LUW,1,10,IOPT,IER)
      IF (IER) 5,2,1
    2 GO TO (3,14), IOPT
    3 WRITE (LUW,201)
  201 FORMAT (T5, 'ENTER--MAX, TIME, RUN(2A4), DATE(2A4)')
    4 CALL FFINP(LUW,4,10,MAX,10,IPOL,23,RUN,23,DATE,IER)
      IF (IER) 5,7,6
    5 WRITE (LUW,202)
  202 FORMAT (T5, 'CALL TO FFINP IN ERROR--CRASH')
      CALL EXIT
    6 WRITE (LUW,203)
  203 FORMAT (T5, 'INPUT TO FFINP IN ERROR--TRY AGAIN')
      GO TO 4
    7 CONTINUE
      WRITE(LUW,204)
  204 FORMAT (T5, 'ENTER BOTTOM COMPOSITION,FEED
     * COMPOSITION'
     1,/,T5,'AND FEED PLATE')
    8 CALL FFINP(LUW,3,11,BTCMP,11,FDCMP,10,FDPLT,IER)
      IF (IER) 5,10,9
    9 WRITE(LUW,205)
  205 FORMAT(T5, 'INPUT TO FFINP IN ERROR--TRY AGAIN')
      GO TO 8
   10 READ (10'1) NLOOP
      WRITE (10'1) NLOOP,KOUNT,MAX,IPOL,BTCMP,FDCMP,FDPLT
      READ (10'4) ((TITLE(I,J), J=1,3), I=1,10)
      WRITE (10'4) ((TITLE(I,J), J=1,3), I= 1,10), (RUN(I),
     * I=1,2),
     1  (DATE(I), I=1,2)
      WRITE(10'8)(ZERO(I),I=1,32)
      WRITE(10'9)(ZERO(I),I=1,32)
      CALL QLEVL(DATA1,1,0,0,IEQ)
      GO TO (11,12,13),IEQ
   11 CONTINUE
      WRITE (LUW,206)
  206 FORMAT (T5, 'DATA1 QUEUED')
      CALL EXIT
   12 CONTINUE
      WRITE (LUW,207)
  207 FORMAT (T5, 'QUEUE TABLE IS FULL')
      CALL EXIT
   13 CONTINUE
      WRITE (LUW,208)
  208 FORMAT (T5, 'PROGRAM WAS QUEUED')
```

...(CONT'D)

```
      CALL EXIT
   14 IT = 9
      CALL CANCL(IT)
      WRITE (LUW,209)
  209 FORMAT(T5, 'DASS1--EXIT')
      CALL EXIT
      END
```

```
C
C                      PROGRAM DATA1
C                      -------------
C
C
C     PURPOSE
C        DATA ACQUISITION FOR MATERIAL AND ENERGY BALANCE
C        CALCULATIONS
C
C     USAGE
C        THIS PROGRAM IS QUEUED BY DASS1 AND 'MAX' DATA
C        POINTS ARE SAMPLED EVERY 'IPOL' SECONDS
C
C     REMARKS
C        USE OF THIS PROGRAM WILL SUSPEND EXECUTION OF
C        OTHER PROGRAMS IN VCORE FOR A PERIOD OF 'MAX*IPOL'
C        AS SPECIFIED BY THE USER IN THE PROGRAM DASS1
C
C     SUBROUTINES AND PROGRAMS REQUIRED
C        DASS1,BALN1,COLDJ,ZERCK,TCONV
C
C     SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C        AIP,SUSPN,QLEVEL,CANCL
C
C


      EXTERNAL BALN1
      EXTERNAL DATA1
      DIMENSION A(32),B(32),TOT(32),SUMSQ(32),ICODE(32)
     *,IC(32)
      DEFINE FILE 10(60,80,U,INXT)

C     MULTIPLEXER NUMBER FOR EACH VARIABLE
C

      DATA ICODE/4102,4101,4100,4104,4103,4105,4106,4109
     *,4107,4108,4110
     *,-8062,-8061,-8060,-8059,-8058,-8057,-8056,-8055,-8054
     *,-8053,-8052
     *,-8051,-8050,-8049,-8048,-8047,-8046,-8045,-8044,-8043
     *,-8042/

C     CALCULATION CODE
C        1=ZERO CHECK PLUS FLOW CALCULATION
C        2=ZERO CHECK ONLY
C        3=TEMPERATURE CALCULATION
C

      DATA IC/6*1,5*2,21*3/
```

...(CONT'D)

```
C      CONA=SPAN/32767
C
       DATA A/7.385E-04,7.843E-04,1.297E-03,4.578E-04,4.791E
      *-04
      *,159.194E-04,4.048E-04,1.520E-04,2.525E-03,-2.525E-03
      *,2.525E-03
      *,21*1111.0/
C      CONB=INTERCEPT
C
       DATA B/6*0.0,88.85,-2.491,20.684,103.420,20.684,21*
      *-18.0/
       READ (10'1) NLOOP, KOUNT, MAX ,IPOL
       READ(10'8)(TOT(I),I=1,32)
       READ(10'9)(SUMSQ(I),I=1,32)
       DO 999 KOUNT = 1,MAX
C      CALCULATE COLD JUNCTION COMPENSATION
C
       CALL COLDJ(TRBT)
       DO 140 I=1,32
C      GET CALCULATION TYPE
C
       ITYPE=IC(I)
C      GET MULTIPLEXER ADDRESS
C
       MPXAD=ICODE(I)
C      READ INPUT FROM MULTIPLEXER
C
       CALL AIP(11000,INPUT,MPXAD)
    20 CALL AIP(0,JTEST)
       GO TO(20,40),JTEST
    40 DATA=INPUT
       GO TO(60,80,100),ITYPE
C      ZERO CHECK PLUS FLOW CALCULATION
C
    60 CALL ZERCK(DATA)
```

...(CONT'D)

```
C       TAKE SQUARE ROOT AND RESCALE INPUT
C

        DATA=SQRT(DATA)*SQRT(32767.0)
        GO TO 110

C       ZERO CHECK ONLY
C

     80 CALL ZERCK(DATA)
        GO TO 110

C       TEMPERATURE CALCULATION
C

    100 CALL TCONV(DATA,INPUT,TRBT)
        GO TO 110
    110 CONTINUE
        DATA=A(I)*DATA+B(I)
        TOT(I)=TOT(I)+DATA
        SUMSQ(I)=SUMSQ(I)+DATA*DATA
    140 CONTINUE

C       WRITE DATA TO FILES '10' EVERY IPOL TIME
C

        WRITE (10'1) NLOOP, KOUNT, MAX
        WRITE(10'8)(TOT(I),I=1,32)
        WRITE(10'9)(SUMSQ(I),I=1,32)

C       CHECK IF ENOUGH POINTS ACCUMULATED
C

        TIME=IPOL
        INT=TIME/0.1
        CALL SUSPN (9,INT)
    999 CONTINUE
        CALL QLEVL(BALN1,1,0,0,IER)
        CALL CANCL(9)
        CALL EXIT
        END
```

```
C
C                    SUBROUTINE COLDJ
C                    ----------------
C
C
C     PURPOSE
C        CALCULATE COLD JUNCTION COMPENSATION FOR
C        TEMPERATURES
C
C     DESCRIPTION OF PARAMETERS
C        TRBT       COLD JUNCTION CONPENSATION
C
C     USAGE
C        CALL COLD(TRBT)
C
C     SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C        AIP
C
C

      SUBROUTINE COLDJ(TRBT)
      DIMENSION MPXAD(2),INPUT(2)
      DATA C,D/9.6447,-291.638/
      DATA MPXAD/-8063,-8064/

C     BRIDGE UNBALANCE=-8063, REFERENCE VOLTAGE=-8064
C

      DO 40 I=1,2
      CALL AIP(11000,INPUT(I),MPXAD(I))
   20 CALL AIP(0,JTEST)
      GO TO(20,40),JTEST
   40 CONTINUE
      TRBT=51.876*FLOAT(INPUT(1))/FLOAT(INPUT(2))+41.0
      TRBT=C*TRBT+D
      RETURN
      END
```

```
C
C                       SUBROUTINE ZERCK
C                       ----------------
C
C
C     PURPOSE
C        ZERO CHECK
C
C     DESCRIPTION OF PARAMETERS
C        DATAN      MULTIPLEXER READING IN BINARY COUNTS
C
C     USAGE
C        CALL ZERCK(DATAN)
C
C

      SUBROUTINE ZERCK(DATAN)
      DATAN=DATAN-6553.4
C     ZERO CHECK INPUT
C

      IF(DATAN)20,20,40
   20 DATAN=0.0
      GO TO 60
C     RESCALE INPUT
C

   40 DATAN=DATAN*(32767.0/(32767.0-6553.4))
   60 CONTINUE
      RETURN
      END
```

```
C
C                    SUBROUTINE TCONV
C                    ----------------
C
C
C      PURPOSE
C         CALCULATE THE TEMPERATURE FROM THE MUTIPLEXER
C         READING AND THE COLD JUNCTION COMPENSATION
C
C      DESCRIPTION OF PARAMETERS
C         TEMP      COMPENSATED TEMPERATURE IN DEG C
C         INPUT     MULTIPLEXER READING IN BINARY COUNTS
C         TRBT      COLD JUNCTION COMPENSATION
C
C      USAGE
C         CALL TCONV(TEMP,INPUT,TRBT)
C
C      RANGE  '0 DEG F TO 300 DEG F'  OR  '-17.78 DEG C TO
C      148.89 DEG C'
C      INCREMENT  25.0 DEG F
C
C      REMARKS
C      METHOD USE CALCULATE COUNT,SLOPE,INTCP
C
C           COUNT(I)=TEMP(DEG F)*65534.0/2000.0
C
C           SLOPE(I)=(10.0/2.0)*(65534.0/2
C           00.0)*((TEMP(N+1)-TEMP(N))
C                    /(V(N+1)-V(N)))
C           V(N)=V(TEMP)-V(75.0 DEG F)
C           V(N)=VOLTAGE IN MV
C
C           INTCP(I)=V(N)*16384.0*2.0/RANGE
C           RANGE=10 MV
C
       SUBROUTINE TCONV(TEMP,INPUT,TRBT)
       REAL INTCP(7)
       DIMENSION COUNT(7),SLOPE(7)
       DATA COUNT/-6914.048,-2359.296,2359.296
      *,7176.192,12091.392,17039.360,22020.096/


       DATA SLOPE/5893.885,5689.236,5573.129
      *,5461.167,5425.497,5389.803,5319.805/


       DATA INTCP/0.0,1638.35,3276.7,4915.05
      *,6553.4,8191.75,9830.1/
```

...(CONT'D)

```
C      ADD COLD JUNCTION COMPENSATION
C
       XNPUT=INPUT+TRBT
C      SEARCH FOR SEGMENT
C
       DO 10 I=1,7
       IF(COUNT(I)-XNPUT) 10,1,1
     1 INDX=I-1
       GO TO 2
    10 CONTINUE
C      CALCULATE TEMPERATURE IN BINARY COUNTS
C
     2 TC=(XNPUT-COUNT(INDX))*SLOPE(INDX)/16384.0+INTCP(INDX)
C      TEMPERATURE BEFORE FINAL CONVERSION TO DEGREES
C
       TEMP=TC/65534.0
       RETURN
       END
```

```
C
C                    PROGRAM BALN1
C                    -------------
C
C
C      PURPOSE
C         PERFORM MATERIAL AND ENERGY BALANCE CALCULATIONS
C         WHEN DATA ACCUMULATION HAS BEEN COMPLETED BY DATA1
C
C      USAGE
C         THIS PROGRAM IS QUEUED BY DATA1
C
C
C

       INTEGER FDPLT
       DIMENSION RUN(2),DATE(2)
       DIMENSION AV(34),VAR(34),DEV(34),TITLE(34,3),UNIT(34
      *,2)
       DEFINE FILE 10(60,80,U,NEXT)
C      DEFINE FUNCTION STATEMENTS
C

       TEMP(T)=(T*9.0/5.0)+32.0

C      CPMG(T)=10.26+1.984E-02*T-0.144E-05*T**2-1.92E-09*T**3

       CPML(T)=2.3305+7.8241E-03*T+37.6978E-06*T**2

C      CPML(T)=0.5573+0.1870E-02*T+0.9011E-05*T**2
C      CPWG(T)=7.880+0.32E-02*T-483.3E-05*T**2

       CPWL(T)=4.1882-5.6902E-04*T+8.4935E-06*T**2

C      CPWL(T)=1.0009-0.1362E-03*T+0.2026E-05*T**2
C      HEATM(T)=(291.2-0.3308*T-0.1526E-02*T**2)*1.8

       HEATW(T)=2510.35-9.27E-01*T-13.06E-04*T**2

C      HEATW(T)=1080.2-0.3785*T-0.5618E-03*T**2

       DENS(T,X)=1.0201-0.5123E-03*T-0.1512E-05*T**2-0.1519E
      *-02*100.*X
      1  -0.8106E-05*(100.*X)**2
       DENSW(T)=1.0050-0.2142E-03*T-0.2508E-05*T**2

C      READ DATA FROM FILES
C
```

...(CONT'D)

```
      READ (10'1) NLOOP,KOUNT,MAX,IPOL,BTCMP,FDCMP,FDPLT
      READ (10'2) ((TITLE(I,J), J=1,3), I=1,12)
      READ (10'3)  ((TITLE(I,J), J=1,3), I=13,24)
      READ (10'4) ((TITLE(I,J), J=1,3), I=25,34), (RUN(I),
     * I=1,2),
     1  (DATE(I), I=1,2)
      READ (10'5) ((UNIT(I,J), J=1,2), I=1,18)
      READ (10'6) ((UNIT(I,J), J=1,2), I=19,34)
      READ(10'8)(AV(I),I=1,32)
      READ(10'9)(VAR(I),I=1,32)

C     COMPUTE AVERAGE AND VARIANCE OF READINGS
C

      DO 1 I=1,32
      VAR(I)=(VAR(I)-AV(I)*AV(I)/KOUNT)/(KOUNT-1)
      DEV(I)=SQRT(VAR(I))
    1 AV(I)=AV(I)/KOUNT

C     INSERT FEED AND BOTTOM COMPOSITIONS IN VECTOR
C

      DO 2 I=1,32
      K=35-I
      IF(K-10)4,3,3
    3 AV(K)=AV(K-2)
      DEV(K)=DEV(K-2)
    2 CONTINUE
    4 DEV(8)=0.0
      DEV(9)=0.0
      AV(8)=BTCMP
      AV(9)=FDCMP

C      CORRECT FLOW SIGNALS FOR TEMPERATURE
C

      F=AV(1)*SQRT(DENS(AV(27),AV(9)/100.0))
      R=AV(2)*SQRT(DENS(AV(26),AV(7)/100.0))
      B=AV(4)*SQRT(DENS(AV(28),AV(8)/100.0))
      D=AV(5)*SQRT(DENS(AV(26),AV(7)/100.0))
      W=AV(6)*SQRT(DENSW(AV(33)))

C     OVERALL MATERIAL BALANCE
C

      FIN=F
      FOUT=D+B
      ERRT=(FOUT/FIN-1.0)*100.0

C     COMPONENT BALANCE
```

...(CONT'D)

```
C
      CIN1=F*AV(9)/100.0
      CIN2=F*(1.0-AV(9)/100.0)
      CT1=D*AV(7)/100.0
      CT2=D*(1.0-AV(7)/100.0)
      CB1=B*AV(8)/100.0
      CB2=B*(1.0-AV(8)/100.0)
      COUT1=CT1+CB1
      COUT2=CT2+CB2
      ERR1=(COUT1/CIN1-1.0)*100.0
      ERR2=(COUT2/CIN2-1.0)*100.0
C     ENERGY BALANCE
C     ENTHALPY BASED ON 0 DEG C
C
C     FEED ENTHALPY IN
C
      HFI1=CPML(AV(30))*AV(9)/100.0*F*AV(30)
      HFI2=CPWL(AV(30))*(1.0-AV(9)/100.0)*F*AV(30)
      HFI=HFI1+HFI2
C     STEAM ENTHALPY IN
C
      HSI=(HEATW(TEMP(AV(24)))+CPWL(AV(24))*AV(24))*AV(3)
C     REFLUX ENTHALPY IN
C
      HRI1=CPML(AV(31))*R*AV(7)/100.0*AV(31)
      HRI2=CPWL(AV(31))*R*(1.0-AV(7)/100.0)*AV(31)
      HRI=HRI1+HRI2
C     COOLING WATER ENTHALPY IN
C
      HWI=CPWL(AV(33))*W*AV(33)
C     TOTAL ENTHALPY IN
C
      HIT=HFI+HSI+HRI+HWI
C     BOTTOM ENTHALPY OUT
C
      HBO1=CPML(AV(14))*B*AV(8)/100.0*AV(14)
      HBO2=CPWL(AV(14))*B*(1.0-AV(8)/100.0)*AV(14)
```

...(CONT'D)

```
      HBO=HBO1+HBO2
C     TOP PRODUCT ENTHALPY OUT
C
      HTO1=CPML(AV(23))*D*AV(7)/100.0*AV(23)
      HTO2=CPWL(AV(23))*D*(1.0-AV(7)/100.0)*AV(23)
      HTO=HTO1+HTO2
C     REFLUX ENTHALPY OUT
C
      HRO1=CPML(AV(23))*R*AV(7)/100.0*AV(23)
      HRO2=CPWL(AV(23))*R*(1.0-AV(7)/100.0)*AV(23)
      HRO=HRO1+HRO2
C     STEAM CONDENSATE ENTHALPY OUT
C
      HSO=CPWL(AV(25))*AV(3)*AV(25)
C     COOLING WATER OUT
C
      HWO=CPWL(AV(34))*W*AV(34)
C     TOTAL ENTHALPY OUT
C
      HOT=HBO+HTO+HRO+HSO+HWO
C     DIFFERENCE IS THE HEAT LOSS
C
      HLOSS=HIT-HOT
C     DOCUMENT STEADY STATE, MATERIAL AND ENERGY BALANCES
C
C     STEADY STATE
C
      WRITE (6,5)
    5 FORMAT ('1',//////)
      WRITE (6,6) (RUN(I), I=1,2), (DATE(I), I=1,2)
    6 FORMAT (T51, 'STEADY STATE DATA'/T52, 'RUN NO ', 2A4
     */T55, 2A4//)
      WRITE (6,900) (TITLE(1,J), J=1,3), F, (UNIT(1,J), J=1
     *,2),
     1              (TITLE(4,J), J=1,3), B, (UNIT(4,J), J=1
     *,2)
```

...(CONT'D)

```
 900 FORMAT (T32, 3A4, T45, F7.3, 1X, 2A4, T62, 3A4, T75,
    * F7.3, 1X
    1,2A4)
     WRITE (6,900) (TITLE(2,J), J=1,3), R, (UNIT(2,J), J=1
    *,2),
    1                (TITLE(5,J), J=1,3), D, (UNIT(5,J), J=1
    *,2)
     WRITE (6,900) (TITLE(3,J), J=1,3), AV(3),  (UNIT(3,J),
    * J=1,2),
    1                (TITLE(6,J), J=1,3), W, (UNIT(6,J), J=1
    *,2)
     WRITE (6,901) FDPLT, (TITLE(9,J), J=1,3), FDCMP,
    * (UNIT(9,J),J=1,2)
 901 FORMAT (T32, 'FEED PLATE', T45, I3, T62, 3A4, T75,
    * F7.3, 1X, 2A4)
     WRITE (6,902) (TITLE(7,J), J=1,3), AV(7), (UNIT(7,J),
    * J=1,2),
    1                (TITLE(8,J), J=1,3), AV(8), (UNIT(8,J),
    * J=1,2)
 902 FORMAT (T32, 3A4, T45, F7.3, 1X, 2A4, T62, 3A4, T75,
    * F7.3, 1X
    1,2A4)
     WRITE (6,903) (TITLE(30,J), J=1,3), AV(30), (UNIT(30
    *,J), J=1,2),
    1                (TITLE(31,J), J=1,3), AV(31), (UNIT(31
    *,J), J=1,2)
 903 FORMAT (T32, 3A4, T45, F7.2, 1X, 2A4, T62, 3A4, T75,
    * F7.2, 1X
    1,2A4)
     WRITE (6,903) (TITLE(24,J), J=1,3), AV(24), (UNIT(24
    *,J), J=1,2),
    1                (TITLE(10,J), J=1,3), AV(10), (UNIT(10
    *,J), J=1,2)
     WRITE (6,60)

C      MATERIAL BALANCE
C

     WRITE (6,10)
  10 FORMAT (T44, 'M A T E R I A L   B A L A N C E'//T55,
    * 'FLOW', T65,
    1 'COMP', T72, 'METHANOL', T84, 'WATER',/T53, '(G/SEC)'
    *, T63,
    2 '(WT PCT)', T72, '(G/SEC)', T83, '(G/SEC)')
     WRITE (6,20) F, AV(9), CIN1, CIN2
  20 FORMAT (/T32, 'FEED', T53, F7.3, T63, F7.3, T72, F7.3,
    * T82, F7.3)
     WRITE (6,30) B, AV(8), CB1, CB2
  30 FORMAT (T32, 'BOTTOM PRODUCT', T53, F7.3, T63, F7.3,
    * T72, F7.3,
```

...(CONT'D)

```
     1   T82, F7.3)
      WRITE (6,40) D, AV(7), CT1, CT2
   40 FORMAT (T32, 'TOP PRODUCT', T53, F7.3, T63, F7.3, T72,
     * F7.3, T82,
     1   F7.3)
      WRITE (6,50) ERRT, ERR1, ERR2
   50 FORMAT (T32, 'CLOSURE ERROR-PC', T53, F5.1, T72, F5.1,
     * T82, F5.1)
      WRITE (6,60)
   60 FORMAT (////)
C       ENERGY BALANCE
C
      WRITE (6,70)
   70 FORMAT (T47, 'E N E R G Y   B A L A N C E'//T57,
     * 'ENTHALPY IN',
     1  T74, 'ENTHALPY OUT'/T59, '(J/SEC)', T76, '(J/SEC)')
      WRITE (6,80) HWI, HWO
   80 FORMAT (/T39, 'COOLING WATER', T59, F7.1, T76, F7.1)
      WRITE (6,90) HRI, HRO
   90 FORMAT (T39, 'REFLUX', T59, F7.1, T76, F7.1)
      WRITE (6,100) HTO
  100 FORMAT (T39, 'TOP PRODUCT', T76, F7.1)
      WRITE (6,110) HFI
  110 FORMAT (T39, 'FEED', T59, F7.1)
      WRITE (6,120) HSI, HSO
  120 FORMAT (T39, 'STEAM', T59, F7.1, T76, F7.1)
      WRITE (6,130) HBO
  130 FORMAT (T39, 'BOTTOM PRODUCT', T76, F7.1)
      WRITE (6,140) HIT, HOT
  140 FORMAT (T39, 'TOTAL', T59, F7.1, T76, F7.1)
      WRITE (6,150) HLOSS
  150 FORMAT (T39, 'HEAT LOSS', T68, F7.1)
C       DOCUMENT ALL STEADY STATE CONDITIONS
C
      WRITE (6,200)
  200 FORMAT ('1', //////)
      WRITE (6,210) KOUNT, (RUN(I), I=1,2), (DATE(I), I=1,2)
  210 FORMAT (T33, 'STEADY STATE CONDITIONS BASED ON', I4, '
     * POINTS'/
     1       T42, 'RUN NO ', 2A4, 1X, 2A4/)
      DO 230 I=1,13
      WRITE (6,220) (TITLE(I,J), J=1,3),AV(I),(UNIT(I,J),
     * J=1,2),DEV(I)
  220 FORMAT (T33, 3A4, '=', F7.3, 1X, 2A4, 3X, 'DEV=',
     * F7.4)
  230 CONTINUE
```

...(CONT'D)

```
    DO 250 I=14,34
    WRITE (6,240)  (TITLE(I,J), J=1,3),AV(I),(UNIT(I,J),
   * J=1,2),DEV(I)
240 FORMAT (T33, 3A4, '=', F7.1, 1X, 2A4, 3X, 'DEV=',
   * F7.4)
250 CONTINUE
    WRITE (6,260)
260 FORMAT ('1')
    CALL EXIT
    END
```

```
C
C                   PROGRAM INFF
C                   ------------
C
C
C      PURPOSE
C         READ INTO DISK STORAGE FILE 'US001' THE TITLES
C         WHICH WILL BE USED FOR PROGRAM BALN1 OUTPUT
C
C      DESCRIPTION OF PARAMETERS
C         TITLE    TITLES FOR BALN1 OUTPUT
C         UNIT     UNITS OF THE CORRESPONDING TITLES
C
C      USAGE
C         FOLLOWING DATA IS REQUIRED FOR THIS PROGRAM INFF
C              FEED FLOW   G/SEC
C              REFLUX FLOW G/SEC
C              STEAM FLOW  G/SEC
C              BOTTOM PROD G/SEC
C              TOP PROD    G/SEC
C              COOL WATER  G/SEC
C              DIST COMP   WT% MEOH
C              BOTTOM COMP WT% MEOH
C              FEED COMP   WT% MEOH
C              COL PRESS   KPA
C              COND LEVEL  KPA
C              REB'R.LEVEL KPA
C              DIFF PRESS  KPA
C              REB'R TEMP  DEG C
C              TRAY 1 TEMP DEG C
C              TRAY 2 TEMP DEG C
C              TRAY 3 TEMP DEG C
C              TRAY 4 TEMP DEG C
C              TRAY 5 TEMP DEG C
C              TRAY 6 TEMP DEG C
C              TRAY 7 TEMP DEG C
C              TRAY 8 TEMP DEG C
C              COND TEMP   DEG C
C              STEAM TEMP  DEG C
C              CON'TE TEMP DEG C
C              REFLUX TEMP DEG C
C              FEED TEMP   DEG C
C              BOTTOM TEMP DEG C
C              REB'R VAPOR DEG C
C              FEED INLET  DEG C
C              REFLUX INLETDEG C
C              COND VAPOR  DEG C
C              WATER INLET DEG C
C              WATER OUTLETDEG C
```

...(CONT'D)

```
C          (START ALL THE ABOVE DATA IN COLUMN 1)
C
C      REMARKS
C         DATA FILE 10 NAMED 'US001' IS REQUIRED
C
C

      INTEGER FDPLT
      DIMENSION TITLE(34,3), UNIT(34,2)
      DEFINE FILE 10(60,80,U,NEXT)
      NLOOP=34
      DO 10 I=1,NLOOP
      READ(5,100) (TITLE(I,J),J=1,3),(UNIT(I,J),J=1,2)
   10 CONTINUE
  100 FORMAT(5A4)
      KOUNT=0
      MAX=0
      IPOL=0
      BTCMP=0.0
      FDCMP=0.0
      FDPLT=0
      WRITE(10'1) NLOOP,KOUNT,MAX,IPOL,BTCMP,FDCMP,FDPLT
      WRITE(10'2) ((TITLE(I,J),J=1,3),I=1,12)
      WRITE(10'3) ((TITLE(I,J),J=1,3),I=13,24)
      WRITE(10'4) ((TITLE(I,J),J=1,3),I=25,34)
      WRITE(10'5) ((UNIT(I,J),J=1,2),I=1,18)
      WRITE(10'6) ((UNIT(I,J),J=1,2),I=19,34)
      CALL EXIT
      END
```

```
C
C                    PROGRAM DWL44
C                    -------------
C
C
C     PURPOSE
C        TO WRITE BOTTOM COMPOSITION TO DDC LOOP 608.
C        PROGRAM HAS OPTION OF ALSO WRITING FEED COMPOSITION
C        TO DDC LOOP 629.  AT PRESENT 629 IS BEING USED
C        AS A DATA ACCUMULATION LOOP FOR REBOILER LEVEL.
C
C     DESCRIPTION OF PARAMETERS
C        ITIM     ELUTION TIME FOR METHANOL AND WATER PEAKS
C        AREA     PEAK ARE FOR METHANOL AND WATER
C        D1       RESPONSE FACTOR FOR METHANOL
C        D2       RESPONSE FACTOR FOR WATER
C        C1       RESPONSE FACTOR FOR METHANOL
C        C2       RESPONSE FACTOR FOR WATER
C        Y        CURRENT BOTTOM OR FEED COMPOSITION
C
C     USAGE
C        QUEQUED BY ANY G.C. JOB ON G.C. #7 USING
C        CALCUTION OPTION #8.
C
C     SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C        GTVLU,GETCE,PTVLU,TIME
C
C

      REAL MEAS(3)
      DIMENSION ITIM(2),AREA(2),LPID(4)
      DATA ERROR/1000.0/
      DATA ICEAT,ICEAA/Z0017,Z001B/

C     REPONSE FACTORS FOR BOTTOM COMPOSITION
C

      DATA D1,D2/1.025,1.000/

C     REPONSE FACTORS FOR FEED COMPOSITION
C

      DATA C1,C2/1.000,1.000/

C     CODE FOR LOOP NUMBERS
C        1 -- RAW BOTTOM COMPOSITION
C        2 -- FILTERED BOTTOM COMPOSITION
C        3 -- TOP COMPOSITION
C        4 -- FEED COMPOSITION
```

...(CONT'D)

```
C
      DATA LPID(1),LPID(2),LPID(3),LPID(4)/Z0608,Z0607,Z0604
     *,Z0629/
C     GET ELUTION TIMES AND PEAKS AREAS FOR METHANOL AND
C     WATER
C
      CALL GETCE(ICEAT,ITIM,1,2)
      CALL GETCE(ICEAA,AREA,2,2)
C     CHECK WHICH COMPOSITION IS TO BE WRITTEN TO THE DDC
C     LOOP
C
      R=AREA(2)/AREA(1)
      IF(R-2.0) 1,1,2
    1 LPIN=LPID(4)
      GO TO 3
    2 LPIN=LPID(1)
C     GET OLD MEASUREMENT FOR BOTTOM CONCENTRATION
C
      IDESC=13
      CALL GTVLU(LPIN,1,YOLD,IER,IDESC)
      GO TO (3,14,15,16),IER
    3 CONTINUE
C     CHECKS FOR PEAKS NOT DETECTED
C
      IF(ITIM(1)) 18,18,4
    4 IF(ITIM(2)) 18,5,7
    5 IF(AREA(1)-ERROR) 18,18,6
    6 AREA(2)=0.0
    7 CONTINUE
      IF(LPIN-LPID(4)) 12,8,8
C     CALCULATE FEED COMPOSITION(PERCENT METHANOL)
C
    8 ATOT=C1*AREA(1)+C2*AREA(2)
      Y=C1*AREA(1)/ATOT*100
      IF(Y) 18,18,9
    9 CONTINUE
      CALL PTVLU(LPID(4),1,Y,IER)
      GO TO (11,14,15,16),IER
   11 GO TO 17
```

...(CONT'D)

```
C     CALCULATE BOTTOM COMPOSITION(PERCENT METHANOL)
C

   12 ATOT=D1*AREA(1)+D2*AREA(2)
      Y=D1*AREA(1)/ATOT*100
      IF(Y) 18,18,13
   13 CONTINUE
      CALL PTVLU(LPID(1),1,Y,IER)
      ITEM=1
      IDESC=13
      DO 20 I=1,3
      CALL GTVLU(LPID(I),ITEM,MEAS(I),IER,IDESC)
      GO TO (10,14,15,16),IER
   10 CONTINUE
   20 CONTINUE

C     WRITE DATA TO TELETYPE #13
C

      LUN=13
      CALL TIME(IHOUR,IMIN,ISEC)
      WRITF(LUN,201 ) IHOUR,IMIN,ISEC
  201 FORMAT(//,T5,'TIME IS  ',I2,'/',I2,'/',I2)
      WRITE(LUN,202) MEAS(1),MEAS(3)
  202 FORMAT(//,T5,'RAW BOTTOM COMPOSITION IS  ',F6.3,'
     * PERCENT',
     */,T5,'TOP COMPOSITION IS  ',F6.3,' PERCENT')
      CALL EXIT
   14 WRITE(LUN,203)
  203 FORMAT(/,T5,'ITEM NOT SPECIFIED IN PVR')
      CALL EXIT
   15 WRITE(LUN,204)
  204 FORMAT(/,T5,'PVR ID NOT FOUND IN TABLE')
      CALL EXIT
   16 WRITE(LUN,205)
  205 FORMAT(/,T5,'INVALID DATA IN CALL')
      CALL EXIT
   17 CONTINUE
      CALL EXIT
   18 CALL PTVLU(LPIN,1,-1.0,IER)
      CALL EXIT
      END
```

```
C
C                    PROGRAM DISTB
C                    -------------
C
C
C      PURPOSE
C         TO START AN OPEN OR CLOSE LOOP RUN.
C
C      DESCRIPTION OF PARAMETERS
C         LUN          LOGICAL UNIT NUMBER
C         IOCLD        TYPE OF DISTURBANCE
C
C      USAGE
C         QUEUED FROM ANY TELETYPE AND ENTER DATA AS
C         REQUIRED.
C
C      EXAMPLE
C         ODISTB
C
C              ENTER TYPE OF DISTURBANCE
C                   1 -- OPEN LOOP
C                   2 -- CLOSE LOOP
C
C         1
C              ENTER VARIABLE FOR OPEN LOOP DISTURBANCE
C
C                   1---FEED FLOW
C                   2---REFLUX FLOW
C                   3---STEAM FLOW
C
C         1
C              ENTER PRESENT VALUE AND
C              MAGNITUDE OF DISTURBANCE(+ OR- PERCENT)
C
C         19.035,-20.0
C
C
C              TIME OF DISTURBANCE IS  15/52/19
C
C      SUBROUTINES AND PROGRAMS REQUIRED
C         ODIST
C
C      SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C         FFINP,GETTY
C

      DEFINE FILE 10(60,80,U,NEXT)
      CALL GETTY(LUN)
      WRITE(LUN,200)
```

...(CONT'D)

```
 200 FORMAT(T5,'ENTER TYPE OF DISTURBANCE',
    */,T10,'1 -- OPEN LOOP',
    */,T10,'2 -- CLOSE LOOP')
   1 CALL FFINP(LUN,1,10,IOCLD,IER)
     IF(IER) 2,4,3
   2 WRITE(LUN,202)
 202 FORMAT(T5,'INPUT TO FFINP IN ERROR--PROGRAM EXITED')
     CALL EXIT
   3 WRITE(LUN,203)
 203 FORMAT(T5,'INPUT TO FFINP IN ERROR--TRY AGAIN')
     GO TO 1
   4 CONTINUE
     GO TO (5,6),IOCLD
   5 CONTINUE
     CALL ODIST(LUN)
     CALL EXIT
   6 CONTINUE
     WRITE(LUN,204)
 204 FORMAT(T5,'CLOSE LOOP --- NOT FINISHED')
     CALL EXIT
     END
```

```
C
C                    SUBROUTINE ODIST
C                    ----------------
C
C
C     PURPOSE
C        TO TO ENTER TYPE OF DISTURBANCE AND THE
C        MAGNITUDE.
C
C     DESCRIPTION OF PARAMETERS
C        PRVAL      PRESENT VALUE
C        DISMG      MAGNITUDE OF DISTURBANCE
C        SETPT      VALUE OF VARIABLE
C        OUTPT      OUTPUT(PCNT) SEND TO DDC LOOP
C
C     SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C        FFINP,PTVLU,TIME,OPER,MMANL
C
C


      SUBROUTINE ODIST(LUN)
      INTEGER LUN,IDIS,IER,IERR
      REAL PRVAL,DISMG,SETPT,OUTPT
      DIMENSION A(3),B(3),LPID(3)
      DATA LPID/Z0601,Z0602,Z0603/
      DATA A/2.4979,2.6385,3.9309/
      DATA B/-1.0961,-0.7962,1.9549/

C     MAKE ALL THREE LOOPS OPERABLE AND ON MANUAL
C

      DO 10 I=1,3
      CALL OPER(LPID(I))
      CALL MMANL(LPID(I))
   10 CONTINUE
      WRITE(LUN,200)
  200 FORMAT(T5,'ENTER VARIABLE FOR OPEN LOOP DISTURBANCE')
      WRITE(LUN,201)
  201 FORMAT(/,T10,'1---FEED FLOW',/,T10,'2---REFLUX FLOW'
     *,/,T10,'3---STEAM FLOW')
    1 CALL FFINP(LUN,1,10,IDIS,IER)
      IF(IER) 2,4,3
    2 WRITE(LUN,203)
  203 FORMAT(T5,'INPUT TO FFINP IN ERROR--PROGRAM EXITED')
      GO TO 15
    3 WRITE(LUN,204)
  204 FORMAT(T5,'INPUT TO FFINP IN ERROR--TRY AGAIN')
      GO TO 1
    4 CONTINUE
```

...(CONT'D)

```
      WRITE(LUN,205)
  205 FORMAT(T5,'ENTER PRESENT VALUE AND'
     1,/,T5,'MAGNITUDE OF DISTURBANCE(+ OR - PERCENT)')
    6 CALL FFINP(LUN,2,11,PRVAL,11,DISMG,IER)
      IF(IER) 7,9,8
    7 GO TO 2
    8 WRITE(LUN,206)
  206 FORMAT(T5,'INPUT TO FFINP IN ERROR--TRY AGAIN')
      GO TO 6

C     NEW VALUE CAN EITHER GO TO THE SETPOINT OR THE OUTPUT
C     OF THE LOOP.  IF TO THE SETPOINT THEN LOOP MUST BE ON
C     AUTOMATIC CONTROL AND IF TO THE OUTPUT LOOP MUST BE ON
C     MANUAL.
C

    9 SETPT=PRVAL*(1.0+DISMG/100.0)
      OUTPT=((SETPT-B(IDIS))/A(IDIS))**2.0

C     VALUE OF ITEM WILL BE
C           1 -- MEASUREMENT
C           2 -- SETPOINT
C           3 -- OUTPUT
C

      ITEM=3
      CALL PTVLU(LPID(IDIS),ITEM,OUTPT,IERR)
      GO TO (11,12,13,14),IERR
   11 CALL TIME(IHOUR,IMIN,ISEC)
      WRITE(LUN,207) IHOUR,IMIN,ISEC
  207 FORMAT(//,T5,'TIME OF DISTURBANCE IS  ',I2,'/',I2,'/'
     *,I2)
      GO TO 15
   12 WRITE(LUN,208)
  208 FORMAT(/,T5,'ITEM NOT SPECIFIED IN PVR')
      GO TO 15
   13 WRITE(LUN,209)
  209 FORMAT(/,T5,'PVR ID NOT FOUND IN TABLE')
      GO TO 15
   14 WRITE(LUN,211)
  211 FORMAT(/,T5,'INVALID DATA IN CALL')
   15 CONTINUE
      RETURN
      END
```

APPENDIX F

F.1 Liquid Mass Holdup

Accurate experimental liquid mass holdup was only necessary for the transient distillation column dynamics, that is, changing the holdup has no effect on the steady state. Since the column shell consists of glass sections between each of the trays, it was easy to measure the clear liquid height on each tray to help determine the liquid mass holdup.

F.2 Method of Measurement

The liquid mass holdup of each stage was calculated in the following manner. Before determining the liquid mass holdup on each stage for the operating conditions the maximum volume per stage was determined as follows. Using only the feed stream, liquid was introduced into the column on the four bottom trays and reboiler. The reboiler level was operated at the normal level with each tray overflowing with the height of the outlet weir being 2 inches. The system was allowed to reach steady state, then the feed pump was turned off and at the same time the bottom control valve was closed. The column was allowed to drain and the total amount of liquid weighed 12,665.2 grams at a specific gravity of 0.9507. Next, only the reboiler was filled to its operating point and drained. The liquid in the reboiler had a mass of 7328.4 grams and a specific gravity of 0.9561. With this information, the volume per tray was calculated to be 1403.4 cm^3 , based on the 4 bottom trays, which corresponds to two inches of liquid on each tray.

Next the column was operated at the normal operating conditions. Measuring the height of liquid on a tray was difficult because of foaming. If the feed, and reflux pumps and steam flow valve were closed at the same time, the foaming ceased. At this point, the height of the liquid on each tray could be measured and a sample of liquid on each tray was obtained. The samples were analyzed by measuring the specific gravity to determine the percent methanol for each tray liquid. Knowing the composition and the temperature of the liquid on each tray, the density at the operating conditions was determined by using equation A-1. Thus, using the height of liquid on a tray, the volume and mass holdup on each tray was determined. Table F.1 summarizes the experimental values. The condenser mass holdup was determined in a similar manner to the reboiler calculations.

Table F.1 Experimental Liquid Mass Holdup

Stage	Density 20 °C	Percent Comp	Temp °C	Density g/cm³	Height inch	Volume cm³	Holdup g
1	------	3.50	95.5	0.9519	-----	7664.4	7296.5
2	0.9842	8.31	86.7	0.9511	1.500	1052.6	1001.1
3	0.9619	24.20	80.2	0.9278	1.812	1271.8	1180.0
4	0.9325	41.10	75.6	0.8966	1.750	1227.9	1101.0
5	0.9094	53.00	72.6	0.8717	1.375	964.8	841.0
6	0.8755	68.30	70.8	0.8347	1.250	877.1	732.1
7	0.8472	79.90	66.4	0.8063	1.250	877.1	707.2
8	0.8272	87.50	64.9	0.7855	0.937	657.8	516.7
9	0.8131	92.50	62.8	0.7721	0.937	657.8	507.9
10	------	96.00	61.5	0.7623	-----	956.8	727.4

Appendix G

G.1 Experimental Data for Chapter 3

Appendix G contains all mass and energy balance reports for each experimental run presented in Chapter 3. For each of the variables to be disturbed, that is, steam, reflux, and feed flow rates, the initial and final steady states values are shown. Since the test were performed as a combination of increases and decreases in the variable of interest, the final steady state values for the response to an increase in flow were the initial steady state values for the decrease in flow rate. The following code is used:

ST - Steam Flow
RE - Reflux Flow
FE - Feed Flow
I - Initial
FI - Final-Initial
F - Final

A plot of tower pressure for each disturbed variable is shown respectively in Figures G.1, G.5, and G.7 for steam, reflux and feed flow. Figures G.2, G.6, and G.8 show the constancy of the other two input variables that are not being disturbed. Figure G.3 shows the steam and reboiler temperatures during a 7.5 % increase in steam flow. Figure G.4 shows the response of product flow rates for a step decrease in steam flow rate. The response of the top

```
                 STEADY STATE DATA
                  RUN NO STI -10%
                     15/05/78


FEED FLOW       18.018 G/SEC     BOTTOM PROD      9.042 G/SEC
REFLUX FLOW     12.100 G/SEC     TOP PROD         8.780 G/SEC
STEAM FLOW      14.535 G/SEC     COOL WATER     512.960 G/SEC
FEED PLATE       4               FEED COMP       50.120 WT% MEOH
DIST COMP       95.692 WT% MEOH  BOTTOM COMP      4.174 WT% MEOH
FEED INLET       63.80 DEG C     REFLUX INLET     52.00 DEG C
STEAM TEMP      106.00 DEG C     COL PRESS         0.02 KPA


           M A T E R I A L   B A L A N C E

                         FLOW      COMP    METHANOL    WATER
                       (G/SEC)  (WT PCT)  (G/SEC)    (G/SEC)

FEED                    18.018    50.120     9.030      8.987
BOTTOM PRODUCT           9.042     4.174     0.377      8.665
TOP PRODUCT              8.780    95.692     8.402      0.378
CLOSURE ERROR-PC        -1.0                -2.7        0.6


             E N E R G Y   B A L A N C E

                      ENTHALPY IN      ENTHALPY OUT
                        (J/SEC)          (J/SEC)

    COOLING WATER      28746.3          51881.5
    REFLUX              1822.9           2130.3
    TOP PRODUCT                          1545.8
    FEED                4119.3
    STEAM              39050.5           6418.3
    BOTTOM PRODUCT                       3510.4
    TOTAL              73739.2          65486.5
    HEAT LOSS                  8252.6
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO STI -10% 15/05/78

FEED FLOW     = 18.946 G/SEC        DEV= 0.0652
REFLUX FLOW   = 13.755 G/SEC        DEV= 0.0718
STEAM FLOW    = 14.535 G/SEC        DEV= 0.4371
BOTTOM PROD   =  9.092 G/SEC        DEV= 0.0575
TOP PROD      =  9.981 G/SEC        DEV= 0.0846
COOL WATER    =512.530 G/SEC        DEV= 1.9214
DIST COMP     = 95.692 WT% MEOH     DEV= 0.1398
BOTTOM COMP   =  4.174 WT% MEOH     DEV= 0.0000
FEED COMP     = 50.120 WT% MEOH     DEV= 0.0000
COL PRESS     =  0.029 KPA          DEV= 0.0509
COND LEVEL    = 33.860 KPA          DEV= 0.2014
REB'R LEVEL   = 94.313 KPA          DEV= 0.2420
DIFF PRESS    =103.420 KPA          DEV= 0.1868
REB'R TEMP    =    93.0 DEG C       DEV= 0.2685
TRAY 1 TEMP   =    82.9 DEG C       DEV= 0.2411
TRAY 2 TEMP   =    76.9 DEG C       DEV= 0.2594
TRAY 3 TEMP   =    73.1 DEG C       DEV= 0.5807
TRAY 4 TEMP   =    70.9 DEG C       DEV= 0.2417
TRAY 5 TEMP   =    69.0 DEG C       DEV= 0.2298
TRAY 6 TEMP   =    64.5 DEG C       DEV= 0.2327
TRAY 7 TEMP   =    63.0 DEG C       DEV= 0.2469
TRAY 8 TEMP   =    61.2 DEG C       DEV= 0.2522
COND TEMP     =    59.1 DEG C       DEV= 0.2377
STEAM TEMP    =   106.0 DEG C       DEV= 0.3345
CON'TE TEMP   =   104.6 DEG C       DEV= 0.2851
REFLUX TEMP   =    45.7 DEG C       DEV= 0.2551
FEED TEMP     =    33.9 DEG C       DEV= 0.2467
BOTTOM TEMP   =    42.4 DEG C       DEV= 0.2225
REB'R VAPOR   =    91.9 DEG C       DEV= 0.2796
FEED INLET    =    63.8 DEG C       DEV= 0.3644
REFLUX INLET=      52.0 DEG C       DEV= 0.8462
COND VAPOR    =    61.9 DEG C       DEV= 0.2512
WATER INLET =      13.4 DEG C       DEV= 0.2180
WATER OUTLET=      24.2 DEG C       DEV= 0.1917
```

```
                    STEADY STATE DATA
                     RUN NO STFI 10%
                       15/05/78


FEED FLOW       18.037 G/SEC     BOTTOM PROD     10.323 G/SEC
REFLUX FLOW     12.128 G/SEC     TOP PROD         7.467 G/SEC
STEAM FLOW      13.447 G/SEC     COOL WATER     511.895 G/SEC
FEED PLATE       4               FEED COMP       50.120 WT% MEOH
DIST COMP       96.401 WT% MEOH  BOTTOM COMP     14.076 WT% MEOH
FEED INLET       63.80 DEG C     REFLUX INLET     51.90 DEG C
STEAM TEMP       99.20 DEG C     COL PRESS         0.03 KPA



              M A T E R I A L   B A L A N C E

                            FLOW      COMP    METHANOL    WATER
                          (G/SEC)  (WT PCT)  (G/SEC)    (G/SEC)

FEED                       18.037    50.120     9.040      8.996
BOTTOM PRODUCT             10.323    14.076     1.453      8.870
TOP PRODUCT                 7.467    96.401     7.198      0.268
CLOSURE ERROR-PC           -1.3                -4.2        1.5



                 E N E R G Y   B A L A N C E

                         ENTHALPY IN      ENTHALPY OUT
                           (J/SEC)          (J/SEC)

      COOLING WATER        29755.8          51346.3
      REFLUX                1816.8           2128.7
      TOP PRODUCT                            1310.7
      FEED                  4123.5
      STEAM                35976.3           5495.1
      BOTTOM PRODUCT                         3620.3
      TOTAL                71672.6          63901.2
      HEAT LOSS                      7771.3
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
        RUN NO STFI 10% 15/05/78

FEED FLOW     = 18.966 G/SEC        DEV= 0.0720
REFLUX FLOW   = 13.803 G/SEC        DEV= 0.0496
STEAM FLOW    = 13.447 G/SEC        DEV= 0.4684
BOTTOM PROD   = 10.472 G/SEC        DEV= 0.0988
TOP PROD      =  8.499 G/SEC        DEV= 0.0676
COOL WATER    =511.502 G/SEC        DEV= 2.0166
DIST COMP     = 96.401 WT% MEOH     DEV= 0.1538
BOTTOM COMP   = 14.076 WT% MEOH     DEV= 0.0000
FEED COMP     = 50.120 WT% MEOH     DEV= 0.0000
COL PRESS     =  0.037 KPA          DEV= 0.1236
COND LEVEL    = 33.960 KPA          DEV= 0.2295
REB'R LEVEL   = 94.028 KPA          DEV= 0.4705
DIFF PRESS    =103.420 KPA          DEV= 0.1868
REB'R TEMP    =    86.1 DEG C       DEV= 0.3242
TRAY 1 TEMP   =    76.0 DEG C       DEV= 0.2692
TRAY 2 TEMP   =    72.8 DEG C       DEV= 0.2510
TRAY 3 TEMP   =    70.6 DEG C       DEV= 0.3474
TRAY 4 TEMP   =    70.4 DEG C       DEV= 0.2427
TRAY 5 TEMP   =    68.5 DEG C       DEV= 0.2547
TRAY 6 TEMP   =    64.1 DEG C       DEV= 0.2390
TRAY 7 TEMP   =    62.7 DEG C       DEV= 0.2620
TRAY 8 TEMP   =    61.1 DEG C       DEV= 0.2654
COND TEMP     =    59.1 DEG C       DEV= 0.2554
STEAM TEMP    =    99.2 DEG C       DEV= 0.6269
CON'TE TEMP   =    97.0 DEG C       DEV= 0.2810
REFLUX TEMP   =    45.3 DEG C       DEV= 0.2320
FEED TEMP     =    34.0 DEG C       DEV= 0.2408
BOTTOM TEMP   =    43.6 DEG C       DEV= 0.2404
REB'R VAPOR   =    85.1 DEG C       DEV= 0.2386
FEED INLET    =    63.8 DEG C       DEV= 0.3345
REFLUX INLET=      51.9 DEG C       DEV= 0.8666
COND VAPOR    =    61.8 DEG C       DEV= 0.2639
WATER INLET   =    13.9 DEG C       DEV= 0.2547
WATER OUTLET=      24.0 DEG C       DEV= 0.2132
```

```
                    STEADY STATE DATA
                     RUN NO STF +10%
                        15/05/78


FEED FLOW       18.035 G/SEC     BOTTOM PROD       9.193 G/SEC
REFLUX FLOW     12.136 G/SEC     TOP PROD          8.731 G/SEC
STEAM FLOW      14.578 G/SEC     COOL WATER      512.675 G/SEC
FEED PLATE       4               FEED COMP        50.120 WT% MEOH
DIST COMP       95.676 WT% MEOH  BOTTOM COMP       4.180 WT% MEOH
FEED INLET       64.10 DEG C     REFLUX INLET      52.30 DEG C
STEAM TEMP      106.40 DEG C     COL PRESS          0.04 KPA



               M A T E R I A L   B A L A N C E

                             FLOW      COMP    METHANOL     WATER
                           (G/SEC)  (WT PCT)  (G/SEC)     (G/SEC)

FEED                        18.035    50.120     9.039       8.996
BOTTOM PRODUCT               9.193     4.180     0.384       8.809
TOP PRODUCT                  8.731    95.676     8.354       0.377
CLOSURE ERROR-PC            -0.6                -3.3         2.1



                 E N E R G Y   B A L A N C E

                         ENTHALPY IN        ENTHALPY OUT
                          (J/SEC)             (J/SEC)

     COOLING WATER        30443.7             53565.4
     REFLUX                1841.1              2154.6
     TOP PRODUCT                               1550.1
     FEED                  4144.9
     STEAM                39175.6              6462.7
     BOTTOM PRODUCT                            3588.9
     TOTAL                75605.4             67321.8
     HEAT LOSS                        8283.6
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
        RUN NO STF +10% 15/05/78

FEED FLOW     = 18.966 G/SEC          DEV= 0.0703
REFLUX FLOW   = 13.799 G/SEC          DEV= 0.0428
STEAM FLOW    = 14.578 G/SEC          DEV= 0.5811
BOTTOM PROD   =  9.246 G/SEC          DEV= 0.0560
TOP PROD      =  9.928 G/SEC          DEV= 0.0978
COOL WATER    =512.304 G/SEC          DEV= 2.1181
DIST COMP     = 95.676 WT% MEOH       DEV= 0.1365
BOTTOM COMP   =  4.180 WT% MEOH       DEV= 0.0000
FEED COMP     = 50.120 WT% MEOH       DEV= 0.0000
COL PRESS     =  0.046 KPA            DEV= 0.0806
COND LEVEL    = 34.163 KPA            DEV= 0.2076
REB'R LEVEL   = 94.105 KPA            DEV= 0.2239
DIFF PRESS    =103.420 KPA            DEV= 0.1868
REB'R TEMP    =   93.5 DEG C          DEV= 0.3213
TRAY 1 TEMP   =   83.3 DEG C          DEV= 0.2663
TRAY 2 TEMP   =   77.3 DEG C          DEV= 0.3059
TRAY 3 TEMP   =   73.5 DEG C          DEV= 0.5131
TRAY 4 TEMP   =   71.2 DEG C          DEV= 0.2594
TRAY 5 TEMP   =   69.3 DEG C          DEV= 0.2620
TRAY 6 TEMP   =   64.9 DEG C          DEV= 0.2505
TRAY 7 TEMP   =   63.4 DEG C          DEV= 0.2525
TRAY 8 TEMP   =   61.5 DEG C          DEV= 0.2666
COND TEMP     =   59.5 DEG C          DEV= 0.2499
STEAM TEMP    =  106.4 DEG C          DEV= 0.3226
CON'TE TEMP   =  105.0 DEG C          DEV= 0.2670
REFLUX TEMP   =   46.4 DEG C          DEV= 0.2363
FEED TEMP     =   34.2 DEG C          DEV= 0.2244
BOTTOM TEMP   =   43.1 DEG C          DEV= 0.2575
REB'R VAPOR   =   92.3 DEG C          DEV= 0.2670
FEED INLET    =   64.1 DEG C          DEV= 0.3449
REFLUX INLET=     52.3 DEG C          DEV= 0.7427
COND VAPOR    =   62.2 DEG C          DEV= 0.2575
WATER INLET =     14.2 DEG C          DEV= 0.2180
WATER OUTLET=     25.0 DEG C          DEV= 0.2685
```

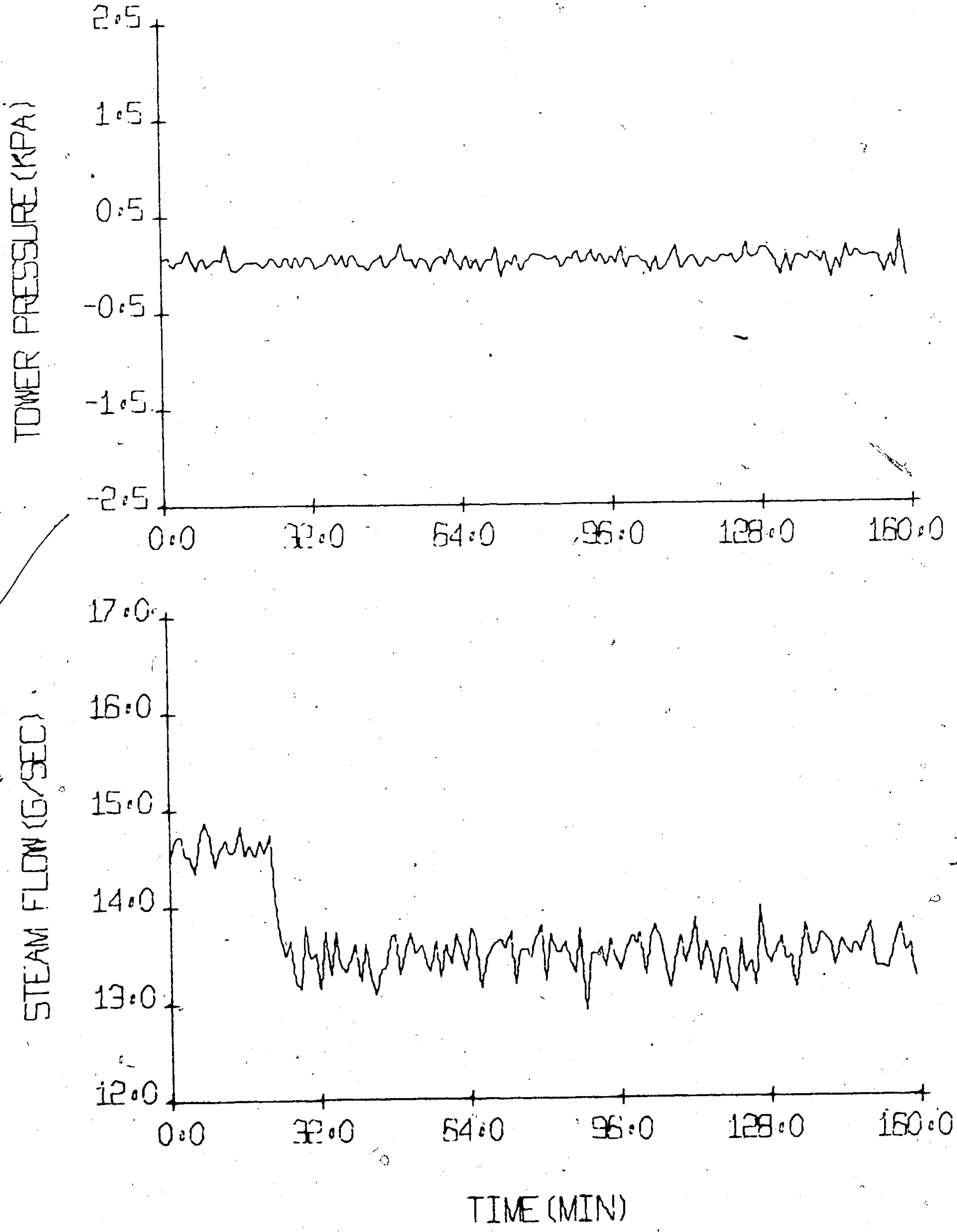

Figure G.1 Response of tower pressure and steam flow during a 7.5% decrease in steam flow.

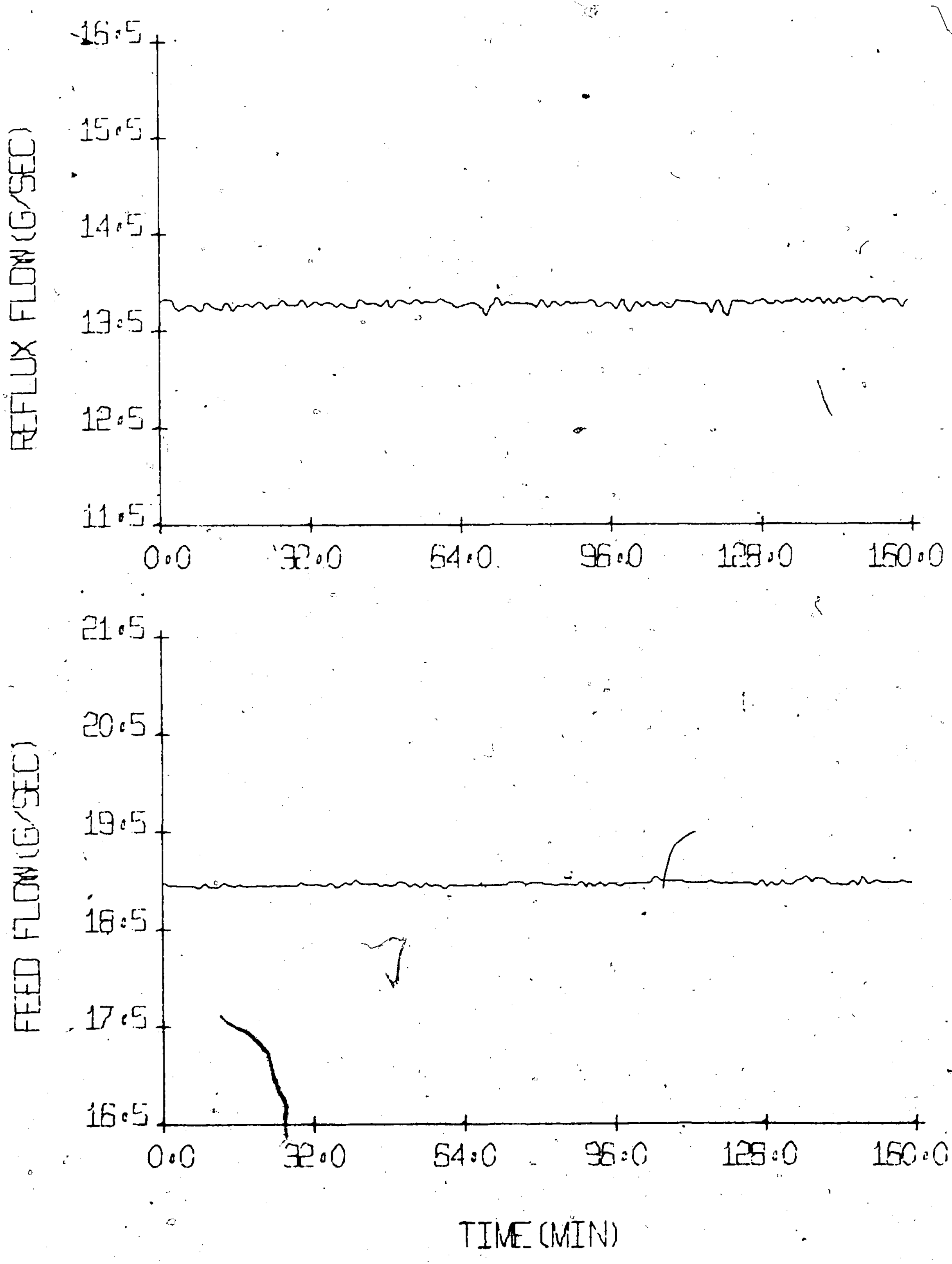

Figure G.2 Response of reflux and feed flow during a 7.5% decrease in steam flow.

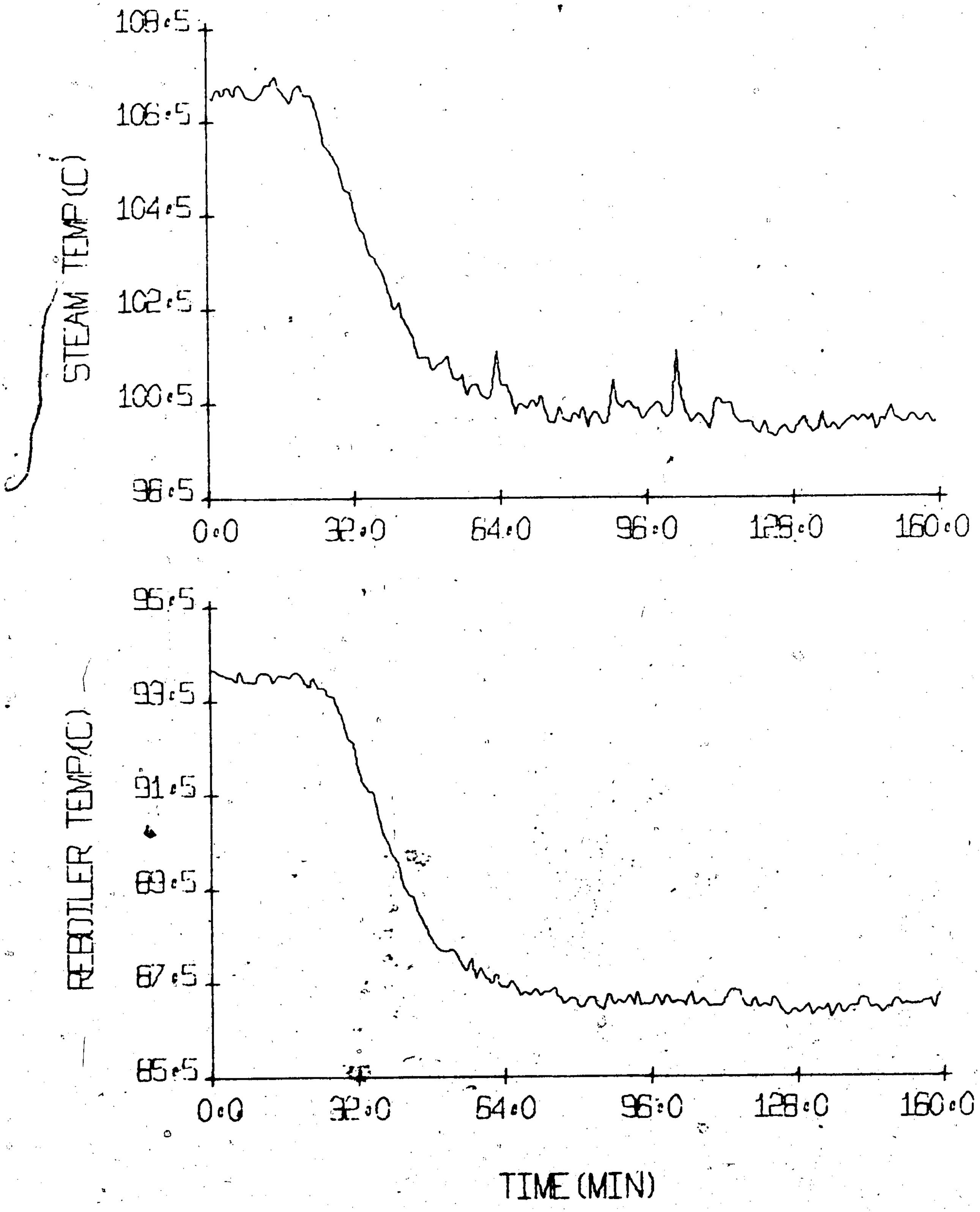

Figure G.3 Response of steam and reboiler temperatures during a 7.5% decrease in steam flow.

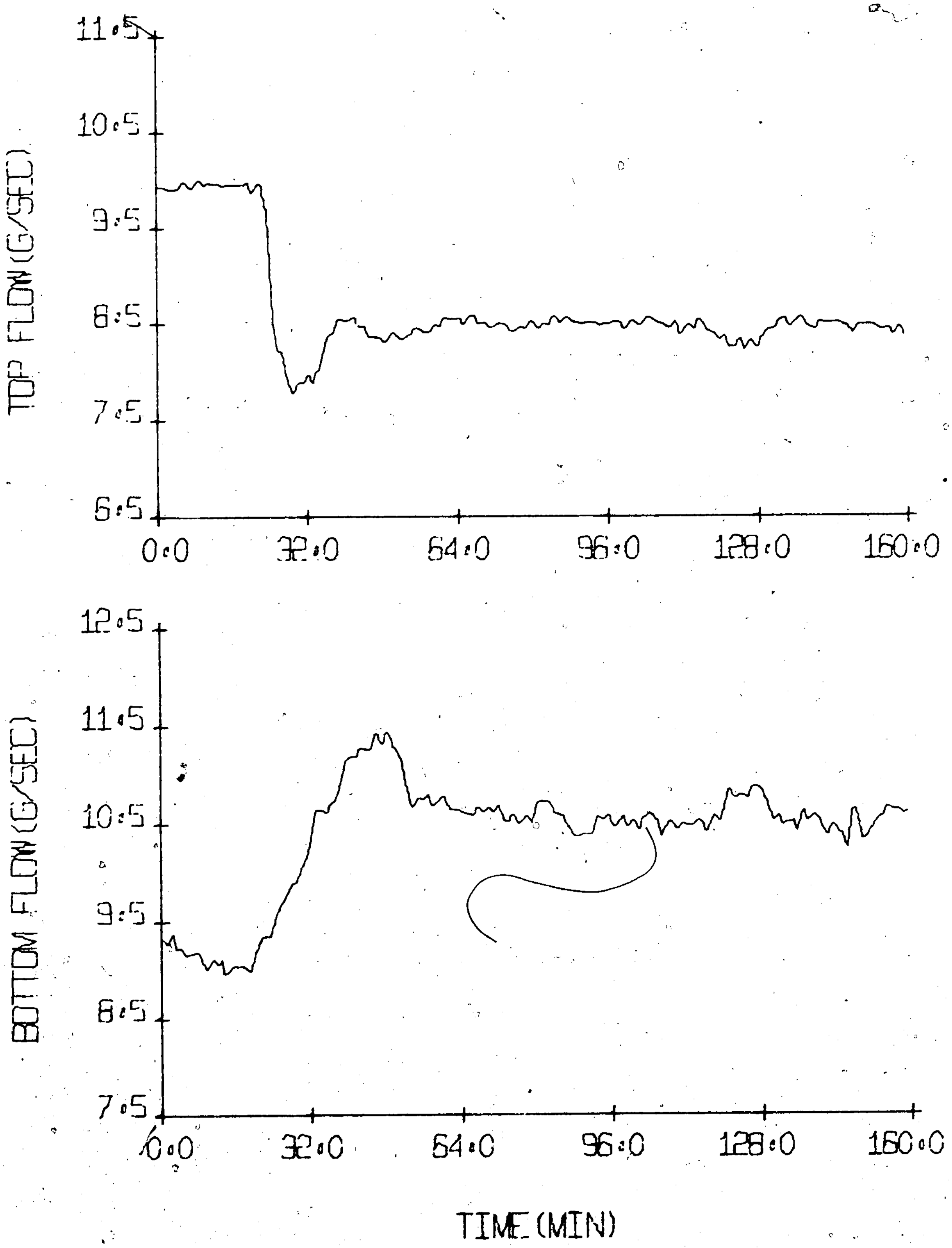

Figure G.4 Response of top and bottom product flows during a 7.5% decrease in steam flow.

```
                    STEADY STATE DATA
                     RUN NO REI +15%
                        15/05/78

FEED FLOW      18.094 G/SEC     BOTTOM PROD      9.231 G/SEC
REFLUX FLOW    12.103 G/SEC     TOP PROD         8.712 G/SEC
STEAM FLOW     14.556 G/SEC     COOL WATER     502.599 G/SEC
FEED PLATE      4               FEED COMP       50.120 WT% MEOH
DIST COMP      96.006 WT% MEOH  BOTTOM COMP      4.399 WT% MEOH
FEED INLET      65.00 DEG C     REFLUX INLET     53.60 DEG C
STEAM TEMP     107.10 DEG C     COL PRESS         0.01 KPA


              M A T E R I A L   B A L A N C E

                         FLOW      COMP     METHANOL    WATER
                       (G/SEC)  (WT PCT)   (G/SEC)    (G/SEC)

FEED                   18.094    50.120     9.069      9.025
BOTTOM PRODUCT          9.231     4.399     0.406      8.825
TOP PRODUCT             8.712    96.006     8.364      0.347
CLOSURE ERROR-PC       -0.8                -3.2        1.6


                E N E R G Y   B A L A N C E

                        ENTHALPY IN     ENTHALPY OUT
                          (J/SEC)          (J/SEC)

     COOLING WATER        31314.9          54401.4
     REFLUX                1888.5           2190.6
     TOP PRODUCT                            1576.8
     FEED                  4223.8
     STEAM                39133.1           6490.9
     BOTTOM PRODUCT                         3626.0
     TOTAL                76560.5          68286.0
     HEAT LOSS                     8274.5
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO REI +15% 15/05/78

FEED FLOW    = 19.033 G/SEC        DEV= 0.0650
REFLUX FLOW  = 13.774 G/SEC        DEV= 0.0361
STEAM FLOW   = 14.556 G/SEC        DEV= 0.5517
BOTTOM PROD  =  9.285 G/SEC        DEV= 0.0462
TOP PROD     =  9.915 G/SEC        DEV= 0.0978
COOL WATER   =502.285 G/SEC        DEV= 2.5156
DIST COMP    = 96.006 WT% MEOH     DEV= 0.1348
BOTTOM COMP  =  4.399 WT% MEOH     DEV= 0.0000
FEED COMP    = 50.120 WT% MEOH     DEV= 0.0000
COL PRESS    =  0.018 KPA          DEV= 0.0819
COND LEVEL   = 34.046 KPA          DEV= 0.1897
REB'R LEVEL  = 93.906 KPA          DEV= 0.2606
DIFF PRESS   =103.420 KPA          DEV= 0.1868
REB'R TEMP   =    94.1 DEG C       DEV= 0.2532
TRAY 1 TEMP  =    83.8 DEG C       DEV= 0.4217
TRAY 2 TEMP  =    78.0 DEG C       DEV= 0.2586
TRAY 3 TEMP  =    74.1 DEG C       DEV= 0.5075
TRAY 4 TEMP  =    72.1 DEG C       DEV= 0.2285
TRAY 5 TEMP  =    70.2 DEG C       DEV= 0.2317
TRAY 6 TEMP  =    65.8 DEG C       DEV= 0.2383
TRAY 7 TEMP  =    64.3 DEG C       DEV= 0.2436
TRAY 8 TEMP  =    62.5 DEG C       DEV= 0.2157
COND TEMP    =    60.5 DEG C       DEV= 0.2247
STEAM TEMP   =   107.1 DEG C       DEV= 0.3265
CON'TE TEMP  =   105.6 DEG C       DEV= 0.2988
REFLUX TEMP  =    46.9 DEG C       DEV= 0.3031
FEED TEMP    =    35.0 DEG C       DEV= 0.2613
BOTTOM TEMP  =    42.9 DEG C       DEV= 0.1993
REB'R VAPOR  =    93.0 DEG C       DEV= 0.2494
FEED INLET   =    65.0 DEG C       DEV= 0.3030
REFLUX INLET=     53.6 DEG C       DEV= 0.6334
COND VAPOR   =    63.2 DEG C       DEV= 0.2300
WATER INLET  =    14.9 DEG C       DEV= 0.2238
WATER OUTLET=     25.9 DEG C       DEV= 0.2484
```

```
                         STEADY STATE DATA
                          RUN NO REFI 15%
                             15/05/78

FEED FLOW      18.032 G/SEC     BOTTOM PROD    10.171 G/SEC
REFLUX FLOW    13.882 G/SEC     TOP PROD        7.737 G/SEC
STEAM FLOW     14.604 G/SEC     COOL WATER    502.522 G/SEC
FEED PLATE      4               FEED COMP      50.120 WT% MEOH
DIST COMP      96.888 WT% MEOH  BOTTOM COMP    11.834 WT% MEOH
FEED INLET      64.60 DEG C     REFLUX INLET    53.00 DEG C
STEAM TEMP     102.00 DEG C     COL PRESS        0.02 KPA


                 M A T E R I A L   B A L A N C E

                            FLOW       COMP    METHANOL    WATER
                          (G/SEC)   (WT PCT)  (G/SEC)    (G/SEC)

FEED                       18.032    50.120     9.037      8.994
BOTTOM PRODUCT             10.171    11.834     1.203      8.968
TOP PRODUCT                 7.737    96.888     7.496      0.240
CLOSURE ERROR-PC           -0.6                -3.7        2.3


                    E N E R G Y   B A L A N C E

                            ENTHALPY IN       ENTHALPY OUT
                             (J/SEC)            (J/SEC)

       COOLING WATER         29001.1            52924.2
       REFLUX                 2128.1             2472.4
       TOP PRODUCT                               1378.0
       FEED                   4180.3
       STEAM                 39139.7             6151.0
       BOTTOM PRODUCT                            3681.5
       TOTAL                 74449.3            66607.3
       HEAT LOSS                       7842.0
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO REFI 15% 15/05/78

FEED FLOW    = 18.967 G/SEC        DEV= 0.0670
REFLUX FLOW  = 15.822 G/ SEC       DEV= 0.0645
STEAM FLOW   = 14.604 G/SEC        DEV= 0.4946
BOTTOM PROD  = 10.300 G/ SEC       DEV= 0.0792
TOP PROD     =  8.819 G/SEC        DEV= 0.0600
COOL WATER   =502.129 G/ SEC       DEV= 8.0620
DIST COMP    = 96.888 WT% MEOH     DEV= 0.1359
BOTTOM COMP  = 11.834 WT% MEOH     DEV= 0.0000
FEED COMP    = 50.120 WT% MEOH     DEV= 0.0000
COL PRESS    =  0.020 KPA          DEV= 0.0464
COND LEVEL   = 33.896 KPA          DEV= 0.1968
REB'R LEVEL  = 92.750 KPA          DEV= 0.3022
DIFF PRESS   =103.420 KPA          DEV= 0.1868
REB'R TEMP   =    88.3 DEG C       DEV= 0.2339
TRAY 1 TEMP  =    77.7 DEG C       DEV= 0.2550
TRAY 2 TEMP  =    73.9 DEG C       DEV= 0.2695
TRAY 3 TEMP  =    71.4 DEG C       DEV= 0.2389
TRAY 4 TEMP  =    70.8 DEG C       DEV= 0.2318
TRAY 5 TEMP  =    68.9 DEG C       DEV= 0.2472
TRAY 6 TEMP  =    64.5 DEG C       DEV= 0.2299
TRAY 7 TEMP  =    63.2 DEG C       DEV= 0.2498
TRAY 8 TEMP  =    61.6 DEG C       DEV= 0.2466
COND TEMP    =    59.9 DEG C       DEV= 0.2388
STEAM TEMP   =   102.0 DEG C       DEV= 0.2819
CON'TE TEMP  =    99.9 DEG C       DEV= 0.2609
REFLUX TEMP  =    46.4 DEG C       DEV= 0.2407
FEED TEMP    =    34.9 DEG C       DEV= 0.2479
BOTTOM TEMP  =    44.4 DEG C       DEV= 0.2009
REB'R VAPOR  =    87.3 DEG C       DEV= 0.2775
FEED INLET   =    64.6 DEG C       DEV= 0.3224
REFLUX INLET=     53.0 DEG C       DEV= 0.6985
COND VAPOR   =    62.2 DEG C       DEV= 0.2549
WATER INLET  =    13.8 DEG C       DEV= 0.2432
WATER OUTLET=     25.2 DEG C       DEV= 0.2852
```

```
                    STEADY STATE DATA
                    RUN NO REF -15%
                       15/05/78

FEED FLOW      18.040 G/SEC     BOTTOM PROD      9.182 G/SEC
REFLUX FLOW    12.070 G/SEC     TOP PROD         8.751 G/SEC
STEAM FLOW     14.572 G/SEC     COOL WATER     511.354 G/SEC
FEED PLATE      4               FEED COMP       50.120 WT% MEOH
DIST COMP      95.963 WT% MEOH  BOTTOM COMP      4.316 WT% MEOH
FEED INLET      64.30 DEG C     REFLUX INLET     52.70 DEG C
STEAM TEMP     106.40 DEG C     COL PRESS         0.02 KPA


               M A T E R I A L   B A L A N C E

                         FLOW      COMP    METHANOL    WATER
                       (G/SEC)  (WT PCT)  (G/SEC)    (G/SEC)

FEED                    18.040    50.120     9.041      8.998
BOTTOM PRODUCT           9.182     4.316     0.396      8.786
TOP PRODUCT              8.751    95.963     8.397      0.353
CLOSURE ERROR-PC        -0.5                -2.7        1.5


                 E N E R G Y   B A L A N C E

                     ENTHALPY IN      ENTHALPY OUT
                       (J/SEC)          (J/SEC)

    COOLING WATER      28656.3          51719.2
    REFLUX              1845.6           2149.2
    TOP PRODUCT                          1558.1
    FEED                4160.5
    STEAM              39159.4           6460.0
    BOTTOM PRODUCT                       3579.6
    TOTAL              73822.1          65466.4
    HEAT LOSS                   8355.6
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO REF -15% 15/05/78

FEED FLOW    = 18.973 G/SEC       DEV= 0.0614
REFLUX FLOW  = 13.730 G/SEC       DEV= 0.0446
STEAM FLOW   = 14.572 G/SEC       DEV= 0.4049
BOTTOM PROD  =  9.235 G/SEC       DEV= 0.0555
TOP PROD     =  9.954 G/SEC       DEV= 0.0615
COOL WATER   =510.926 G/SEC       DEV= 1.9450
DIST COMP    = 95.963 WT% MEOH    DEV= 0.1553
BOTTOM COMP  =  4.316 WT% MEOH    DEV= 0.0000
FEED COMP    = 50.120 WT% MEOH    DEV= 0.0000
COL PRESS    =  0.027 KPA         DEV= 0.0885
COND LEVEL   = 34.230 KPA         DEV= 0.1829
REB'R LEVEL  = 93.843 KPA         DEV= 0.2627
DIFF PRESS   =103.420 KPA         DEV= 0.1868
REB'R TEMP   =    93.4 DEG C      DEV= 0.2516
TRAY 1 TEMP  =    83.3 DEG C      DEV= 0.2811
TRAY 2 TEMP  =    77.3 DEG C      DEV= 0.2261
TRAY 3 TEMP  =    73.5 DEG C      DEV= 0.2333
TRAY 4 TEMP  =    71.4 DEG C      DEV= 0.2474
TRAY 5 TEMP  =    69.5 DEG C      DEV= 0.2462
TRAY 6 TEMP  =    65.1 DEG C      DEV= 0.2534
TRAY 7 TEMP  =    63.6 DEG C      DEV= 0.2677
TRAY 8 TEMP  =    61.7 DEG C      DEV= 0.2528
COND TEMP    =    59.7 DEG C      DEV= 0.2689
STEAM TEMP   =   106.4 DEG C      DEV= 0.3245
CON'TE TEMP  =   105.0 DEG C      DEV= 0.2734
REFLUX TEMP  =    46.0 DEG C      DEV= 0.2458
FEED TEMP    =    34.5 DEG C      DEV= 0.2399
BOTTOM TEMP  =    42.8 DEG C      DEV= 0.2463
REB'R VAPOR  =    92.3 DEG C      DEV= 0.2683
FEED INLET   =    64.3 DEG C      DEV= 0.3596
REFLUX INLET=     52.7 DEG C      DEV= 0.8071
COND VAPOR   =    62.4 DEG C      DEV= 0.2275
WATER INLET  =    13.4 DEG C      DEV= 0.2612
WATER OUTLET=     24.2 DEG C      DEV= 0.1620
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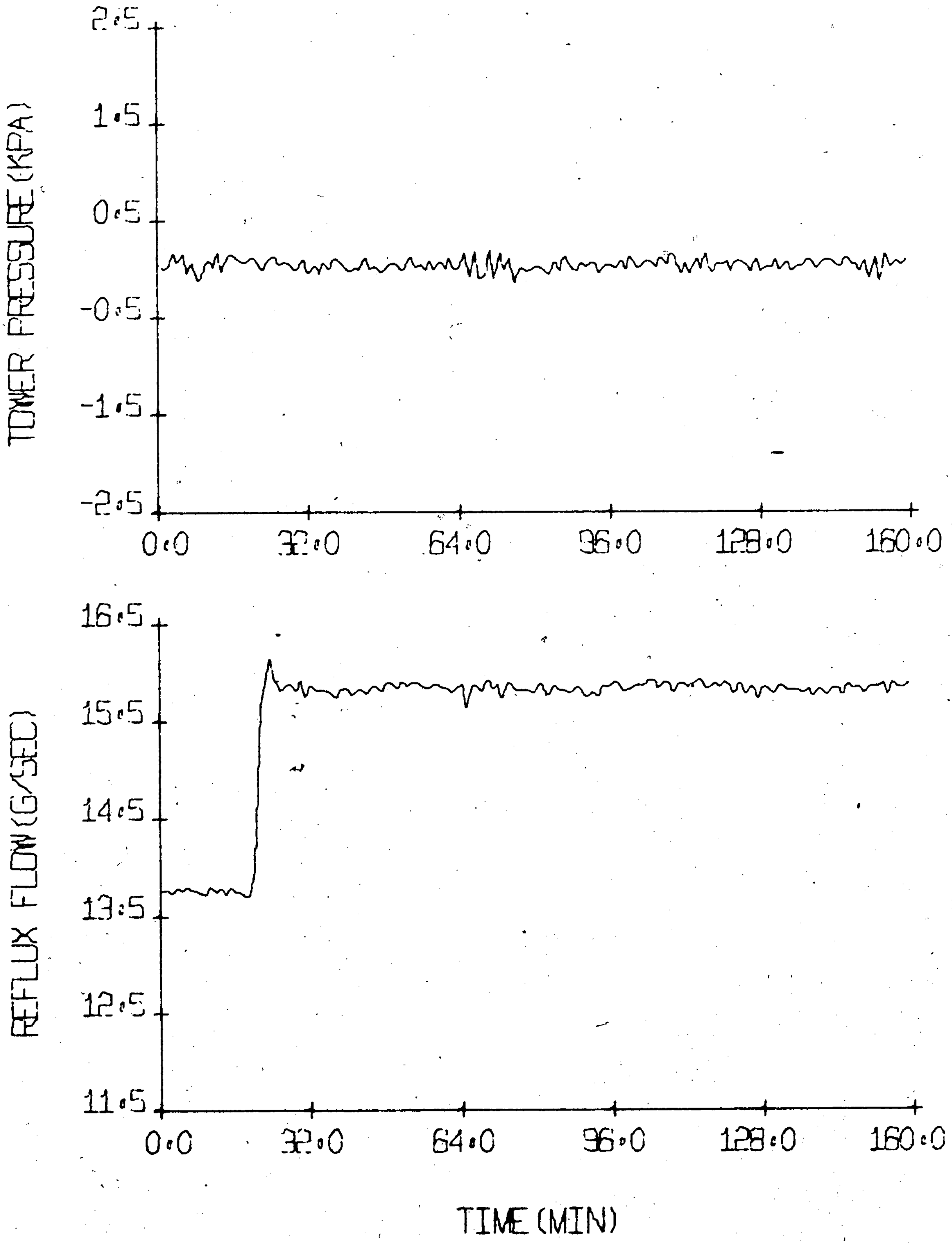

Figure G.5 Response of tower pressure and reflux flow during a 15% increase in reflux flow.

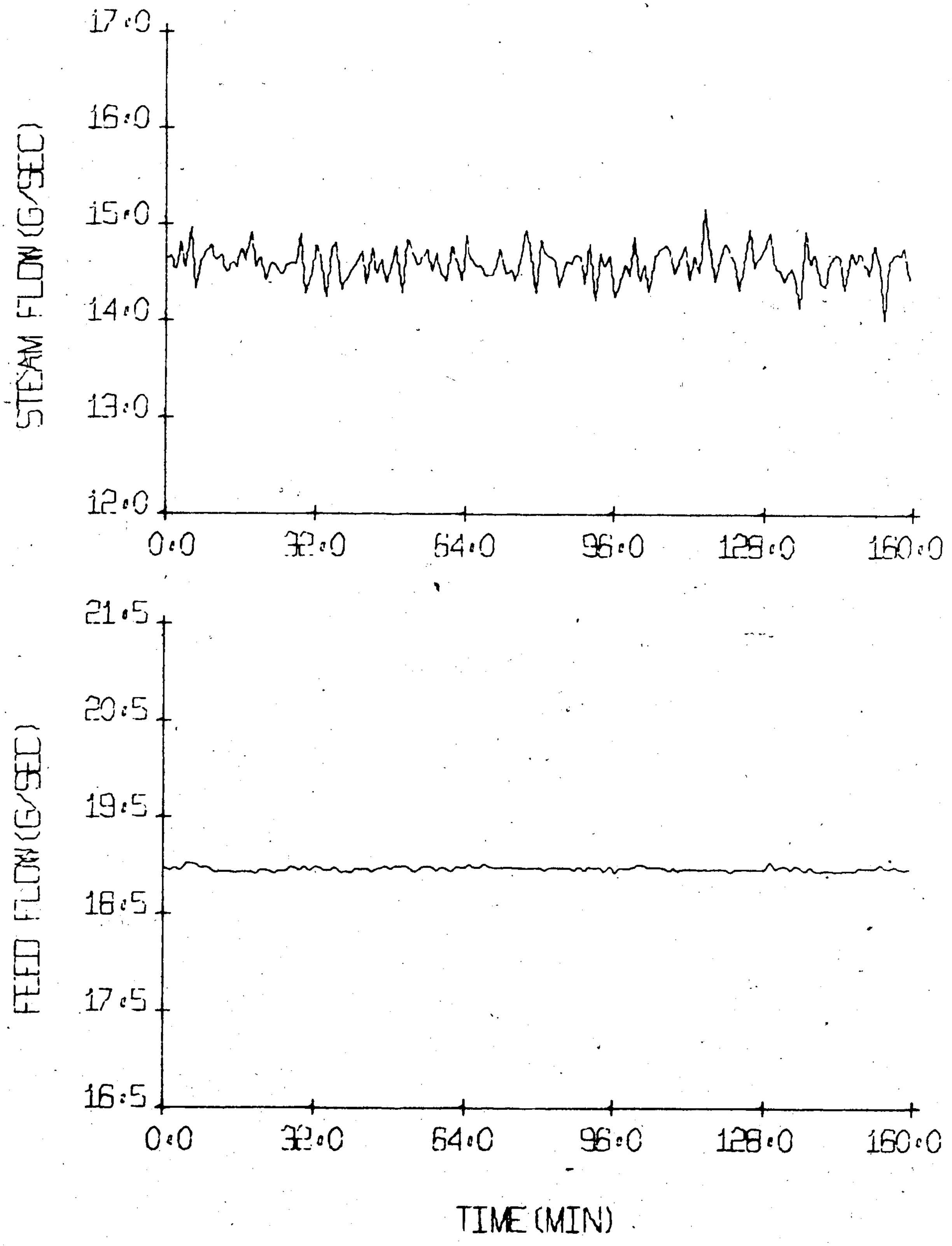

Figure G.6 Response of steam and feed flow during a 15% increase in reflux flow.

```
                        STEADY STATE DATA
                         RUN NO FEI +20%
                            16/05/78

FEED FLOW       18.034 G/SEC       BOTTOM PROD       8.481 G/SEC
REFLUX FLOW     12.154 G/SEC       TOP PROD          9.156 G/SEC
STEAM FLOW      14.708 G/SEC       COOL WATER      501.470 G/SEC
FEED PLATE       4                 FEED COMP        51.615 WT% MEOH
DIST COMP       95.831 WT% MEOH    BOTTOM COMP       4.671 WT% MEOH
FEED INLET       63.90 DEG C       REFLUX INLET      52.40 DEG C
STEAM TEMP      106.30 DEG C       COL PRESS          0.00-KPA


               M A T E R I A L   B A L A N C E

                           FLOW       COMP     METHANOL     WATER
                         (G/SEC)    (WT PCT)   (G/SEC)     (G/SEC)

FEED                      18.034     51.615      9.308       8.726
BOTTOM PRODUCT             8.481      4.671      0.396       8.085
TOP PRODUCT                9.156     95.831      8.774       0.381
CLOSURE ERROR-PC          -2.2                  -1.4        -2.9


                 E N E R G Y   B A L A N C E

                        ENTHALPY IN       ENTHALPY OUT
                          (J/SEC)           (J/SEC)

     COOLING WATER        29987.7           53651.0
     REFLUX                1846.7            2160.8
     TOP PRODUCT                             1627.8
     FEED                  4109.5
     STEAM                39522.5            6507.5
     BOTTOM PRODUCT                          3289.3
     TOTAL                75466.6           67236.5
     HEAT LOSS                      8230.0
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO FEI +20% 16/05/78

FEED FLOW     = 18.999 G/SEC        DEV= 0.0652
REFLUX FLOW   = 13.824 G/SEC        DEV= 0.0463
STEAM FLOW    = 14.708 G/SEC        DEV= 0.4927
BOTTOM PROD   =  8.531 G/SEC        DEV= 0.0535
TOP PROD      = 10.414 G/SEC        DEV= 0.0904
COOL WATER    =501.114 G/SEC        DEV= 2.5110
DIST COMP     = 95.831 WT% MEOH     DEV= 0.1220
BOTTOM COMP   =  4.671 WT% MEOH     DEV= 0.0000
FEED COMP     = 51.615 WT% MEOH     DEV= 0.0000
COL PRESS     =  0.003 KPA          DEV= 0.0856
COND LEVEL    = 33.721 KPA          DEV= 0.2185
REB'R LEVEL   = 92.518 KPA          DEV= 0.3476
DIFF PRESS    =103.420 KPA          DEV= 0.1868
REB'R TEMP    =   93.0 DEG C        DEV= 0.3078
TRAY 1 TEMP   =   82.5 DEG C        DEV= 0.2750
TRAY 2 TEMP   =   76.5 DEG C        DEV= 0.2793
TRAY 3 TEMP   =   72.7 DEG C        DEV= 0.3876
TRAY 4 TEMP   =   70.8 DEG C        DEV= 0.2561
TRAY 5 TEMP   =   69.0 DEG C        DEV= 0.2470
TRAY 6 TEMP   =   64.7 DEG C        DEV= 0.2622
TRAY 7 TEMP   =   63.3 DEG C        DEV= 0.2464
TRAY 8 TEMP   =   61.4 DEG C        DEV= 0.2464
COND TEMP     =   59.6 DEG C        DEV= 0.2803
STEAM TEMP    =  106.3 DEG C        DEV= 0.3338
CON'TE TEMP   =  104.8 DEG C        DEV= 0.2680
REFLUX TEMP   =   46.5 DEG C        DEV= 0.2748
FEED TEMP     =   33.8 DEG C        DEV= 0.2513
BOTTOM TEMP   =   42.4 DEG C        DEV= 0.2221
REB'R VAPOR   =   91.9 DEG C        DEV= 0.2629
FEED INLET    =   63.9 DEG C        DEV= 0.2951
REFLUX INLET=     52.4 DEG C        DEV= 0.7997
COND VAPOR    =   62.1 DEG C        DEV= 0.2725
WATER INLET   =   14.3 DEG C        DEV= 0.2479
WATER OUTLET=     25.6 DEG C        DEV= 0.2406
```

```
                    STEADY STATE DATA
                     RUN NO FEFI 20%
                        16/05/78

FEED FLOW       21.535 G/SEC     BOTTOM PROD     11.826 G/SEC
REFLUX FLOW     12.139 G/SEC     TOP PROD         9.414 G/SEC
STEAM FLOW      14.644 G/SEC     COOL WATER     503.330 G/SEC
FEED PLATE       4               FEED COMP       51.615 WT% MEOH
DIST COMP       95.918 WT% MEOH  BOTTOM COMP     14.091 WT% MEOH
FEED INLET       63.60 DEG C     REFLUX INLET     52.40 DEG C
STEAM TEMP      100.40 DEG C     COL PRESS         0.04 KPA


             M A T E R I A L   B A L A N C E

                          FLOW      COMP     METHANOL    WATER
                        (G/SEC)  (WT PCT)   (G/SEC)    (G/SEC)

FEED                     21.535    51.615    11.115     10.419
BOTTOM PRODUCT           11.826    14.091     1.666     10.159
TOP PRODUCT               9.414    95.918     9.030      0.384
CLOSURE ERROR-PC         -1.3                -3.7        1.1


                E N E R G Y   B A L A N C E

                         ENTHALPY IN      ENTHALPY OUT
                           (J/SEC)          (J/SEC)

     COOLING WATER         27786.1          51958.4
     REFLUX                 1843.8           2139.5
     TOP PRODUCT                             1659.3
     FEED                   4881.4
     STEAM                 39208.0           6041.2
     BOTTOM PRODUCT                          4141.8
     TOTAL                 73719.5          65940.5
     HEAT LOSS                      7779.0
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
        RUN NO FEFI 20% 16/05/78

FEED FLOW     = 22.687 G/SEC        DEV= 0.0959
REFLUX FLOW   = 13.807 G/SEC        DEV= 0.0426
STEAM FLOW    = 14.644 G/SEC        DEV= 0.5244
BOTTOM PROD   = 12.000 G/SEC        DEV= 0.0556
TOP PROD      = 10.708 G/SEC        DEV= 0.0927
COOL WATER    =502.894 G/SEC        DEV= 1.9717
DIST COMP     = 95.918 WT% MEOH     DEV= 0.1253
BOTTOM COMP   = 14.091 WT% MEOH     DEV= 0.0000
FEED COMP     = 51.615 WT% MEOH     DEV= 0.0000
COL PRESS     =  0.040 KPA          DEV= 0.0460
COND LEVEL    = 34.488 KPA          DEV= 0.1948
REB'R LEVEL   = 93.202 KPA          DEV= 0.3534
DIFF PRESS    =103.420 KPA          DEV= 0.1868
REB'R TEMP    =   86.0 DEG C        DEV= 0.2809
TRAY 1 TEMP   =   76.0 DEG C        DEV= 0.2525
TRAY 2 TEMP   =   72.8 DEG C        DEV= 0.2464
TRAY 3 TEMP   =   70.4 DEG C        DEV= 0.3658
TRAY 4 TEMP   =   70.1 DEG C        DEV= 0.2465
TRAY 5 TEMP   =   68.7 DEG C        DEV= 0.2564
TRAY 6 TEMP   =   64.6 DEG C        DEV= 0.2493
TRAY 7 TEMP   =   63.2 DEG C        DEV= 0.2227
TRAY 8 TEMP   =   61.2 DEG C        DEV= 0.2237
COND TEMP     =   59.2 DEG C        DEV= 0.2590
STEAM TEMP    =  100.4 DEG C        DEV= 0.4955
CON'TE TEMP   =   97.9 DEG C        DEV= 0.2852
REFLUX TEMP   =   46.0 DEG C        DEV= 0.2449
FEED TEMP     =   33.8 DEG C        DEV= 0.2476
BOTTOM TEMP   =   44.6 DEG C        DEV= 0.4653
REB'R VAPOR   =   85.0 DEG C        DEV= 0.2734
FEED INLET    =   63.6 DEG C        DEV= 0.3582
REFLUX INLET=     52.4 DEG C        DEV= 0.8021
COND VAPOR    =   61.9 DEG C        DEV= 0.2194
WATER INLET   =   13.2 DEG C        DEV= 0.2366
WATER OUTLET=     24.7 DEG C        DEV= 0.2547
```

```
                    STEADY STATE DATA
                     RUN NO FEF -20%
                        16/05/78

FEED FLOW       18.001 G/SEC      BOTTOM PROD       8.527 G/SEC
REFLUX FLOW     12.114 G/SEC      TOP PROD          9.062 G/SEC
STEAM FLOW      14.800 G/SEC      COOL WATER      514.416 G/SEC
FEED PLATF       4                FEED COMP        51.615 WT% MEOH
DIST COMP       95.869 WT% MEOH   BOTTOM COMP       4.171 WT% MEOH
FEED INLET       63.50 DEG C      REFLUX INLET      52.10 DEG C
STEAM TEMP      106.30 DEG C      COL PRESS          0.02 KPA


             M A T E R I A L   B A L A N C E

                           FLOW      COMP     METHANOL     WATER
                         (G/SEC)  (WT PCT)   (G/SEC)     (G/SEC)

FEED                      18.001    51.615     9.291       8.710
BOTTOM PRODUCT             8.527     4.171     0.355       8.172
TOP PRODUCT                9.062    95.869     8.688       0.374
CLOSURE ERROR-PC          -2.2                -2.6        -1.8


                E N E R G Y   B A L A N C E

                       ENTHALPY IN       ENTHALPY OUT
                         (J/SEC)           (J/SEC)

     COOLING WATER       27968.3           52028.8
     REFLUX               1827.7            2131.1
     TOP PRODUCT                            1594.2
     FEED                 4073.3
     STEAM               39769.7            6554.6
     BOTTOM PRODUCT                         3317.9
     TOTAL               73639.1           65626.8
     HEAT LOSS                     8012.2
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO FEF -20% 16/05/78

FEED FLOW    = 18.962 G/SEC       DEV= 0.0731
REFLUX FLOW  = 13.775 G/SEC       DEV= 0.0492
STEAM FLOW   = 14.800 G/SEC       DEV= 0.3981
BOTTOM PROD  =  8.571 G/SEC       DEV= 0.0383
TOP PROD     = 10.305 G/SEC       DEV= 0.0864
COOL WATER   =513.956 G/SEC       DEV= 1.8237
DIST COMP    = 95.869 WT% MEOH    DEV= 0.1549
BOTTOM COMP  =  4.171 WT% MEOH    DEV= 0.0000
FEED COMP    = 51.615 WT% MEOH    DEV= 0.0000
COL PRESS    =  0.029 KPA         DEV= 0.0659
COND LEVEL   = 33.790 KPA         DEV= 0.1693
REB'R LEVEL  = 92.815 KPA         DEV= 0.2176
DIFF PRESS   =103.420 KPA         DEV= 0.1868
REB'R TEMP   =   93.2 DEG C       DEV= 0.2777
TRAY 1 TEMP  =   82.8 DEG C       DEV= 0.2867
TRAY 2 TEMP  =   76.7 DEG C       DEV= 0.2654
TRAY 3 TEMP  =   72.8 DEG C       DEV= 0.4970
TRAY 4 TEMP  =   70.4 DEG C       DEV= 0.2599
TRAY 5 TEMP  =   68.7 DEG C       DEV= 0.2489
TRAY 6 TEMP  =   64.4 DEG C       DEV= 0.2232
TRAY 7 TEMP  =   63.0 DEG C       DEV= 0.2503
TRAY 8 TEMP  =   61.2 DEG C       DEV= 0.2173
COND TEMP    =   59.1 DEG C       DEV= 0.2457
STEAM TEMP   =  106.3 DEG C       DEV= 0.3382
CON'TE TEMP  =  104.9 DEG C       DEV= 0.2169
REFLUX TEMP  =   45.7 DEG C       DEV= 0.2819
FEED TEMP    =   33.4 DEG C       DEV= 0.2322
BOTTOM TEMP  =   41.2 DEG C       DEV= 0.2473
REB'R VAPOR  =   92.0 DEG C       DEV= 0.2709
FEED INLET   =   63.5 DEG C       DEV= 0.3051
REFLUX INLET=    52.1 DEG C       DEV= 0.9328
COND VAPOR   =   61.8 DEG C       DEV= 0.2442
WATER INLET =    13.0 DEG C       DEV= 0.2271
WATER OUTLET=    24.2 DEG C       DEV= 0.3893
```

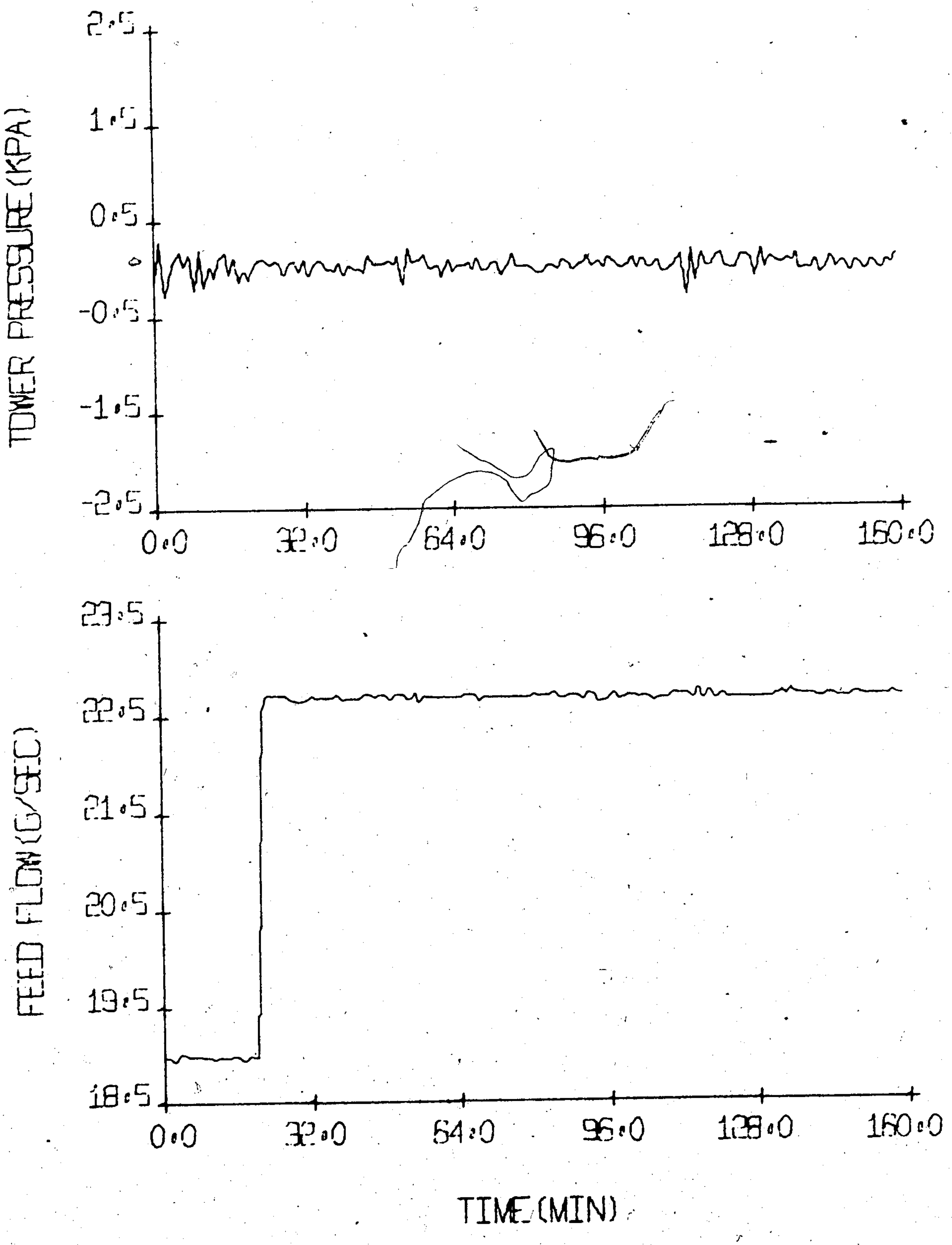

Figure G.7 Response of tower pressure and feed flow during a 20% increase in feed flow.

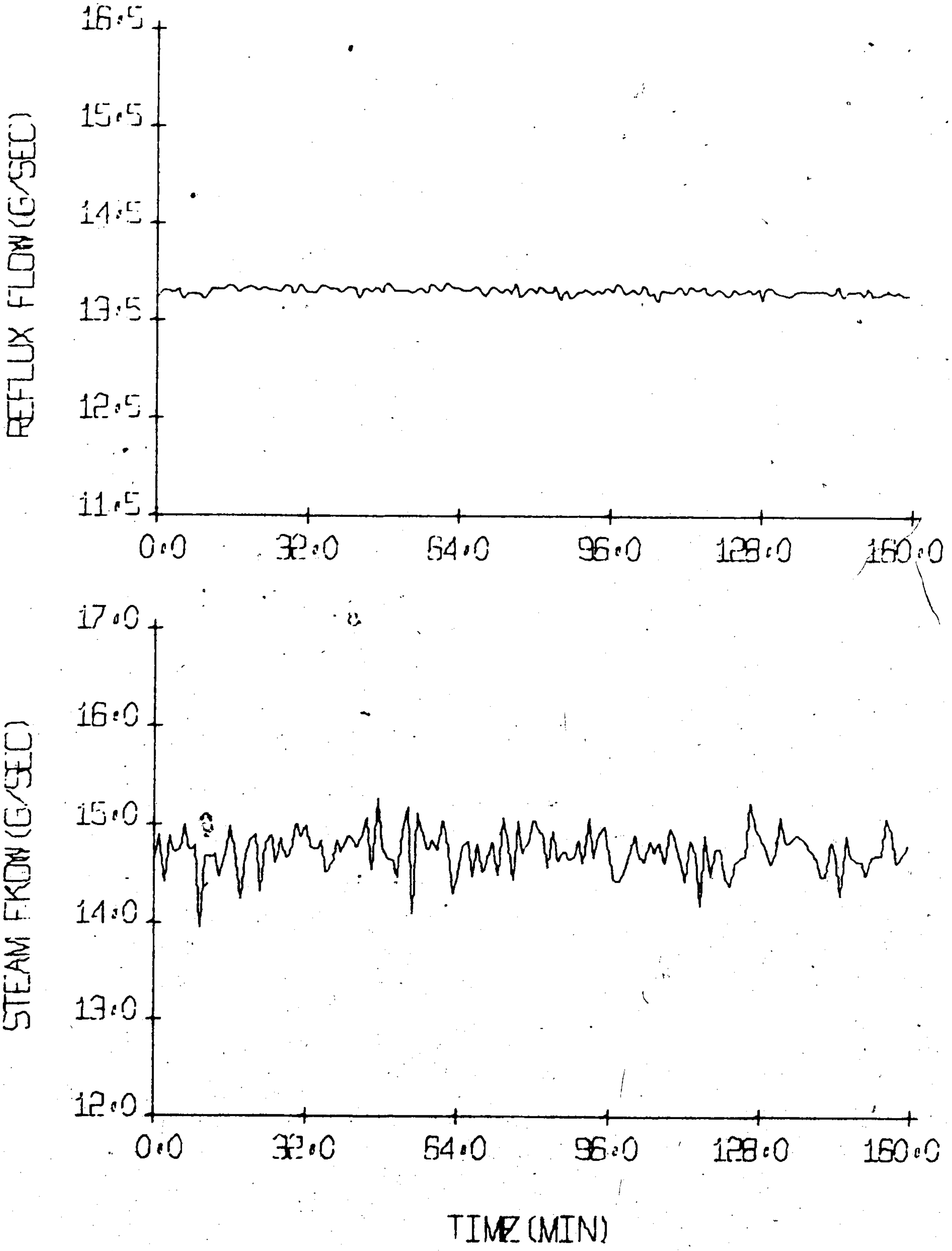

Figure G.8 Response of reflux and steam flow during a 20% increase in feed flow.

product flow rate is approximately twice as fast as the bottom product flow rate. The following Table G.1 shows the time delay in the bottom composition for open loop tests in Chapter 3.

Table G.1 Time Delay in Xb for Open Loop Tests

Figure	Time Delay (minutes)
3.2	5.40
3.4	5.67
3.5	10.88
3.7	8.96
3.8	5.12
3.9	4.48
3.10	6.88
3.11	8.00
3.12	4.58
3.13	8.96
3.14	5.27

APPENDIX H

H.1 Experimental Data for Chapter 4

This appendix contains the programs used to perform the PRBS studies of Chapter 4, that is, INII, PRBS1 and PRBS2. Again the programs are adequately documented and no description will be given. Also, the mass and energy balance was performed on the column before each PRBS test to ensure steady state conditions had been reached. The coding used on the computer outputs is

R - Reflux
S - Steam
31D - double 31 PRBS
63 - 63 PRBS
127 - 127 PRBS

where; PRBS31DS denotes a double 31 PRBS using steam flow rate.

```
C
C                      PROGRAM INII
C                      ------------
C
C
C     PURPOSE
C        TO ENTER IN FILES THE LENGTH OF THE PRBS
C        STUDY UNDER CONSIDERATION IE. EITHER A 63,
C        A DOUBLE 31 OR A 127 SEQUENCE.
C
C

      REAL PRBS(130)
      INTEGER NUM
      DEFINE FILE 10(60,80,U,NEXT)
      NUM=127
      READ(5,101)(PRBS(I),I=1,NUM)
  101 FORMAT(10F8.1)

C     THE NUMBER OF WRITE STATEMENTS DEPENDS ON THE
C     LENGTH OF THE PRBS RUN
C

      WRITE(10'51)(PRBS(I),I=1,25)
      WRITE(10'52)(PRBS(I),I=26,50)
      WRITE(10'53)(PRBS(I),I=51,75)
      WRITE(10'54)(PRBS(I),I=76,100)
      WRITE(10'55)(PRBS(I),I=101,125)
      WRITE(10'56)(PRBS(I),I=126,NUM)
      WRITE(6,101)(PRBS(I),I=1,NUM)
      STOP
      END
```

```
C
C                  PROGRAM PRBS1
C                  -------------
C
C
C     PURPOSE
C        TO PROVIDE THE REQUIRED DATA FOR PROGRAM
C        'PRBS2' AND START EXECUTION OF THAT PROGRAM.
C        ALSO TO PRINT THE STARTING TIME OF THE PRBS
C        STUDY.
C
C     DESCRIPTION OF PARAMETERS
C        LPID      DDC LOOPS
C        LUN       LOGICAL UNIT NUMBER
C        ICLE      CYCLE INTERVAL
C        IDIS      VARIABLE UNDER STUDY
C        KCNT      COUNTER
C        NUM       LENGTH OF PRBS STUDY
C        PRVAL     STEADY-STATE VALUE OF VARIABLE
C        DMAG      MAGNITUDE OF DISTURBANCE(PCNT)
C
C     USAGE
C        QUEUED FROM ANY TELETYPE AND ENTER DATA AS
C        REQUIRED.
C
C     EXAMPLE
C         QPRBS1
C
C               ENTER VARIABLE FOR PRBS STUDY
C                    1 -- REFLUX
C                    2 -- STEAM
C
C         1
C               ENTER STEADY STATE VALUE AND MAGNITUDE OF
C               DISTURBANCE
C
C         13.781,15.0
C               ENTER INTERVAL(640 OR 1280) AND
C               RUN TIME(62,63 OR 127)
C
C         1280,127
C               PRBS STUDY STARTED AT  14/58/47
C
C     REMARKS
C        PROGRAM INII MUST FIRST BE EXECUTED.
C
C     SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C        OPER,GETTY,FFINP,SCPTB,CYCLE,TIME
C
```

...(CONT'D)

```
C
      INTEGER LUN,ICLE,IDIS,KCNT,LPID(4)
      REAL PRVAL,DMAG
      DEFINE FILE 10(60,80,U,NEXT)
      DATA LPID/Z0602,Z0603,Z0608,Z0604/
      KCNT=0
C     MAKE LOOPS OPERABLE
C
      DO 10 I=1,4
      CALL OPER(LPID(I))
   10 CONTINUE
      CALL GETTY(LUN)
      WRITE(LUN,201)
  201 FORMAT(T5,'ENTER VARIABLE FOR PRBS STUDY'
     *,/,T10,'1 -- REFLUX'
     *,/,T10,'2 -- STEAM')
    1 CALL FFINP(LUN,1,10,IDIS,IER)
      IF(IER) 2,4,3
    2 WRITE(LUN,202)
  202 FORMAT(T5,'INPUT TO FFINP IN ERROR -- PROGRAM EXITED')
      CALL EXIT
    3 WRITE(LUN,203)
  203 FORMAT(T5,'INPUT TO FFINP IN ERROR -- TRY AGAIN')
      GO TO 1
    4 CONTINUE
      WRITE(LUN,204)
  204 FORMAT(T5,'ENTER STEADY STATE VALUE AND MAGNITUDE OF
     * DISTURBANCE')
    5 CALL FFINP(LUN,2,11,PRVAL,11,DMAG,IER)
      IF(IER) 2,7,6
    6 WRITE(LUN,203)
      GO TO 5
    7 CONTINUE
      WRITE(LUN,205)
  205 FORMAT(T5,'ENTER INTERVAL(640 OR 1280) AND '
     *,/,T5,'RUN TIME(40,70 OR 130)')
    8 CALL FFINP(LUN,2,10,ICLE,10,NUM,IER)
      IF(IER) 2,11,9
    9 WRITE(LUN,203)
      GO TO 8
   11 CONTINUE
      WRITE(10'60) LUN,ICLE,IDIS,KCNT,NUM,PRVAL,DMAG
C     PUT TIMER IN THE CYCLIC PROGRAM AUTO RESTART TABLE
C
      ITIM=7
```

...(CONT'D)

```
      ILEV=8
      IBIT=2
      CALL SCPTB(1,ITIM,ILEV,IBIT,1,ICLE,IER)
      CALL CYCLE(ILEV,IBIT,ITIM,ICLE)
C     WRITE TIME RUN WAS STARTED
C
      CALL TIME(IHOUR,IMIN,ISEC)
      WRITE(LUN,206) IHOUR,IMIN,ISEC
  206 FORMAT(T5,'PRBS STUDY STARTED AT  ',I2,'/',I2,'/',I2)
      CALL EXIT
      END
```

```
C
C                   PROGRAM PRBS2
C                   -------------
C
C
C     PURPOSE
C        TO EXECUTE THE PRBS NUMBERS AT THE
C        SPECIFIED INTERVAL.
C
C     DESCRIPTION OF PARAMETERS
C        SAME AS LIST FOR PRBS1 PLUS
C        ITEM      DDC LOOP WORD IE.
C                  1 -- SETPOINT OF LOOP
C                  2 -- MEASUREMENT OF LOOP
C                  3 -- OUTPUT OF LOOP
C        A         SLOPE OF RELATIONSHIP BETWEEN
C                  FLOW AND OUTPUT.
C        B         INTERCEPT FOR RELATIONSHIP BETWEEN
C                  FLOW AND OUTPUT.
C        SETPT     VALUE OF VARIABLE
C        OUTPT     OUTPUT(PCNT) SEND TO DDC LOOP
C
C     SYSTEM SUBROUTINES(IBM 1800) REQUIRED
C        GTVLU,PTVLU,CANCL,SCPTB,TIME
C
C

      INTEGER LUN,ICLE,IDIS,KCNT,NUM,LPID(4)
      REAL PRVAL,PRBS(130),DATA(4),A(2),B(2)
      DEFINE FILE 10(60,80,U,NEXT)
      DATA A/2.6385,3.9309/
      DATA B/-0.7962,1.9549/
      DATA LPID/Z0602,Z0603,Z0608,Z0604/
      READ(10'60) LUN,ICLE,IDIS,KCNT,NUM,PRVAL,DMAG

C     THE NUMBER OF READ STATEMENTS DEPENDS ON THE
C     LENGHT OF THE PRBS RUN.
C

      READ(10'51)(PRBS(I),I=1,25)
      READ(10'52)(PRBS(I),I=26,50)
      READ(10'53)(PRBS(I),I=51,75)
      READ(10'54)(PRBS(I),I=76,100)
      READ(10'55)(PRBS(I),I=101,125)
      READ(10'56)(PRBS(I),I=126,NUM)
      KCNT=KCNT+1

C     GET PRESENT VALUES OF VARIABLES FROM DDC LOOPS
C
```

...(CONT'D)

```
C          Z0602      REFLUX FLOW
C          Z0603      STEAM FLOW
C          Z0608      BOTTOM COMPOSITION(PCNT MEOH)
C          Z0604      TOP COMPOSITION(PCNT MEOH)
C

      ITEM=1
      IDESC=13
      DO 10 I=1,4
      CALL GTVLU(LPID(I),ITEM,DATA(I),IER,IDESC)
      GO TO (10,6,7,8),IER
   10 CONTINUE
      WRITE(LUN,201) DATA(IDIS),KCNT,DATA(3),DATA(4)
  201 FORMAT(//,T5,'VARIABLE OF INTEREST IS  ',F7.3,'  AT
     * KCNT= ',I4
     *,/,T5,'BOTTOM COMPOSITION IS  ',F5.2
     *,/,T5,'TOP COMPOSITION IS  ',F5.2)
      IF(KCNT-NUM) 1,1,9

C     CALCULATE NEW VALUE TO SEND TO DDC LOOP UNDER PBRS
C     STUDY
C

    1 CONTINUE
      SETPT=PRVAL*(1.0+PRBS(KCNT)*DMAG/100.0)

C     STEAM AND REFLUX FOLLOW A SQUARE ROOT RELATIONSHIP
C     BETWEEN
C     FLOW AND OUTPUT TO THE VALVE.
C

      OUTPT=((SETPT-B(IDIS))/A(IDIS))**2.0
      ITEM=3
      CALL PTVLU(LPID(IDIS),ITEM,OUTPT,IER)
      GO TO (5,6,7,8), IER
    5 CONTINUE
      WRITE(10'60) LUN,ICLE,IDIS,KCNT,NUM,PRVAL,DMAG
      CALL EXIT
    6 WRITE(LUN,202)
  202 FORMAT(T5,'ITEM NOT SPECIFIED IN PVR')
    7 WRITE(LUN,203)
  203 FORMAT(T5,'PVR ID NOT FOUND IN TABLE')
      GO TO 5
    8 WRITE(LUN,204)
  204 FORMAT(T5,'INVALID DATA IN CALL')
      GO TO 5
    9 CONTINUE
      CALL CANCL(7)

C     REMOVE TIMER FROM THE CYCLIC PROGRAM AUTO RESTART
```

...(CONT'D)

```
C     TABLE
C

      CALL SCPTB(0,7,8,2,1,ICLE,IE)
      CALL TIME(IHOUR,IMIN,ISEC)
      WRITE(LUN,205) IHOUR,IMIN,ISEC
  205 FORMAT(//,T5,'PBRS STUDY FINISHED AT  ',I2,'/',I2,'/'
     *,I2)
      CALL EXIT
      END
```

```
                         STEADY STATE DATA
                          RUN NO PRBS31DR
                             13/05/78

FEED FLOW       18.025 G/SEC     BOTTOM PROD        9.196 G/SEC
REFLUX FLOW     11.718 G/SEC     TOP PROD           8.782 G/SEC
STEAM FLOW      14.490 G/SEC     COOL WATER       508.441 G/SEC
FEED PLATE       4               FEED COMP         50.937 WT% MEOH
DIST COMP       95.860 WT% MEOH  BOTTOM COMP        4.045 WT% MEOH
FEED INLET       65.00 DEG C     REFLUX INLET       53.40 DEG C
STEAM TEMP      107.50 DEG C     COL PRESS          -0.02 KPA



              M A T E R I A L   B A L A N C E

                             FLOW      COMP     METHANOL    WATER
                           (G/SEC)  (WT PCT)   (G/SEC)    (G/SEC)

FEED                        18.025    50.937     9.181      8.843
BOTTOM PRODUCT               9.196     4.045     0.372      8.824
TOP PRODUCT                  8.782    95.860     8.419      0.363
CLOSURE ERROR-PC            -0.2                -4.2        3.8



                 E N E R G Y   B A L A N C E

                      ENTHALPY IN        ENTHALPY OUT
                        (J/SEC)            (J/SEC)

     COOLING WATER      31466.5            54821.4
     REFLUX              1821.4             2122.1
     TOP PRODUCT                            1590.4
     FEED                4196.3
     STEAM              38965.1             6493.0
     BOTTOM PRODUCT                         3622.8
     TOTAL              76449.4            68650.0
     HEAT LOSS                    7799.3
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO PRBS31DR 13/05/78

FEED FLOW     = 18.982 G/SEC         DEV= 0.0720
REFLUX FLOW   = 13.332 G/SEC         DEV= 0.0462
STEAM FLOW    = 14.490 G/SEC         DEV= 0.5196
BOTTOM PROD   =  9.251 G/SEC         DEV= 0.0484
TOP PROD      =  9.992 G/SEC         DEV= 0.0866
COOL WATER    =508.116 G/SEC         DEV= 1.8331
DIST COMP     = 95.860 WT% MEOH      DEV= 0.1127
BOTTOM COMP   =  4.045 WT% MEOH      DEV= 0.0000
FEED COMP     = 50.937 WT% MEOH      DEV= 0.0000
COL PRESS     = -0.020 KPA           DEV= 0.1266
COND LEVEL    = 34.289 KPA           DEV= 0.5759
REB'R LEVEL   = 93.422 KPA           DEV= 0.5081
DIFF PRESS    =103.420 KPA           DEV= 0.1868
REB'R TEMP    =   94.3 DEG C         DEV= 0.3224
TRAY 1 TEMP   =   84.2 DEG C         DEV= 0.2555
TRAY 2 TEMP   =   78.2 DEG C         DEV= 0.2628
TRAY 3 TEMP   =   77.7 DEG C         DEV= 0.2513
TRAY 4 TEMP   =   72.2 DEG C         DEV= 0.2536
TRAY 5 TEMP   =   70.5 DEG C         DEV= 0.2262
TRAY 6 TEMP   =   66.0 DEG C         DEV= 0.2253
TRAY 7 TEMP   =   64.6 DEG C         DEV= 0.2408
TRAY 8 TEMP   =   62.6 DEG C         DEV= 0.2400
COND TEMP     =   60.5 DEG C         DEV= 0.2666
STEAM TEMP    =  107.5 DEG C         DEV= 0.2779
CON'TE TEMP   =  106.1 DEG C         DEV= 0.3128
REFLUX TEMP   =   47.0 DEG C         DEV= 0.2265
FEED TEMP     =   35.3 DEG C         DEV= 0.2387
BOTTOM TEMP   =   44.1 DEG C         DEV= 0.2510
REB'R VAPOR   =   93.2 DEG C         DEV= 0.2703
FEED INLET    =   65.0 DEG C         DEV= 0.2764
REFLUX INLET=     53.4 DEG C         DEV= 0.5569
COND VAPOR    =   63.3 DEG C         DEV= 0.2341
WATER INLET   =   14.8 DEG C         DEV= 0.2178
WATER OUTLET=     25.8 DEG C         DEV= 0.6124
```

```
                       STEADY STATE DATA
                        RUN NO PRBS31DS
                          12/05/78

FEED FLOW      18.008 G/SEC     BOTTOM PROD      9.294 G/SEC
REFLUX FLOW    11.766 G/SEC     TOP PROD         9.179 G/SEC
STEAM FLOW     14.944 G/SEC     COOL WATER     508.637 G/SEC
FEED PLATE      4               FEED COMP       50.937 WT% MEOH
DIST COMP      96.093 WT% MEOH  BOTTOM COMP      4.050 WT% MEOH
FEED INLET      65.50 DEG C     REFLUX INLET     53.80 DEG C
STEAM TEMP     108.60 DEG C     COL PRESS         0.01 KPA


            M A T E R I A L   B A L A N C E

                          FLOW      COMP     METHANOL    WATER
                        (G/SEC)  (WT PCT)   (G/SEC)    (G/SEC)

FEED                     18.008    50.937     9.173      8.835
BOTTOM PRODUCT            9.294     4.050     0.376      8.917
TOP PRODUCT               9.179    96.093     8.821      0.358
CLOSURE ERROR-PC          2.5                 0.2        4.9


              E N E R G Y   B A L A N C E

                        ENTHALPY IN      ENTHALPY OUT
                          (J/SEC)          (J/SEC)

     COOLING WATER        31691.1          55267.4
     REFLUX                1843.5           2155.0
     TOP PRODUCT                            1681.3
     FEED                  4228.7
     STEAM                40212.8           6774.6
     BOTTOM PRODUCT                         3697.4
     TOTAL                77976.2          69575.9
     HEAT LOSS                     8400.2
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO PRBS31DS 12/05/78

FEED FLOW     = 18.964 G/SEC       DEV= 0.0591
REFLUX FLOW   = 13.396 G/SEC       DEV= 0.0512
STEAM FLOW    = 14.944 G/SEC       DEV= 0.4613
BOTTOM PROD   =  9.349 G/SEC       DEV= 0.0597
TOP PROD      = 10.451 G/SEC       DEV= 0.0668
COOL WATER    =508.320 G/SEC       DEV= 1.8294
DIST COMP     = 96.093 WT% MEOH    DEV= 0.1579
BOTTOM COMP   =  4.050 WT% MEOH    DEV= 0.0000
FEED COMP     = 50.937 WT% MEOH    DEV= 0.0000
COL PRESS     =  0.015 KPA         DEV= 0.1681
COND LEVEL    = 34.106 KPA         DEV= 0.4984
REB'R LEVEL   = 93.744 KPA         DEV= 0.5945
DIFF PRESS    = 96.920 KPA         DEV= 1.8209
REB'R TEMP    =   95.2 DEG C       DEV= 0.2484
TRAY 1 TEMP   =   85.2 DEG C       DEV= 0.2818
TRAY 2 TEMP   =   79.4 DEG C       DEV= 0.2591
TRAY 3 TEMP   =   75.2 DEG C       DEV= 0.2464
TRAY 4 TEMP   =   72.9 DEG C       DEV= 0.2392
TRAY 5 TEMP   =   71.2 DEG C       DEV= 0.2718
TRAY 6 TEMP   =   66.8 DEG C       DEV= 0.2525
TRAY 7 TEMP   =   65.3 DEG C       DEV= 0.2621
TRAY 8 TEMP   =   63.3 DEG C       DEV= 0.2532
COND TEMP     =   61.1 DEG C       DEV= 0.2089
STEAM TEMP    =  108.6 DEG C       DEV= 0.3437
CON'TE TEMP   =  107.3 DEG C       DEV= 0.2910
REFLUX TEMP   =   47.5 DEG C       DEV= 0.2100
FEED TEMP     =   35.2 DEG C       DEV= 0.2506
BOTTOM TEMP   =   44.1 DEG C       DEV= 0.2270
REB'R VAPOR   =   94.0 DEG C       DEV= 0.2705
FEED INLET    =   65.5 DEG C       DEV= 0.3171
REFLUX INLET=     53.8 DEG C       DEV= 0.2650
COND VAPOR    =   64.0 DEG C       DEV= 0.2309
WATER INLET   =   14.9 DEG C       DEV= 0.1973
WATER OUTLET=     26.0 DEG C       DEV= 0.2420
```

```
                    STEADY STATE DATA
                     RUN NO PRBS-63R
                        13/05/78

FEED FLOW      17.980 G/SEC     BOTTOM PROD      9.351 G/SEC
REFLUX FLOW    11.742 G/SEC     TOP PROD         9.163 G/SEC
STEAM FLOW     14.819 G/SEC     COOL WATER     507.817 G/SEC
FEED PLATE      4               FEED COMP       50.937 WT% MEOH
DIST COMP      95.908 WT% MEOH  BOTTOM COMP      4.340 WT% MEOH
FEED INLET      65.30 DEG C     REFLUX INLET     53.70 DEG C
STEAM TEMP     107.90 DEG C     COL PRESS         0.02 KPA



              M A T E R I A L   B A L A N C E

                           FLOW      COMP    METHANOL    WATER
                         (G/SEC)  (WT PCT)  (G/SEC)    (G/SEC)

FEED                      17.980    50.937    9.158      8.821
BOTTOM PRODUCT             9.351     4.340    0.405      8.945
TOP PRODUCT                9.163    95.908    8.788      0.374
CLOSURE ERROR-PC           2.9                0.3        5.6



                E N E R G Y   B A L A N C E

                        ENTHALPY IN       ENTHALPY OUT
                          (J/SEC)           (J/SEC)

     COOLING WATER        28033.8           51785.4
     REFLUX                1837.2            2143.6
     TOP PRODUCT                             1672.6
     FEED                  4207.6
     STEAM                39859.5            6666.3
     BOTTOM PRODUCT                          3693.8
     TOTAL                73938.2           65961.9
     HEAT LOSS                      7976.3
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO PRBS-63R 13/05/78

FEED FLOW     = 18.936 G/SEC        DEV= 0.0672
REFLUX FLOW   = 13.360 G/SEC        DEV= 0.0413
STEAM FLOW    = 14.819 G/SEC        DEV= 0.4378
BOTTOM PROD   =  9.407 G/SEC        DEV= 0.0627
TOP PROD      = 10.425 G/SEC        DEV= 0.0796
COOL WATER    =507.377 G/SEC        DEV= 1.9587
DIST COMP     = 95.908 WT% MEOH     DEV= 0.1368
BOTTOM COMP   =  4.340 WT% MEOH     DEV= 0.0000
FEED COMP     = 50.937 WT% MEOH     DEV= 0.0000
COL PRESS     =  0.023 KPA          DEV= 0.1496
COND LEVEL    = 34.013 KPA          DEV= 0.5204
REB'R LEVEL   = 93.393 KPA          DEV= 0.4828
DIFF PRESS    =103.230 KPA          DEV= 0.5350
REB'R TEMP    =    94.6 DEG C       DEV= 0.2470
TRAY 1 TEMP   =    84.4 DEG C       DEV= 0.2672
TRAY 2 TEMP   =    78.6 DEG C       DEV= 0.2197
TRAY 3 TEMP   =    64.3 DEG C       DEV= 0.3091
TRAY 4 TEMP   =    72.7 DEG C       DEV= 0.2379
TRAY 5 TEMP   =    70.9 DEG C       DEV= 0.2403
TRAY 6 TEMP   =    66.4 DEG C       DEV= 0.2397
TRAY 7 TEMP   =    65.0 DEG C       DEV= 0.2549
TRAY 8 TEMP   =    63.1 DEG C       DEV= 0.2645
COND TEMP     =    60.9 DEG C       DEV= 0.2456
STEAM TEMP    =   107.9 DEG C       DEV= 0.3305
CON'TE TEMP   =   106.5 DEG C       DEV= 0.3514
REFLUX TEMP   =    46.8 DEG C       DEV= 0.2404
FEED TEMP     =    35.4 DEG C       DEV= 0.2196
BOTTOM TEMP   =    43.5 DEG C       DEV= 0.2264
REB'R VAPOR   =    93.4 DEG C       DEV= 0.2828
FEED INLET    =    65.3 DEG C       DEV= 0.3243
REFLUX INLET=      53.7 DEG C       DEV= 0.4522
COND VAPOR    =    63.8 DEG C       DEV= 0.2539
WATER INLET   =    13.2 DEG C       DEV= 0.2305
WATER OUTLET=      24.4 DEG C       DEV= 0.2017
```

```
                        STEADY STATE DATA
                         RUN NO PRBS-63S
                            13/05/78

FEED FLOW      18.022 G/SEC    BOTTOM PROD     9.325 G/SEC
REFLUX FLOW    11.723 G/SEC    TOP PROD        9.116 G/SEC
STEAM FLOW     14.856 G/SEC    COOL WATER    508.460 G/SEC
FEED PLATE      4              FEED COMP      50.397 WT% MEOH
DIST COMP      95.895 WT% MEOH BOTTOM COMP     3.850 WT% MEOH
FEED INLET      65.30 DEG C    REFLUX INLET    53.50 DEG C
STEAM TEMP     108.10 DEG C    COL PRESS        0.04 KPA


                 M A T E R I A L   B A L A N C E

                           FLOW      COMP    METHANOL    WATER
                         (G/SEC)  (WT PCT)  (G/SEC)    (G/SEC)

FEED                      18.022    50.397    9.082      8.939
BOTTOM PRODUCT             9.325     3.850    0.359      8.966
TOP PRODUCT                9.116    95.895    8.741      0.374
CLOSURE ERROR-PC           2.3                0.1        4.4


                   E N E R G Y   B A L A N C E

                        ENTHALPY IN      ENTHALPY OUT
                          (J/SEC)          (J/SEC)

     COOLING WATER        29980.9          53337.2
     REFLUX                1826.0           2135.8
     TOP PRODUCT                            1660.7
     FEED                  4224.9
     STEAM                39963.9           6689.4
     BOTTOM PRODUCT                         3699.2
     TOTAL                75995.9          67522.6
     HEAT LOSS                     8473.3
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO PRBS-63S 13/05/78

FEED FLOW     = 18.968 G/SEC         DEV= 0.0680
REFLUX FLOW   = 13.339 G/SEC         DEV= 0.0409
STEAM FLOW    = 14.856 G/SEC         DEV= 0.4706
BOTTOM PROD   =  9.378 G/SEC         DEV= 0.0472
TOP PROD      = 10.372 G/SEC         DEV= 0.1542
COOL WATER    =508.084 G/SEC         DEV= 1.5857
DIST COMP     = 95.895 WT% MEOH      DEV= 0.1727
BOTTOM COMP   =  3.850 WT% MEOH      DEV= 0.0000
FEED COMP     = 50.397 WT% MEOH      DEV= 0.0000
COL PRESS     =  0.048 KPA           DEV= 0.0826
COND LEVEL    = 33.067 KPA           DEV= 0.4471
REB'R LEVEL   = 94.100 KPA           DEV= 0.4494
DIFF PRESS    =103.345 KPA           DEV= 0.5648
REB'R TEMP    =   94.9 DEG C         DEV= 0.2701
TRAY 1 TEMP   =   84.8 DEG C         DEV= 0.2731
TRAY 2 TEMP   =   78.9 DEG C         DEV= 0.2912
TRAY 3 TEMP   =   73.9 DEG C         DEV= 0.3290
TRAY 4 TEMP   =   72.6 DEG C         DEV= 0.2530
TRAY 5 TEMP   =   70.8 DEG C         DEV= 0.2462
TRAY 6 TEMP   =   66.5 DEG C         DEV= 0.2563
TRAY 7 TEMP   =   64.9 DEG C         DEV= 0.2211
TRAY 8 TEMP   =   63.0 DEG C         DEV= 0.2494
COND TEMP     =   60.8 DEG C         DEV= 0.2715
STEAM TEMP    =  108.1 DEG C         DEV= 0.3093
CON'TE TEMP   =  106.6 DEG C         DEV= 0.3023
REFLUX TEMP   =   47.0 DEG C         DEV= 0.2503
FEED TEMP     =   35.6 DEG C         DEV= 0.2433
BOTTOM TEMP   =   43.7 DEG C         DEV= 0.2403
REB'R VAPOR   =   93.7 DEG C         DEV= 0.2782
FEED INLET    =   65.3 DEG C         DEV= 0.3367
REFLUX INLET=     53.5 DEG C         DEV= 0.3556
COND VAPOR    =   63.6 DEG C         DEV= 0.2518
WATER INLET =     14.1 DEG C         DEV= 0.2035
WATER OUTLET=     25.1 DEG C         DEV= 0.2634
```

```
                  STEADY STATE DATA
                   RUN NO PRBS127R
                      24/05/78

FEED FLOW      17.988 G/SEC     BOTTOM PROD      9.068 G/SEC
REFLUX FLOW    12.113 G/SEC     TOP PROD         9.141 G/SEC
STEAM FLOW     14.741 G/SEC     COOL WATER     503.501 G/SEC
FEED PLATE      4               FEED COMP       51.615 WT% MEOH
DIST COMP      95.983 WT% MEOH  BOTTOM COMP      4.740 WT% MEOH
FEED INLET      63.90 DEG C     REFLUX INLET     53.40 DEG C
STEAM TEMP     105.40 DEG C     COL PRESS         0.02 KPA



             M A T E R I A L   B A L A N C E

                            FLOW     COMP    METHANOL    WATER
                          (G/SEC)  (WT PCT)  (G/SEC)    (G/SEC)

FEED                       17.988   51.615     9.284      8.703
BOTTOM PRODUCT              9.068    4.740     0.429      8.638
TOP PRODUCT                 9.141   95.983     8.774      0.367
CLOSURE ERROR-PC            1.2               -0.8        3.4



               E N E R G Y   B A L A N C E

                         ENTHALPY IN      ENTHALPY OUT
                           (J/SEC)          (J/SEC)

     COOLING WATER         31371.1          55129.9
     REFLUX                 1881.8           2143.3
     TOP PRODUCT                             1617.4
     FEED                   4099.1
     STEAM                 39589.4           6470.8
     BOTTOM PRODUCT                          3504.9
     TOTAL                 76941.5          68866.5
     HEAT LOSS                       8075.0
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
         RUN NO PRBS127R 24/05/78

FEED FLOW     = 18.956 G/SEC        DEV= 0.0809
REFLUX FLOW   = 13.784 G/SEC        DEV= 0.0501
STEAM FLOW    = 14.741 G/SEC        DEV= 0.7458
BOTTOM PROD   =  9.126 G/SEC        DEV= 0.0608
TOP PROD      = 10.402 G/SEC        DEV= 0.0847
COOL WATER    =503.187 G/SEC        DEV= 3.8705
DIST COMP     = 95.983 WT% MEOH     DEV= 0.1599
BOTTOM COMP   =  4.740 WT% MEOH     DEV= 0.0000
FEED COMP     = 51.615 WT% MEOH     DEV= 0.0000
COL PRESS     =  0.021 KPA          DEV= 0.0463
COND LEVEL    = 34.383 KPA          DEV= 0.1482
REB'R LEVEL   = 93.922 KPA          DEV= 0.2795
DIFF PRESS    =103.420 KPA          DEV= 0.1868
REB'R TEMP    =   92.7 DEG C        DEV= 0.2469
TRAY 1 TEMP   =   82.4 DEG C        DEV= 0.2468
TRAY 2 TEMP   =   76.5 DEG C        DEV= 0.2588
TRAY 3 TEMP   =   72.8 DEG C        DEV= 0.5456
TRAY 4 TEMP   =   70.8 DEG C        DEV= 0.2553
TRAY 5 TEMP   =   68.9 DEG C        DEV= 0.2677
TRAY 6 TEMP   =   64.7 DEG C        DEV= 0.2309
TRAY 7 TEMP   =   63.3 DEG C        DEV= 0.2757
TRAY 8 TEMP   =   61.5 DEG C        DEV= 0.2448
COND TEMP     =   59.4 DEG C        DEV= 0.2711
STEAM TEMP    =  105.4 DEG C        DEV= 0.2247
CON'TE TEMP   =  104.0 DEG C        DEV= 0.2919
REFLUX TEMP   =   46.8 DEG C        DEV= 0.2323
FEED TEMP     =   34.6 DEG C        DEV= 0.2300
BOTTOM TEMP   =   43.6 DEG C        DEV= 0.2284
REB'R VAPOR   =   91.6 DEG C        DEV= 0.2473
FEED INLET    =   63.9 DEG C        DEV= 0.2891
REFLUX INLET=     53.4 DEG C        DEV= 0.2608
COND VAPOR    =   62.1 DEG C        DEV= 0.2669
WATER INLET   =   14.9 DEG C        DEV= 0.2351
WATER OUTLET=     26.2 DEG C        DEV= 0.2776
```

```
                    STEADY STATE DATA
                     RUN NO PRBS127S
                        24/05/78

FEED FLOW      18.020 G/SEC     BOTTOM PROD      8.976 G/SEC
REFLUX FLOW    12.086 G/SEC     TOP PROD         9.136 G/SEC
STEAM FLOW     14.666 G/SEC     COOL WATER     514.453 G/SEC
FEED PLATE      4               FEED COMP       50.839 WT% MEOH
DIST COMP      95.618 WT% MEOH  BOTTOM COMP      4.360 WT% MEOH
FEED INLET      64.50 DEG C     REFLUX INLET     54.00 DEG C
STEAM TEMP     106.30 DEG C     COL PRESS         0.01 KPA



              M A T E R I A L   B A L A N C E

                          FLOW      COMP    METHANOL    WATER
                        (G/SEC)   (WT PCT)  (G/SEC)    (G/SEC)

FEED                     18.020    50.839     9.161      8.859
BOTTOM PRODUCT            8.976     4.360     0.391      8.585
TOP PRODUCT               9.136    95.618     8.736      0.400
CLOSURE ERROR-PC          0.5                -0.3        1.4



                E N E R G Y   B A L A N C E

                         ENTHALPY IN     ENTHALPY OUT
                           (J/SEC)         (J/SEC)

    COOLING WATER          33772.5         57403.2
    REFLUX                  1906.2          2164.1
    TOP PRODUCT                             1635.9
    FEED                    4160.3
    STEAM                  39409.7          6495.3
    BOTTOM PRODUCT                          3506.9
    TOTAL                  79248.8         71205.5
    HEAT LOSS                       8043.3
```

```
STEADY STATE CONDITIONS BASED ON  50 POINTS
        RUN NO PRBS127S 24/05/78

FEED FLOW    = 18.974 G/SEC       DEV= 0.0607
REFLUX FLOW  = 13.747 G/SEC       DEV= 0.0476
STEAM FLOW   = 14.666 G/SEC       DEV= 0.5249
BOTTOM PROD  =  9.033 G/SEC       DEV= 0.0410
TOP PROD     = 10.392 G/SEC       DEV= 0.0864
COOL WATER   =514.192 G/SEC       DEV= 2.5593
DIST COMP    = 95.618 WT% MEOH    DEV= 0.1027
BOTTOM COMP  =  4.360 WT% MEOH    DEV= 0.0000
FEED COMP    = 50.839 WT% MEOH    DEV= 0.0000
COL PRESS    =  0.019 KPA         DEV= 0.0455
COND LEVEL   = 33.080 KPA         DEV= 0.1962
REB'R LEVEL  = 94.017 KPA         DEV= 0.2703
DIFF PRESS   =103.420 KPA         DEV= 0.1868
REB'R TEMP   =    93.6 DEG C      DEV= 0.3152
TRAY 1 TEMP  =    83.6 DEG C      DEV= 0.2263
TRAY 2 TEMP  =    77.4 DEG C      DEV= 0.2452
TRAY 3 TEMP  =    73.5 DEG C      DEV= 0.3513
TRAY 4 TEMP  =    71.3 DEG C      DEV= 0.2388
TRAY 5 TEMP  =    69.6 DEG C      DEV= 0.2723
TRAY 6 TEMP  =    65.3 DEG C      DEV= 0.2560
TRAY 7 TEMP  =    63.8 DEG C      DEV= 0.2534
TRAY 8 TEMP  =    62.0 DEG C      DEV= 0.2678
COND TEMP    =    59.9 DEG C      DEV= 0.3038
STEAM TEMP   =   106.3 DEG C      DEV= 0.3313
CON'TE TEMP  =   104.9 DEG C      DEV= 0.2403
REFLUX TEMP  =    47.5 DEG C      DEV= 0.3320
FEED TEMP    =    35.2 DEG C      DEV= 0.2317
BOTTOM TEMP  =    44.4 DEG C      DEV= 0.2517
REB'R VAPOR  =    92.5 DEG C      DEV= 0.2648
FEED INLET   =    64.5 DEG C      DEV= 0.3333
REFLUX INLET=     54.0 DEG C      DEV= 0.2559
COND VAPOR   =    62.6 DEG C      DEV= 0.2504
WATER INLET  =    15.7 DEG C      DEV= 0.2549
WATER OUTLET=     26.7 DEG C      DEV= 0.2450
```

APPENDIX I

I.1 Continuous Form Approximations

In the analysis of the PRBS data collected, the parameters of the pulse transfer functions were determined. The data was approximated by a first order plus time delay system. The pulse transfer function can be related to the continuous elements of the loop using the following relationship

$$z^{-1}\{\underline{\underline{P}}(s)\underline{\underline{G}}(s)\} = \underline{\underline{P}}(z)\underline{\underline{G}}(z) \qquad \text{I-1}$$

The above relationship holds true (55) for the case when either $\underline{P}(s)$ or $\underline{G}(s)$ contains only delay terms. With the above relationship the continuous form can be approximated from the pulse transfer function (Equation 4.1) for the sampling interval used. The time delay in the first order system is assumed to be some integer multiple of the sampling interval. The relation between the parameters of the discrete and continuous transfer functions can be found in Table I of reference (66). If a first order zero hold is used then the time constant and gain for the first order plus time delay continuous system is

$$Tc = -Pt/\ln(A) \qquad \text{I-2}$$

```
G  = B/(1.0 - A)
```