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TERMINAL COMPOSITION CONTROL OF A BINARY DISTILLATION COLUMN

BY

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ABSTRACT

Terminal composition control of a binary distillation column operated under computer control has been studied for disturbances in feed flow rate. Conventional two point control, whereby the overhead composition is controlled by the reflux flow rate and bottoms composition by means of reboiler heat duty, was demonstrated to be unsatisfactory. The unsatisfactory behaviour of this control system is a direct result of the interaction between the two control loops due to the dynamic coupling inherent in a distillation process. Two alternate control systems, namely a noninteracting control system and a ratio control system, whereby the overhead composition is controlled by manipulating the reflux to overhead vapour rate ratio were proposed in order to eliminate the deleterious effects of the dynamic coupling.

The control systems were evaluated using a 9 inch diameter, 8 tray pilot scale distillation column interfaced with an IBM 1800 digital computer. The results show that a very significant improvement in the control of both product compositions is achieved by using the noninteracting system or the ratio control system as compared to the behaviour that results using conventional two point feedback control. However, the control performance obtained using the noninteracting was only marginally better than that obtained using the ratio control scheme.

The noninteracting control system was constructed from a simplified transfer function representation of the distillation column dynamics, determined from a series of on-line pulse tests. It was found that the dynamic behaviour of the distillation column, about an operating point, could be adequately represented by a set of first order plus dead time transfer functions.

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TABLE OF CONTENTS

TABLE OF CONTENTS (Continued)

PAGE

LIST OF TABLES

LIST OF FIGURES

LIST OF FIGURES (Continued)

1. INTRODUCTION

The distillation process is extensively used in the chemical processing industry to separate a feed stream into product streams of a specified composition. In order that a distillation column, subject to disturbances in a changeable environment, yield the specified product compositions, some form of control system is necessary.

The control of a distillation column to produce a single stream at a specified composition has received wide spread application. On the other hand, the control of more than one product composition presents some special difficulties due to the dynamic nature of the process. Consider a two point control system whereby the overhead and bottom compositions are maintained at their desired values by manipulation of the reflux and steam flows respectively. It is readily apparent that manipulation of the reflux flow will affect not only the top product composition but also the bottom composition as well. Likewise, the reboiler heat duty will also affect both of the end compositions. Indeed, the magnitude of the interaction between the two control loops, that is the effect of the reflux flow on the bottom composition and steam flow on the top composition are of the same magnitude as the direct effects of the manipulated inputs on the controlled variables. Thus an attempt to control the two product compositions in this manner will not meet with success, as control of the top product composition will cause disturbances to enter the bottom composition control loop and vice versa.

The synthesis of a control system embodies the appropriate selection of the controlled and manipulated variables based on the objectives of the control system as well as the allowable degrees of freedom for the process. An analysis of the degrees of freedom for a distillation process is presented in Section 1.1-D. If the objective of the control system is production of one or more products of a specified quality then the main controlled variables would clearly be the specified product compositions. However, in many instances it is not possible to obtain a direct measure of the product quality. As a result, temperatures on intermediate trays, or the column pressure, or the reflux/feed ratio are often selected as the controlled variables. Reflux rate, reboiler heat duty, and product flow rates are the most common manipulated variables.

1.1 LITERATURE REVIEW:

The control of distillation columns has received much attention in the recent chemical engineering literature. It is not possible to discuss all the literature pertaining to distillation column control. Rather, only those studies relevant to dual quality control, i.e. the control of both end compositions will be outlined in detail. For descriptions of control systems other than dual quality control, the reader is referred to several excellent reviews in the literature (1,2, 3,4).

1.1-A Dual Quality Control

In a recent series of papers Rijnsdorp (5,6,7) discusses the interaction in two variable control systems for distillation processes. It is concluded that the interaction inherent in the process will be deleterious to controllability in many instances. An alternate control scheme, using ratio control between the reflux and top vapour flows is suggested for reducing the effects of the interactions. The top product quality controller is used to adjust the ratio setting. Owing to the ratio control top product composition will be insensitive to variations in vapour flow, i.e. reboiler heating. This control system is discussed in more detail in Chapter 4.

In a recent theoretical study of distillation column dynamics, Davison (74) illustrates that severe interaction may exist in a binary distillation control system between temperature, or composition and pressure control loops and between the top and bottom temperature or composition control loops.

Distillation column control loops are inherently sluggish due to the large time constants within the process. An approach to improving the performance of such systems has been suggested by Rosenbrock (8) and extended by Gordon-Clark (9), Davison (10) and Gould (11). In the formulation presented by these authors, it is assumed that the distillation process can be adequately modelled by the state equation

$$\underline{x} = \underline{A}\,\underline{x} + \underline{B}\,\underline{U}$$

The matrix $\underline{A}$ may be written

$$\underline{A} = \underline{V}\,\underline{\lambda}\,\underline{V}'$$

where $\underline{V}$ is the matrix of eigenvectors of $\underline{A}$ and $\underline{V}'$ is the matrix of eigenvectors of $\underline{A}^T$. If the control vector $\underline{u}$ is chosen such that $\underline{u} = \underline{K}\,\underline{V}'\,\underline{x}$ and $\underline{B} = \underline{V}$, then the controlled system becomes

$$\underline{x} = \underline{V}\,[\underline{\lambda} + \underline{K}]\,\underline{V}'\,\underline{x}$$

Thus the eigenvalues of the system may be made to take any desired value by specification of the gain matrix K. If K is a diagonal matrix, it can be shown that the shifting of any one eigenvalue does not influence the other eigenvalues of the system. Thus the result will be a form of noninteracting control.

However, several restrictions will be encountered when an attempt is made to apply the above theory. The control matrix B will not usually be identical to V. Gould (11) has shown that this is not a serious restriction and may be overcome. A more serious restriction will be encountered in a system with inaccessible state variables, as would be the case in most complex industrial processes. In addition in a regulatory control system, it would be necessary to add integral control, the integral contributions to control would then interact.

1.1-B Multivariable Process Control

The inherent interaction present in a distillation column control system would suggest the application of multivariable control theory to symthesize a two point control system for the process. There are basically two different types of design methods available for multivariable control systems. The first method (noninteracting control) consists of placing a compensator or feedforward controller in the feedback path to compensate for the effects of interaction between control loops. The second technique is based on optimal control theory and consists of optimizing some objective function given a plant model. The effects of interaction are not eliminated in this case rather they are taken into account or used to improve system performance by means of the optimization process.

Synthesis of a Noninteracting Control System

Dynamic inter-coupling between process variables in a multivariable system prevents the direct application of single variable control theory. However, if the interacting effects could be cancelled by some form of compensator then single variable techniques could be employed to construct a suitable control system.

Consider the system shown in Figure 1.1.

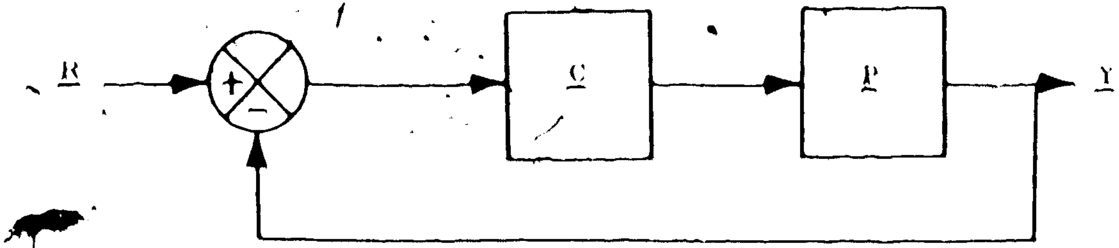

FIGURE 1.1

The closed loop transfer function for this system is given by $\underline{G} = (\underline{I} + \underline{P}\,\underline{C})^{-1}\,\underline{P}\,\underline{C}$ By definition the system is noninteracting if the matrix $\underline{G}$ is diagonal. If $\underline{G}$ is diagonal, one input will effect only one output; hence noninteraction is achieved. It can be shown that if the open loop transfer matrix $\underline{P}\,\underline{C}$ is diagonal then the closed loop transfer matrix will also be diagonal. Thus it appears that the noninteracting controller can be synthesized by open loop methods. The method for open loop synthesis of a noninteracting control system as given by several workers (12,13,14) will not be outlined. Consider that the elements of the open loop transfer matrix $\underline{G}$ are given by

$$G_{IJ} = \sum_{K=1}^{N} P_{IK} C_{KJ}$$

The design procedure involves selecting the elements C_{KJ} so that the off diagonal elements of G are zero,

i.e.

$$\sum_{K=1}^{N} P_{IK} C_{KJ} = 0 \qquad I \neq J$$

If the form of the diagonal elements G_{II} are specified from frequency response considerations or other methods, the elements of the compensator can be computed. The compensator must be chosen so as to yield a stable system. Morgan (13) and Chen (14) have outlined a set of constraints that must be placed on the compensators to insure stability. In addition to stability considerations the form of the specified open loop transfer function G_{ii} must be such that the elements of the compensator $\underline{C}$ are physically realizable. This is a trial and error task, indeed the desired form for G_{ii} may not be possible due to the physical realizability constraint.

Zalkind (15) gives a method for obtaining noninteracting control that is somewhat easier to realize than the above method. This method consists of placing two compensators in series as illustrated in Figure 1.2; the compensator $\underline{D}$ is introduced to cancel the interaction effects and the controller $\underline{C}$ can then be constructed by single variable methods to control the individual loops.

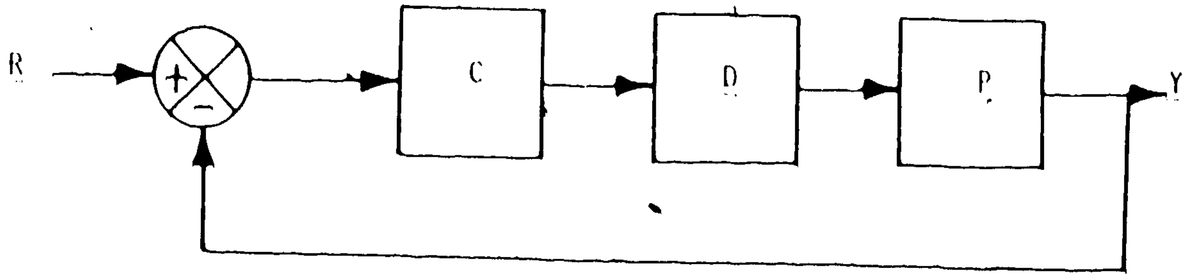

FIGURE 1.2

In Figure 1.2 $\underline{C}$ is a diagonal matrix of controller transfer functions. The compensator $\underline{D}$ can be determined from

$$\underline{DP} = \text{diagonal matrix}$$

and the diagonal elements $d_{ii} = 1$

Once again the compensator $\underline{D}$ may be difficult if not impossible to realize depending on the form of $\underline{P}$.

A considerable amount of work has been done on the synthesis of multivariable control systems to yield a specified closed loop response. The various methods of synthesis when the closed loop response is specified have been discussed by several workers (16-28). Basically the method consists of specifying a suitable form of the closed loop response $\underline{Y}$. However, it should be noted that the form of $\underline{Y}$ must be such that the compensator $\underline{C}$ is physically realizable.

$$\underline{Y} = (\underline{I} + \underline{P}\,\underline{C})^{-1}\,\underline{PC}$$

solving for C gives

$$C = P^{-1} G(I - G)^{-1}$$

This method has been outlined by Kavanagh (19,20).

Horowitz (16,22) also presents a synthesis procedure for multivariable systems. The objectives of this procedure are rejection of disturbances and compensation for changing plant parameters.

Other investigators (24,25,26) have discussed the synthesis of noninteracting controllers and stability of a multivariable system by the use of linear transformations.

Foster (29) and Greenfield (30) describe methods which may be used to decouple a multivariable system by means of a canonical transformation. Feedforward and feedback controllers are then added to each independent subsystem.

Kinnen (32) describes a method of multivariable control system synthesis based on a root locus analysis. This method is particularly applicable to systems where design specifications such as rise time, overshoot, etc. are important rather than noninteraction. Design parameters can be chosen to restrict the interaction.

As discussed previously, the application of noninteracting control theory (11-30) may give rise to an overly complex system. On account of this and the fact that it may not always be possible to achieve the desired performance, Rosenbrock (33) suggested an alternate synthesis technique. A structural synthesis technique is outlined which will handle the nonminimum phase effects caused by process dead times and the existence of right-half-plane zeros.

Another disadvantage of the previous conventional analysis for obtaining noninteracting control is the loss of structural information about the process in the complex form of the controller. Thus a design method that would retain the structural characteristics of the process may be desirable in some cases. Techniques developed by Mathias (35), Greenfield (36,37) and Morgan (28) are designed to retain structural information about the process in the controller.

A multivariable controller synthesis using time domain procedures will often be more desirable than a frequency domain approach. This is especially true for a system with a large number of states as the frequency domain approach would yield an overly complex controller. Morgan (38) and Falb (39) have introduced a technique for attaining noninteraction by means of state variable feedback. The method is similar in principle to the frequency domain techniques already described. For a process which can be described by

$$\dot{\underline{x}}(t) = \underline{A}\,\underline{x}(t) + \underline{B}\,\underline{u}(t)$$

and if control action of the form

$$\underline{u} = \underline{H}\,\underline{x} + \underline{k}\,R$$

is applied, then the expression for the controlled plant is given by

$$\dot{\underline{x}} = (\underline{A} + \underline{B}\,\underline{H})\,\underline{x} + \underline{B}\,\underline{k}\,\underline{R}$$

Thus for complete noninteraction among the states and the reference inputs the matrices $[\underline{A} + \underline{B}\,\underline{H}]$ and $\underline{B}\,\underline{k}$ must be diagonal.

Little work has been done on sampled data versions of noninteracting controllers. Fodor (40) has summarized the synthesis and limitations of a sampled compensator for noninteracting control. The major conclusion of the work is that a system can only be made noninteracting at the sampling intervals.

The synthesis of a decoupling compensator for a distillation column is discussed by Luyben (41). The series type compensator recommended by Zalkind (15) is suggested for "two-point" control. In another article on the multivariable control of phase separation processes Kleinpeter (42) illustrates the application of the structural formulation of Greenfield and Ward to a vapor liquid separation cell. The results serve to illustrate the advantages of a structural formulation over the terminal approach for noninteracting control.

Synthesis of an Interacting Control System

Since the interaction is an intrinsic property of a multivariable system, some workers have suggested that improved performance would result if the interaction is allowed to remain unimpeded or used rather than eliminated.

The majority of methods that appear in the literature (43-52) for interacting system synthesis are based on optimal control theory. Given a dynamic model of the plant the usual approach is to optimize some objective function via dynamic programming or calculus of variations. However, it has only been in a more recent study by Newell (76) that an optimal control technique has been applied to a regulatory control system with unmeasured load disturbances.

Mesarovic (53,54) has shown that an interacting system can give improved performance over an uncoupled system. From knowledge of the disturbance inputs an interacting domain is constructed in which the performance of an interacting system is superior to that of a decoupled control system.

Unlike their noninteracting counterparts, optimal control systems are more easily formulated in discrete time. Optimal control of discretized systems is the subject of much attention in the literature. The reader is referred to books by Tou (55) and Lapidus (56).

In a paper of specific interest Heiderman (52) has illustrated the design of an optimal discrete controller for a system with load changes. The technique is applied to a distillation column. However, the design method is restricted in that the disturbances must be measured.

A paper by MacFarlane (77), published after completion of this study, presents a survey of multivariable feedback control theory. This paper reviews the available design techniques and compares the advantages and disadvantages of the design methods.

1.1.C Process Control Via Gas Chromatograph

It has long been recognized that a gas chromatograph (GC) is a reliable and accurate tool for chemical analysis in the laboratory and the same should apply for its use in process control systems. Although there is widespread use of the instrument for on stream analysis, there is little evidence of process control loops being closed with a GC. There are several reasons for this, the discontinuous signal and measurement lag may seriously affect the controllability of the system, however, probably the most serious obstacle is the computation of the stream compositions from the voltage output of the gas chromatograph.

Villaboo (57,58) outlines the advantages and application of gas chromatographs in industry; while Vankampen (21) discusses the difficulties of process control via a gas chromatograph and outlines the dynamic problems connected with the introduction of a gas chromatograph analyzer into a process control loop. The effect of sampling time on system stability is examined by means of a polar diagram. Satisfactory control of a pentane-propylene splitter was obtained with a ratio of sampling time to process time constant at .18. Froode (59,60) outlines the use of a digital computer - gas chromatograph system for process control. The design of the computer system is discussed and examples of industrial applications are given.

The use of a small digital computer to monitor gas chromatographs surmounts the previously mentioned difficulties of computing the compositions and incorporating the results in a control system. Dedicated digital computer systems have become popular for monitoring laboratory gas chromatographs. Several authors (61,62,63) have described the use of such systems. However, little has been reported on an integrated gas chromatograph process control system using a digital computer.

A computer monitor system for gas chromatographs must provide for the following functions. It must be possible to initiate a gas chromatograph analysis by means of an external or internal interrupt within the computer. In the case of a process control system the computer must have the capability of initiating the GC run and controlling the sample injection on a repetitive basis. The system must be capable of reading the voltage output from the GC bridge and converting the signal to a digital value at rates up to 5 times a second. After the data acquisition or portion thereof is complete the monitor system must be capable of analyzing the data. Lichtenstein (61) and Briggs (62) have outlined the necessary features of the data analysis program. The analysis of the data consists of peak recognition, integration of the area under the peak and calculation of base line corrected areas.

The method most commonly used for peak recognition is examination of the first and second derivatives. The derivatives may be computed based on the least squares estimates as outlined by Savitzky (64). The start of a peak is recognized when the first derivative becomes greater than some specified deadband. The termination of a peak is reached after the sign of the first derivative changes from negative to positive. Examination of the second derivative is the usual means of determining if the base line is reached at the end of a peak. Base line corrected areas are computed by saving the coordinates of the start and end of the peak and then fitting a least squares line through sections of the base line immediately before and after the peak. The superfluous areas are then subtracted and the compositions computed from a calibration of the instrument.

1.1-D Degrees of Freedom for Distillation Columns

The available degrees of freedom for control of a distillation process is subject to varied opinion in the literature. The basis of the differences is the manner in which the various holdups within the column are treated.

Howard (65) states that the conventional steady state analysis technique is incorrect when applied to dynamical systems where the holdups of material may vary. The degrees of freedom for the unsteady state process may be derived in the usual manner by including the holdup in the same fashion as a normal stream. The holdup is analogous to a normal process stream in that it has the same variables except that a mass of material is involved instead of a flow rate.

An alternate approach is given by Bertrand (66) and Murrill (67) who tabulate the variables and relationships for a distillation process. This approach considers only the steady state, the holdups being ignored. n + 6 degrees of freedom will result from this approach, whereas the approach of Howard (65) shows n + 8 degrees of freedom. This difference, in the number of degrees of freedom, resulted because Murrill in his analysis assumed that the liquid levels in the condensor and reboiler were maintained constant by level control loops while Howard considered both levels as variables. Since the reboiler and condensor levels must be controlled the most logical approach seems to be to consider the column to have n + 6 degrees of freedom over and above specification of the reboiler and condensor levels. It should also be noted that if the approach of Howard is followed the top and bottom product flow rates could be controlled rather than the levels. If this is the case, the control of the holdup vessels would likely be over specified.

1.2 PROBLEM FORMULATION

As outlined previously the control of the composition of both the top and bottom product compositions from a distillation column presents special difficulties due to the interaction or dynamic intercoupling that exists between the control loops. As a result of these interaction effects, the direct application of single loop control theory will not result in satisfactory control of the column. There are basically two solutions to this problem of constructing a "two point" control system for a distillation process. One alternative is the application of control systems as suggested by Rosenbrock (8) and Rijnscorp (5,6,7), the second alternative being the design of a multivariable control system for the process.

It is the objective of this study to design and implement a noninteracting control system on a distillation column. The performance of the noninteracting system will be compared with a conventional two point beedback control system and the system suggested by Rijnsdorp (5,6,7). In order to design the noninteracting system, it will first be necessary to construct a simplified dynamic model of the process.

2. EXPERIMENTAL

2.1 EXPERIMENTAL PROGRAM

As outlined in Chapter 1 the objective of the present study is the comparison of systems for dual quality control of a distillation column. This comparison was carried out by implementation of the control systems on a pilot scale distillation column. The synthesis of a noninteracting controller outlined in the previous chapter requires a knowledge of the dynamic behaviour of the process. In particular a linear transfer function model that relates the manipulated inputs to the process outputs is required. The first step of the experimental program was to perform dynamic tests on the process to establish the necessary dynamic relationships. After establishing a suitable model for the process the control systems were implemented and their performance in rejecting disturbances compared.

2.2 EXPERIMENTAL EQUIPMENT

The experimental studies were performed on a pilot scale distillation column interfaced with an IBM 1800 data acquisition and control computer. The distillation column employed in this study was an eight tray nine inch diameter column with a total condensor and a basket type reboiler. The column trays consisted of four 2-1/4 x 1-7/8 inch bubble caps mounted in a square pattern. Solution holdups in the reboiler and on each tray were estimated by Svreck (68) at .25 and .065 cubic feet respectively. For further details on the column construction the reader is referred to studies by Svreck (68) and Pacey (69). A compilation of operating procedures and equipment associated with the distillation column is on file with the Department of Chemical and Petroleum Engineering (70). A binary mixture of methanol and water was the system used in the distillation column. The computer system provided the means for the rapid data acquisition and implementation of the varied control systems necessary for the completion of this project.

Since the dynamic models to be employed in the controller synthesis were assumed to be linear, it was necessary to select an operating point about which the dynamics of the column could be adequately represented by a linear model. The standard operating conditions about which the model was constructed are given in Table 2.1. Since the column was somewhat overdesigned, it was necessary to strike a compromise in the selection of the base operating condition. This compromise was necessary in order to maintain a reasonable composition profile throughout the column and at the same time obtain a measureable composition in the bottom product stream. The column also had a tendency to "flood" due to foam formation on the trays. A lower feed rate than one would expect from design considerations was necessary to eliminate the flooding condition.

2.3 COMPOSITION MEASUREMENT

In an earlier study on the same distillation column Svreck (68) obtained measurements of the compositions by measuring the capacitance of the solution. This method proved to be adequate for solutions of high methanol content but gave erratic results at low methanol concentrations. An alternate method, namely a gas chromatograph, was selected instead of the capacitance method for measuring the bottom composition.

TABLE 2.1

TYPICAL STEADY STATE OPERATING CONDITIONS

	Flow (lbs./min.)	Composition (wt. % MEOH)
Overhead product	1.18	96.0
Reflux	1.95	96.0
Bottom product	1.27	.5
Feed	2.45	46.5
Steam	1.71	-

	Temperature (°F)
Reflux	151.7
Feed	168.0
Steam	233.0
Condensate	227.5
Reboiler	209.6
Plate 1	20[illegible]6
2	194.4
3	181.2
4	172.9
5	164.1
6	156.8
7	152.1
8	148.5
Condensor	143.9

2.4 CONTROL LOOPS

In order to insure that no outside disturbances influenced the dynamic tests and control studies all inputs to the column were monitored and/or controlled. A schematic diagram of the distillation column and control loops is shown in Figure 2.1.

The flow rates and temperatures of the feed and reflux streams were controlled as well as the steam flow rate to the reboiler. To insure a constant pressure operation the condensor pressure was controlled by means of the cooling water flow rate and the differential pressure across the tower was monitored. Liquid levels in the condensor and reboiler were controlled by manipulation of the top and bottom product flow rates respectively. Examination of the control system will show that eight variables have been specified which is consistent with the degrees of freedom analysis of Chapter 1. The two levels, the column pressure as well as the feed and reflux temperatures were controlled by analog controllers. Control of the top and bottom compositions was accomplished using the IBM 1800 computer. The required control actions (i.e. the reflux and steam flow rates) were cascaded to the setpoints of the flow controllers.

2.5 COMPUTER INTERFACE

The values of all variables described above as well as the plate temperatures and temperatures of the flow streams were available for logging by the IBM 1800 computer system. Pneumatic signals were converted to analog signals suitable for the computer by air to current transducers. Control signals originating from the computer were converted to analog signals by a digital to analog converter and transduced

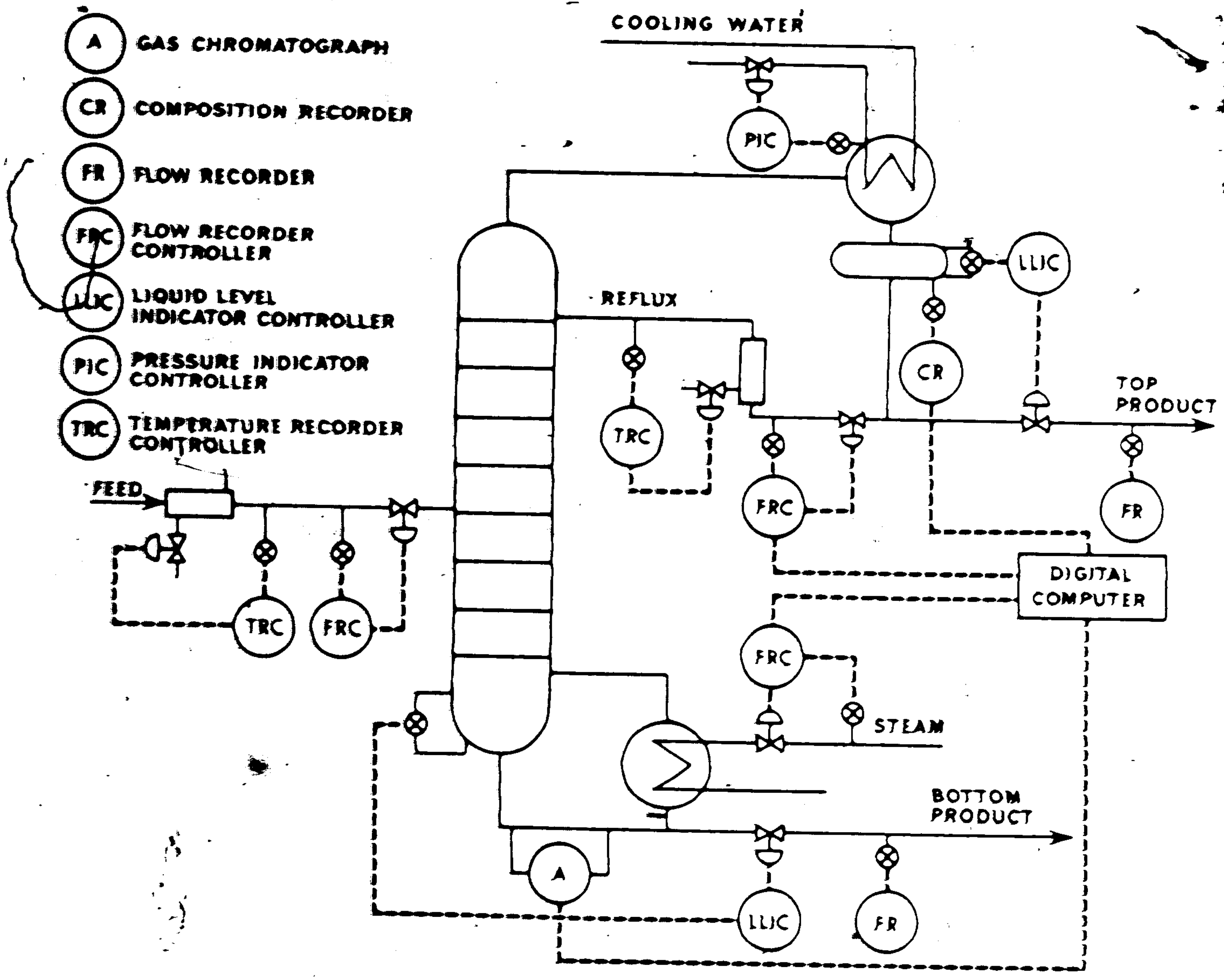

FIGURE 2.1. DISTILLATION COLUMN AND CONTROL LOOPS

to a current signal by current output stations (zero order hold devices) then further converted to pneumatic signals suitable for the control valve. A feature of the IBM 1800 computer was the IBM supplied direct digital control (DDC) program. With the exception of acquiring data for material and energy balance calculations, all communication with the process was done through this system. Each process control loop and data acquisition point is assigned a "loop record". The DDC program then allows the user to automatically read the analog input signals, calculate the control actions and output signals to the process at specified intervals. The DDC program also had the capability of collecting and storing a large amount of data over a specified time. This feature was used to advantage in acquiring data for construction of the process transfer functions.

Due to the large amount of data required for the material and energy balance calculations a separate program was written which would read the desired analog input signals and convert these signals to engineering units for use in the material and energy balance calculations. This procedure was necessary in order to allow other users access to the facilities of DDC at the same time.

2.6 GAS CHROMATOGRAPH

The gas chromatograph used in the analysis of the bottoms composition was a modified version of the Beckman Series C Analyzer. The analyzer was equipped with a sampling valve of one microlitre capacity. Sample injection was initiated on command from the computer by means of an external contact output (ECO). Closure of the ECO activated a 115 volt relay which i[illegible]rn activated a solenoid valve.

The solenoid valve controlled the operation of the sample valve by directing the application of air pressure to a diaphragm within the sample valve. Closing the ECO would thus result in movement of the 1 microlitre sample volume from the carrier gas stream to the liquid sample stream. On opening the ECO a short time later the sample volume would return to the carrier gas stream and the liquid sample would be flushed into the chromatograph column. Liquid sample flow was maintained by a pump circulating solution from the reboiler through the chromatograph and back to the reboiler. A tabulation of the operating conditions for the gas chromatograph is given in Table 2.2.

The 0 to 20 millivolt output from the chromatograph was read by the computer at one second intervals and stored by the DDC program for later analysis. The data acquisition phase lasted approximately 150 seconds at which time the data was analyzed and a new cycle initiated. Details of the programs for the data analysis are given in Chapter 5.

2.7 EXPERIMENTAL PROCEDURE

During the performance of the dynamic tests the column was brought to the desired steady state using the local analog controllers. Once at the desired steady state the data acquisition phase was initiated at the computer. After approximately fifteen minutes of steady state data was acquired the dynamic test was initiated by manual adjustment of the desired manipulated variable. The data acquisition was allowed to continue until the proces[illegible]nce again reached a steady state, at which time the data was punched onto cards for off line analysis.

TABLE 2.2

CHROMATOGRAPH OPERATING CONDITIONS

Temperature		120°C
Helium Pressure		30 psig
Column Length		5.5 ft
Type		Porapak Q
Cycle Time		150 sec
Retention Time	Water	60 sec
	Methanol	100 sec

The procedure followed during the control studies was to start up the distillation column under analog control. Once the desired steady state was achieved the system was switched so that the analog reflux and steam flow controllers received their setpoints from the computer. The control algorithms could then be initiated and the process disturbed by changing the feed flow rate.

The operation of the distillation column is illustrated on the McCabe-Thiele diagram in Appendix 3-A. The thermodynamic consistency of the temperature composition measurements are examined in Appendix 3-B.

3. DYNAMIC MODELS

The greatest single difficulty in the application of modern control theory to a chemical process and in particular distillation columns is the construction of a suitable dynamic model for the system. It is well established that a distillation process can be described by a set of nonlinear material and energy balance equations. However, a model of such complexity is of little use in a process control system. The nature of the equations, especially the equilibrium relationship, make it difficult to obtain an adequate model by linearizing the describing differential equations. In any case, if a model is constructed by linearizing a component material balance associated with each stage, it would then be necessary to reduce the order of the resulting state equation. There are well known techniques (71) for reducing the number of state variables in such an equation based on elimination of the states with the eigenvalues furthest from the origin. However, this approach results in a further approximation especially if the order of the state equation must be drastically reduced as would be the case in a large industrial column. In addition, the dominant eigenvalues may not necessarily be associated with the states of interest, i.e. the terminal compositions. Thus dynamic testing would appear to be a more reasonable approach to obtaining a simplified dynamic model than the theoretical approach.

The nonlinearity of the equilibrium relationship would appear to make the strongest contribution to the nonlinear behaviour of a distillation column. This is especially true if the column is operated such that the end compositions are in the "pinch" of the equilibrium curve. Composition excursions from the steady state towards the ends of the equilibrium curve would be severely limited whereas changes in the other direction would not be so limited. Thus in order to construct a meaningful model from dynamic test data, it is necessary to restrict the magnitude of the input disturbance such that the process is not unduly effected by the inherent nonlinearities. This is valid if the model is to be used in a control system where the terminal compositions are under "tight" control.

The procedure adopted in this study was to perform a set of dynamic tests on the distillation column in order to establish the relationships between the two end compositions and the inputs (reflux flow, steam flow, feed flow and feed composition). Since the object of these tests was to establish a process model about the specified operating point a pulse test was employed rather than a step input which would tend to yield information about the dynamics on excursions away from the operating point. A rectangular pulse was used for determining the model parameters.

The process was assumed to be adequately represented by a combined first order time constant plus a time delay transfer function model relating the process outputs to each input. The model parameters were then determined by fitting the time domain response of the model to the recorded transient data.

The "fit" was accomplished by employing Rosenbrock's direct search technique (72) using a least squares criterion. In order to simplify the search for the optimum parameters the process time delay was first calculated. This was accomplished by locating the point with the largest slope on the initial side of the response curve. The time delay was then determined from the intercept of the line of maximum slope and the initial steady state value.

The above procedure assumes a linear process behaviour about the operating point. The validity of this assumption and the use of the simplified model can be seen by comparing the model predictions with the experimental data.

The resulting model parameters are given in Tables 3.1 - 3.4 with the area under the associated input pulse. The values of the process gains are tabulated in (wt. %)/(lbs./min) except in the case of the feed composition runs where the units are (wt. %)/(wt. %). The time constants and time delays are tabulated in values of minutes. Further data concerning the dynamic tests (magnitude of the forcing pulse, and steady state operating conditions) are tabulated in Appendix 2A.

The transient response of the process is compared with that predicted by the model for several tests in Figures 3.1 to 3.6. In all cases the process and model responses are in good agreement illustrating that the simplified model adequately describes the behaviour of the system about the designated operating point.

In order to arrive at a set of transfer functions for the process, the average model parameters were computed based on the tabulated parameters. The time delay associated with the bottom product composition does not include the 2.5 minute analysis time of the gas chromatograph, however, it does include the time delay in the sampling lines. Thus it is necessary to subtract an estimated .5 minute delay from the tabulated bottom product time delay to arrive at the true process transfer function model. The transfer function model of the process is given by equations 3.1 and 3.2.

TABLE 3.1

REFLUX PULSE TESTS RESULTS

Run No	Pulse Area	TOP COMPOSITION			BOTTOM COMPOSITION		
		Gain	Time Constant	Time Delay	Gain	Time Constant	Time Delay
20	4.2	10.1	16.3	.5	6.7	12.0	7.9
21	-3.3	15.6	13.5	1.1	-	-	-
22	3.9	11.1	17.6	1.1	6.4	11.3	7.4
23	2.2	12.9	19.4	.8	6.8	9.5	7.3
24	-1.5	14.5	16.8	1.6	-	-	-

TABLE 3.2

STEAM PULSE TESTS RESULTS

Run No	Pulse Area	TOP COMPOSITION Gain	TOP COMPOSITION Time Constant	TOP COMPOSITION Time Delay	BOTTOM COMPOSITION Gain	BOTTOM COMPOSITION Time Constant	BOTTOM COMPOSITION Time Delay
20	-2.7	-17.6	23.6	2.4	-20.8	15.5	3.8
24	-2.0	-15.7	22.0	2.7	-21.1	13.6	3.7
25	1.8	-22.9	17.1	3.2	-	-	-
27	-1.6	-19.4	21.3	3.4	-16.2	14.1	2.8

TABLE 3.3

FEED FLOW PULSE TESTS RESULTS

		TOP COMPOSITION			BOTTOM COMPOSITION		
Run No	Pulse Area	Gain	Time Constant	Time Delay	Gain	Time Constant	Time Delay
20	6.9	2.7	13.9	6.1	4.6	13.2	3.8
21	-3.7	4.5	10.7	10.1	-	-	-
24	4.3	4.3	20.0	8.0	5.2	13.2	3.0

TABLE 3.4

FEED COMPOSITION PULSE TESTS RESULTS

		TOP COMPOSITION			BOTTOM COMPOSITION		
Run No	Pulse Area	Gain	Time Constant	Time Delay	Gain	Time Constant	Time Delay
20	137	.04	26.7	8.3	.097	10.8	10.2
21	- 71	.43	17.2	6.4	-	-	-
22	87	.18	25.2	9.3	.190	14.2	8.1
23	-115	.22	22.1	6.7	-	-	-

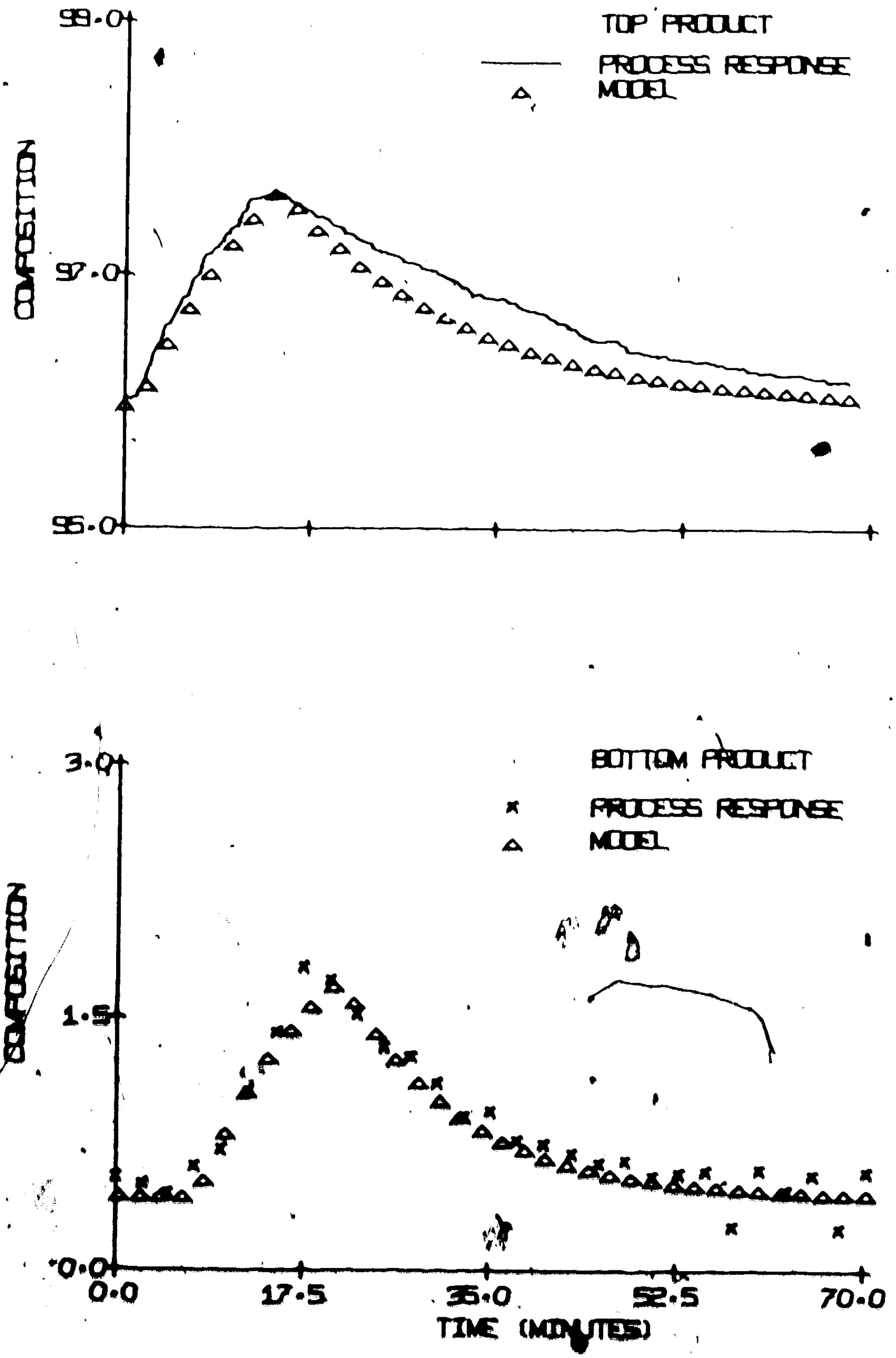

FIGURE 3.1

PRODUCT COMPOSITION RESPONSE TO A PULSE IN REFLUX FLOW [RUN 22]

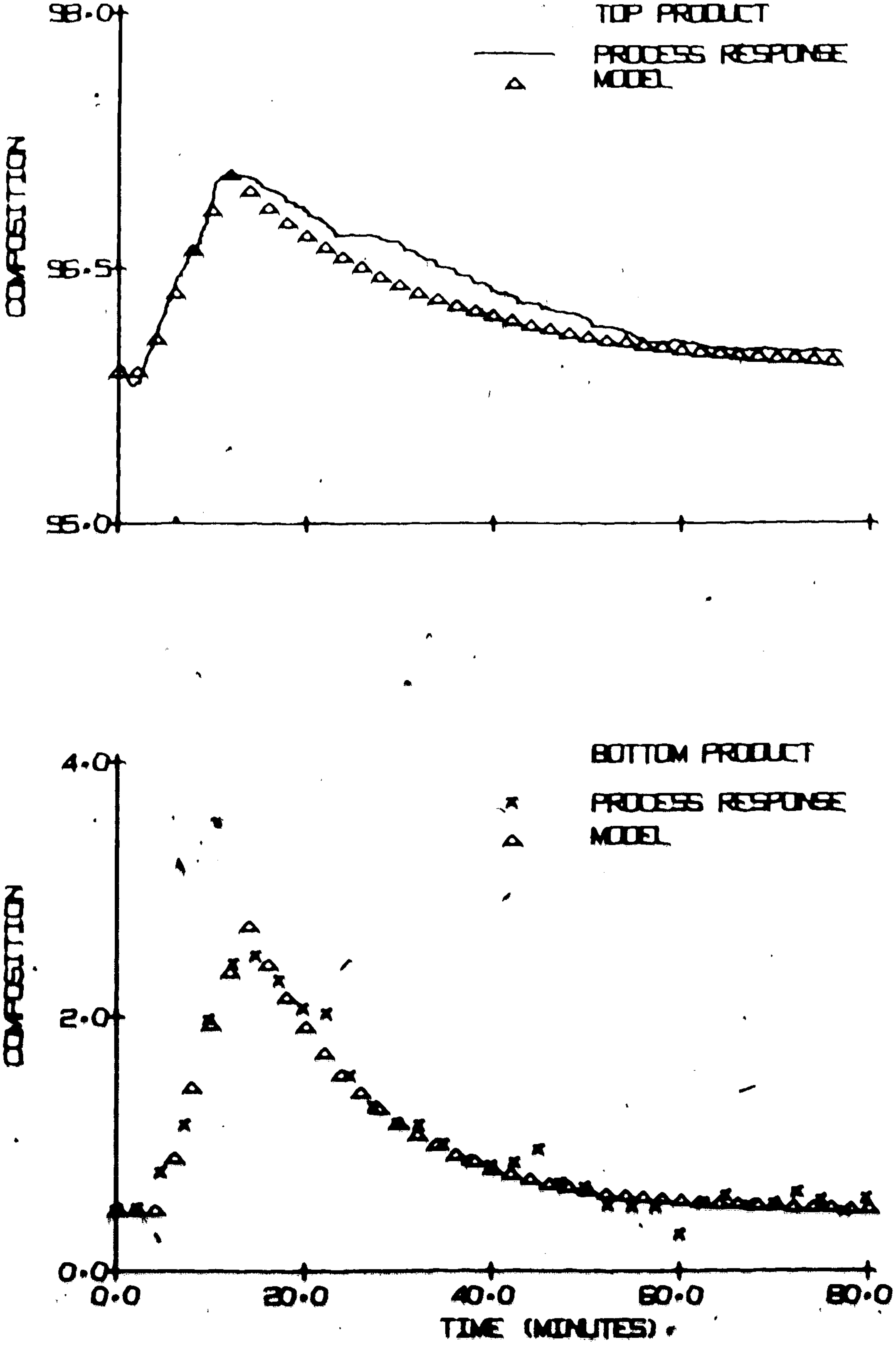

FIGURE 3.2

PRODUCT COMPOSITION RESPONSE TO A PULSE IN STEAM FLOW [RUN 24]

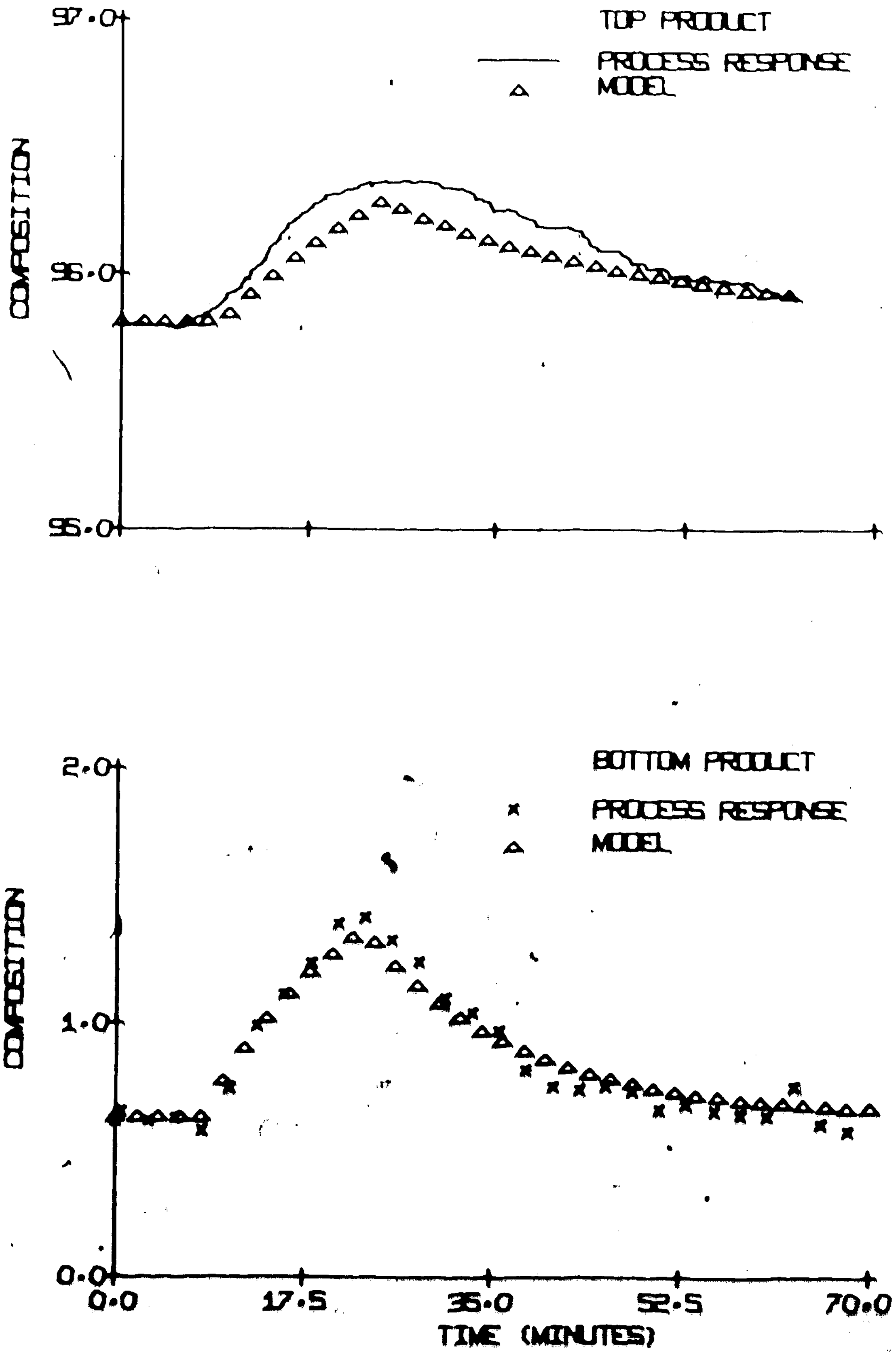

FIGURE 3.3

PRODUCT COMPOSITION RESPONSE TO A PULSE IN FEED COMPOSITION [RUN 22]

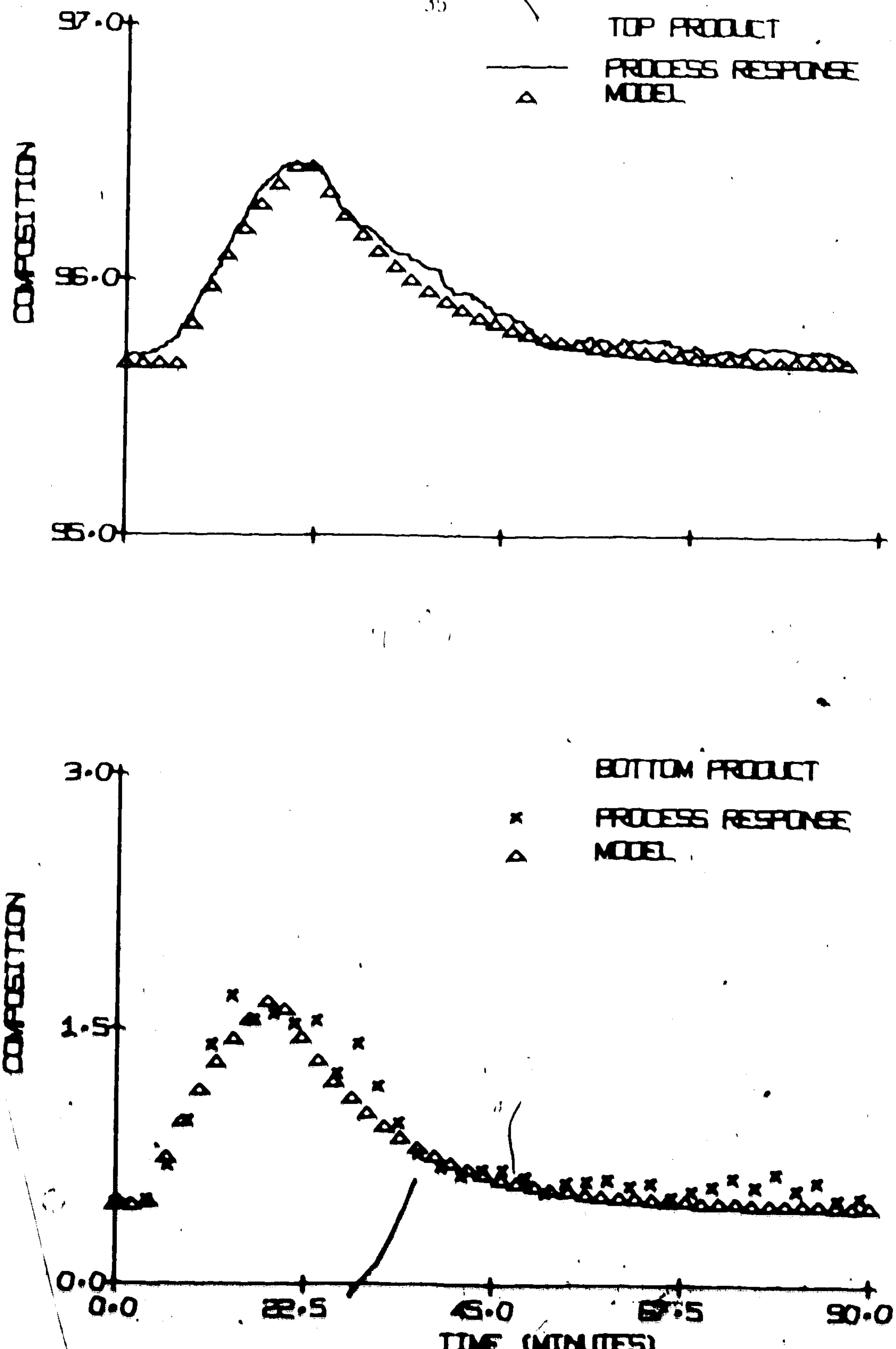

FIGURE 3.4

PRODUCT COMPOSITION RESPONSE TO A PULSE IN FEED FLOW [RUN 20]

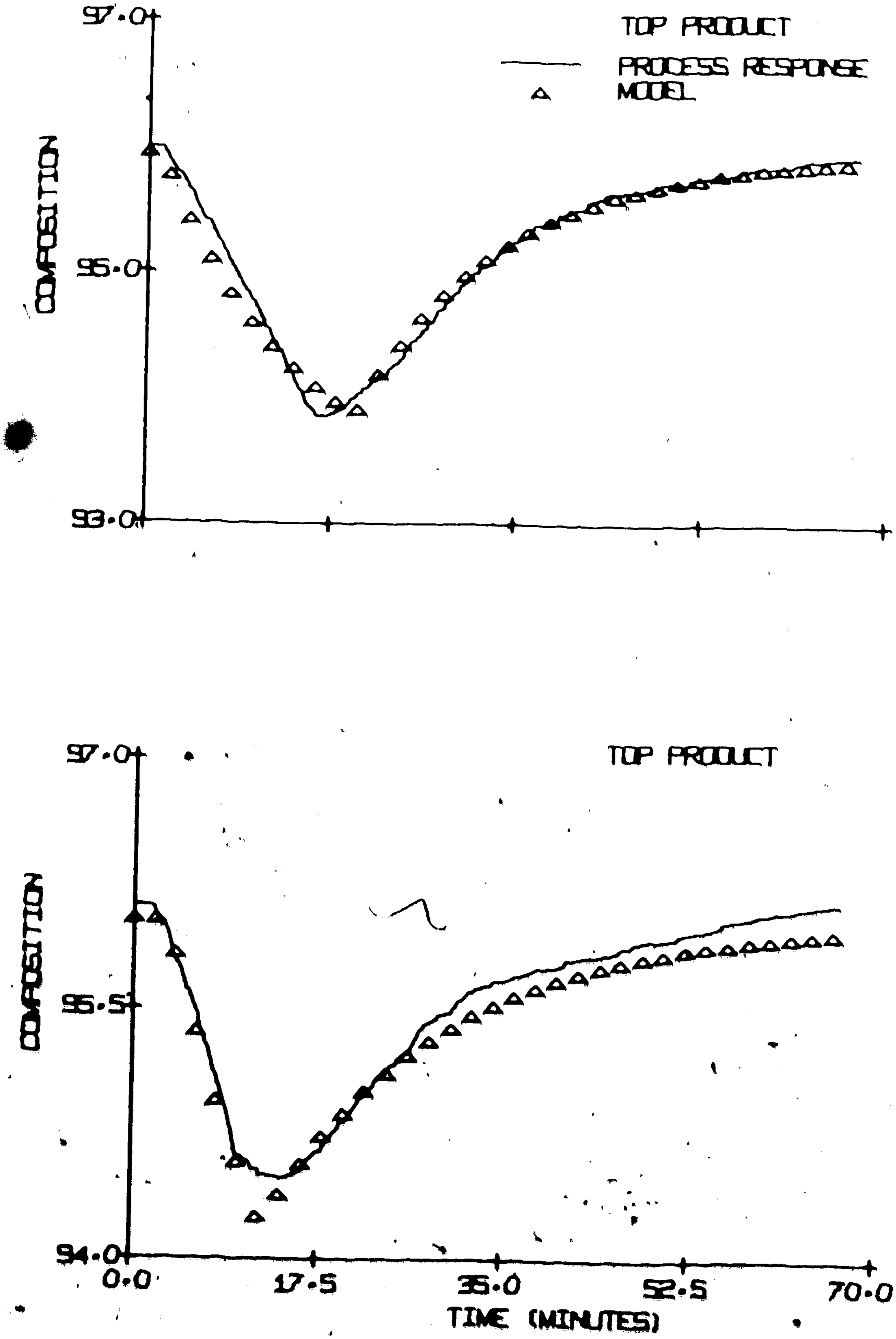

FIGURE 3.5

TOP PRODUCT COMPOSITION RESPONSE TO A PULSE IN REFLUX FLOW (TOP - RUN 21) AND STEAM FLOW (BOTTOM - RUN 25)

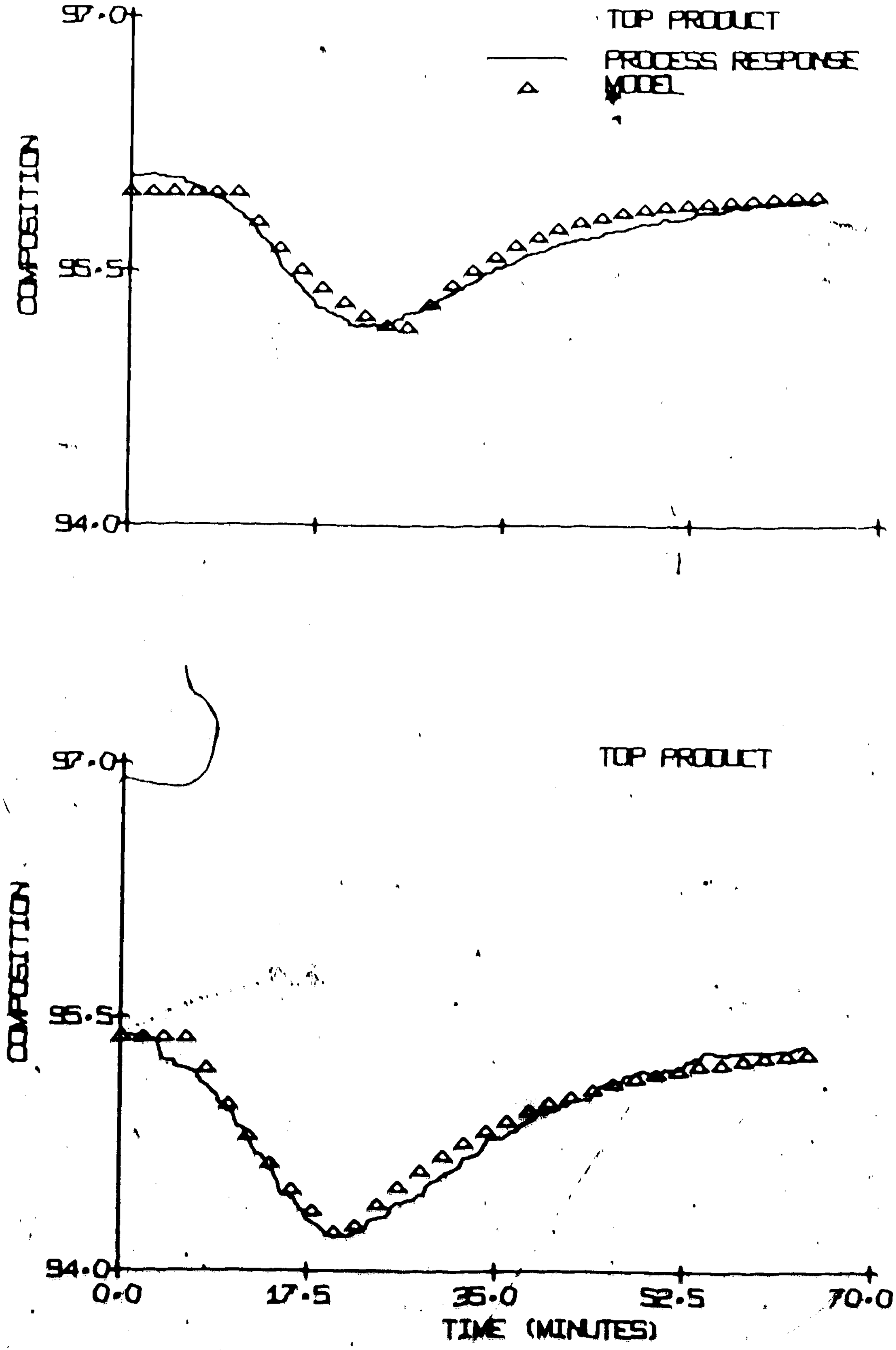

FIGURE 3.6

TOP PRODUCT COMPOSITION RESPONSE TO A PULSE IN FEED COMP.
(TOP - RUN 21) AND FEED FLOW (BOTTOM - RUN 21)

The percent deviation of the model parameters from the mean value for each set of runs is recorded in Table 3.5. For the control inputs (reflux and steam flows) the agreement is good, the % deviation being less than 15% in all cases except one.

$$\begin{bmatrix} x_D \\ \\ x_B \end{bmatrix} = \begin{bmatrix} \dfrac{12.8e^{-1.s}}{16.7s + 1} & \dfrac{-18.9e^{-2.9s}}{21.s + 1} \\ \\ \dfrac{6.6e^{-7.s}}{10.9s + 1} & \dfrac{-19.4e^{-2.9s}}{14.4s + 1} \end{bmatrix} \begin{bmatrix} R \\ \\ S \end{bmatrix} \qquad (3.1)$$

$$\begin{bmatrix} x_D \\ \\ x_B \end{bmatrix} = \begin{bmatrix} \dfrac{3.8e^{-8.1s}}{14.9s + 1} & \dfrac{.22e^{-7.7s}}{22.8s + 1} \\ \\ \dfrac{4.9e^{-3.4s}}{13.2s + 1} & \dfrac{.14e^{-9.2s}}{12.1s + 1} \end{bmatrix} \begin{bmatrix} F \\ \\ x_F \end{bmatrix} \qquad (3.2)$$

The larger deviations in the case of the top product composition gain to changes in reflux and steam flow rates can be attributed to the nonlinear behaviour of the column. From Tables 3.1 and 3.2 it can be observed that a somewhat larger gain results when the system is forced out of the "pinch" than occurs when the composition is forced into the "pinch".

In the case of the disturbance inputs (feed flow and composition), more data is required to arrive at an adequate model. Since the feed flow and feed composition data is superfluous with respect to the study this data will not be given further consideration.

A comparison of the results of the pulse tests in Tables 3.1

TABLE 3.5

PERCENT DEVIATION OF MODEL PARAMETERS

	TOP COMPOSITION			BOTTOM COMPOSITION		
	K	τ	τ_D	K	τ	τ_D
Reflux Flow	14.0	8.7	28.4	2.3	8.7	3.2
Steam Flow	12.0	9.3	12.5	10.2	5.1	12.0
Feed Flow	19.5	23.0	17.0	6.1	0	8.7
Feed Comp.	50.0	14.0	14.5	29.0	13.6	11.5

and 3.2 show that similar model parameters resulted from pulses of different magnitudes and in opposite directions. Thus it would appear reasonable to assume a linear process behaviour with respect to the effects of reflux and steam flows on the end compositions over the small operating range considered.

3.1 RELATIVE GAIN MATRIX

The correct pairing of manipulated and controlled variables in an interacting control system may be accomplished through the use of a relative gain matrix (75). The object is to control a given process output with the manipulated input that will have the greatest influence on it. The relative gain matrix is determined from a "normalized" matrix of the open loop process gains.

From equation 3.1 $\underline{K}$ is the matrix of the open loop gains.

$$\underline{K} = \begin{bmatrix} 12.8 & -18.9 \\ 6.6 & -19.4 \end{bmatrix}$$

An intermediate matrix $\underline{L}$ is then formed

$$\underline{L} = (\underline{K}^{-1})^T.$$

The relative gain matrix $\underline{K}^1$ is then formulated by multiplying corresponding elements of $\underline{K}$ and $\underline{L}$

$$\text{i.e.} \quad K^1_{ij} = K_{ij} L_{ij}.$$

For the case at hand the resulting matrix of relative gains between the manipulated variables (R and S) and the controlled process outputs (X_D and X_B) is :

$$\begin{array}{c|cc} & R & S \\ \hline X_D & 2.01 & -1.01 \\ X_B & -1.01 & 2.01 \end{array}$$

This illustrates that the overhead composition (X_D) and bottoms composition (X_B) should be controlled by the reflux (R) and steam (S) flow rates respectively.

4. DUAL PRODUCT QUALITY CONTROL SYSTEMS

The most readily apparent scheme for accomplishing dual product quality control of a distillation process, shown schematically in Figure 4.1, is control of the overhead composition by manipulation of reflux flow and bottom composition control by altering reboiler heating duty. However, it is evident on examination of the dynamic nature of the process (Equation 3.1) that a strong interaction will exist between the two control loops. Indeed the effect of the bottom product control on the overhead composition is larger than the direct effect of the manipulated variable, reflux rate, on the overhead composition.

Thus any control system for the direct control of the overhead and bottom product composition must be based on a method that will compensate for, or reduce, the effects of the interaction between the overhead and bottom composition control loops.

4.1 RATIO CONTROL

A novel approach for reducing the effects of the interaction between the control loops has been suggested by Rijnsdorp et al (5,6,7). The control system that was suggested is illustrated in Figure 4.2. In this scheme the overhead composition controller is cascaded to the setpoint of a ratio controller which controls the ratio of the overhead vapour rate to the reflux flow rate by manipulation of the reflux rate. A disturbance to the column which causes the overhead composition to change will cause a change in the desired ratio of the two flows, which for a constant overhead vapour rate will cause a corresponding change in the reflux flow. On the other hand the ratio controller will automatically compensate for a change in overhead vapour rate by a

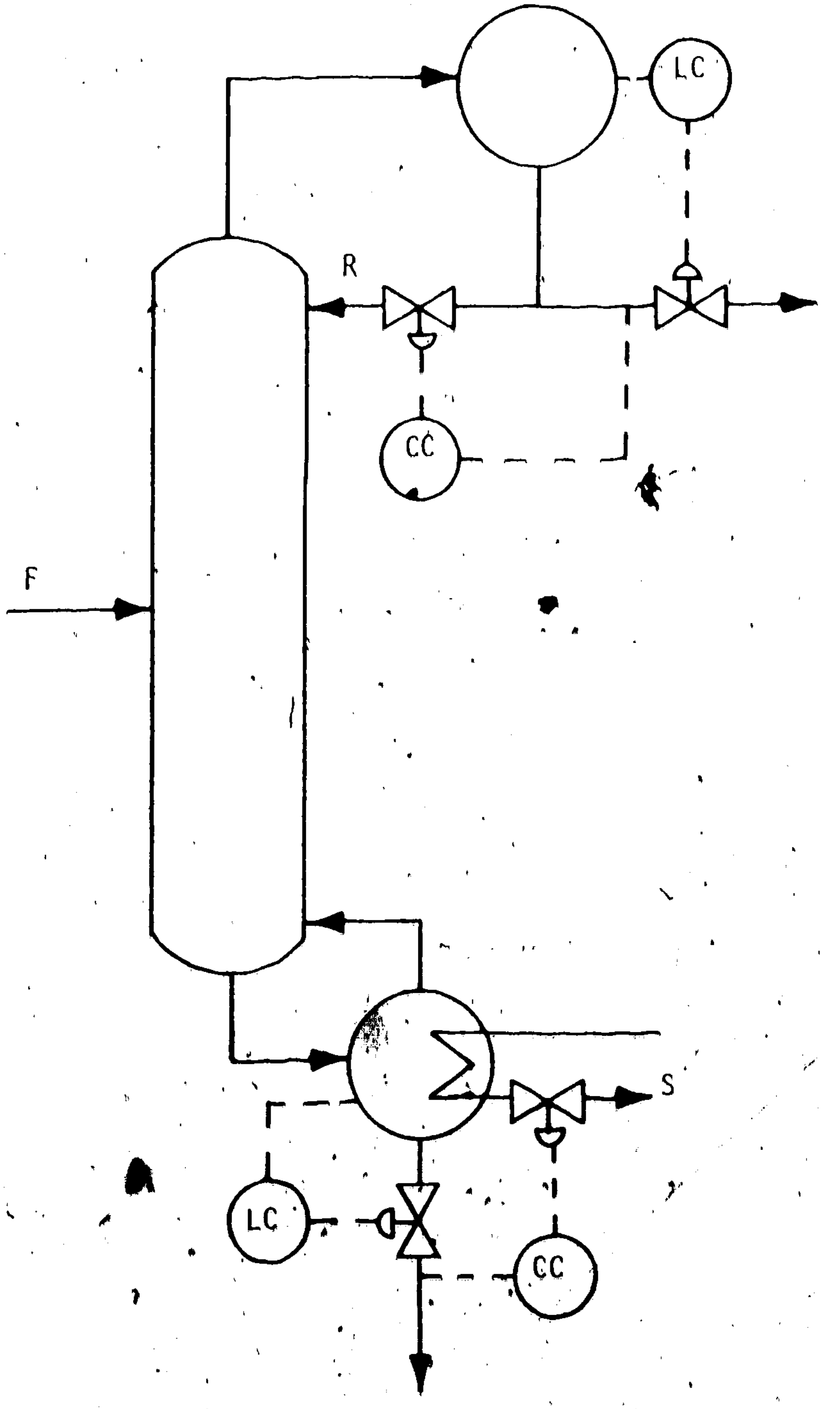

FIGURE 4.1

SCHEMATIC DIAGRAM OF THE CONVENTIONAL CONTROL SYSTEM

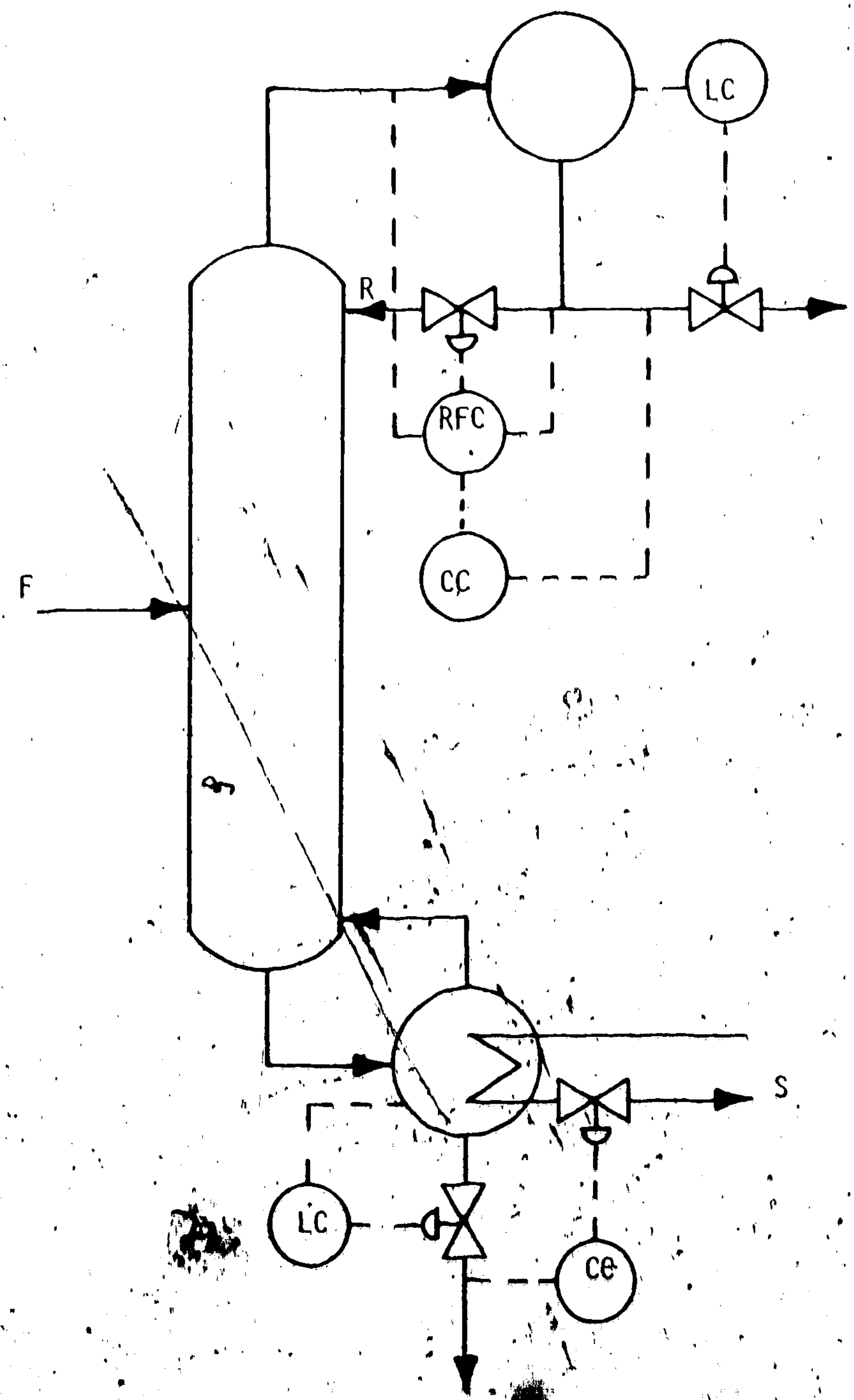

FIGURE 4.2

SCHEMATIC DIAGRAM OF THE RATIO CONTROL SYSTEM corresponding change in the reflux flow rate. Since the effect of the overhead vapour rate on the top composition reflects the effect of the reboiler heating rate on the overhead composition, the top composition loop has been rendered insensitive to control action in the bottom product loop. However, no account is taken of the effect of the reflux flow on the bottom product control loop. The effect of this omission on the performance of the control system remains to be evaluated.

4.2 MULTIVARIABLE CONTROL

A distillation column is a somewhat unique multivariable system due to the opposing actions of the two manipulated variables. Reflux flow has a positive effect on both end compositions; that is an increase in reflux flow will cause an increase in the more volatile components in both overhead and bottom compositions. Steam flow has the opposite effect as an increase in steam flow will cause a resultant decrease of the more volatile component in both end compositions. Consider a disturbance in feed flow or feed composition entering a distillation column where the top product composition is maintained constant by the reflux flow (top control loop) and the bottom product quality controlled by the steam flow to the reboiler (bottom control loop). Such a disturbance will have similar effects on both end compositions. If the disturbance is a step increase in feed flow for example, the top and bottom product compositions will increase. If the column is controlled as illustrated in Figure 4.1 the action of the controllers to maintain the end compositions at their required values would be to decrease the reflux rate and increase the steam rate. A decrease in reflux flow will cause a decrease in the bottom composition which is the desired action of the bottom control loop. Similarly an increase in the steam flow will cause a resultant decrease in the overhead composition which is also the objective of the top control loop. However, the above analysis assumes that no oscillations will occur in the end compositions. If oscillations do occur the interaction between the control loops could oppose the effect of the manipulated variable and thus sustain the oscillations. Therefore, it would appear that it would be extremely difficult to tune the control loops if the interacting effects are ignored. The logical solution to the above proplem is to design a control system that will compensate for the interaction. This would serve to alleviate the difficulties of loop tuning and prevent one control loop from being disrupted by the action of the other. However, the above analysis has illustrated that the overall effect of the interaction, if controlled, is to enhance the action of the individual control loops. It may then be suggested that a more "optimal" method is one in which the interacting effects are used to enhance the performance of the control system.

For the most part, the design techniques for multivariable control systems, reviewed in chapter 1, are concerned with servo control problem rather than the regulatory control problem. Only in a more recent study has Newell (76) applied optimal control theory to a regulatory control System with unmeasured load disturbances.

The approach for providing noninteracting control outlined by Zacklind (15) and Luyben (41), discussed in Chapter 1, appear to be the most applicable to the problem at hand. The basis of the method these workers proposed is to compensate each control loop for the control actions taken in the other loops of the system. In essence feedforward control is applied to compensate for the effects of the interaction.

The block diagram representation of the control system is shown in Figure 4.3. The process is assumed to be adequately represented by

$$\begin{bmatrix} x_1(s) \\ x_2(s) \end{bmatrix} = \begin{bmatrix} P_{11}(s) & P_{12}(s) \\ P_{21}(s) & P_{22}(s) \end{bmatrix} \begin{bmatrix} R(s) \\ S(s) \end{bmatrix}$$

where the P_{IJ} are a combination of a first order transfer function and a dead time. The process model used in the synthesis of the control equation is given in equation 3.1.

The design equations for the noninteracting compensator are constructed by the following analysis. The objectives of the compensator are to eliminate the effects of the top and bottom control actions on the overhead and bottom compositions respectively.

Therefore

$$D_{21}(s)\, P_{22}(s) + P_{21}(s) = 0$$

and

$$D_{12}(s)\, P_{11}(s) + P_{12}(s) = 0$$

in general

$$D_{IJ}(s)\, P_{II}(s) + P_{IJ}(s) = 0$$

Thus the compensators D_{IJ} may be constructed from the equation

$$D_{IJ}(s) = -P_{IJ}(s)/P_{II}(s) \qquad 4.1$$

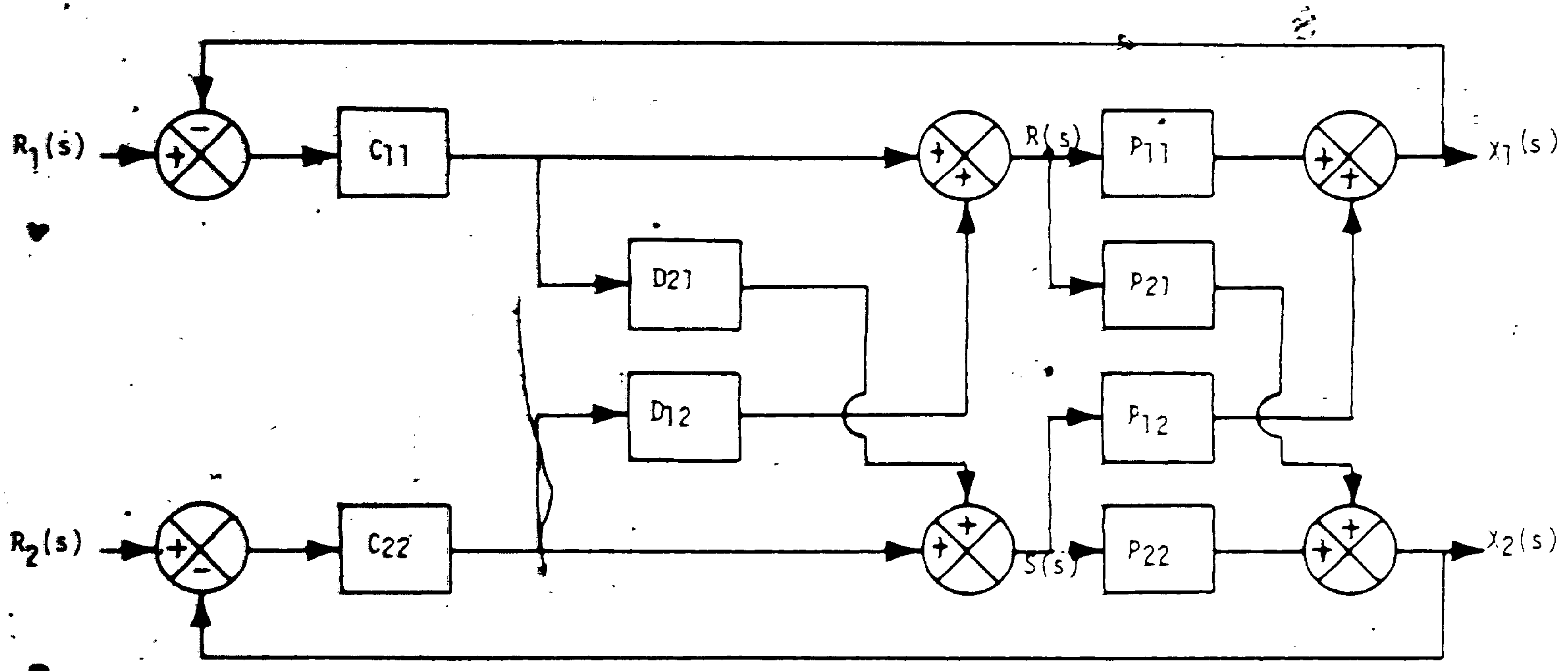

FIGURE 4.3

BLOCK DIAGRAM OF THE NONINTERACTING CONTROL SYSTEM

It is assumed in the above analysis that the dead time associated with the interacting function (P_{IJ}) is greater than or equal to the dead time associated with the direct transfer function (P_{II}). This is necessary for the compensator $\hat{D}_{IJ}$ to be physically realizable. This assumption is usually valid as one would expect the dead time of the interacting transfer function to be greater than that associated with the direct transfer function.

5. IMPLEMENTATION

The performance of the control systems described in Chapter 4 were evaluated on the pilot scale distillation column. Implementation of these control systems was accomplished utilizing the IBM 1800 digital computer. The control computations were performed in programs executed in variable core on the 1800 at a specified periodic interval. The DDC program was employed as a convenient interface between the control algorithms and the process. The computed control actions to correct any error in the end compositions were inserted into the outputs of the specified DDC loop records. The outputs of these DDC loop records were cascaded to the setpoints of the analog flow controllers. All control computations were performed at 2.5 minute intervals as dictated by the cycle time of the gas chromatograph.

5.1 RATIO CONTROL SYSTEM

The ratio control system studied was discussed in Chapter 4. In this system the bottom composition is controlled directly by a proportional plus integral (PI) controller manipulating the steam flow rate. The overhead composition on the other hand is controlled by cascading the output of the PI composition controller to the setpoint of a ratio controller. The ratio controller maintained the required ratio of overhead vapour rate to reflux flow rate. Since a direct measurement of the overhead vapour flow was not available, it was necessary to estimate the flow rate from a material balance about the condenser. If the level in the condenser is assumed to be constant, the vapour rate may be computed as the sum of the reflux and top product flows. The assumption of a constant level in the condenser is not unreasonable as the level was under "tight" control during these studies. In addition, the period between control actions allowed sufficient time for the level to recover from any sudden changes in reflux flow made in the previous control interval. An average value of the top product flow rate was computed over the 15 seconds prior to the control action, using data accumulated, once per second, in a DDC ring buffer. Once an estimate of the overhead vapour rate was computed, the correct reflux flow could be computed on the basis of the specified ratio.

5.2 NONINTERACTING CONTROL SYSTEM

The block diagram representation of the noninteraction control system employed was shown in Figure 4.3. In this scheme the outputs of the proportional integral controllers are fed to the noninteracting compensators. The function of the compensators is to compute the necessary action to compensate for the effects of the single loop control actions on the other loop. The output to the process is the sum of a compensator output and the control action of the proportional plus integral controller. To facilitate programming on the digital computer, it was desirable to express the required compensator (Equation 4.1) in a different equation format. Since a digital computer in a control system can be treated as a sampled data controller with a zero order hold device on the output, it is necessary to combine the process transfer functions of Equation 3.1 with the transfer function of a zero order hold in computing the sampled data form of the noninteracting compensator.

It follows that

$$H(s)\, P_{IJ}(s) = \frac{1 - e^{-sT}}{s} \; \frac{K_{IJ}\, e^{-L_{IJ}s}}{T_{IJ}\, s + 1}$$

taking the z transform and substituting $a_{IJ} = \frac{1}{T_{IJ}}$

$$H\, P_{IJ}(z) = \frac{K_{IJ}\, z^{-L_{IJ}} \left[1 - e^{-a_{IJ}T} \right]}{z - e^{-a_{IJ}T}}$$

In terms of the z transformed elements the compensator becomes

$$D_{IJ}(z) = -\, H\, P_{IJ}(z) / H\, P_{II}(z)$$

$$D_{IJ}(z) = \frac{B\, z^{-L} \left[z - e^{-a_{II}T} \right]}{z - e^{-a_{IJ}T}} \qquad 5.1$$

where $L = L_{IJ} - L_{II}$

$$B = -\frac{K_{IJ} \left[1 - e^{-a_{IJ}T} \right]}{K_{II} \left[1 - e^{-a_{II}T} \right]}$$

in difference equation format Equation 5.1 becomes

$$w_o(n) = B\, w_i(n-L) - B\, e^{-a_{II}T}\, w_i(n-L-1) - e^{-a_{IJ}T}\, w_o(n-1) \qquad 5.2$$

where w_o is the compensator output and w_i the input to the compensator.

The above equation is restricted in that the delay time must be an integer number of sample periods. This is not desirable if a long sample period is involved. To circumvent this restriction, the output was computed in two time steps. The time delay may be subdivided into two intervals, the first interval representing the fractional sample period delay, the second representing the whole number of sample periods remaining in the time delay. The compensator output can then be computed in two time steps using the output after the first time step as an initial condition for the second time step.

The numeric formulation of the noninteracting compensator and the ratio controller are outlined in Appendix 4. In addition the sensitivity of the estimated overhead vapour rate (by a material balance about the condenser) to changes in the condenser level is examined in Appendix 4.

6. PROGRAMS

The control studies and associated data acquisition were performed by a number of programs executed on the IBM 1800 computer. For the purpose of description these programs may be divided into four categories (online programs, initialization and data retrieval programs, material balance programs, and auxilliary programs). A description of the auxilliary program category is given in Appendix 1.

6.1 ONLINE PROGRAMS

MAINLINE CHRTM

The online programs performed the gas chromatograph data analysis and the control computations. These functions were controlled by the interrupt core load CHRTM, while the separate functions were performed by individual subroutines.

On entry to CHRTM the ring buffer acquiring data from the gas chromatograph is deactivated and an ECO is closed to initiate a new sample injection to the GC. The core address of the ring buffer containing the GC data is located and the data loaded into a vector for analysis. Following computation of the composition as an area percent by subroutine GC the appropriate calibration is applied to convert the result to weight percent. The ECO is then opened to complete the sample injection phase and the ring buffer is activated. Subroutine CNTRL is called to implement the control action.

SUBROUTINE GC

Subroutine GC is called by CHRTM to analyze the accumulated data from the gas chromatograph. The start and end points of each chromatograph peak are located by subroutine PEAK. The area under each peak is computed using Simpson's rule. If it is determined that the base line is reached at a peak end, base line corrected areas are computed for all peaks occurring since the last time an "on the base line" condition was detected. The base line corrected areas were computed by fitting a least squares line to adjacent sets of base line data. The superfluous area under the base line is then subtracted from the corresponding peak area. The corrected peak areas are assigned to the designated time bands and the area percent associated with each time band computed.

SUBROUTINE PEAK

Subroutine PEAK is called from subroutine GC to locate the start and end of each peak in the chromatogram. A peak start is recognized when the first derivative of the data becomes greater than a specified deadband. This event must be confirmed for a given number of successive data points (3) before a peak start is accepted. Similar logic is employed to locate the end of a peak; that is, the computed first derivative must become more positive than a specified deadband. This event must also be confirmed a designated number of times (3) before a firm decision is made. A test is made to determine if the base line is reached at the end of each peak.

The derivatives used in peak detection are computed from the least squares formula given by Savitzky (64). The seven point formula used in this routine is

$$\begin{aligned} Y' = {} & (3\,Y(I+3) + 2Y(I+2) + Y(I+1) \\ & - Y(I-1) - 2Y(I-2) - 3Y(I-3))/28H \end{aligned}$$

line is reached at a peak end, base line corrected areas are com-
for all peaks occurring since the last time an "on the base line"
:ion was detected. The base line corrected areas were computed
:ting a least squares line to adjacent sets of base line data.
iperfluous area under the base line is then subtracted from the
sponding peak area. The corrected peak areas are assigned to
:signated time bands and the area percent associated with each
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and end of each peak in the chromatogram. A peak start is
ized when the first derivative of the data becomes greater than
:ified deadband. This event must be confirmed for a given number
:cessive data points (3) before a peak start is accepted. Similar
is employed to locate the end of a peak; that is, the computed
derivative must become more positive than a specified deadband.
event must also be confirmed a designated number of times (3)
a firm decision is made. A test is made to determine if the
line is reached at the end of each peak.

The derivatives used in peak detection are computed from the
squares formula given by Savitzky (64). The seven point formula
in this routine is

$$Y' = (3\,Y(I+3) + 2Y(I+2) + Y(I+1) - Y(I-1) - 2Y(I-2) - 3Y(I-3))/28H$$

in CNTRL. An average value of the top product flow rate is computed from data obtained in a ring buffer. The overhead vapour rate is then computed as the sum of the reflux and top product flow rates. This new value of the overhead vapour rate is used to compute the new reflux flow rate necessary to yield the specified ratio.

SUBROUTINE NICMP

Subroutine NICMP was employed during the noninteracting control studies to compute the noninteracting compensation from equation 5.2. The output of the proportional integral controllers were fed to subroutine NICMP which calculated the compensating action based on these inputs. The control output to the process is then computed as the sum of a compensating term and a control action term.

To circumvent the restriction of an integer sample period delay in equation 5.2 the compensation was computed in two time steps as outlined in section 5.2.

6.2 INITIATION AND DATA RETRIEVAL PROGRAMS

MAINLINE CSTRT

CSTRT was used to perform the following functions on request from the 1816 keyboard.

1) Initiate repetitive execution of the gas chromatograph/control program CHRTM.
2) Terminate repetitive execution of CNTRL.
3) Initiate control action by setting a flag that informs CNTRL that the control action is to be implemented. The disk files are also initialized to start the control run.

4) Terminate control action.

5) Change controller constants and/or model parameters.

MAINLINE DUMP

The nonprocess program DUMP was employed to retrieve the data from the disk files at the end of a control test. Measurements of the top and bottom compositions and controller outputs are dumped to the line printer or both the line printer and cards according to a specified option.

6.3 MATERIAL AND ENERGY BALANCE PROGRAMS

MAINLINE DASS

DASS initiates the data acquisition for the material and energy balance computations. The current feed composition and bottom product composition are entered via the 1816 keyboard. The disk files are initialized and repetitive queuing of the data acquisition program DATAC at a specified interval is initiated.

MAINLINE DATAC

Program performs the data acquisition for the material and energy balance calculations. The program is queued a specified number of times at a given interval. A running total and sum of squares of the measurements are stored in the disk file.

The data acquisition is accomplished by reading the multiplexer point directly and applying the appropriate conversion routines to the ADC reading. Three different calculation options are available, flow calculation, zero check, and thermocouple conversion (with cold junction compensation). These calculation routines parallel those in the DDC program.

The cold junction compensation and temperature conversion were accomplished following the example outlined in the IBM 1800 Functional Characteristics Manual (73). Thermocouple calibration data is stored as a series of straight line segments in the program.

MAINLINE BALNC

BALNC uses the data acquired by DATAC to complete the material and energy balance calculations. The average and variance of the accumulated data is computed. The flow rates are corrected for temperature and composition. The material balance and a summary of the operating conditions are printed on the line printer.

7. RESULTS

The object of the control phase of the present study was to compare experimentally different methods for dual product quality control of a distillation column. The performances of the control systems discussed in Chapter 4, that is ordinary proportional integral control, overhead composition control by manipulation of the reflux to overhead vapour rate ratio and the noninteracting control system were studied on the pilot scale distillation column. The control loops were tuned for [illegible] integral of the absolute error (IAE). This enabled a comparison to be drawn between the ability of the different systems to compensate for a disturbance to the column. The control runs performed are summarized in Table 7.1. The steady state operating conditions for all control runs are documented in Appendix 2.1.

7.1 SINGLE LOOP CONTROL

The performances of the individual control loops that is proportional integral control of only the top composition and proportional integral control of only the bottom composition were studied. This allowed a comparison to be drawn between the performance of the single loop control with the behaviour of the system with both loops operative which results in interaction. The two control loops were tuned individually for a minimum IAE response to a disturbance. Figures 7.1 and 7.2 illustrate the performance of these two single loop control systems following disturbances in the feed flow rate. The performance of the single loop control systems even with the associated long sampling periods and dead time is quite adequate. The larger IAE term associated with the bottom composition control is due largely to the relative

TABLE 7.1

SUMMARY OF CONTROL RUNS

RUN NUMBER	DISTURBANCE IN FEED FLOW (lbs/min)	CONTROL SCHEME
S-1	.34	Control of only top composition
S-2	.34	Control of only bottom composition alone
PI-1	.34	Proportional-Integral control of both end compositions
PI-2	-.36	"
PI-3	.34	"
PI-4	-.36	"
PI-5	-.37	"
R-1	.34	Ratio control of overhead composition
R-2	-.39	"
R-3	-.38	"
R-4	.34	"

Table 7.1 (cont'd)

RUN NUMBER	DISTURBANCE IN FEED FLOW (lbs/min)	CONTROL SCHEME
C-1	.34	Noninteracting control
C-2	- .39	"
C-3	- .40	"
C-4	.33	"

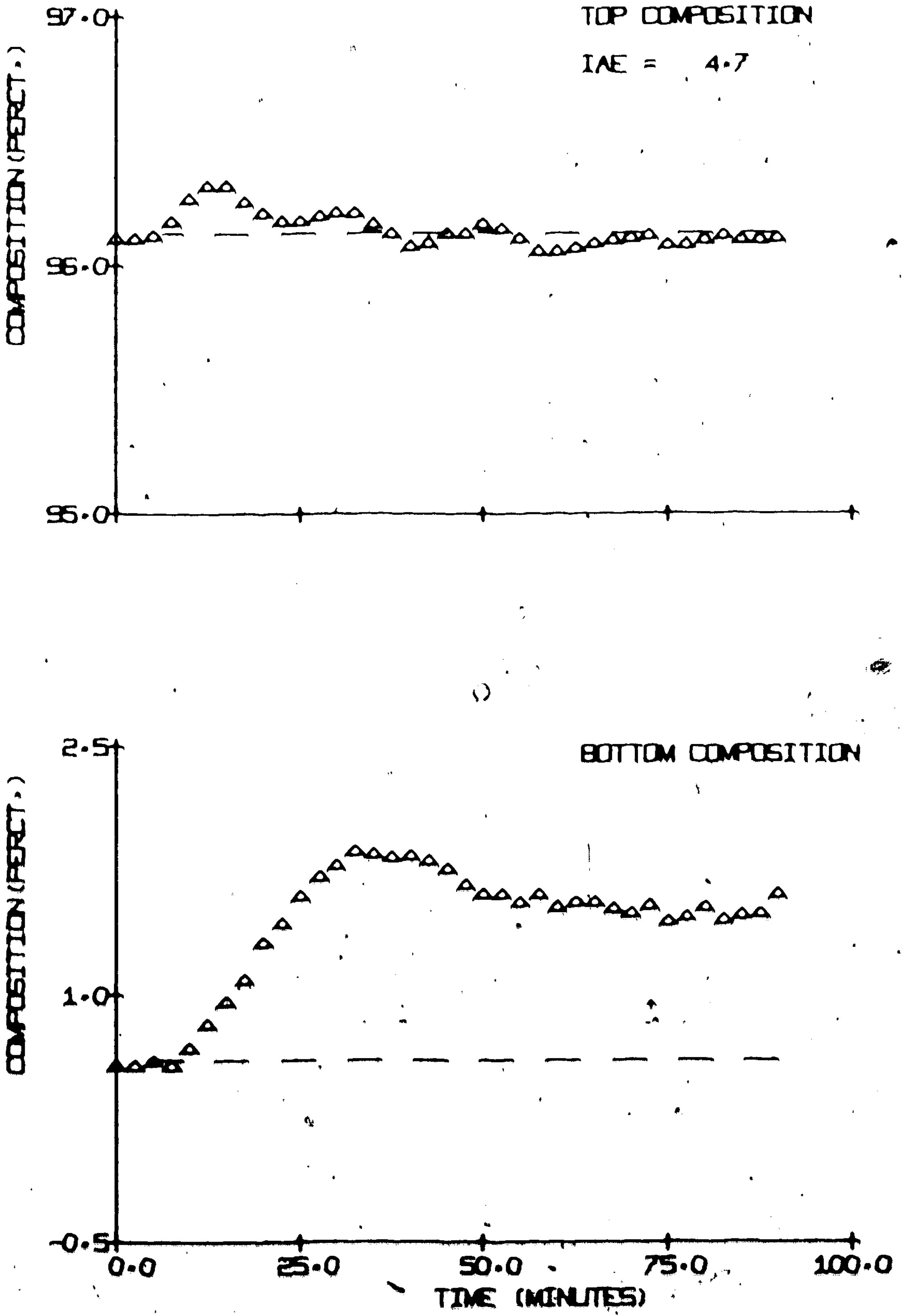


FIGURE 7.1 CONTROL OF ONLY THE TOP COMPOSITION [15% INCREASE IN FEED FLOW, RUN S-1]

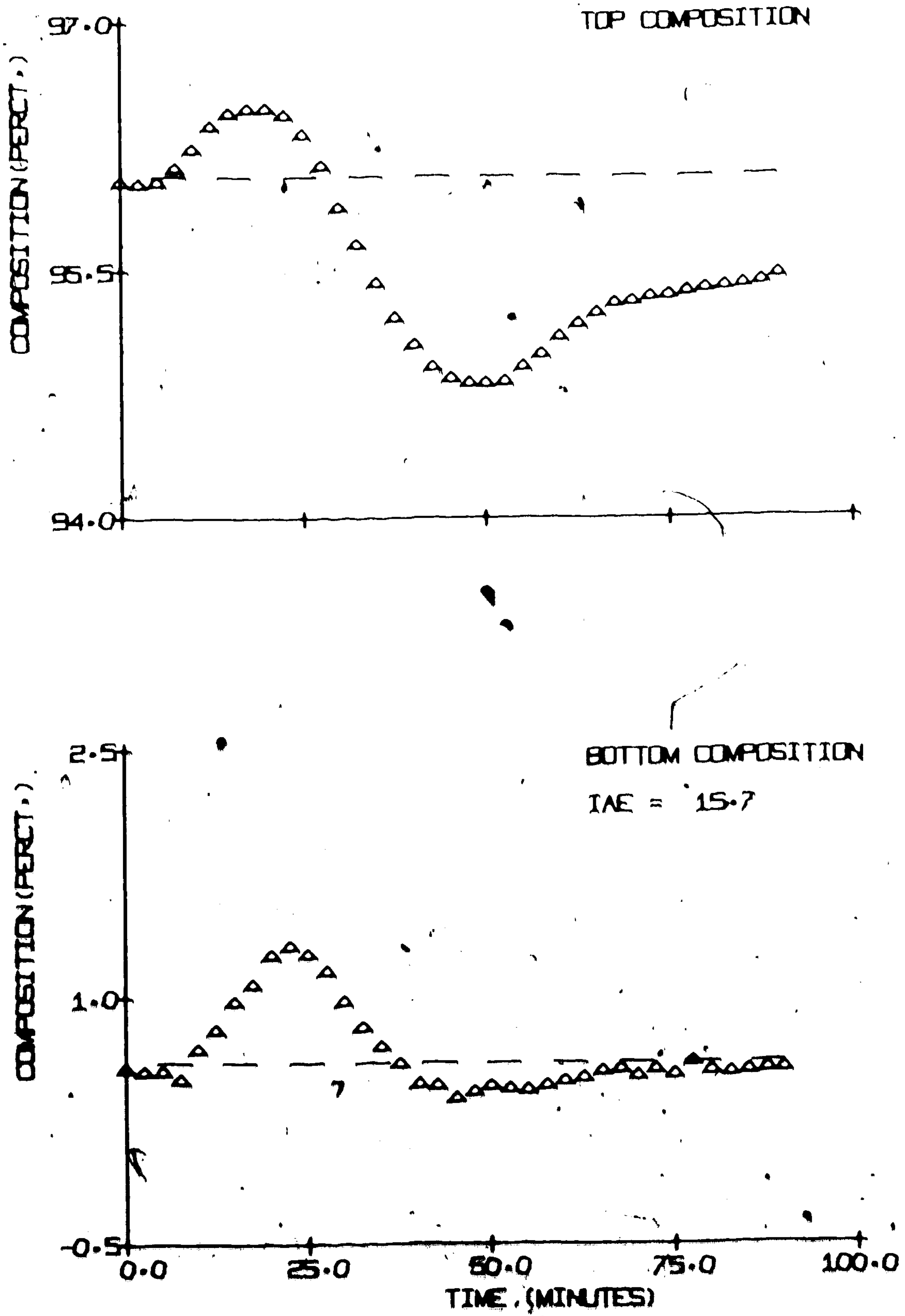


FIGURE 7.2 CONTROL OF ONLY THE BOTTOM COMPOSITION [15% INCREASE IN FEED FLOW, RUN S-2]

instability of the bottom composition control loop. The rather "drastic" effect of the bottom product control on the overhead product quality is the most important observation to be drawn from these tests. This observation underlines the deleterious effect of the interaction on the controllability of the overhead product composition.

7.2 PROPORTIONAL PLUS INTEGRAL CONTROL OF BOTH PRODUCT COMPOSITIONS

From observations of the results in the previous section one would expect the interaction if neglected, to cause a serious deterioration in the performance of the control system when an attempt is made to control both of the end compositions. In order to establish if indeed this was the case an attempt was made to control both of the product qualities using the controller settings determined for control of only one composition. The effect of attempting simultaneous control of both product compositions on the performance of the control system is immediately evident from Figures 7.3 and 7.4. The sustained oscillations that result are the predominant characteristic of the system's response to a disturbance. No IAE values are given for these tests as the system did not reach a steady state in a reasonable time. An IAE comparison would thus be meaningless. The disturbing effect of the interaction, steam rate, on the overhead composition is clearly evident from results given in Figures 7.3 and 7.4. As the overhead product composition, under the control of the reflux flow, begins to recover from the initial disturbance, a second disturbance effects the control loop causing a large deviation of the composition in the opposite direction to the initial disturbance. This second disturbance is the effect of the control of the bottom composition, i.e. steam rate on the top composition. The end result is a sustained oscillation.

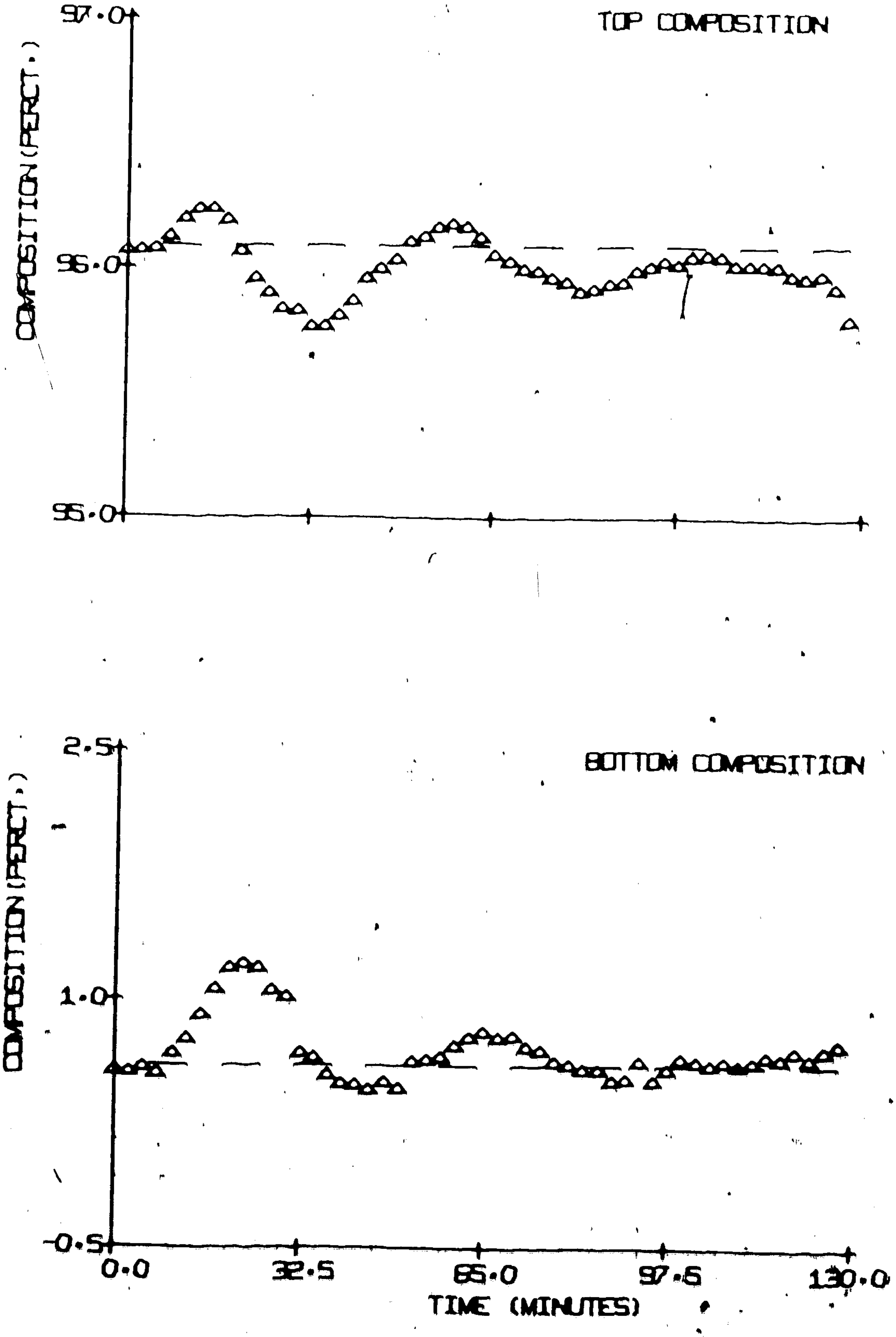
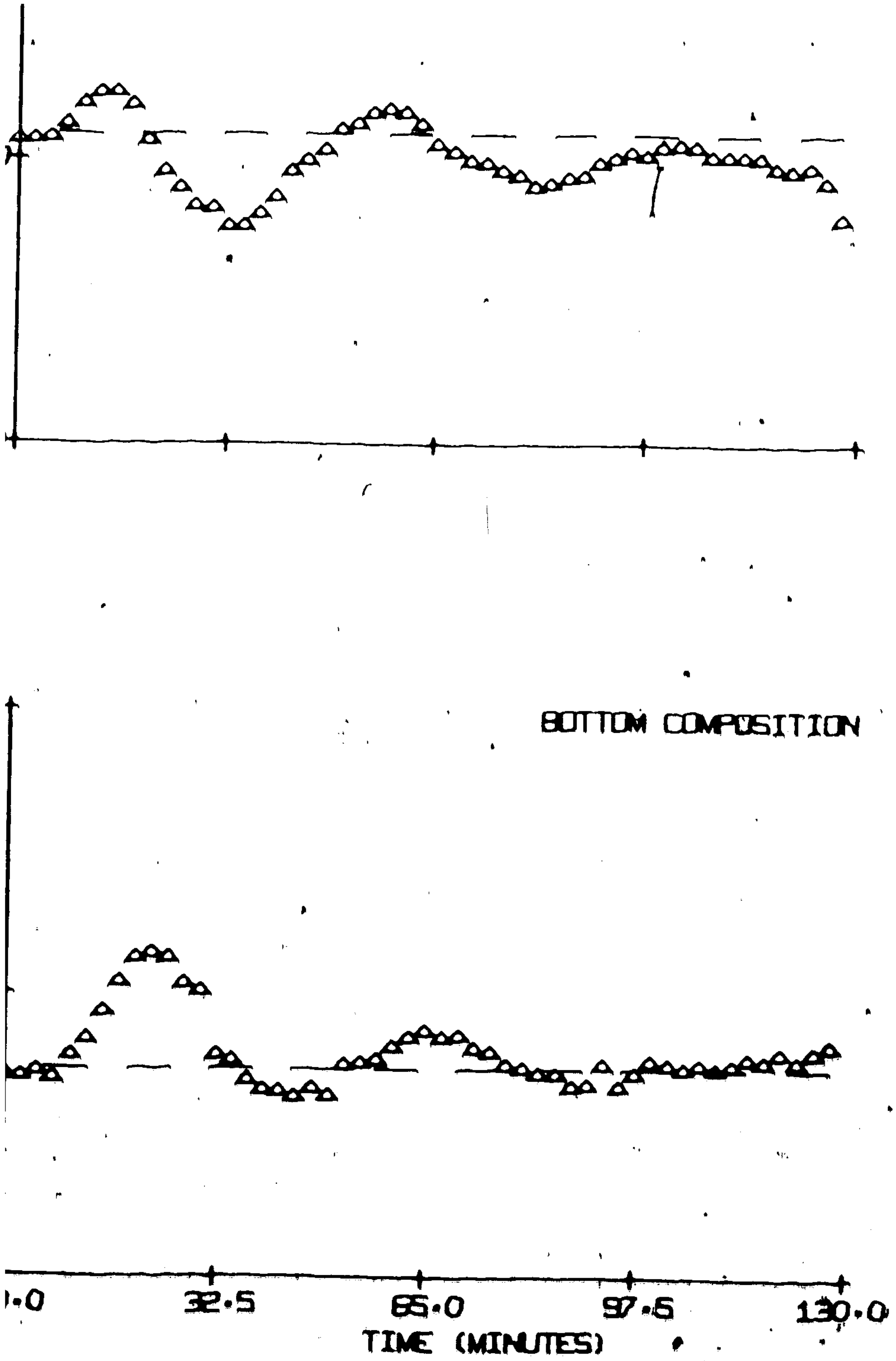


FIGURE 7.3 PROPORTIONAL INTEGRAL CONTROL OF BOTH END COMPOSITIONS
[15% INCREASE IN FEED FLOW, RUN PI-1]



RE 7.3 PROPORTIONAL INTEGRAL CONTROL OF BOTH END COMPOSITIONS
[15% INCREASE IN FEED FLOW, RUN PI-1]

The performance of the system is clearly undesirable. On examination of the control behaviour of the loops it can be reasoned that a much "slower" control is necessary especially in the bottom composition control loop if the oscillations are to be reduced. On the basis of this reasoning an attempt was made with both control loops operating to tune the individual proportional integral controllers for a minimum IAE recovery from a disturbance. Figures 7.5, 7.6 and 7.7 illustrate the slow control that is necessary to achieve a satisfactory recovery from a disturbance. A complete tabulation of the controller constants employed in the control tests is given in Table 7.2. A substantial reduction in control action in the bottom loop was necessary. The proportional and integral modes in the bottom control loop were 3.0 and 2.2 times less respectively than the corresponding constants employed when only the bottom composition controller was active. This large reduction in bottom loop control action is necessary to prevent the overhead composition control loop from being overly disrupted by the interaction. Recall that equation 3.1 shows that the overhead composition is more sensitive to steam rate than reflux rate. The reduction in controller constants is especially important in the case of the proportional mode and less important with regard to the integral term as sudden changes in the steam rate are less acceptable than a gradual change.

In addition to reduced control action in the bottom control loop, it was also necessary to reduce the controller constants in the overhead composition controller to prevent a sustained oscillation. The associated proportional and integral constants were 1.7 and 2.2 times less respectively than when only the overhead composition

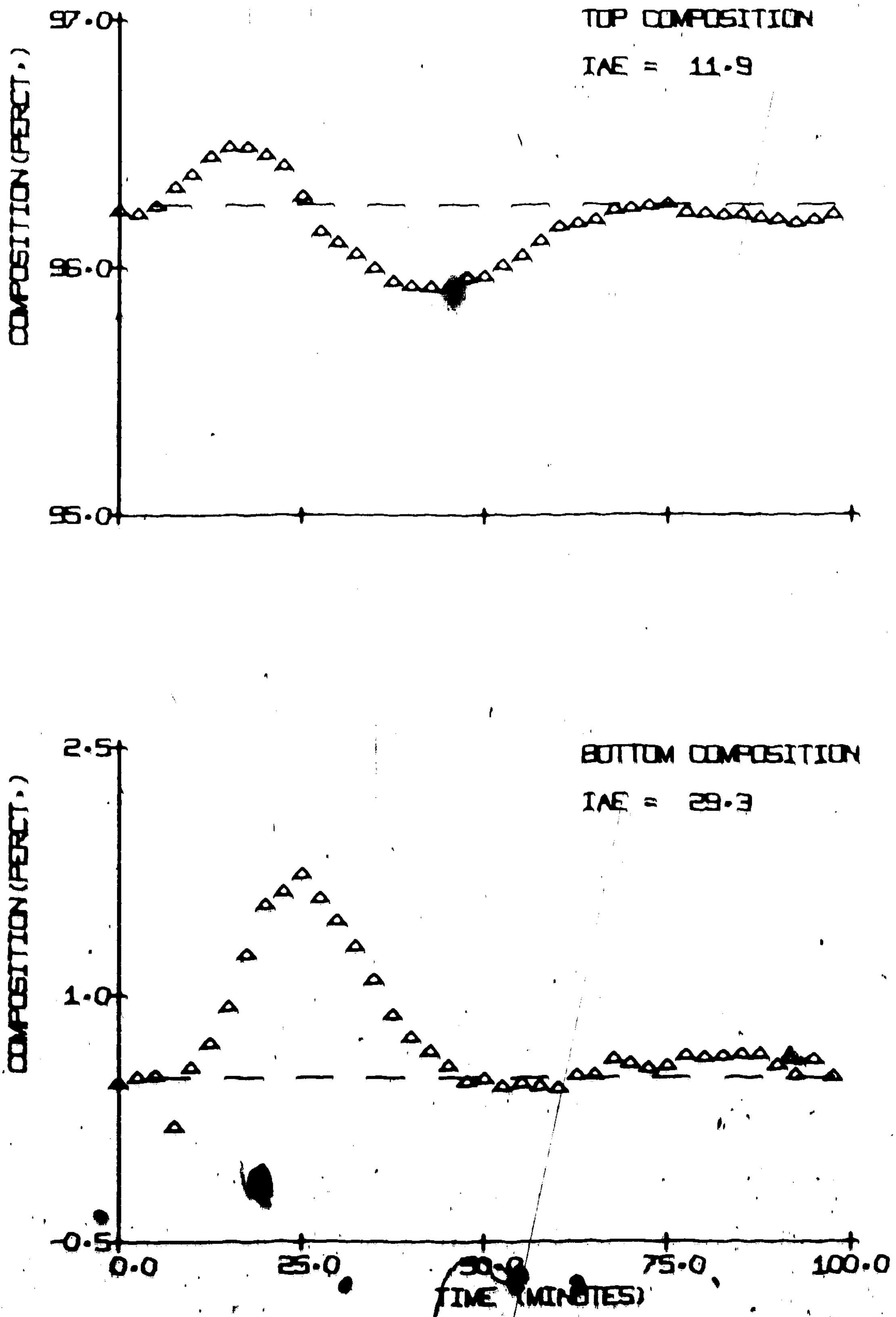


FIGURE 7.5 PROPORTIONAL INTEGRAL CONTROL OF BOTH END COMPOSITIONS [15% INCREASE IN FEED FLOW, RUN PI-3]

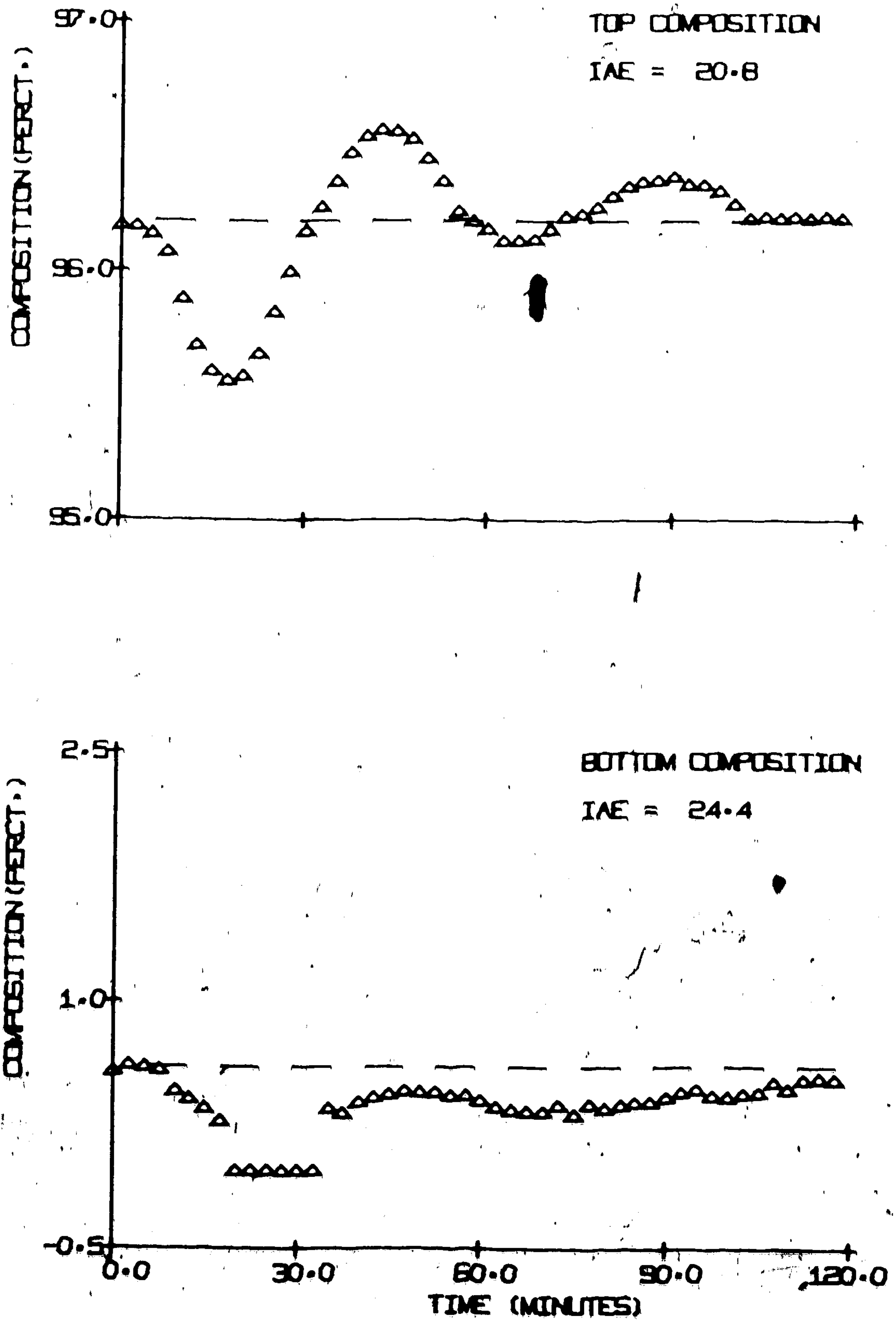


FIGURE 7.6 PROPORTIONAL INTEGRAL CONTROL OF BOTH END COMPOSITIONS [15% DECREASE IN FEED FLOW, RUN PI-4]

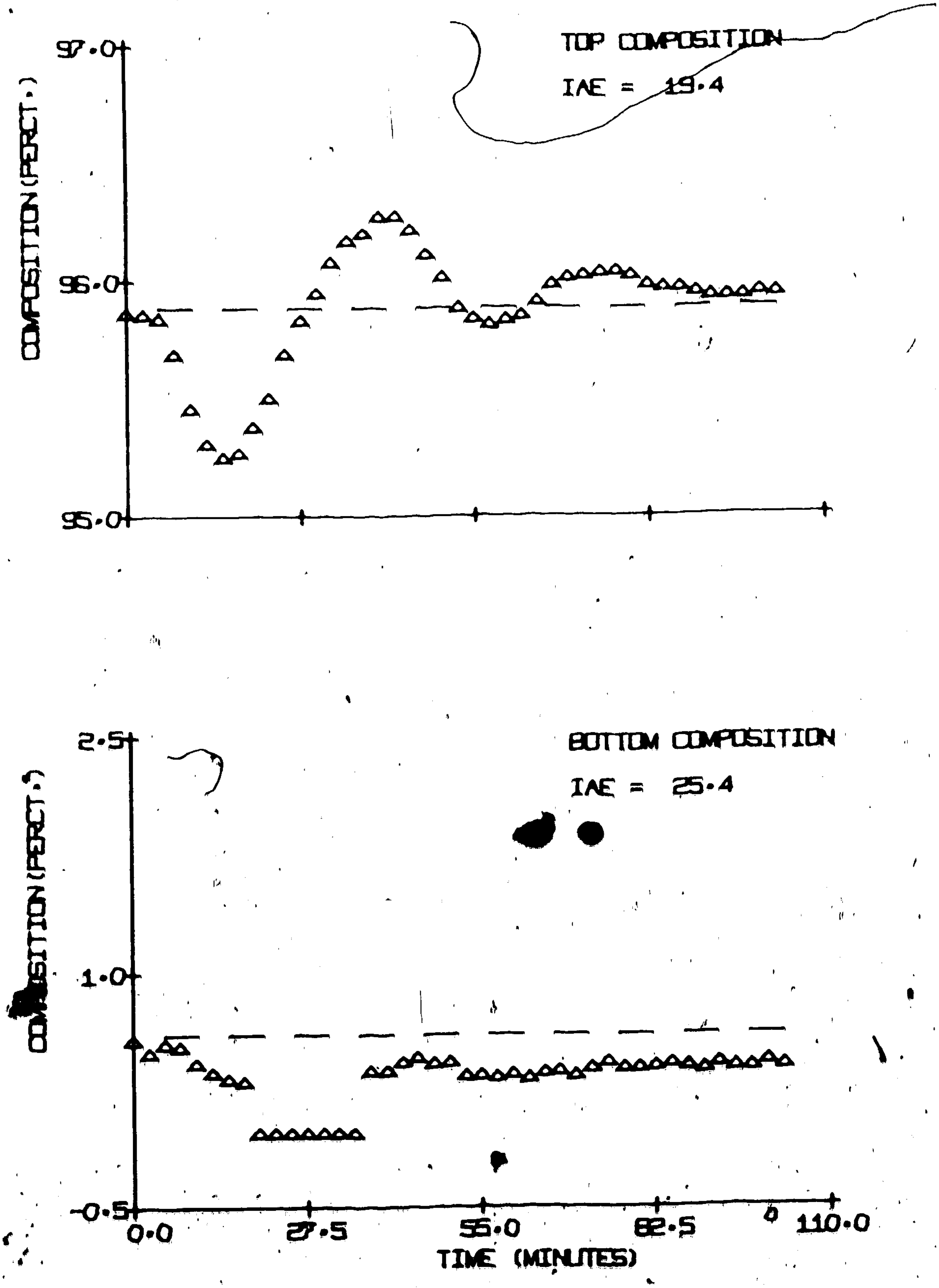


FIGURE 7.7 PROPORTIONAL INTEGRAL CONTROL OF BOTH END COMPOSITIONS [15% DECREASE IN FEED FLOW, RUN PI-5]

TABLE 7.2

RUN NO.	DISTURBANCE (lbs/min)	TOP LOOP K_p	TOP LOOP K_I	BOTTOM LOOP K_p	BOTTOM LOOP K_I	IAE TOP LOOP	IAE BOTTOM LOOP
S-1	.34	-.35	-.10	0.	0.	4.7	-
S-2	.34	0.	0.	.12	.035	-	15.8
PI-1	.34	-.35	-.10	.12	.035	-	-
PI-2	-.36	-.35	-.10	.12	.035	-	-
PI-3	.34	-.20	-.045	.040	.015	11.9	29.3
PI-4	-.36	-.20	-.045	.040	.015	20.8	24.4
PI-5	-.37	-.20	-.045	.040	.015	19.4	25.4
R-1	.34	-.090	-.015	.10	.025	5.1	18.5
R-2	-.39	-.090	-.015	.10	.025	11.8	17.4
R-3	-.38	-.090	-.015	.10	.025	12.9	25.7
R-4	.34	-.090	-.015	.10	.025	5.8	21.3
C-1	.34	-.35	-.10	.12	.035	4.1	16.7
C-2	-.39	-.35	-.10	.12	.035	14.6	17.0
C-3	-.40	-.35	-.10	.12	.035	14.6	17.0
C-4	.33	-.35	-.10	.12	.035	4.1	19.9

controller was active.

A comparison of the IAE figures for the control runs made with only one control loop active against those obtained when both control loops were active clearly shows a deterioration in performance in the latter case. Indeed, the IAE approximately doubles when both control loops are active.

7.3 RATIO CONTROL

In this control system the overhead composition is controlled by means of the reflux to overhead vapour rate ratio. As discussed in section 4.1 this ratio control of the overhead composition reduces the sensitivity of this composition to the interaction. It is at once evident from the results illustrated in Figures 7.8, 7.9, 7.10 and 7.11 that the performance of this control system is a vast improvement over that attainable with conventional proportional integral control. It can be seen from these results that the overhead composition has been rendered insensitive to control action taken in the bottom loop. This observation can be drawn when Figures 7.8 - 7.11 are compared with Figures 7.3 - 7.7 where the overhead composition appeared to be affected by a second disturbance due to the interaction as outlined in section 7.2.

A comparison of the IAE's resulting from tests R-1 to R-4 against those obtained with ordinary PI control in runs PI-3 to PI-5 ~~show that~~ this ratio control system has succeeded in reducing the IAE associated with the top composition loop by a factor of about 2. In addition, an improvement is also noted in control of the bottom composition. This improvement is due to the faster control that may be employed in the bottom control loop. This increase in control action

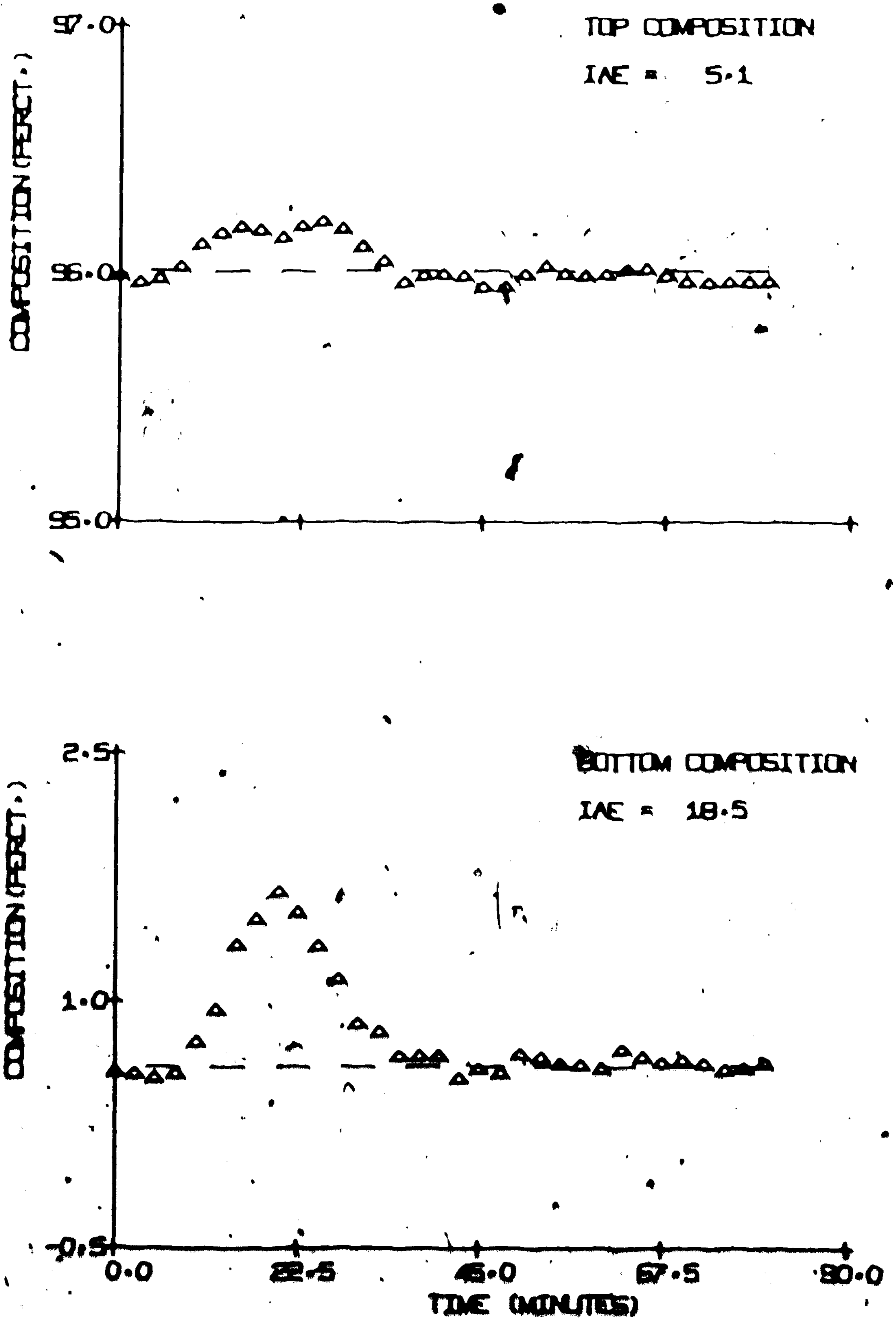


FIGURE 7.8 RATIO CONTROL OF THE TOP COMPOSITION [15% INCREASE IN FEED FLOW, RUN R-1]

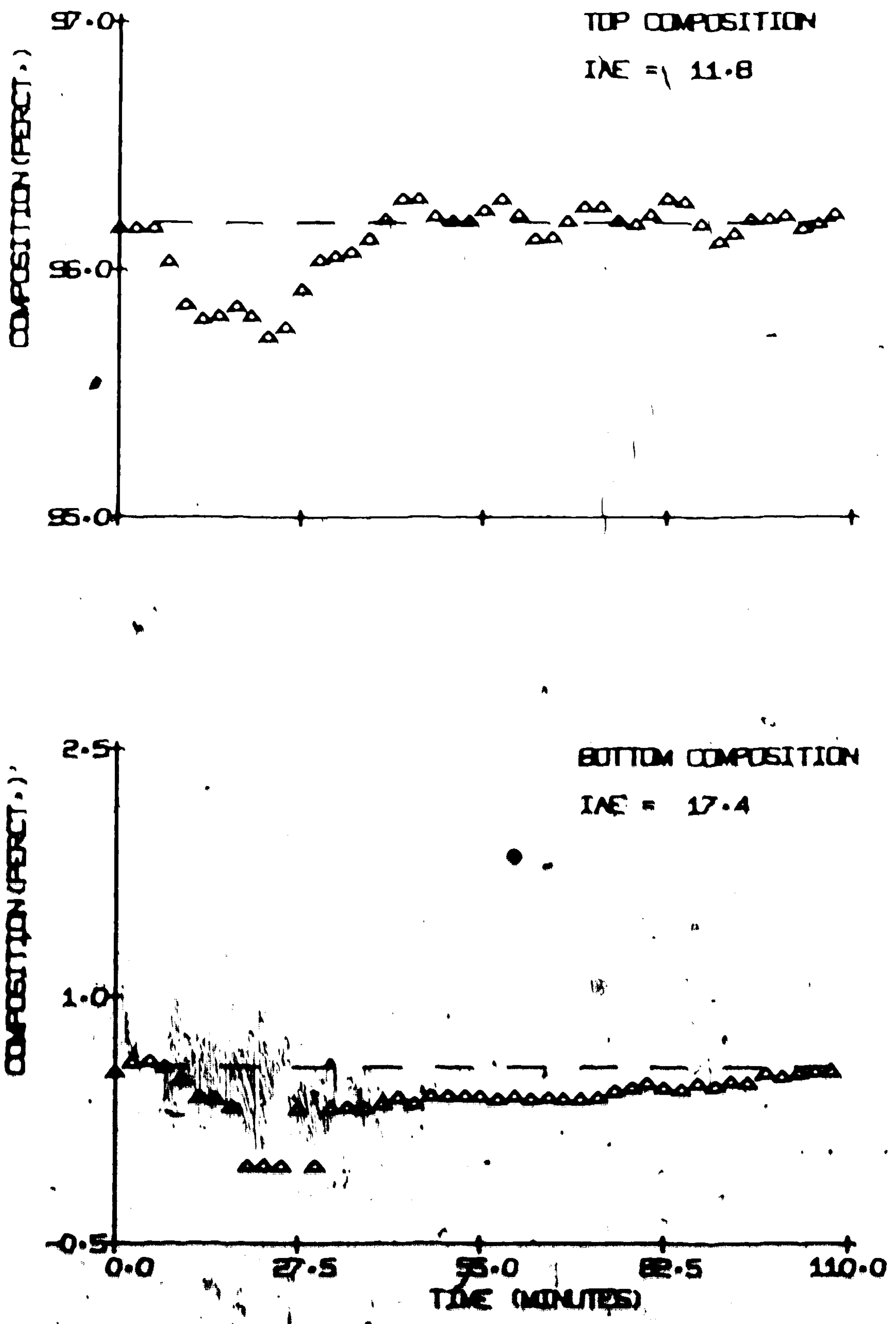


FIGURE 4.9 RATIO CONTROL OF THE TOP COMPOSITION [15% DECREASE IN FEED FLOW, RUN R-2]

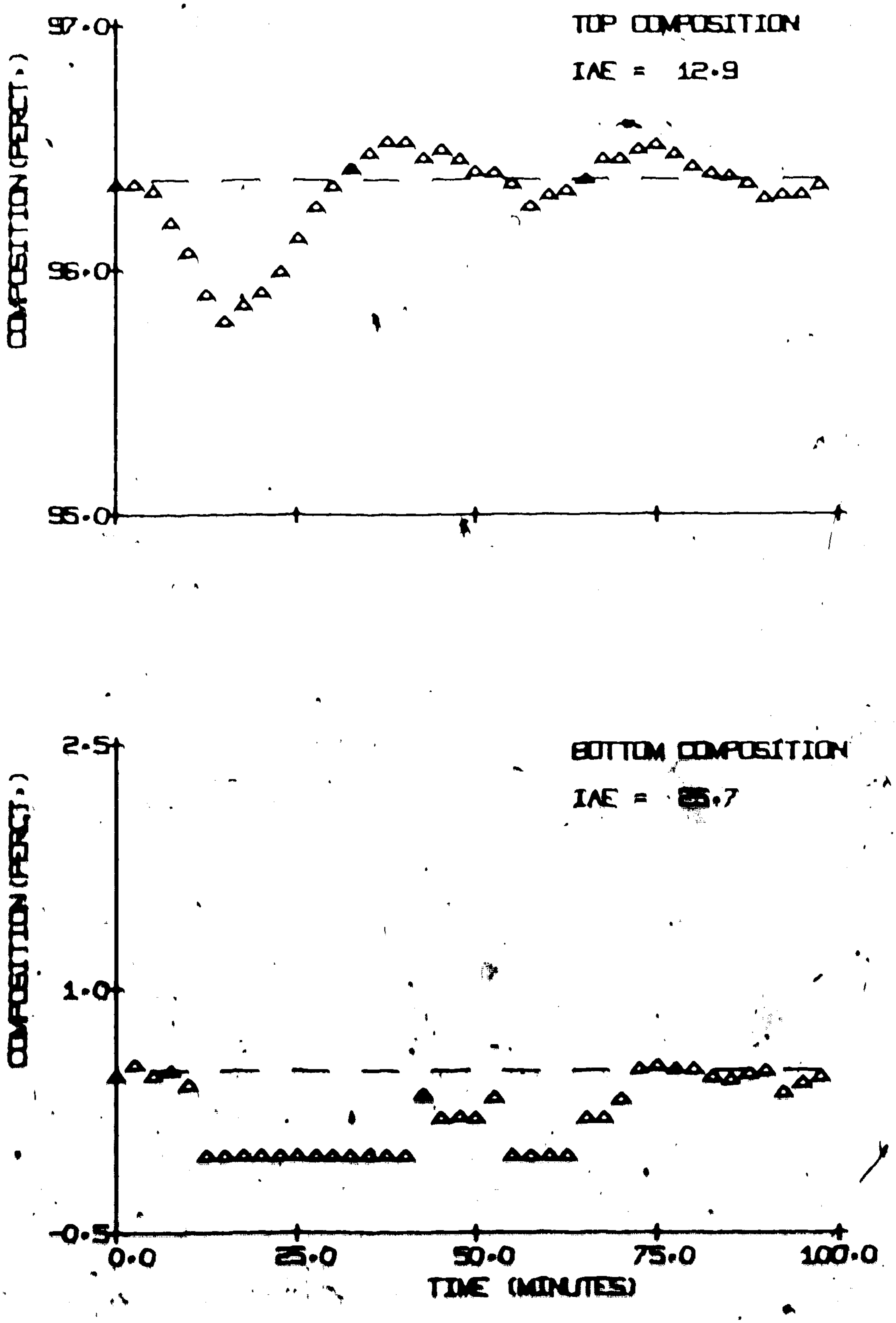


FIGURE 7.10 RATIO CONTROL OF THE TOP COMPOSITION [15% DECREASE IN FEED FLOW, RUN R-3]

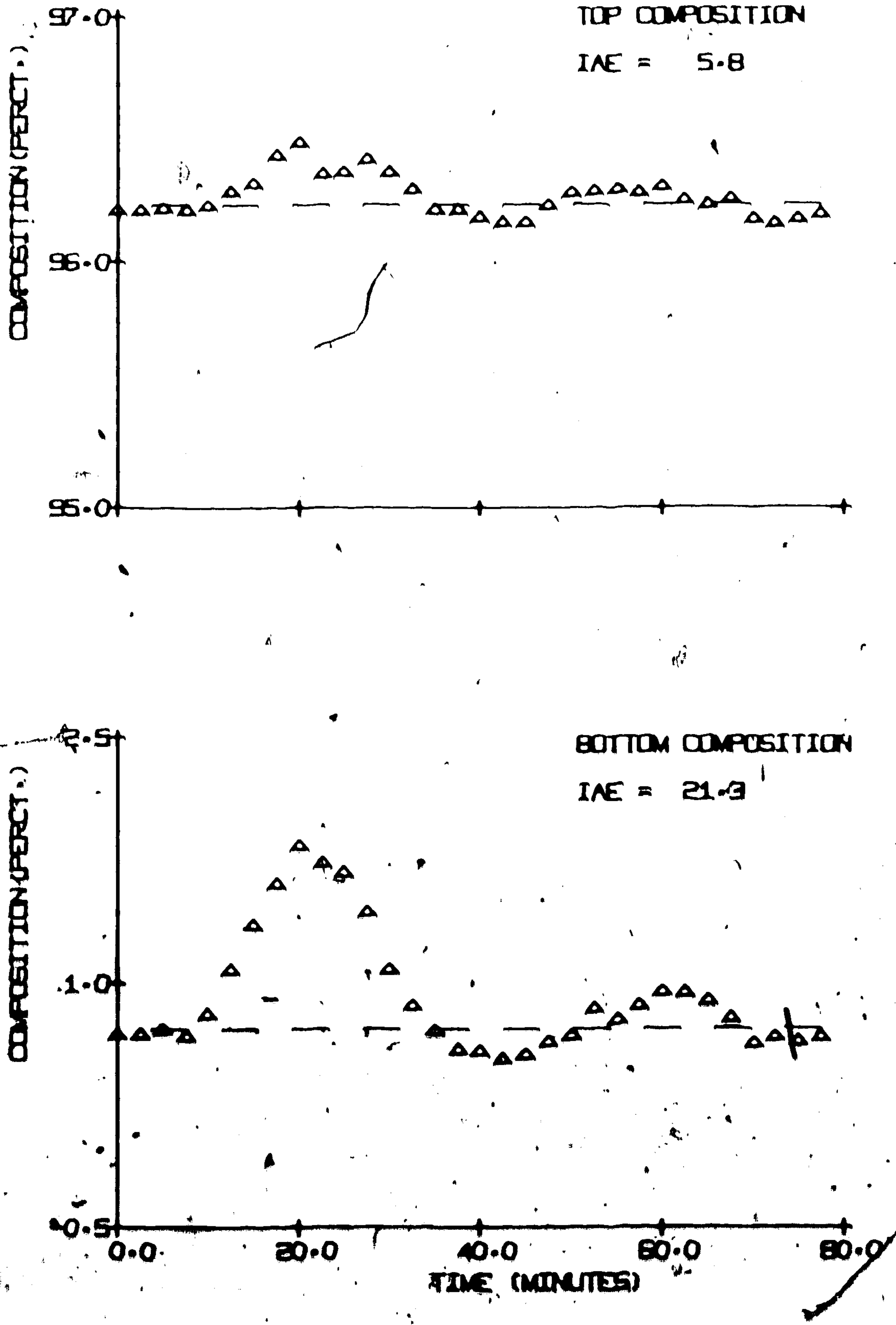


FIGURE 7.11 RATIO CONTROL OF THE TOP COMPOSITION [15% INCREA[illegible] IN FEED FLOW, RUN R-4]

that may be taken in the bottom loop can be attributed to ratio control of the overhead composition. Care should be taken when comparing the performances of the different control systems on the basis of the IAE values for the bottom control loop as the composition analysis becomes unreliable below 0.3% weight percent methanol.

A potential drawback of the ratio control system is that this system ignores the effects of control action in the top loop on the bottom composition. This does not appear to seriously effect the controllability of the column used in this study, but it should be noted that the reflux has a relatively small effect on the bottom product composition in this case. Also the stability of the bottom loop is such that tight control is impossible. This makes is difficult to evaluate the effect of the interaction on controllability in this loop. Thus it is difficult to evaluate the effect changes in reflux flow could have on the bottom control loop in other distillation control systems.

7.4 NONINTERACTING CONTROL

The noninteracting control system was successful in achieving a significant improvement in control over the conventional proportional integral control system. The performance of this control system following a disturbance in feed flow is illustrated in Figures 7.12, 7.13, 7.14 and 7.15. The noninteracting system reduced the IAE in the top composition control loop by a factor of 2 over that obtained with conventional control. In addition, there was a significant improvement in control of the bottom composition as the IAE in this loop was reduced from 29.3 to 16.7 in a typical case.

In comparing the performance of the ratio control scheme with the noninteracting system, both systems result in a similar "goodness"

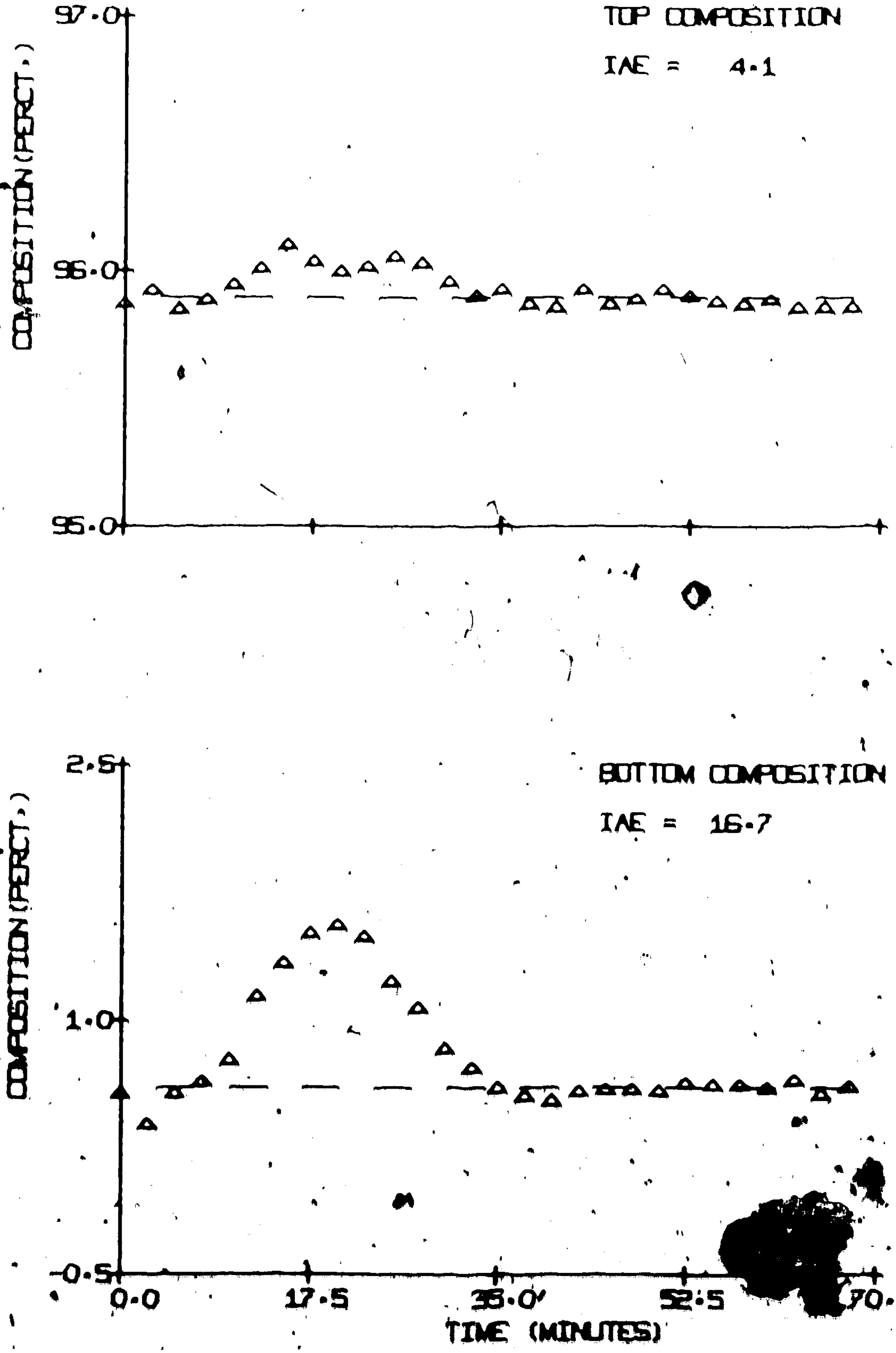


FIGURE 7.12 NONINTERACTING CONTROL [15% INCREASE IN FEED FLOW, RUN C-1]

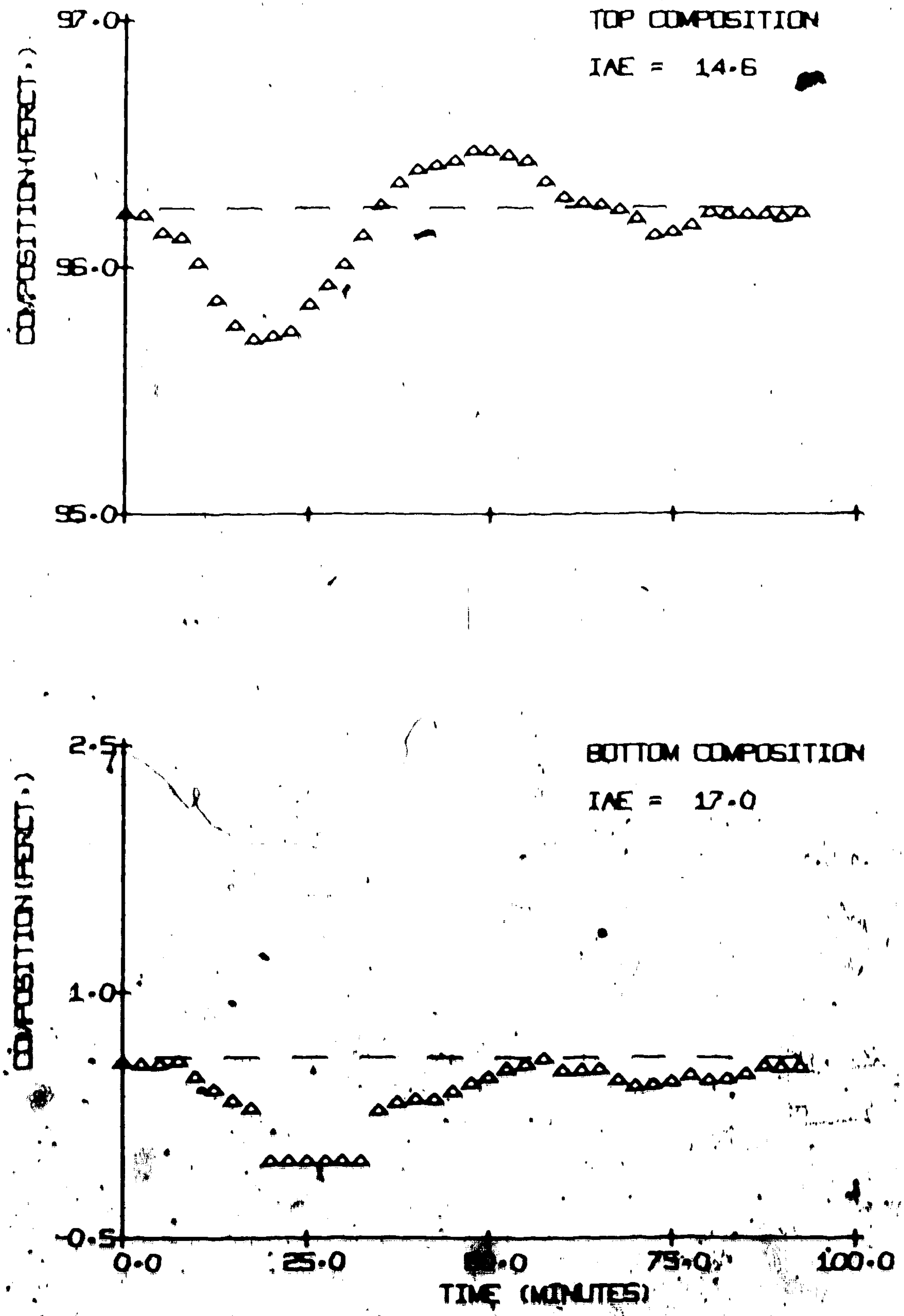


FIGURE 7.13 NONINTERACTING CONTROL [15% DECREASE IN FEED FLOW, RUN C-2]

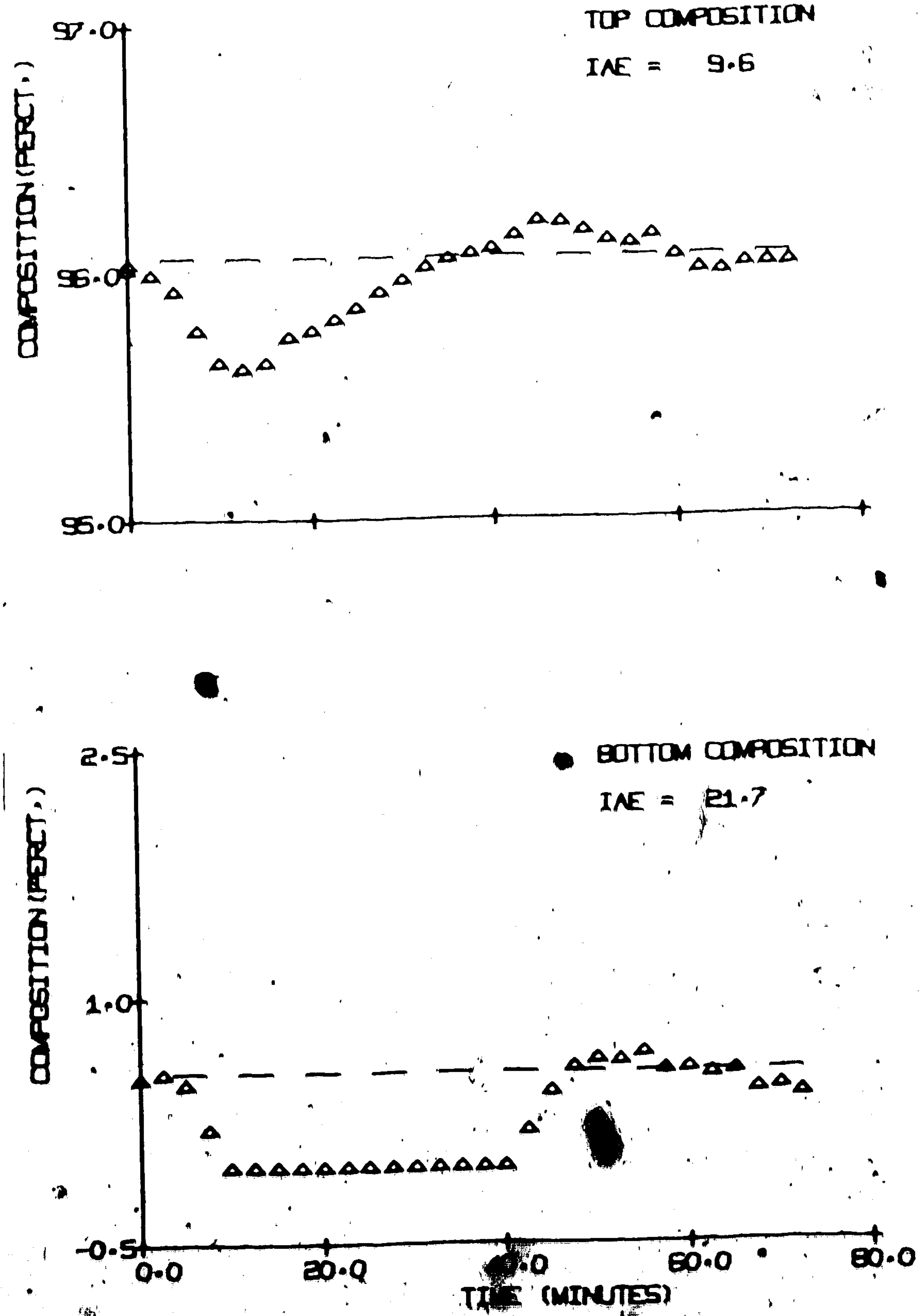


FIGURE 7.14 NONINTERACTING CONTRO[illegible]15% DECREASE IN FEED FLOW, RUN C-3]

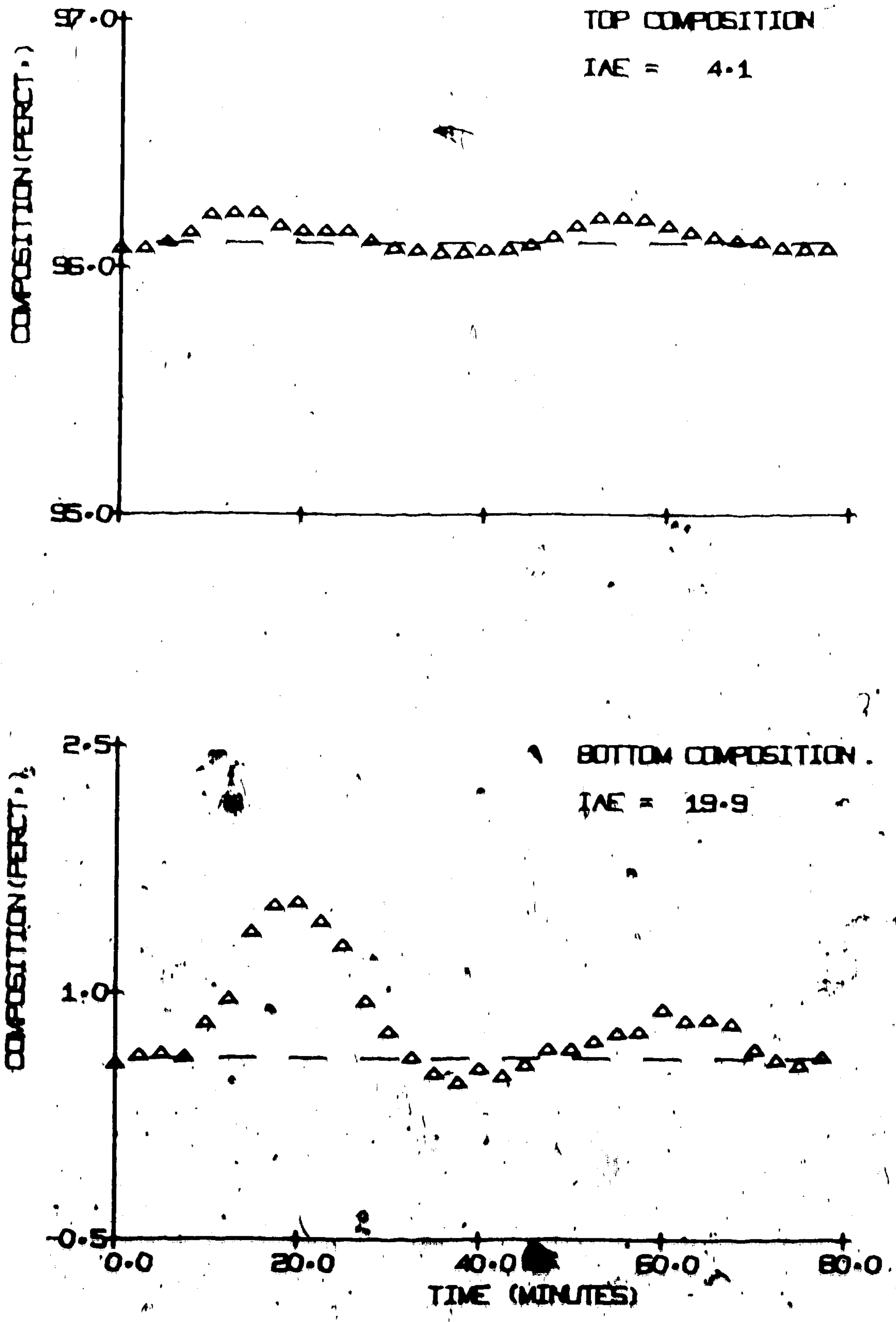


FIGURE 7.15 NONINTERACTING CONTROL [15% INCREASE IN FEED FLOW, RUN C-4]

of control as illustrated by the IAE values for different tests. There is a marginal improvement in control in the bottom composition loop with noninteracting control over that obtained with the ratio control scheme.

A comparison of Figures 7.12 - 7.15 with 7.3 and 7.4 illustrates the effect of the noninteracting compensation. The overhead composition is no longer disrupted by control action taken in the bottom loop as was evident in Figures 7.3 and 7.4. The sustained oscillations evident in Figures 7.3 and 7.4 have been eliminated. In fact control of both end compositions together with noninteracting compensation yields as good a performance in the individual control loops as was attainable when only one end composition was controlled.

The controller constants that yielded a minimum IAE response with noninteracting compensation were the same as those employed to obtain a minimum IAE response when two loops were controlled separately in runs S-1 and S-2. This suggests that noninteracting compensation allows independent operation of the two control loops.

It is reasonable to suggest that noninteracting compensation would show an even greater improvement in the control behaviour of the column if the bottom composition control loop was more stable. The main objective of noninteracting compensation and the ratio control scheme is a reduction of the effect of control action taken in the bottom loop on the overhead composition. With a relatively unstable bottom product control loop, the amount that the control action in this loop can be increased with the use of an advanced control algorithm is limited. Thus the full benefit of the noninteracting and ratio control systems may not be realized in this case.

CONCLUSIONS

1. A gas chromatograph has been demonstrated to be applicable to a feedback control system where the ratio of the sampling period to the process time constant was .17.
2. As a design basis for a noninteracting control system, the dynamic behaviour of a binary distillation column (for excursions in product composition of up to 1. wt. percent about an operating point) may be adequately represented by a set of first order plus dead time transfer functions. Averaging the model parameters obtained from a series of on line pulse tests coupled with a fit of the transient responses in the time domain provided a satisfactory means of evaluating these transfer functions.
3. The direct control of both the overhead and bottoms composition of a distillation column is not recommended unless a means of reducing the interaction is employed. It has been illustrated that the interaction, if ignored, can cause a serious deterioration in the performance of the control system. One must be prepared to accept either extremely slow control or an oscillatory performance in such a system. In addition, as a direct result of the interaction between the control loops, it is unduly difficult to tune the individual controllers.
4. Control of the overhead composition by manipulation of the reflux to overhead vapour ratio provides an effective means for reducing the interacting effect of the steam flow on the over-

head composition. The use of this ratio control system results in a significant improvement in the performance of the overhead composition control loop as compared to a conventional system.

5. The control system designed on the basis of eliminating interactions yielded a much improved control of the overhead composition and a somewhat improved performance in the bottom composition loop over a conventional system. In addition, the problem of tuning the individual control loops has been greatly reduced, as the noninteracting compensation allows the loops to be tuned independently.
6. In the case studied in this work, there is little difference between the performance of the ratio control system and the noninteracting system. The additional improvement in control with a noninteracting system may be offset by difficulties encountered in model construction. On the other hand, if the qualities of multiple side streams are to be controlled by their respective drawoff rates a form of noninteracting compensation is necessary. In this respect a noninteracting control system will receive wider application if difficulties in model construction are not prohibitive.

RECOMMENDATIONS FOR FUTURE WORK

1. The construction of an adequate dynamic model is the single greatest difficulty in the application of modern control theory to a distillation process. Although much work has been done in

this area a completely satisfactory method for determining a simplified dynamic model does not exist. As has been illustrated in the present study a set of transfer functions determined by on-line dynamic testing yields a satisfactory description of the dynamic behaviour of the process. However, dynamic testing is usually not desirable in an industrial process and is at best time consuming. Consequently, a method of simplifying the higher order dynamic models prevalent in the literature is required. There are well known techniques for reducing the order of a state variable system based on elimination of these states with the least significant eigenvalues. However, there is little evidence of an investigation of the accuracy of a model when the order of the system has been drastically reduced. Such an investigation would be especially significant in the case of a distillation column where the eigenvalues are usually of the same order of magnitude. Thus a meaningful study would be an investigation of the representation of column dynamics by a reduced dynamic model of the process and subsequent study of its performance as a basis for implementing advanced control systems.

2. The problem of formulating a dynamic model for noninteracting compensation can be largely overcome by using a simple gain compensation in the place of the dynamic compensator. The gains may be determined from simple step tests on the process or a proven steady model of the process.

3. The ratio control system employed in this study was successful in reducing the effects of the interaction by manipulation of the

reflux to overhead vapour ratio. That is any change in the overhead vapour rate (which reflects the interacting effect on the top product) is automatically compensated for by a corresponding change in reflux flow. As was the case in this work, it is often impossible to measure the overhead vapour rate. This would suggest the use of an alternate system whereby the overhead composition is controlled by the top product drawoff rate, the condensor level being controlled by the reflux flow. Thus any change in the overhead vapour rate would be automatically compensated for by a corresponding change in reflux flow in order to maintain the condensor level constant. A similar system may be employed in control of the bottom product composition whereby the composition is controlled by the product drawoff and the reboiler level is controlled by manipulation of the steam rate. This is the material balance control scheme proposed by Shinskey (75).

4. Rosenbrock (8) and Davison (10) have described a method for two point control of a distillation column based on a model analysis technique as described in Chapter 1. It would appear worthwhile to investigate the performance of [illegible]tem and compare the performance attained with that obt[illegible]ble using other multivariable control systems.

5. As suggested in the review of the literature in Chapter 1, perhaps the control system with the most potential for dual quality control of a distillation process is one based on optimal control theory. Although it is difficult to formulate an "optimal" control policy for a regulatory system, investigation of this area would appear to be warranted.

NOMENCLATURE

$\underline{A}$	coefficient matrix for the state equation
a_{ij}	element of the matrix $\underline{A}$.
A	analyzer
$\underline{B}$	coefficient matrix for the state equation
$\underline{C}$	matrix of controller transfer functions
C_{ij}	individual controller transfer function
CC	composition controller
$\underline{D}$	noninteracting compensator matrix
D_{ij}	noninteracting compensator.element
E	error signal
F	distillation column feed flow rate
FRC	flow recorder controller
$\underline{G}$	closed loop transfer function matrix
G_{ij}	element of the closed loop transfer function matrix
H(s)	zero order hold transfer function
$\underline{H}$	controller matrix
$\underline{K}$	gain matrix
K_{ij}	individual gain element
K_p	proportional gain
K_I	integral gain
LC	level controller
n	number of components in a stream
$\underline{P}$	process transfer function matrix
P_{ij}	element of the process transfer function matrix
R	distillation column reflux flow
$\underline{R}$	reference input vector

RFC	ratio flow controller
S	distillation column steam flow
T	sample period
T'_{ij}	process time constant
$\underline{U}$	vector of control variables
$\underline{V}$	matrix of eigenvectors
$\underline{X}$	state vector
x_B	distillation column bottom product composition
x_D	distillation column overhead product composition
x_F	distillation column feed composition
$\underline{\lambda}$	matrix of eigenvalues

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APPENDIX 1

DESCRIPTION OF THE OFF LINE PROGRAMS

TRANSFER FUNCTION PROGRAM (TRSFCT)

The function of the transfer function program was to fit the transient response data from a pulse test to a first order plus dead time transfer function model.

The first function of the program was to smooth the transient data using an exponential filter. The transient data was then converted to perturbation values by subtracting a computed average steady state value. Since the process did not always return precisely to the initial steady state following the pulse test, it was necessary to force the pulse to close by insuring that the last 20 data points were indeed at the initial steady state. This procedure was necessary to avoid the large time constants that would result from any minor disturbance that would cause a steady state offset.

To determine the parameters of the model the time delay was first determined by locating the point of maximum slope on the initial side of the response curve. The time delay was then determined from the intercept of the line of maximum slope and the initial steady state value. The process gain and time constant were then determined by searching for the values of these parameters that minimized the sum of the squares of the errors between the actual response and that predicted by model. The response of a first order transfer function process $(\frac{K}{\tau S+1})$ to a pulse input of width a and height b is given by

$$Kb\,[1 - e^{-\frac{t}{\tau}} - U(t - a)(1 - e^{-\frac{t-a}{\tau}}]$$

The search for the optimum process parameters was accomplished by employing Rosenbrock's search technique (72).

The program requires the following input data cards.

1. NCASE. the number of cases to be run.
2. For each case require card 1 → ICONT = 1 if the pulse input function is not part of the data input.

   NPTS = number of points of transient data.

   ITYPE = 1 for a positive gain process.

   -1 for a negative gain process.

   NUMBR = number of vectors of data as input.

   Card 2 → NVCTR = 1 for each vector of data that is to be fit to a model.

   Card 3 → title

   Card 4 → PWDTH = pulse width in minutes.

   PHGHT = pulse height in process units.

   (IF ICONT = 1 the pulse width and height are calculated from the input data)

   IVCTR = vector number of the pulse input data (used only if ICONT = 1)

   NPSST = number of steady state data points, prior to the start of the pulse.

   NWIDE = pulse width as a number of data points.

   Card 5 → FLTCST = filter constant.

   Cards 6 + → NPTS of transient data for each vector of data (preceded by 4 title cards which are ignored.)

AUXILLARY TRANSFER FUNCTION PROGRAM (GCCON)

This program is functionally similar to the transfer function program TRSFCT. The difference being that GCCON is employed to fit the transient data from the gas chromatograph to the first order plus dead time transfer function model. The data from the gas chromatograph was necessarily acquired at a much slower rate than the other transient data. This necessitates several differences in the computer program to fit the transient data to the process model.

1) A two point formula for the first derivative versus the five point formula used in the TRSFCT program was employed in calculating the deadtime.
2) The last five points of transient data were forced to be at the initial steady state value in order to force the pulse to close.
3) The transient data was not filtered.

In all other respects GCCONS is identical to TRSFCT.

DATA SORT PROGRAM (GTDAT)

The function of GTDAT was to sort the data acquired during the dynamic tests, convert the data to the appropriate units and punch the data to cards in a suitable format for the transfer function programs.

The number of loops, number of data points, time interval between data points and a flag (IGC) indicating that the gas chromatograph analysis of the bottom product composition is to be outputted, for each of the data acquisition loops, the instrument range (A), offset (B) steady state temperature and composition are read from data cards. The steady state temperature and composition are used in conjunction with a flow loop to correct the flow rate for composition and temperature based on the steady state operating conditions. For other than flow loops the steady state temperature and composition should be specified as zero.

The data acquired by the DDC system is obtained from the data acquisition file by means of the system subroutine GBDAT. The data is converted from unscaled engineering units into actual engineering units by the transformation.

$$x = A/32767 \times Ix + B$$

where Ix is the unscaled value and x is the scaled value in engineering units. For flow loops the flow is corrected for the steady state composition and temperature. The data is then punched to cards. The gas chromatograph analysis is also outputted for the dynamic test if the IGC parameter is non zero.

APPENDIX 2

```
     PULSE   IN   REFLUX   FLOW(RUN 20)



PULSE  DATA
...........

          PULSE HEIGHT                  .29    LBS./MIN

          PULSE WIDTH                  14.7    MINUTES



STEADY  STATE  OPERATING  CONDITIONS
.....................................

          TOP COMPOSITION              95.5    WT. PERCT.
          BOTTOM COMPOSITION            .51    WT. PERCT.

          FEED FLOW                    2.45    LBS./MIN
          FEED COMPOSITION             46.6    WT. PERCT.
          REFLUX FLOW                  1.91    LBS./MIN
          STEAM FLOW                   2.02    LBS./MIN
          TOP PRODUCT FLOW             1.18    LBS./MIN
          BOTTOM PRODUCT FLOW          1.28    LBS./MIN

MATERIAL  BALANCE  ERROR  OF  CLOSURE
.....................................

          OVERALL                       0.6    PERCT.
          METHANOL                     -0.5    PERCT.
```

```
          PULSE  IN  REFLUX  FLOW(RUN 21)



      PULSE  DATA
      ...........

             PULSE HEIGHT                 -.18    LBS./MIN

             PULSE WIDTH                  18.7    MINUTES



      STEADY  STATE  OPERATING  CONDITIONS
      ....................................


             TOP COMPOSITION              95.9.   WT. PERCT.

             BOTTOM COMPOSITION           .50     WT. PERCT.


             FEED FLOW                    2.45    LBS./MIN

             FEED COMPOSITION             46.6    WT. PERCT.

             REFLUX FLOW                  1.94    LBS./MIN

             STEAM FLOW                   2.00    LBS./MIN

             TOP PRODUCT FLOW             1.18    LBS./MIN

             BOTTOM PRODUCT FLOW          1.27    LBS./MIN


      MATERIAL  BALANCE  ERROR  OF  CLOSURE
      .....................................

             OVERALL                      -0.3    PERCT.

             METHANOL                     -0.7    PERCT.
```

PULSE IN REFLUX FLOW(RUN 22)

PULSE DATA

PULSE HEIGHT	.29	LBS./MIN
PULSE WIDTH	13.3	MINUTES

STEADY STATE OPERATING CONDITIONS

TOP COMPOSITION	96.0	WT. PERCT.
BOTTOM COMPOSITION	.55	WT. PERCT.
FEED FLOW	2.46	LBS./MIN
FEED COMPOSITION	46.6	WT. PERCT.
REFLUX FLOW	1.93	LBS./MIN
STEAM FLOW	2.01	LBS./MIN
TOP PRODUCT FLOW	1.17	LBS./MIN
BOTTOM PRODUCT FLOW	1.30	LBS./MIN

MATERIAL BALANCE ERROR OF CLOSURE

OVERALL	0.5	PERCT.
METHANOL	-1.2	PERCT.

```
PULSE  IN  REFLUX  FLOW(RUN 23)


PULSE  DATA
...........

         PULSE HEIGHT                        .28    LBS./MIN

         PULSE WIDTH                        8.00    MINUTES


STEADY  STATE  OPERATING  CONDITIONS
....................................


         TOP COMPOSITION                    95.0    WT. PERCT.

         BOTTOM COMPOSITION                  .50    WT. PERCT.


         FEED FLOW                          2.46    LBS./MIN

         FEED COMPOSITION                   45.5    WT. PERCT.

         REFLUX FLOW                        1.96    LBS./MIN

         STEAM FLOW                         1.99    LBS./MIN

         TOP PRODUCT FLOW                   1.12    LBS./MIN

         BOTTOM PRODUCT FLOW                1.30    LBS./MIN


MATERIAL  BALANCE  ERROR  OF  CLOSURE
.....................................

         OVERALL                            -1.4    PERCT.

         METHANOL                           -3.5    PERCT.
```

```
        PULSE  IN  REFLUX  FLOW(RUN 24)


    PULSE  DATA
    ...........

         PULSE HEIGHT                    -.19    LBS./MIN
         PULSE WIDTH                     7.70    MINUTES


    STEADY  STATE  OPERATING  CONDITIONS
    ....................................

         TOP COMPOSITION                 95.7    WT. PERCT.
         BOTTOM COMPOSITION               .48    WT. PERCT.

         FEED FLOW                       2.45    LBS./MIN
         FEED COMPOSITION                47.5    WT. PERCT.
         REFLUX FLOW                     1.93    LBS./MIN
         STEAM FLOW                      2.02    LBS./MIN
         TOP PRODUCT FLOW                1.17    LBS./MIN
         BOTTOM PRODUCT FLOW             1.22    LBS./MIN


    MATERIAL  BALANCE  ERROR  OF  CLOSURE
    .....................................

         OVERALL                         -2.4    PERCT.
         METHANOL                        -3.4    PERCT.
```

PULSE IN STEAM FLOW(RUN 20)

PULSE DATA

PULSE HEIGHT	-.26	LBS./MIN
PULSE WIDTH	10.7	MINUTES

STEADY STATE OPERATING CONDITIONS

TOP COMPOSITION	95.6	WT. PERCT.
BOTTOM COMPOSITION	.53	WT. PERCT.
FEED FLOW	2.45	LBS./MIN
FEED COMPOSITION	46.6	WT. PERCT.
REFLUX FLOW	1.92	LBS./MIN
STEAM FLOW	2.02	LBS./MIN
TOP PRODUCT FLOW	1.20	LBS./MIN
BOTTOM PRODUCT FLOW	1.30	LBS./MIN

MATERIAL BALANCE ERROR OF CLOSURE

OVERALL	1.9	PERCT.
METHANOL	0.9	PERCT.

PULSE IN STEAM FLOW(RUN 24)

PULSE DATA

PULSE HEIGHT	-.21	LBS./MIN
PULSE WIDTH	9.33	MINUTES

STEADY STATE OPERATING CONDITIONS

TOP COMPOSITION	95.9	WT. PERCT.
BOTTOM COMPOSITION	.60	WT. PERCT.
FEED FLOW	2.47	LBS./MIN
FEED COMPOSITION	47.0	WT. PERCT.
REFLUX FLOW	1.96	LBS./MIN
STEAM FLOW	2.01	LBS./MIN
TOP PRODUCT FLOW	1.15	LBS./MIN
BOTTOM PRODUCT FLOW	1.31	LBS./MIN

MATERIAL BALANCE ERROR OF CLOSURE

OVERALL	-0.5	PERCT.
METHANOL	-1.1	PERCT.

PULSE IN STEAM FLOW(RUN 25)

PULSE DATA

PULSE HEIGHT	.19	LBS./MIN
PULSE WIDTH	9.07	MINUTES

STEADY STATE OPERATING CONDITIONS

TOP COMPOSITION	96.0	WT. PERCT.
BOTTOM COMPOSITION	0.6	WT. PERCT.
FEED FLOW	2.46	LBS./MIN
FEED COMPOSITION	47.0	WT. PERCT.
REFLUX FLOW	1.95	LBS./MIN
STEAM FLOW	2.01	LBS./MIN
TOP PRODUCT FLOW	1.17	LBS./MIN
BOTTOM PRODUCT FLOW	1.26	LBS./MIN

MATERIAL BALANCE ERROR OF CLOSURE

OVERALL	-1.2	PERCT.
METHANOL	-2.5	PERCT.

```
        PULSE  IN  STEAM  FLOW(RUN 27)



    PULSE  DATA
    ...........

              PULSE HEIGHT                 -.27    LBS./MIN

              PULSE WIDTH                  5.87    MINUTES




    STEADY  STATE  OPERATING  CONDITIONS
    ...................................



              TOP COMPOSITION              95.8    WT. PERCT.

              BOTTOM COMPOSITION            .48    WT. PERCT.



              FEED FLOW                    2.45    LBS./MIN

              FEED COMPOSITION             47.6    WT. PERCT.

              REFLUX FLOW                  1.93    LBS./MIN

              STEAM FLOW                   2.02    LBS./MIN

              TOP PRODUCT FLOW             1.16    LBS./MIN

              BOTTOM PRODUCT FLOW          1.23    LBS./MIN


    MATERIAL  BALANCE  ERROR  OF  CLOSURE
    ...................................

              OVERALL                      -2.4    PERCT.

              METHANOL                     -3.9    PERCT.
```

PULSE IN FEED FLOW(RUN 20)

PULSE DATA

PULSE HEIGHT	.46	LBS./MIN
PULSE WIDTH	15.2	MINUTES

STEADY STATE OPERATING CONDITIONS

TOP COMPOSITION	95.7	WT. PERCT.
BOTTOM COMPOSITION	.51	WT. PERCT.
FEED FLOW	2.45	LBS./MIN
FEED COMPOSITION	46.6	WT. PERCT.
REFLUX FLOW	1.94	LBS./MIN
STEAM FLOW	2.02	LBS./MIN
TOP PRODUCT FLOW	1.20	LBS./MIN
BOTTOM PRODUCT FLOW	1.27	LBS./MIN

MATERIAL BALANCE ERROR OF CLOSURE

OVERALL	0.4	PERCT.
METHANOL	0.9	PERCT.

```
      PULSE  IN  FEED  FLOW(RUN 21)



PULSE  DATA
...........

        PULSE HEIGHT                 -.24   LBS./MIN
        PULSE WIDTH                  15.5   MINUTES



STEADY  STATE  OPERATING  CONDITIONS
....................................

        TOP COMPOSITION              95.9   WT. PERCT.
        BOTTOM COMPOSITION            .75   WT. PERCT.

        FEED FLOW                    2.46   LBS./MIN
        FEED COMPOSITION             46.6   WT. PERCT.
        REFLUX FLOW                  1.95   LBS./MIN
        STEAM FLOW                   2.00   LBS./MIN
        TOP PRODUCT FLOW             1.13   LBS./MIN
        BOTTOM PRODUCT FLOW          1.29   LBS./MIN

MATERIAL  BALANCE  ERROR  OF  CLOSURE
....................................

        OVERALL                      -1.6   PERCT.
        METHANOL                     -4.7   PERCT.
```

PULSE IN FEED FLOW(RUN 24)

PULSE DATA

PULSE HEIGHT	.44	LBS./MIN
PULSE WIDTH	9.87	MINUTES

STEADY STATE OPERATING CONDITIONS

TOP COMPOSITION	95.7	WT. PERCT.
BOTTOM COMPOSITION	.40	WT. PERCT.
FEED FLOW	2.45	LBS./MIN
FEED COMPOSITION	47.6	WT. PERCT.
REFLUX FLOW	1.93	LBS./MIN
STEAM FLOW	2.03	LBS./MIN
TOP PRODUCT FLOW	1.17	LBS./MIN
BOTTOM PRODUCT FLOW	1.18	LBS./MIN

MATERIAL BALANCE ERROR OF CLOSURE

OVERALL	-3.8	PERCT.
METHANOL	-3.1	PERCT.

```
PULSE  IN FEED  COMPOSITION(RUN 20)



PULSE  DATA
..........

          PULSE HEIGHT                  8.7    WT. PERCT.

          PULSE WIDTH                  15.7    MINUTES



STEADY  STATE  OPERATING  CONDITIONS
..................................

          TOP COMPOSITION              95.5    WT. PERCT.

          BOTTOM COMPOSITION            .52    WT. PERCT.


          FEED FLOW                    2.46    LBS./MIN

          FEED COMPOSITION             45.5    WT. PERCT.

          REFLUX FLOW                  1.94    LBS./MIN

          STEAM FLOW                   1.99    LBS./MIN

          TOP PRODUCT FLOW             1.13    LBS./MIN

          BOTTOM PRODUCT FLOW          1.28    LBS./MIN


MATERIAL  BALANCE  ERROR  OF  CLOSURE
..................................

          OVERALL                      -2.1    PERCT.

          METHANOL                     -2.9    PERCT.
```

```
PULSE  IN  FEED  COMPOSITION(RUN 21)


PULSE  DATA
...........

          PULSE HEIGHT                  -4.9    WT. PERCT.

          PULSE WIDTH                   14.4    MINUTES


STEADY  STATE  OPERATING  CONDITIONS
....................................


          TOP COMPOSITION               95.3    WT. PERCT.

          BOTTOM COMPOSITION             .48    WT. PERCT.


          FEED FLOW                     2.46    LBS./MIN

          FEED COMPOSITION              45.5    WT. PERCT.

          REFLUX FLOW                   1.94    LBS./MIN

          STEAM FLOW                    1.99    LBS.[illegible]N

          TOP PRODUCT FLOW              1.13    LBS./MIN

          BOTTOM PRODUCT FLOW           1.26    LBS./MIN


MATERIAL  BALANCE  ERROR  OF  CLOSURE
....................................

          OVERALL                       -2.4    PERCT.

          METHANOL                      -2.6    PERCT.
```

```
PULSE IN FEED COMPOSITION(RUN 22)



PULSE DATA
..........

     PULSE HEIGHT                5.9    WT. PERCT.

     PULSE WIDTH                14.9    MINUTES



STEADY STATE OPERATING CONDITIONS
.................................


     TOP COMPOSITION            95.8    WT. PERCT.

     BOTTOM COMPOSITION          .60    WT. PERCT.


     FEED FLOW                  2.45    LBS./MIN

     FEED COMPOSITION           48.3    WT. PERCT.

     REFLUX FLOW                1.96    LBS./MIN

     STEAM FLOW                 2.00    LBS./MIN

     TOP PRODUCT FLOW           1.15    LBS./MIN

     BOTTOM PRODUCT FLOW        1.17    LBS./MIN


MATERIAL BALANCE ERROR OF CLOSURE
.................................

     OVERALL                    -5.1    PERCT.

     METHANOL                   -6.2    PERCT.
```

```
PULSE  IN  FEED  COMPOSITION(RUN 23)



PULSE  DATA
...........

        PULSE HEIGHT                -7.7   WT. PERCT.

        PULSE WIDTH                 14.9   MINUTES



STEADY  STATE  OPERATING  CONDITIONS
....................................


        TOP COMPOSITION             95.9   WT. PERCT.

        BOTTOM COMPOSITION            .6   WT. PERCT.


        FEED FLOW                   2.46   LBS./MIN

        FEED COMPOSITION            48.3   WT. PERCT.

        REFLUX FLOW                 1.06   LBS./MIN

        STEAM FLOW                  2.00   LBS./MIN

        TOP PRODUCT FLOW            1.15   LBS./MIN

        BOTTOM PRODUCT FLOW         1.27   LBS./MIN


MATERIAL  BALANCE  ERROR  OF  CLOSURE
.....................................

      - OVERALL                     -1.7   PERCT.

        METHANOL                    -6.7   PERCT.
```

```
[illegible] OP LOOP ALONE(RUN S1)


DISTURBANCE
...........

  STEP IN FEED FLOW                +.34   LBS./MIN.


STEADY  STATE  OPERATING  CONDITIONS
....................................


                                 INITIAL      FINAL

  TOP COMPOSITION                 96.1        96.1   WT. PERCT.

  BOTTOM COMPOSITION              .60         1.52   WT. PERCT.


  FEED FLOW                       2.47        2.81   LBS./MIN.

  FEED COMPOSITION                47.8        47.8   WT. PERCT.

  REFLUX FLOW                     1.87        1.74   LBS./MIN.

  STEAM FLOW                      2.02        2.02   LBS./MIN.

  TOP PRODUCT FLOW                1.18        1.39   LBS./MIN.

  BOTTOM PRODUCT FLOW             1.21        1.41   LBS./MIN.


MATERIAL  BALANCE  ERROR  OF  CLOSURE
.....................................

  OVERALL                         -3.1        0.0    PERCT.

  METHANOL                        -3.1        1.5    PERCT.
```

```
        CONTROL OF BOTTOM LOOP ALONE(RUN S2)


     DISTURBANCE
     ...........
       STEP IN FEED FLOW                   +.34   LBS./MIN.


     STEADY  STATE  OPERATING  CONDITIONS
     ....................................

                                          INITIAL      FINAL
      TOP COMPOSITION                      96.0        95.5    WT. PERCT.
      BOTTOM COMPOSITION                   .60         .60     WT. PERCT.

      FEED FLOW                            2.47        2.81    LBS./MIN.
      FEED COMPOSITION                     47.8        47.8    WT. PERCT.
      REFLUX FLOW                          1.88        1.87    LBS./MIN.
      STEAM FLOW                           2.02        2.13    LBS./MIN.
      TOP PRODUCT FLOW                     1.22        1.40    LBS./MIN.
      BOTTOM PRODUCT FLOW                  1.22        1.39    LBS./MIN.

     MATERIAL BALANCE ERROR OF CLOSURE
     .................................
      OVERALL                              -0.9        -0.8    PERCT.
      METHANOL                         [illegible]     0.0     PERCT.
```

```
   P - I CONTROL (RUN PI1)


DISTURBANCE
...........

  . STEP IN FEED FLOW             +.34   LBS./MIN.


STEADY  STATE  OPERATING  CONDITIONS
.....................................


                                 INITIAL      FINAL

  TOP COMPOSITION                 96.1          -      WT. PERCT.
  BOTTOM COMPOSITION              .60           -      WT. PERCT.

  FEED FLOW                       2.47          -      LBS./MIN.
  FEED COMPOSITION                47.8          -      WT. PERCT.
  REFLUX FLOW                     1.87          -      LBS./MIN.
  STEAM FLOW                      2.02          -      LBS./MIN.
  TOP PRODUCT FLOW                1.22          -      LBS./MIN.
  BOTTOM PRODUCT FLOW             1.21          -      LBS./MIN.

MATERIAL  BALANCE  ERROR  OF  CLOSURE
.....................................

  OVERALL   -                 [illegible]1.4    -      PERCT.
  METHANOL                        0.2           -      PERCT.
```

```
  P - I CONTROL(RUN PI2)



DISTURBANCE
...........
  STEP IN FEED FLOW                     -.36   LBS./MIN.




STEADY  STATE  OPERATING  CONDITIONS
....................................



                                      INITIAL      FINAL
  TOP COMPOSITION                      96.0          -       WT. PERCT.
  BOTTOM COMPOSITION                   .60           -       WT. PERCT.


  FEED FLOW                            2.46          -       LBS./MIN.
  FEED COMPOSITION                     47.8          -       WT. PERCT.
  REFLUX FLOW                          1.86          -       LBS./MIN.
  STEAM FLOW                           2.02          -       LBS./MIN.
  TOP PRODUCT FLOW                     1.22          -       LBS./MIN.
  BOTTOM PRODUCT FLOW                  1.21          -       LBS./MIN.


MATERIAL  BALANCE  ERROR  OF  CLOSURE
.....................................
  OVERALL                              -1.4          -       PERCT.
  METHANOL                             -0.3          -       PERCT.
```

```
P - I CONTROL(RUN PI3)


DISTURBANCE
...........

  STEP IN FEED FLOW              +.34   LBS./MIN.


STEADY  STATE  OPERATING  CONDITIONS
..................................


                                 INITIAL     FINAL
  TOP COMPOSITION                 96.1        96.1   WT. PERCT.
  BOTTOM COMPOSITION              .60         .60    WT. PERCT.

  FEED FLOW                       2.46        2.80   LBS./MIN.
  FEED COMPOSITION                47.8        47.8   WT. PERCT.
  REFLUX FLOW                     1.87        1.96   LBS./MIN.
  STEAM FLOW                      2.02        1.86   LBS./MIN.
  TOP PRODUCT FLOW                1.22        1.37   LBS./MIN.
  BOTTOM PRODUCT FLOW             1.22        1.38   LBS./MIN.

MATERIAL  BALANCE  ERROR  OF  CLOSURE
..................................

  OVERALL                         -1.0        -1.5   PERCT.
  METHANOL                        -0.2        -0.7   PERCT.
```

```
     P - I CONTROL(RUN PI4)



   DISTURBANCE
   ...........

    STEP IN FEED FLOW                       -.36    LBS./MIN.



   STEADY  STATE  OPERATING  CONDITIONS

   ....................................



                                        INITIAL      FINAL

    TOP COMPOSITION                     96.2         96.2    WT. PERCT.

    BOTTOM COMPOSITION                  .60          .60     WT. PERCT.


    FEED FLOW                           2.47         2.08    LBS./MIN.

    FEED COMPOSITION                    47.8         47.8    WT. PERCT.

    REFLUX FLOW                         1.86         1.85    LBS./MIN.

    STEAM FLOW                          2.02         1.88    LBS./MIN.

    TOP PRODUCT FLOW                    1.21         1.00    LBS./MIN.

    BOTTOM PRODUCT FLOW                 1.24         1.06    LBS./MIN.


   MATERIAL  BALANCE  ERROR  OF  CLOSURE

   .....................................

    OVERALL                             -0.7         -1.0    PERCT.

    METHANOL                            -0.8         -2.4    PERCT.
```

```
     P - I CONTROL(RUN PI5)



   DISTURBANCE
   ...........

     STEP IN FEED FLOW                     -.37   LBS./MIN.



   STEADY  STATE  OPERATING  CONDITIONS
   ...................................


                                     INITIAL      FINAL
     TOP COMPOSITION                   95.9        95.9   WT. PERCT.
     BOTTOM COMPOSITION                 .60         .60   WT. PERCT.


     FEED FLOW                         2.48        2.08   LBS./MIN.
     FEED COMPOSITION                  47.8        47.8   WT. PERCT.
     REFLUX FLOW                       1.86        1.84   LBS./MIN.
     STEAM FLOW                        2.03        1.88   LBS./MIN.
     TOP PRODUCT FLOW                  1.18        1.00   LBS./MIN.
     BOTTOM PRODUCT FLOW               1.23        1.06   LBS./MIN.


   MATERIAL BALANCE ERROR OF CLOSURE
   .................................

     OVERALL                           -2.6        -1.0   PERCT.
     METHANOL                          -3.8        -2.3   PERCT.
```

```
STURBANCE
.........

STEP IN FEED FLOW                -.37   LBS./MIN.



EADY  STATE  OPERATING  CONDITIONS
..................................


                              INITIAL      FINAL
TOP COMPOSITION                95.9         95.9   WT. PERCT.
BOTTOM COMPOSITION             .60          .60    WT. PERCT.


FEED FLOW                      2.48         2.08   LBS./MIN.
FEED COMPOSITION               47.8         47.8   WT. PERCT.
REFLUX FLOW                    1.86         1.84   LBS./MIN.
STEAM FLOW                     2.03         1.88   LBS./MIN.
OP PRODUCT FLOW                1.18         1.00   LBS./MIN.
BOTTOM PRODUCT FLOW            1.23         1.06   LBS./MIN.


ERIAL  BALANCE  ERROR  OF  CLOSURE
..................................
OVERALL                        -2.6         -1.0   PERCT.
ETHANOL                        -3.8         -2.3   PERCT.
```

RATIO CONTROL(RUN R2)

DISTURBANCE
....

STEP IN FEED FLOW -.39 LBS./MIN.

STEADY STATE OPERATING CONDITIONS
....

	INITIAL	FINAL	
TOP COMPOSITION	96.2	96.2	WT. PERCT.
BOTTOM COMPOSITION	.60	.60	WT. PERCT.
FEED FLOW	2.47	2.08	LBS./MIN.
FEED COMPOSITION	47.8	47.8	WT. PERCT.
REFLUX FLOW	1.86	1.76	LBS./MIN.
STEAM FLOW	2.02	1.85	LBS./MIN.
TOP PRODUCT FLOW	1.24	.98	LBS./MIN.
BOTTOM PRODUCT FLOW	1.25	1.03	LBS./MIN.

MATERIAL BALANCE ERROR OF CLOSURE
....

OVERALL	0.5	-3.2	PERCT.
METHANOL	1.3	-4.0	PERCT.

```
RATIO CONTROL(RUN R3)


DISTURBANCE
...........
 STEP IN FEED FLOW                 -.38   LBS./MIN.


STEADY  STATE  OPERATING  CONDITIONS
....................................


                                 INITIAL      FINAL
 TOP COMPOSITION                  96.4        96.4    WT. PERCT.
 BOTTOM COMPOSITION                .55         .55    WT. PERCT.


 FEED FLOW                        2.46        2.07    LBS./MIN.
 FEED COMPOSITION                 47.8        47.8    WT. PERCT.
 REFLUX FLOW                      1.90        1.67    LBS./MIN.
 STEAM FLOW                       2.01        1.80    LBS./MIN.
 TOP PRODUCT FLOW                 1.22        1.05    LBS./MIN.
 BOTTOM PRODUCT FLOW              1.24        1.07    LBS./MIN.


MATERIAL  BALANCE  ERROR  OF  CLOSURE
.....................................
 OVERALL                          -0.0        2.7     PERCT.
 METHANOL                          0.8        3.2     PERCT.
```

```
RATIO CONTROL(RUN 34)


DISTURBANCE
...........

  STEP IN FEED FLOW                          +.34   LBS./MIN.



STEADY  STATE  OPERATING  CONDITIONS
.................................


                                           INITIAL      FINAL
  TOP COMPOSITION                          96.1         96.1    WT. PERCT.
  BOTTOM COMPOSITION                       .60          .60     WT. PERCT.

  FEED FLOW                                2.47         2.80    LBS./MIN.
  FEED COMPOSITION                         47.8         47.8    WT. PERCT.
  REFLUX FLOW                              1.87         2.04    LBS./MIN.
  STEAM FLOW                               2.01         2.03    LBS./MIN.
  TOP PRODUCT FLOW                         1.24         1.35    LBS./MIN.
  BOTTOM PRODUCT FLOW                      1.24         1.43    LBS./MIN.

MATERIAL  BALANCE  ERROR  OF  CLOSURE
.................................

  OVERALL                                  0.5          -0.7    PERCT.
  METHANOL                                 1.3          -3.[illegible]  PERCT.
```

```
NONINTERACTING CONTROL(RUN G1)


DISTURBANCE
...........

 STEP IN FEED FLOW                    +.34    LBS./MIN.



STEADY  STATE  OPERATING  CONDITIONS
.................................


                                     INITIAL        FINAL
 TOP COMPOSITION                      95.9          95.9     WT. PERCT.
 BOTTOM COMPOSITION                   .60           .60      WT. PERCT.

 FEED FLOW                            2.47          2.81     LBS./MIN.
 FEED COMPOSITION                     47.8          47.8     WT. PERCT.
 REFLUX FLOW                          1.87          1.96     LBS./MIN.
 STEAM FLOW                           2.02          2.18     LBS./MIN.
 TOP PRODUCT FLOW                     1.23          1.42     LBS./MIN.
 BOTTOM PRODUCT FLOW                  1.22          1.42     LBS./MIN.


MATERIAL  BALANCE  ERROR  OF  CLOSURE
.................................
 OVERALL                              -0.4          1.0      PERCT.
 METHANOL                             0.9           0.0      PERCT.
```

```
NONINTERACTING CONTROL(RUN C2)


DISTURBANCE
...........

  STEP IN FEED FLOW                -.39   LBS./MIN.


STEADY  STATE  OPERATING  CONDITIONS
...................................


                                INITIAL      FINAL

  TOP COMPOSITION                96.2        96.2    WT. PERCT.
  BOTTOM COMPOSITION             .60         .60     WT. PERCT.

  FEED FLOW                      2.47        2.08    LBS./MIN.
  FEED COMPOSITION               47.8        47.8    WT. PERCT.
  REFLUX FLOW                    1.87        1.70    LBS./MIN.
  STEAM FLOW                     2.02        1.82    LBS./MIN.
  TOP PRODUCT FLOW               1.24        1.08    LBS./MIN.
  BOTTOM PRODUCT FLOW            1.20        1.02    LBS./MIN.

MATERIAL  BALANCE  ERROR  OF  CLOSURE
...................................

  OVERALL                        1.2         0.6     PERCT.
  METHANOL                       1.4         4.5     PERCT.
```

```
     NONINTERACTING CONTROL(RUN C3)



  DISTURBANCE
  ...........

    STEP  IN  FEED  FLOW                 -.40   LBS./MIN.




  STEADY  STATE  OPERATING  CONDITIONS

  ....................................


                                      INITIAL      FINAL

    TOP  COMPOSITION                   96.0        96.0    WT. PERCT.

    BOTTOM  COMPOSITION                 .55         .55    WT. PERCT.


    FEED  FLOW                         2.47        2.07    LBS./MIN.

    FEED  COMPOSITION                  47.8        47.8    WT. PERCT.

    REFLUX  FLOW                       1.91        1.78    LBS./MIN.

    STEAM  FLOW                        2.03        1.84    LBS./MIN.

    TOP  PRODUCT  FLOW                 1.23        1.00    LBS./MIN.

    BOTTOM  PRODUCT  FLOW              1.25        1.06    LBS./MIN.


  MATERIAL  BALANCE  ERROR  OF  CLOSURE

  ....................................

    OVERALL                             0.5        -0.8    PERCT.

    METHANOL  ~~                        0.8        -2.7    PERCT.~~
```

```
    NONINTERACTING CONTROL(RUN C4)



DISTURBANCE
...........

   STEP IN FEED FLOW                   +.33   LBS./MIN.



STEADY  STATE  OPERATING  CONDITIONS

....................................


                                      INITIAL       FINAL

   TOP COMPOSITION                     96.1         96.1    WT. PERCT.

   BOTTOM COMPOSITION                  .60          .60     WT. PERCT.


   FEED FLOW                           2.47         2.80    LBS./MIN.

   FEED COMPOSITION                    47.8         47.8    WT. PERCT.

   REFLUX FLOW                         1.88         1.93    LBS./MIN.

   STEAM FLOW                          2.02         2.17    LBS./MIN.

   TOP PRODUCT FLOW                    1.23         1.44    LBS./MIN.

   BOTTOM PRODUCT FLOW                 1.22         1.41    LBS./MIN.


MATERIAL  BALANCE  ERROR  OF  CLOSURE

....................................

   OVERALL                             -0.3         -1.8    PERCT.

   METHANOL                            0.7          3.8     PERCT.
```

DATA ACQUISITION LOOP RECORDS

```
                 ID = 0170        FEED FLOW
01   6210   205C+04102   3340   8400+07214   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0171        REFLUX FLOW
01   6010   205F+04101   3340   8480+08364   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0172        STEAM FLOW
01   6010   205F+04100   3340   8480+05740   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0173        BOTTOM PRODUCT FLOW
01   6010   205F+04104   3340   8480+05072   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0174        TOP PRODUCT FLOW
01   6010   205F+04103   3340   8480+05382   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0175        TOP COMPOSITION
01   6410   404C+04106   0000   9100-01310+09879   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0176        BOTTOM COMPOSITION
01   0210   407C   0177   0000   9100+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110
```

DATA ACQUISITION LOOP RECORDS

```
                 ID = 0177       BOTTOM COMPOSITION(DUMMY)
01   4010   204F+20000   0000   9180+20000+00000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0178       FEED COMPOSITION
0[illegible]   0010   207F   0179   0000   9180+20000+00000   0000   0000   7FFF
02   7FFF   2110+10000+00000   2110


                 ID = 0179       FEED COMPOSITION
01   4010   204F+20000   0000   9180+20000+00000   0000   0000   7FFF
02   7FFF   2110+10000+00000   2110


                 ID = 0180       REBOILER TEMPERATURE
01   7010   21BF+00130   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0181       PLATE 1 TEMPERATURE
01   7010   21BF+00131   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0182       PLATE 2 TEMPERATURE
01   7010   21BF+00132   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                 ID = 0183       PLATE 3 TEMPERATURE
01   7010   21BF+00133   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110
```

DATA ACQUISITION LOOP RECORDS

```
                ID = 0185        PLATE 5 TEMPERATURE
01   7010   21BF+00135   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                ID = 0186        PLATE 6 TEMPERATURE
01   7010   21BF+00136   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                ID = 0187        PLATE 7 TEMPERATURE
01   7010   21BF+00137   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110


                ID = 0188        PLATE 8 TEMPERATURE
01   7010   21BF+00138   0000   A280+20000   0000   0000   0000   7FFF
02   7FFF   2110   7FFF   0000   2110
```

GAS CHROMATOGRAPH LOOP RECORDS

```
                ID = 0201        GAS CHROMATOGRAPH
01  4010  0004  004D  0000  BF00  7FFF  0000  0000  0000  0000
02  0000  0000  7FFF  0000  0000


                ID = 0202        GAS CHROMATOGRAPH
01  067F  0000  0201  0000  0000  0000
```

RING BUFFERS FOR DATA ACQUISITION

```
                ID = 0210        FEED FLOW
01   0611  4208  0170  0000  0000  0200

                ID = 0211        REFLUX FLOW
01   0611  4208  0171  0000  0000  0200

                ID = 0212        STEAM FLOW
01   0611  4208  0172  0000  0000  0200

                ID = 0213        BOTTOM PRODUCT FLOW
01   061B  4208  0173  0000  0000  0200

                ID = 0214        TOP PRODUCT FLOW
01   061B  4208  0174  0000  0000  0200

                ID = 0215        TOP COMPOSITION
01   061B  4208  0175  0000  0000  0200

                ID = 0216        BOTTOM COMPOSITION
01   0612  4208  0176  0000  0000  0200

                ID = 0217        FEED COMPOSITION
01   0612  4208  0178  0000  0000  0200
```

RING BUFFERS FOR DATA ACQUISITION

```
                 ID = 0220        REBOILER TEMPERATURE
 01   0613   4208   0180   0000   0000   0200

                 ID = 0221        PLATE 1 TEMPERATURE
 01   0613   4208   0181   0000   0000   0200

                 ID = 0222        PLATE 2 TEMPERATURE
 01   0613   4208   0182   0000   0000   0200

                 ID = 0223        PLATE 3 TEMPERATURE
 01   0614   4208   0183   0000   0000   0200

                 ID = 0224        PLATE 4 TEMPERATURE
 01   0614   4208   0184   0000   0000   0200

                 ID = 0225        PLATE 5 TEMPERATURE
 01   0614   4208   0185   0000   0000   0200

                 ID = 0226        PLATE 6 TEMPERATURE
 01   0614   4208   0186   0000   0000   0200

                 ID = 0227        PLATE 7 TEMPERATURE
 01   0615   4208   0187   0000   0000   0200
```

RING BUFFERS FOR DATA ACQUISITION

```
                ID = 0228        PLATE 8 TEMPERATURE
01   0615   4208   0188   0000   0000   0200
```

CONTROL LOOP RECORDS

```
                ID = 0298        TOP PRODUCT CONTROL
01  621B  004C+04106  2AF0  9100-01310+09879  0000  0000  7FFF
02  7FFF  2990  7FFF  0000  2990  7FFF  0000  2990  116A  0000
03  00C0  0000  0000  000[illegible]  001A  003A

                ID = 0299        BOTTOM BRODUCT CONTROL
01  021B  007C  0177  0000  9100+20000  0000  0000  0000  7FFF
02  7FFF  2990  7FFF  0000  2910  7FFF  0000  2990  126A  0000
03  00C0  0000  0000  0000  0004  0056

                ID = 0296        TOP PRODUCT FLOW
01  6010  005F+04103  3340  8480+05382  0000  0000  0000  7FFF
02  7FFF  2110+02691+00000  2110

                ID = 0297        RING BUFFER TOP PRODUCT FLOW
01  0616  0008  0296  0000  0000  0200

                ID = 0170        FEED FLOW
01  621B  205C+04102  3340  8400+07214  0000  0000  0000  7FFF
02  7FFF  2990  7FFF  0000  2990  7FFF  0000  2990  106A  0000
03  0080  0000  0000  0000  0000  0000
```

APPENDIX 3

3-1) McCABE-THIELE DIAGRAM

3-2) THERMODYNAMIC CONSISTENCY OF
TEMPERATURE AND COMPOSITION DATA

1. McCABE-THIELE DIAGRAM

The McCabe-Thiele diagram, Figure 3-1, is constructed employing vapour-liquid equilibrium data given by Perry (78). The operating lines for the two sections of the column can be written as:

Section above the feed plate:

$$Yn = \frac{R}{R + D} X_{n + 1} + \frac{DX_D}{R + D}$$

substituting for R, D and XD from Table 2.1 yields:

$$Yn = .625Xn + 1 + 36.2$$

Section below feed plate:

$$Y_{m + 1} = \frac{R + F}{R + D} X_{m + 1} - BX_B$$

substituting for R, F, D, B, XB from Table 2.1 yields:

$$Y_{m + 1} = 1.41 X_{m + 1} - .635$$

(The temperatures of the feed and reflux streams were controlled to yield saturated liquids.)

The McCabe-Thiele diagram illustrates the operation of the column in terms of theoretical stages. The extremities of the column clearly operate in the "pinched" region of the vapour-liquid equilibrium relationship.

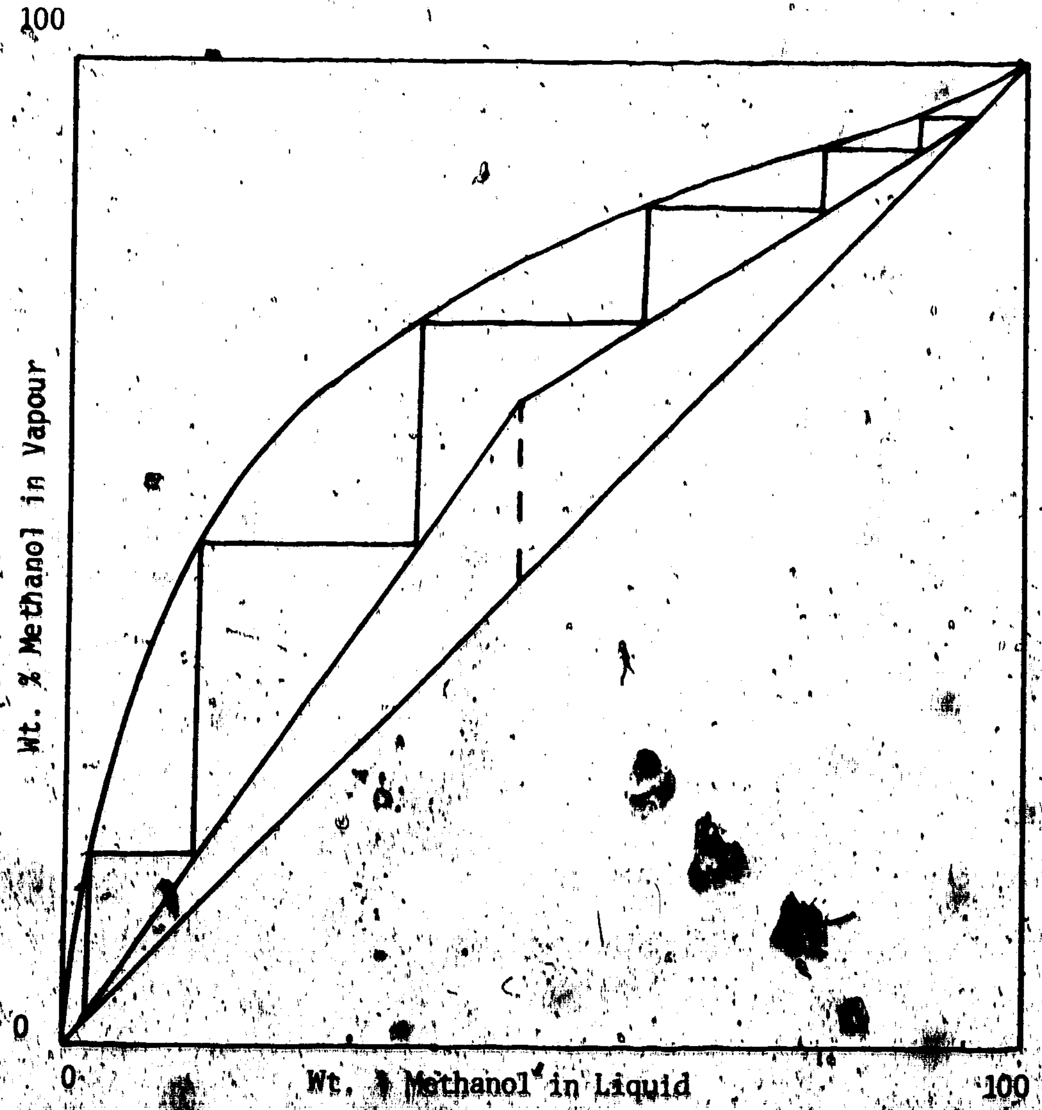

Figure 3-1
McCabe-Thiele Diagram

2. THERMODYNAMIC CONSISTENCY OF THE TEMPERATURE COMPOSITION MEASUREMENTS

When simultaneous composition and temperature measurements of a saturated liquid are available the measurements may be validated by comparison with published data. Temperature - composition data for a saturated solution of methanol and water (at 1 atmosphere total pressure) was obtained from Perry (78).

In the case of this study the condensor effluent was subcooled leaving the reboiler as the only location where temperature and composition measurements of the saturated liquid were available. A comparison of the published data with the operating data shows a composition of 1.3 wt. % MEOH at the operating temperature of the reboiler (209.6°F) versus 0.5 wt. % MEOH measured in the reboiler effluent. (The operating pressure of the distillation column was -.1 inches of H_2O.) Considering the purpose for which the composition measurements were used, a discrepancy of 0.8 wt. % is not unreasonable.

APPENDIX 4

CONTROL ALGORITHMS

NONINTERACTING COMPENSATOR:

The numeric forms of the noninteracting compensator difference equations are derived by substituting the model parameters from equation 3.1 into equation 5.2.

i Compensator D12:

$$Wo(n) = 1.20\ Wi\ (n-1.9) - 1.04\ Wi\ (n-2.9) - .887\ Wo(n-1)$$

ii Compensator D21:

$$Wo(n) = .437\ Wi\ (n-4.1) - .368\ Wi\ (n-5.1) - .796\ Wo(n-1)$$

RATIO CONTROL SYSTEM

This control system controlled the overhead product composition by manipulating the reflux to overhead vapour rate (V) ratio. The output of the proportional plus integral composition controller (θ_0) was used to set the desired R/V ratio

i.e. $\frac{R}{V} = \theta_0$

from which the desired reflux rate was computed.

i.e. $R = \theta_0 V,$

The overhead vapour rate was calculated from a material balance about the condenser (a constant condenser level is assumed).

i.e. $V = R+D$

The sensitivity of this calculation to a change in liquid level in the condensor can be evaluated from the change in condenser holdup per unit change in condenser level.

Cross sectional area of condenser

= .137 square feet.

Span of level instrument

= 0.5 feet.

Density of solution

= 50 lbs./cubic feet (S.G. of 96% methanol water solution at 144^{0}F = .80, ref. Perry (78))

Thus a change of 1% in the condenser level is equivalent to a change in liquid holdup of .034 lbs. This is equivalent to 1.1% of the normal overhead vapour rate.