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CONTROL OF A BINARY DISTILLATION COLUMN: AN EXPERIMENTAL EVALUATION OF FEEDFORWARD AND COMBINED FEEDFORWARD-FEEDBACK CONTROL SCHEMES

BY

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A THESIS

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ABSTRACT

The controlled behavior of a fully instrumented eight tray, nine inch diameter glass distillation column, separating a mixture of methanol and water, has been studied using feedback, feedforward and combined feed-forward-feedback control. Single point overhead product composition control was implemented by manipulation of the reflux flow rate compensating for disturbances in the feed flow rate. The dynamics of the column, determined using the pulse testing techniques, were successfully represented by first order plus time delay models, while the process gains were determined using the transient response testing procedure. Due to the nonlinear process responses exhibited by the column, two process models were determined, one for increasing and the other for decreasing overhead product composition responses.

The feedforward operating characteristics, measured experimentally, were found to be linear, resulting in a constant feedforward gain over the range of feed flow rates studied. The feedforward gain was subsequently found to exhibit a slight functional relationship with the feed flow rate during the on-line evaluation of the feedforward controllers.

Calculation of the frequency response of the feedforward controller using the frequency responses determined from the open loop pulse tests showed that the process responses could be characterized by simple transfer function models. However, the feedforward controller frequency response values calculated from the simple models used to predict the transient response did not agree with those calculated directly from the pulse tests.

Gain, gain plus time delay and gain plus time lag feedforward controllers; gain and gain plus time lag feedforward controllers with feedback trim were implemented using an IBM-1800 digital computer. The effectiveness of these control schemes was compared to that obtained using a conventional feedback controller for feed flow rate disturbances. Both increases and decreases away from and returning to a reference steady state were employed.

The results indicate that:

(a) gain feedforward control is not as effective as feedback control

(b) dynamic feedforward control in the form of either a time delay or first order time lag results in an improved control behavior compared with that obtained using only feedback control

(c) the addition of feedback trim to a feed-forward controller, besides eliminating offsets in the controlled response due to errors in the gain determination or immeasurable disturbances, also improved the effectiveness of the combined control scheme over that obtained using only the same feedforward controller.

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CHAPTER 1

INTRODUCTION

Distillation columns are one of petrochemical industry's most versatile and important unit operations. There are relatively few processes available which do not involve at least one distillation column. For a large number of these processes the distillation column is at the heart of the process, where the ability to maintain close control over the distillation column reflects the ability to maintain economic control over that process. This is exemplified in such processes as catalytic cracking, where even small increases in the recovered gasoline yield and the ability to maintain it can markedly increase the profitability of the unit.

Despite the extensive use of distillation columns in industry, they remain among the most difficult unit operations to control, due to the long characteristic time constants associated with the response of the column outputs. Since distillation columns are also generally among the final processing stages of a product, they, too, are subject to all the disturbances which may occur throughout the process.

The quality of the products produced from distillation columns is normally controlled using feedback controllers, which operate either on a direct measurement of the product quality or on an intermediate measurement, such as temperature, from which the product quality may be inferred. Many methods have been developed to select which intermediate variable gives the best control. However, the control of this intermediate variable does not generally guarantee that the product quality will remain constant, unless the exact relationship between the inferred and controlled variables is known exactly. For example, in a number of processes, temperature is used as an inferential variable supposed to reflect the composition. However, since temperature will only be an exact reflection of the composition for a binary system at constant pressure, control of a multicomponent product using a temperature to infer composition will not guarantee a constant composition product. The best measurement to use in a feedback control scheme is the measured value of the property of the product stream which it is desired to maintain constant.

It is impossible to completely eliminate a transient deviation in the controlled variable using only a feedback controller because, as the name implies, a deviation from the set point must occur before any corrective action can be introduced. The amount of the deviation that will result with a feedback controller, as measured by the integral of the absolute error (IAE), is proportional to the product of the proportional band, the reset time and the disturbance size (115). This product is generally large for most industrial columns due to the long time constants associated with these columns.

The feedback controller does not know, a priori, what control action to implement for a disturbance, but continually adjusts its manipulative variable until the measured value of the controlled variable and its set point are in agreement. In essence, the feedback controller eliminates the effect of a disturbance using a trial and error solution, characterized by the oscillatory behavior which often results.

The advantages of using feedback control are that essentially no knowledge of the process dynamics is required and that corrective action can be implemented no matter which input variable caused the disturbance. The major disadvantages include the slowness of response, the oscillatory control action and the large integral of the absolute error (IAE) obtained.

From the preceding discussion, it would seem obvious that if the corrective action could be introduced when the disturbance entered the column rather than after the disturbance has made itself evident in the product quality, a much more satisfactory control system would result. This is precisely the purpose of a feedforward controller. Feedforward control can, theoretically at least, completely eliminate the effect of a disturbance on the product quality, if the effect of the disturbance on the product quality can be described exactly. Since, in most cases, these relationships are very difficult to model, very simple functions, such as a first order lag (79), a first order lead-lag (115), a second order lag plus dead time (66) and a first order lead plus a second order lag (82) have been used to approximate these relationships. The most stringent constraint on these model descriptions is placed on the steady state, since small inaccuracies in the feedforward gain can be amplified into quite sizable offsets in the controlled variable due to the relatively large gains associated with the open loop responses. The values of the parameters involved in these approximate controller models are chosen to minimize the deviation (IAE) in the product quality. If the parameters of the approximate models are not chosen carefully, the controller may exhibit too much lead or lag action, resulting in a controlled response that would be less satisfactory than that obtainable using only a feedback controller.

The main advantage of feedforward control is its speed of response, with the lead action being obtained by introducing the corrective action on the basis of the measured disturbances entering the column rather than on the disturbances leaving the column. The major drawback of fee -forward control is that it has no indication whether or not it has taken the correct control action, making the effectiveness of the control dependant upon the accuracy of the feedforward model and the absence of unmeasured disturbances.

Since their advantages seem to complement each other, it is obvious that the feedforward and feedback schemes should be combined. The combined control scheme should exhibit the fast initial corrective action due to the lead action of the feedforward controller, and also the long-term removal of any offsets by feedback action no matter what their source. The presence of the feedforward controller does not make the process less difficult to control using feedback control; the gain in effectiveness is obtained because the feedback controller must do less work to incrementally trim the corrective action taken by the feedforward controller. In order to eliminate any offset, integral feedback control is the minimum necessary. Whether proportional or derivative control are beneficial depends upon the need to correct for only offsets due to errors in the model description or for deviations caused by immeasurable disturbances.

For a simple distillation column, the main measurable disturbances include the feed flow rate, the feed composition, the feed enthalpy and the reflux enthalpy. Special-purpose controllers have been proposed to control both of the latter (57). The disturbances which are generally not measured include the steam enthalpy, the cooling water enthalpy and the heat loss to the atmosphere. Both the reflux flow rate and the steam flow rate are considered as the major manipulative variables, although, depending upon the control scheme chosen, the overhead and bottoms flow rate and the feed plate location have also been considered. The overhead and bottoms product compositions are generally considered as the main controlled variables.

The majority of the published industrial effort (57, 66,68,82,90,115) appears to have been directed to developing feedforward controllers which compensate for feed flow rate disturbances, while most of the academic attention (18,30,56,69,70,71,79,105,133) has been focused on developing feedforward controllers compensating for feed composition disturbances. Feed flow rates tend to vary much faster and more often than does the feed composition and, in fact, both Shinskey (115) and Lupfer (57,68) suggest that feedback control designed on the basis of feed flow rate disturbances should be sufficient to compensate for the gradual feed composition changes. They do stress, however, that no standard solution can be applied to each situation, resulting in an individual analysis and probably a unique solution for each case.

Although industry seems to have accepted feedforward control as a standard control algorithm, surprisingly few results have been published. While Foxboro (34) has reported the installation of over 100 feedforward control systems, of which more than half have been applied to distillation columns, Scrimgeour (113) has recently reported that of over 150 feedforward control systems installed on distillation columns throughout Canada and the United States, only four applications have been described in the literature of the past eight years. The economic results reported (66,68,82,90,115) indicate that the gains possible in areas such as closer composition control, increased throughput or decreased utility usage can pay for the installation of a feedforward controller in an extremely short period of time.

The present study was initiated in order to obtain an experimental comparison of the effectiveness of a single-point feedforward controller with that obtained using a standard feedback controller subjected to disturbances in the feed flow rate. This information would hopefully fill an apparent void in the literature. It was also oped that a simple method for estimating the approximate feedforward controller parameters could be established, assuming only a minimal knowledge of the dynamic behavior of the column.

Although this study was not carried out on the complex multicomponent, multiplate, multiproduct columns characteristic of industrial columns, it was hoped that the principles outlined would help bridge the gap between theory and application.

CHAPTER 2

LITERATURE SURVEY

2.1 Introduction

The literature survey considered pertinent to this study will be presented under the following topics:

i) distillation dynamic models

ii) feedback control

iii) feedforward control

iv) pulse testing techniques

v) complex curve fitting techniques.

2.2 Distillation Dynamic Models

The various methods which have been employed to represent the complex relationships involved in distillation column dynamics will be reviewed. Williams (137) and Rademaker and Rijnsdorp (99) have presented excellent descriptions of the complete set of difference-differential equations necessary to describe plate dynamics. These relationships result from the combination of equations describing the material and energy balances, the phase equilibrium, the plate efficiency, the plate hydraulics, the plate mixing and the physical properties. These authors discuss the validity of some of the assumptions which are commonly used in order to simplify the set of equations so that they can be solved more readily. In addition, Gould (37) and Rijnsdorp (100) emphasize the importance of some of the secondary effects such as hydraulic delays and mixing on the ultimate dynamic response. Other review articles available include those of Archer and Rothfus (2), Rosenbrock (107), and the annual reviews of distillation literature published in Industrial and Engineering Chemistry (51).

Examination of the voluminous literature pertaining to the dynamics of distillation columns indicates that the information may be classified according to the assumptions made and the method of solution utilized. These topics include

(a) analytical solution of the linearized material balance equations
(b) analog computer solution of the linearized material balance equations
(c) frequency response stepping solution of the linearized material balance equations using a digital computer
(d) numerical integration of the nonlinear material and energy balance equations using a digital computer

(e) approximate models where the parameters are determined from the steady-state conditions .

(f) approximate models where the parameters are determined experimentally

(g) miscellaneous methods.

The articles of Armstrong et al (3,4,5,140,143) illustrate the early attempts to obtain analytical solutions to the linearized material balance equations. A number of simplifying assumptions, including the existence of a linear vapour-liquid equilibrium relation, must be made in order to obtain the analytical solution. These authors have compared the results of their analyses to experimental transients with moderate success. The largest deviations appeared in the portion of the response for short times.

Lamb et al (60,105) and Gerster et al (6,7,81,129) have presented a method of solving the linearized material balance equations developed at the University of Delaware using an analog computer. This procedure allows the use of a vapour-liquid equilibrium relation which has been linearized for each tray. The major disadvantage of this approach is the large analog computer which is generally required to solve even relatively small problems. The results predicted by this model compare quite favourably with some experimentally determined transients. Lamb and Rippen (105) have also described a method of solution of the same set of equations in the frequency domain using a stepping procedure on a digital computer. These calculations result in a series of Bode diagrams which illustrate the relationships between the input and output variables. The frequency response can then be fit using a transfer function of the appropriate degree of complexity or simplicity. A more recent description of this procedure has also been presented by Bollinger (11).

This model has been used extensively to describe the plant transfer matrix describing the response of a distillation column in subsequent control systems analyses by ~~Luyben~~ (69,70,71,72,73,80,123), Janis (56) and Wardle and Wood (132) among others. Wood (141) and Cadman et al (20,21,22) have also adapted this model to describe the dynamic relationships involved in multicomponent distillation.

Huckaba and co-workers (30,48,49) have presented a detailed theoretical and experimental evaluation of a nonlinear model. This procedure uses a specially developed predictor-corrector numerical integration algorithm to solve the set of differential equations, including the nonlinear material and energy balances, and nonlinear phase equilibrium and enthalpy relations. Similar models have been used in other dynamic studies by Svrcek (128), Davison (28) and Peiser and Grover (39, 95). These latter workers have included a description of the plate hydraulics, plate holdups, reboiler and feed preheater dynamics to analyze an industrial column which was experiencing unstable bottoms product control. The suspected cause of the problem, namely flooding, could not be established using a steady state model, but was subsequently verified by solution of the dynamic model. The modifications proposed, evaluated using the dynamic model, improved the situation and resulted in stable bottoms product control.

The calculation of the dynamic responses of distillation columns using the models outlined previously, generally requires a large general-purpose digital computer. In many cases, this approach to control systems cannot be justified nor can the large general-purpose computers be used for real-time on-line process control. It seems obvious then, that methods of either predicting or determining approximate models for the column is very necessary. A number of workers have attempted to correlate the parameters of simplified transfer functions based on resp ses calculated from linearized models similar to that presented by Gerster et al, generally with limited success. These studies include the work performed by Gilliland and Mohr (36), Mohr (87), Moczek, Otto and Williams (85,86), Bhat and Williams (10) and Wahl (133, 134). Wahl's approximate model is based on the solution of the model of Gerster et al, using the stepping procedure of Rippen et al for a large number of initial steady states. Examination of all the Bode diagrams indicated that the column responses could be represented by simple transfer functions of the form:

a) feed composition

for all plates $$\frac{X_N}{X_F} = \frac{K_{NX}\ (\tau_Z s+1)}{(\tau_1 s+1)\ (\tau_3 s+1)} \tag{2.1}$$

b) feed flow

for all plates below the feed point $$\frac{X_N}{F} = \frac{K_{NF}}{(\tau_1 s+1)} \tag{2.2}$$

for all plates above the feed point $$\frac{X_N}{F} = \frac{K_{NF}\ (\tau_Z s+1)}{(\tau_1 s+1)\ (\tau_3 s+1)} \tag{2.3}$$

c) vapour flow

for all plates $$\frac{X_N}{V} = \frac{K_{NV}}{(\tau_1 s+1)} \tag{2.4}$$

d) reflux flow

for all plates time response
$$\frac{X_N}{R} = \frac{K_{NR}}{(\tau_1 s+1)} \qquad (2.5)$$

for all plates high frequency response
$$\frac{X_N}{R} = \frac{K_{NR}}{(\tau_1 s+1)}\left\{\frac{1}{(\tau_H s+1)}\right\}^N \qquad (2.6)$$

The time constants ($\tau_1, \tau_3, \tau_Z, \tau_H$) calculated from these simple transfer functions have been presented as graphical correlations based on the initial steady state conditions. Both Wahl (133) and Beaverstock (8) have used this method to model a distillation column for an analysis of a feedback control system. Williams et al (10, 85,86) assumed the response of the column could be represented by a second order plus time delay transfer function. They have developed a relationship between these time constants and an overall time constant, called the inventory time, for the distillation column with its plate holdups lumped into an effective holdup. This method was used to analyze the control of a large industrial column producing very pure products (86).

Osborne et al (92) have proposed dividing the columns into sections and writing a set of linearized equations describing the material balance by lumping all the individual tray holdups in each section into a single overall value. The vapour and liquid material balance equations for each section are coupled by the mass transfer occurring between the phases. This mass transfer constant can be determined from the steady state relationships while the holdup parameter must be adjusted to fit the experimental response. By adjusting parameters, this method could predict experimentally determined responses as well as any other.

Approximate models determined using various standard dynamic testing procedures have been evaluated for a number of different industrial columns (50,98,112,114, 136,142). The models developed have been presented as Bode diagrams with no particular attempts made to fit these responses to an approximate transfer function. Rademaker (98) subjected a tall turbogrid column to both frequency and transient response tests to determine the response of the column to various input disturbances. These measured responses compared favourably with the responses predicted from a model outlined previously by Rademaker and Rijnsdorp (99). Wood and Robins (142) have compared the response of a 30 tray industrial column measured using step, sinusoidal and stochastic variations in various input variables to the responses predicted using the linearized perturbation model of

Gerster et al. The comparison was considered sufficiently successful to justify its use in a subsequent control study (132). Janis (56) also arrived at a similar conclusion by comparing the responses of a distillation column determined using the pulse testing technique with those predicted from the linearized model of Gerster et al.

Numerous miscellaneous methods have been proposed to determine the dynamic response of a distillation column. These include, among others, solution of the linearized material balance equations using flow graph analysis (52) and matrix algebra (64).

Levy et al (64) have presented an interesting comparison of the major time constants of three different model formulations based on an eigenvalue analysis. These models consisted of equations established from

a) component, mass and enthalpy balances

b) component and mass balances

c) component balance.

It is interesting to note that for all cases the dominant time constant is associated with the accumulation of chemical species which appears to be equally disturbed throughout the column. The second largest time constant of the more comprehensive model (a) indicates that the mode is associated with the thermal behavior of the accumulator. The temperature dependence of this mode disappears as the condenser and reboiler holdups are made smaller. This analysis indicates under what type of conditions the solution of the linearized component balance equation might fail to provide an adequate representation of the response of the column.

Although most of the column descriptions developed to date have concerned binary systems, a few articles have been presented for the multicomponent case (19,111, 141).

Examination of the results presented previously indicate that the numerical methods are advanced enough to solve the problem for either the linear or nonlinear case. Although the set of equations describing the dynamics can be easily developed, their solution is made difficult by the lack of sufficient knowledge concerning the evaluation of the secondary effects, such as dynamic plate efficiencies, plate hydraulics and plate mixing. Under most conditions the contributions of these effects are small, but can become important limiting conditions. The majority of the studies reported here involved the solution of linear models. However, a great number of industrial columns, ie product purification columns, operate in a nonlinear range where extremely pure products are produced. More work should be done to determine the effects of these nonlinearities. The large models are not generally useful for control system analysis due to the substantial amount of computer time required to obtain a solution. More work should be done to provide workable methods to predict approximate dynamic relations for the column, which could be used in either the analysis of, or as models in, an on-line computer control system.

2.3 Feedback Control

Since the theory of feedback control has been fully developed in most standard process control texts (16,27, 39,40,115), only those references of specific interest to distillation column control will be mentioned.

Numerous authors (44,74,101,108,109,138,139) have reviewed the various available methods which are commonly used to maintain control of the material and energy balances in the column. It is generally accepted that the product quality is much more sensitive to changes in the material balance rather than changes in the energy balance. Two methods are available to control the material balance. The first method, designated as direct, involves controlling the overhead product flow rate and allowing the reflux flow rate to be reset by the level in the reflux accumulator. The second method, designated as indirect, involves controlling the reflux flow rate and allowing the level in the reflux ~~accumu~~lator to reset the overhead product flow rate. Both Shinskey (115,116,117,118) and Nisenfeld (89,90) have advocated the direct approach, while Buckley (17) seems to prefer the indirect approach. Using both a steady state and dynamic simulation, McCune and Gallier (84) have presented a comparison of both approaches to the material balance control. They concluded that the direct approach would be the most preferred configuration. Despite these conflicts, the literature contains many references illustrating the popularity of both the direct and indirect material balance control schemes.

Lupfer et al (67) have outlined the development of various special-purpose controllers to control the internal reflux, feed enthalpy and the reboiler heat and bottom level.

If a direct analysis of the overhead or bottoms composition cannot be made, then control may still be obtained using any other measurable variable to which the controlled variables can be related. The control of product composition by control of an intermediate tray temperature has been analyzed by Jafri et al (55), Luyben (75,76,120), Shoneman and Gerster (119), Harriott (39) and Chanh (23). Chanh has compared the predicted controlled response of an eight tray pilot distillation column with that measured experimentally using various control plates.

The simultaneous control of both overhead and bottoms product compositions using only feedback control is complicated by the interactions experienced between the two manipulative variables, usually reflux and steam flows. Rijnsdorp (102,103,104), Luyben (73) and Berry (9) have illustrated various methods available, which are capable of decoupling the interaction, making the two-point feedback control scheme feasible. A recent article by Maselli and Miller (83) suggests that the interactions existing in a two-point feedback control scheme can be minimized by utilizing a control scheme at one end of the column, consisting of an intermediate plate temperature controller whose cascade set point is reset from the overhead composition controller.

2.4 Feedforward Control

Feedforward control has been successfully applied to numerous unit operations other than distillation columns, including reactors (42,77,80,131), driers (34), neutralizers (34), boilers (34) and evaporators (34,53,88). This section, however, will briefly describe only a few of the applications related to distillation column control.

The general theory concerning feedforward control has been described in a number of recent control texts (38, 45,79,115). Shinskey (116,117) has derived a simplified feedforward controller, based on a direct material balance control scheme, which maintains the ratios of D/F and S/F constant for disturbances in the feed flow rate and composition. Adequate dynamic compensation was obtained using a first order lead/second order lag dynamic compensator. This model was used by MacMullen and Shinskey (82) and Nisenfeld and Stravinski (90) as the basis for developing feedforward controllers for a large industrial superfractionator and an industrial azeotropic distillation column.

Lupfer et al (57,58,65,66,68) developed a simple feedforward controller which was used to control two industrial butane splitters. The model calculates the bottom product flow as a function of the feed flow and composition from the overall material balance. Steady state simulations were performed to obtain a relation between the reflux flow rate and the feed flow rate and composition. The reflux was then incorporated into the model using a regression equation which provided the best fit to the simulation results. A dynamic compensator consisting of a second order lag plus a time delay was added to the feed flow. Since the changes in feed composition generally occur with a low frequency, no dynamic compensation was considered necessary for a feed composition disturbance. Bornard et al (14) have applied a first order lead/lag feedforward controller to a large industrial superfractionator based on a model determined experimentally using a transient response analysis. Other industrial applications of feedforward control to fractionators have been presented by Skillern and Williams (121), Roach (108), Svrcek and Wilson (129) and Maselli and Miller (83).

Rippen and Lamb (105) have utilized their stepping procedure in conjunction with the linearized model of Gerster et al to determine the two-point feedforward controllers required to maintain both overhead and bottoms product compositions constant by manipulation of the reflux and steam flows to correct for disturbances in the feed flow rate and the feed composition. The feedforward controllers, presented as Bode diagrams, were fit with combinations of first and second order lag terms. The control of the distillation column, described by Gerster's model, was simulated on an analog computer using the approximate feedforward controllers. King et al (59) have obtained an expression for the steady state gain based on the solution of the algebraic equations of Rippen's stepping procedure.

Bollinger and Lamb (12) have formalized the method required to synthesize an ideal feedforward controller given the plant transfer matrix. This procedure was later extended to include feedback trim of the feedforward controller (13).

Luyben and his co-workers (69,70,71,72,73,78,79,124) have made extensive use of the procedures presented by Rippen and Lamb and Bollinger and Lamb throughout their numerous analyses of feedforward control systems. These studies have dealt with the determination of feedforward controllers required for two-point composition control by manipulation of reflux and steam flows (69,79), reflux flow and feed plate location (72,123), as well as the determination of feedforward controllers for columns which exhibit inverse responses (71). Luyben has fit the feedforward controllers used in these studies with simple combinations of first and second order lags. Simulation results obtained by Stull (127), using the linearized model of Gerster et al, indicate that very simple dynamics, such as first order lag for feed flow disturbances and no dynamic compensation for feed composition disturbances, are adequate for controlling most columns. Stiso (125), using the comprehensive model of Huckaba et al, has also shown the advantage of using simple dynamic compensation over static feedforward control.

Janis (56) has compared the feedforward controllers predicted using the linearized model of Gerster et al and calculated using Rippen and Lamb's stepping procedure with those calculated from the open loop responses determined experimentally using the pulse testing procedure. The two methods gave good agreement for low frequencies; with major deviations occurring at frequencies above approximately 1.6 radians/dimensionless time. (The dimensionless time was defined as total holdup/reflux flow).

Cadman et al (19,20,21) have also used the stepping procedure of Rippen and Lamb to calculate the open loop dynamic response of a multicomponent distillation column from the linearized model of Gerster et al. These open loop models were then used to calculate feedforward controllers for two-point control, and single point overhead and bottoms composition control using the method outlined by Bollinger and Lamb. The effectiveness of the various feedforward control schemes were studied on a simulated five plate distillation column separating a ternary mixture.

A number of other attempts have been made to experimentally illustrate the effectiveness of feedforward control as applied to a distillation column (1,18,30,31, 79,89). During these studies, the feedforward controller was not directly interfaced with the process. The control strategy was calculated for a specified disturbance off-line, usually on a large digital or analog computer, using the various feedforward models proposed, and then implemented manually.

A noteworthy exception was the study presented by Wardle and Wood (132). These workers have studied the effectiveness of feedforward control as applied to an industrial scale column. Both single-point control of overhead composition by manipulation of either reflux or boilup flow rate, and two-point control of overhead and bottoms composition by manipulation of both reflux and boilup flow rates were examined for disturbances in the feed composition only. The feedforward controllers were calculated from a linearized model similar to that of Gerster et al which had previously been verified experimentally (142), using the stepping technique of Rippen and Lamb. The Bode diagrams which were calculated were fit with polynomial transfer functions; third order lag for the single-point controller; third order lag plus first order lead for the two-point controller, using the complex curve fitting procedure outlined by Levy (63). The polynomial transfer function controllers were implemented using pneumatic analog elements. A comparison of the static and dynamic feedforward control was made by comparing the controlled response to the column's open loop behavior.

Under feedforward control, small offsets occurred in the overhead composition indicating that the determination of the steady state gain of the feedforward controller was not sufficiently accurate. The large initial error obtained under static feedforward control could be sharply reduced using the calculated dynamic feedforward controller.

Very few articles have been published concerning the implementation of combined feedforward-feedback control systems. Both Shinskey (115) and Lupfer (66) mention that feedback trim has been added to some feedforward control systems but give very few details. Luyben (70) has analyzed a feedforward controller with feedback trim from an intermediate plate temperature controller, but gives only scant simulation results. Maselli and Miller (83) have outlined a philosophy of feedforward control with feedback trim which is used at Sun Oil. This method uses a feedback trim from an intermediate tray temperature controller whose set point is reset from an overhead composition analyzer. This technique has proven successful on a number of columns, including deethanizers, butane splitters, depropanizers and debutanizers. However, very little operating data is given.

Wardle and Wood (132) added feedback trim to the single-point feedforward controller controlling the overhead vapour composition. They found that in addition to eliminating the steady state offset caused by inaccuracies in the model, it also acted as a dynamic element in the control loop, reducing the need for any further dynamic compensation.

In summary, it can be stated that only sparse experimental studies are available in the literature describing the effectiveness of feedforward control systems. Although some industries appear to have accepted feedforward control as a standard control strategy (14,83, 121), the details of the majority of the systems installed have not been published (113). Therefore, reliance must be placed on the many university studies which are mainly based on computer simulations for the majority of the information necessary for the synthesis and implementation of such systems.

2.5 Pulse Testing

The theses of Wildman (135) and Lees (62), the recent paper by Pollack and Johnson (97) and the monograph of Hougen (46) concerning the pulse testing procedure were consulted during this study. No attempt to survey the literature in this area was undertaken.

2.6 Complex Curve Fitting Techniques

Stroble (126) and Bosley and Lees (15) have recently surveyed the various methods available to fit transfer functions in the frequency domain. Chanh (23) has also made preliminary comparison between several complex curve fitting techniques, including that presented by Levy (63), and fitting of transfer functions to time data using the search procedure proposed by Rosenbrock (110). He concluded that curve fitting in the time domain was the preferred method. When compared amongst themselves, no obvious advantage could be claimed for any one of the frequency domain complex curve fitting techniques studied.

CHAPTER 3

PROBLEM FORMULATION

3.1 Introduction

This section contains a brief description of:

i) the pulse testing technique and the subsequent curve fitting of the frequency response values

ii) the general expression for ideal feedforward control and the modification necessary for its application

iii) the use of various loop record functions available in the direct digital control (DDC) program to implement the different forms of the feedforward controllers used.

3.2 Pulse Testing

Since a complete mathematical description of the pulse testing procedure has been given elsewhere (62,135), only a brief discussion will follow. By definition, the transfer function of a process, assuming the initial conditions are zero, is given by the Laplace transform of the output divided by the Laplace transform of the input.

$$G = \frac{\mathcal{L}(y(t))}{\mathcal{L}(x(t))} = \frac{\int_0^\infty y(t)e^{-st}dt}{\int_0^\infty x(t)e^{-st}dt} \tag{3.1}$$

The frequency response of the transfer function may be evaluated by substituting

$$s = i\omega \tag{3.2}$$

into Equation (3.1), giving

$$G(i\omega) = \frac{\int_0^\infty y(t)e^{-i\omega t}dt}{\int_0^\infty x(t)e^{-i\omega t}dt} \tag{3.3}$$

Substituting the Euler relation for $e^{-i\omega t}$

$$e^{-i\omega t} = \cos(\omega t) - i\sin(\omega t) \tag{3.4}$$

and replacing the infinite integration limit with a finite value T, Equation (3.3) may be reduced to a complex expression of the form

$$G(i\omega) = \frac{AA + iBB}{CC + iDD} \tag{3.5}$$

where

$$AA = \int_0^T y(t)\sin(\omega t)dt \tag{3.5a}$$

$$BB = \int_0^T y(t)\cos(\omega t)dt \tag{3.5b}$$

$$CC = \int_0^T x(t)\sin(\omega t)dt \tag{3.5c}$$

$$DD = \int_0^T x(t)\cos(\omega t)dt \tag{3.5d}$$

An on-line pulse testing analysis program (PTAP) (135) is available on the IBM-1800, which calculates the frequency response over the frequency range of interest using Equation (3.5) from the transient response time records for the input and output variables. The program also contains the option of evaluating the integrals using three different methods:

(a) Filon's method

(b) Trapezoidal rule

(c) Fast Fourier transform

as well as either listing the frequency response values or plotting them as a Bode diagram. Generally, the transient pulse test data was also saved on cards for later off-line analysis using the pulse test programs.

Wildman (135) also outlined a series of guidelines, which should be followed during the design of a pulse testing experiment. His recommendations were:

(a) obtain an approximate value for the dominant time constant (τ) of the process response to be studied;

(b) select a pulse duration that is of the same order of magnitude as the approximate time constant;

(c) employ a pulse height of such magnitude that a significant measurable response results, yet sufficiently small to cause a minimum excitation of any nonlinearities in the process response;

(d) use a sharp pointed pulse if possible, however, a rectangular pulse shape should be adequate for most industrial testing;

(e) the frequency range to be studied should include the decade above and below the decade containing the frequency value given by the reciprocal of the approximate time constant;

(f) the time record of the response should contain at least five approximate time constants of data in addition to approximately one time constant of initial steady state data;

(g) the sample interval for the data is given by

$$\Delta t = \frac{0.1\pi}{\omega_{max}} \tag{3.6}$$

while the total number of data points to be collected is calculated from

$$NPTS = \frac{5\tau}{\Delta t} \tag{3.7}$$

3.3 Frequency Response Curve Fitting

The mathematical development, which follows, is based on the outline presented by Chen and Haas (24) of the method proposed by Levy (63). The transfer function is defined in the frequency domain as a ratio of polynomial expressions of the form

$$G(i\omega) = \frac{a_0+a_1(i\omega)+a_2(i\omega)^2+\ldots+a_n(i\omega)^n}{b_0+b_1(i\omega)+b_2(i\omega)^2+\ldots+b_m(i\omega)^m} \qquad (3.8)$$

In order for the transfer function to be physically realizable, the order of the numerator must be less than or equal to the order of the denominator. Equation (3.8) can be rearranged to express both the numerator and denominator as a single complex number.

$$G(i\omega) = \frac{NUM(i\omega)}{DEN(i\omega)} = \frac{\alpha+i\beta\omega}{\sigma+i\gamma\omega} \qquad (3.9)$$

where

$$\alpha = (a_0-a_2\omega^2+a_4\omega^4-\ldots) \qquad (3.9a)$$

$$\beta = (a_1-a_3\omega^2+a_5\omega^4-\ldots) \qquad (3.9b)$$

$$\sigma = (b_0-b_2\omega^2+b_4\omega^4-\ldots) \qquad (3.9c)$$

$$\gamma = (b_1-b_3\omega^2+b_5\omega^4-\ldots) \qquad (3.9d)$$

If $F(i\omega)$ is a function, having a real part $R(i\omega)$ and an imaginary part $I(i\omega)$, which is defined to fit the experimental data exactly

$$F(i\omega) = R(i\omega) + i\,I(i\omega) \tag{3.10}$$

then an error function can be defined as the difference between the perfect fit transfer function $F(i\omega)$ and the 'best fit' transfer function $G(i\omega)$ as

$$E(i\omega) = F(i\omega) - G(i\omega) \tag{3.11}$$

$$= F(i\omega) - \frac{NUM(i\omega)}{DEN(i\omega)} \tag{3.11a}$$

Multiplying both sides of Equation (3.11) by $DEN(i\omega)$ gives

$$E(i\omega)\,DEN(\omega) = DEN(i\omega)\,F(\omega) - NUM(\omega) \tag{3.12}$$

which may now be expressed as a complex number

$$E(i\omega)\,DEN(\omega) = r(\omega) + is(\omega) \tag{3.13}$$

The magnitude of the modified error term is

$$|\,E(i\omega)\;DEN(\omega)\,| = \sqrt{r^2(\omega) + s^2(\omega)} \tag{3.14}$$

A weighted error function, E, is defined by squaring the modified error magnitude and summing it over all frequencies of interest.

$$E = \sum_{k=0}^{n+m} |DEN(\omega_k)\;E(i\omega_k)|^2 = \sum_{k=0}^{n+m} (r^2(\omega_k)+s^2(\omega_k)) \tag{3.15}$$

$$= \sum_{k=0}^{n+m} (\sigma_k R_k - \omega_k \gamma_k I_k - \alpha_k)^2 + (\omega_k \gamma_k R_k + \sigma_k I_k - \omega_k \beta_k)^2 \tag{3.16}$$

Equation (3.16) can be differentiated with respect to each of the unknown coefficients and the results set to zero, which will minimize the error function. A set of linear algebraic equations result of the form

$$\underline{\underline{WW}}\ \underline{COEF} = \underline{\underline{VV}} \tag{3.17}$$

whose solution may be obtained using standard matrix procedures, giving the coefficients for the polynomial transfer function. A program called LEVY has been written to solve Equation (3.17) for the best fit coefficients given

(a) the experimental amplitude ratio, phase angle and frequency values

(b) the number of experimental points

(c) the degree of the polynomial model required, n, m.

A relatively unsuccessful attempt was also made to include the determination of a best fit time delay. The time delay does not contribute to the amplitude ratio, only to the phase angle of the frequency response. Given the magnitude of the time delay, τ_D, the experimental phase angle was modified by

$$\phi_k = \phi_k - \tau_D \omega_k \tag{3.18}$$

The amplitude ratio and modified phase angle values were curve fit using LEVY. The program would attempt a simple search to find the 'best fit' value of the time delay yielding the minimum residual variance in the amplitude ratio data. However, the results obtained indicated that the evaluation of the time delay by this procedure was unreliable.

The major drawback in using the complex curve fitting procedure to estimate transfer function models is the evaluation of the significance of the fit. How good must the fit be in the frequency domain to yield a good approximation to the actual time response? Shoneman and Gerster (119) have recently suggested that fitting the amplitude ratio to within 10 decibels and the phase angle to within 40°, gave an adequate representation of the transient response. Although Wardle and Wood (132) considered the transfer functions fit using the Levy procedure to be sufficiently accurate to use in their control system analysis, they did not outline the criteria upon which this judgement was made.

3.4 Ideal Feedforward Control

Bollinger and (12) have presented a generalized procedure for calculating ideal feedforward controllers given the plant transfer matrix. Assuming the distillation column dynamics can be represented by a series of linear

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This equation describes how the reflux should be adjusted for disturbances in various input variables. Since only feed flow disturbances were studied during this project, the ideal feedforward controller of interest is described by

$$G_{ff} = - \frac{G_{PF}}{G_{PR}} \tag{3.23}$$

By a similar analysis, the feedforward controllers can also be derived, which would maintain both the overhead and bottoms product compositions constant. In this case, the model description (Equation (3.19)), can be simplified to

$$\begin{bmatrix} X_D \\ X_B \end{bmatrix} = \begin{bmatrix} G_{PF} & G_{PX} & G_{PT} & G_{PL} \\ G'_{PF} & G'_{PX} & G'_{PT} & G'_{PL} \end{bmatrix} \begin{bmatrix} F \\ X_f \\ T_f \\ L \end{bmatrix} \tag{3.24}$$

Substitution of the control strategy, (ie $X_D = X_B = 0$) and solving for the manipulated variables R and S gives

$$\begin{bmatrix} R \\ S \end{bmatrix} = - \begin{bmatrix} G_{PR} & G_{PS} \\ G'_{PR} & G'_{PS} \end{bmatrix}^{-1} \begin{bmatrix} G_{PF} & G_{PX} & G_{PT} & G_{PL} \\ G'_{PF} & G'_{PX} & G'_{PT} & G'_{PL} \end{bmatrix} \begin{bmatrix} F \\ X_f \\ T_f \\ L \end{bmatrix} \tag{3.25}$$

The ideal feedforward controller describing how both reflux and steam flow should be adjusted for feed flow disturbances is

$$G_{ff} = - \begin{bmatrix} \dfrac{G_{PF}\ G'_{PS} - G_{PF}\ G_{PR}}{G_{PR}\ G'_{PS} - G_{PS}\ G'_{PR}} \\ \\ \dfrac{G_{PR}\ G'_{PF} - G_{PF}\ G_{PS}}{G_{PR}\ G'_{PS} - G_{PS}\ G'_{PR}} \end{bmatrix} \qquad (3.26)$$

3.5 Implementation of the Ideal Feedforward Controller

The block diagram of the feedforward control scheme implemented during this project is given in Figure 3.1. The response of the reflux control loop to a change in its setpoint is given by

$$G_R = \frac{R}{R_{sp}} = \frac{G_{RC}\ G_{RV}}{1+G_{RC}\ G_{RV}\ G_{RM}} \qquad (3.27)$$

In order to satisfy the control criteria, ($e'_D = 0$), it is obvious that the effect of a disturbance in feed flow on the overhead composition, given by

$$X_D = G_{PF}\ F \qquad (3.28)$$

must be eliminated by an equal and opposite reaction from the controller

$$- \left[\frac{G_{PR} \; G'_{PF} - G_{PF} \; G'_{PS}}{G_{PR} \; G'_{PS} - G_{PS} \; G'_{PR}} \right] \quad (3.26)$$

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:liminated by an equal and opposite reaction from
'oller

$$X_D = \frac{G_{FM}}{G_{FR}} G_{ff} \frac{R}{R_{sp}} G_{PR} F \tag{3.29}$$

This may be expressed as

$$G_{PF} \quad F = - \frac{G_{FM}}{G_{RM}} G_{ff} \quad G_R \quad G_{PR} \quad F \tag{3.30}$$

Solving Equation (3.30) for the feedforward controller, G_{ff}, gives

$$G_{ff} = - \frac{G_{PF}}{G_{PR}} \frac{G_{RM}}{G_{FM}} \frac{1}{G_R} \tag{3.31}$$

This expression can be further simplified using the following assumptions based on the observations of the response of the reflux control loop and the flow measurement transducers:

(a) the gain of the closed loop transfer function of the reflux flow control loop to a set point change is 1.0,

(b) the dynamics of the closed loop response of the reflux flow control loop is fast compared to the response of the column,

(c) the dynamics of the flow measurement transducers are also fast with respect to the response of the column.

Equation (3.31) can then be reduced to

$$G_{ff} = - \frac{G_{PF}}{G_{PR}} \frac{K_{RM}}{K_{FM}} \tag{3.32}$$

In order to implement the ideal feedforward controller, the steady-state gain must be modified to reflect the different calibration ranges exhibited by the reflux and feed flow rate transducers.

3.6 Loop Record Functions Available in Direct Digital Control (DDC) Program

This section will describe some of the functions available in the direct digital control (DDC) program used to implement the feedforward and feedback controllers. Since a complete description of the control program exists elsewhere (29), only those functions of direct interest will be outlined.

3.6.1 Exponential Filter

$$\text{OUT} = (1-\text{FF})\ \text{INPT} + \text{FF MEAS} \tag{3.33}$$

where OUT - filtered value of measurement at this sample instant

INPT - raw unfiltered measurement at this sample instant

MEAS - filtered value of measurement at the previous sample instant

FF - filter constant $(0 \leq \text{FF} \leq 1)$

3.6.2 Proportional plus Integral Controller

$$\text{ERROR} = (\text{MEAS} - \text{SETPT}) + \text{BIAS} \tag{3.34}$$

$$DINTR = R\ KC\ KI\ ERROR + (REST1:REST2) \quad (3.35)$$

$$DPROP = R\ KC\ ERROR \quad (3.36)$$

$$OUT = DPROP + DINTR \quad (3.37)$$

where OUT - calculated controller output

DPROP- proportional controller contribution to control output

DINTR- integral controller contribution to control output

R - indicates whether controller is reverse acting

KC - proportional controller gain

KI - integral controller constant

REST1:REST2 - sum of past integral contributions to the control output (reset)

MEAS - current filtered measurement

SETPT- set point for control loop

BIAS - output bias

3.6.3 Proportional Controller

The proportional controller may be obtained as a special case of the proportional plus integral controller with the integral constant (KI) set to zero.

3.6.4 Ratio Controller

$$OUT = RATIO\ \ MCF\ \ (MEAS - SETPT) - BIAS \quad (3.38)$$

where RATIO - desired ratio

MCF - miscalibration factor, ratio of gains of wild measurement to controlled measurement

MEAS - current filtered wild measurement

SETPT - ratio of zero offset to gain of wild measurement (generally zero)

BIAS - ratio of zero offset to gain of controlled measurement (generally zero)

OUT - set point of controlled measurement

3.6.5 Engineering Units Conversion

$$OUT = A \;\; MEAS + B \tag{3.39}$$

where OUT - value of the current measurement in engineering units

MEAS - current value of filtered measurement

A - twice span of calibration

B - calibration offset

3.6.6 Data Accumulation Loop Record

The DDC program will allow the accumulation of historical data in a special type of loop record, with the oldest value in the last position. This loop can then be used to give a time delay.

EAS - current filtered wild measurement

ETPT - ratio of zero offset to gain of wild measurement (generally zero)

IAS - ratio of zero offset to gain of controlled measurement (generally zero)

UT - set point of controlled measurement

ngineering Units Conversion

$$= A \quad MEAS \quad + B \tag{3.39}$$

UT - value of the current measurement in engineering units

IEAS - current value of filtered measurement

- twice span of calibration

- calibration offset

)ata Accumulation Loop Record

: DDC program will allow the accumulation of histo-
ita in a special type of loop record, with the
/alue in the last position. This loop can then be
give a time delay.

TABLE 3.1 ALGORITHMS FOR LOOP RECORDS USED TO IMPLEMENT THE CONTROL SCHEMES

loop record number	loop record functions used
0170	data acquisition of feed flow
0175	data acquisition of overhead composition
0250	input from measurement of 0170 time delay algorithm
0251	input from either measurement of 0170 or oldest value in 0250 proportional plus integral control algorithm exponential filter algorithm
0252	input from output of 0251 ratio algorithm
0260	input from output of 0252 proportional plus integral control algorithm
0261	input from measurement of 0175 proportional plus integral control
0262	input from output of 0260 proportional plus integral control algorithm output to current output station

This function is implemented on the IBM-1800 using the proportional plus integral controller function represented by

$$G_c(t) = KC(1+KI\Delta t\Sigma\varepsilon) \tag{3.41}$$

Figure 3.2 illustrates the values of the important loop record parameters for the feedback controller. The composition measurement is obtained from the multiplexer. The calculated output was applied to the current output station (COS), whose current output acts as the process set point. A feedback controller requires only a single loop record.

3.7.2 Gain Feedforward Control

The transfer function describing a static feedforward controller is given by

$$G_{ff} = K_{ff} \tag{3.42}$$

Four loop records were required to implement this controller using the loop record functions illustrated in Figure 3.3. The first loop record calculates the deviation of the feed flow rate disturbance from its reference steady state value using the proportional control algorithm with a unit gain. The second loop, set up using the ratio algorithm, calculates the necessary reflux correction required. The

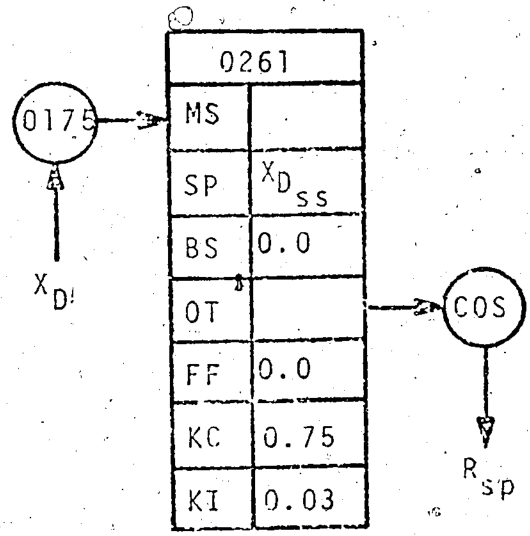

FIGURE 3.2 LOOP RECORD ORGANIZATION FOR FEEDBACK CONTROL

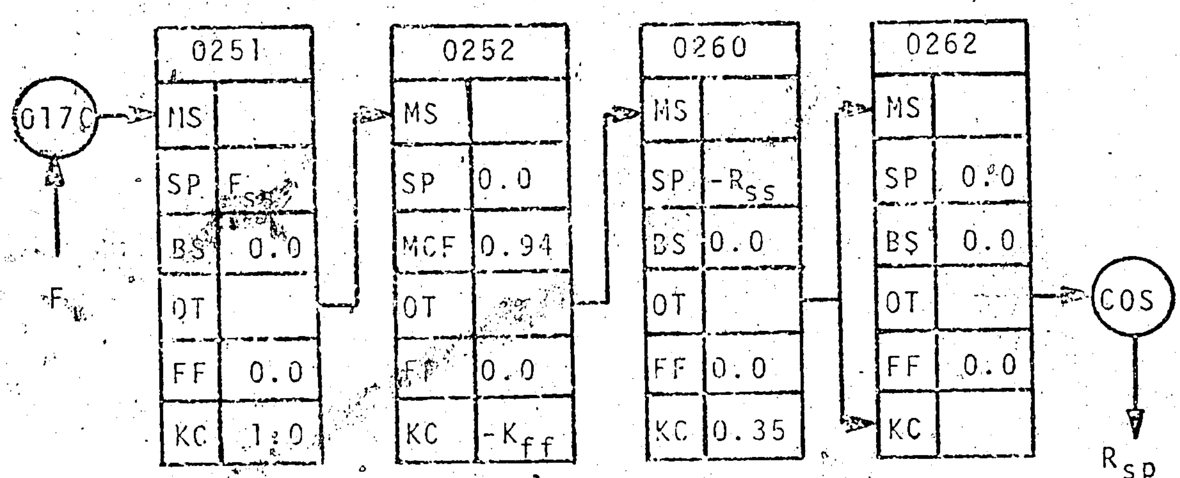

FIGURE 3.3 LOOP RECORD ORGANIZATION FOR GAIN FEEDFORWARD CONTROL

third loop, set up as a proportional controller, adds the reflux correction to the reflux set point. This loop also adjusts the value of reflux prior to calculating the square of the reflux in the fourth loop. The output of the fourth loop, a signal which is proportional to the reflux squared, is cascaded as the set point of the reflux flow controller.

3.7.3 Gain plus Time Delay Feedforward Control

The transfer function describing this controller is given by

$$G_{ff} = K_{ff} e^{-\tau_D s} \tag{3.43}$$

Five loop records are required to simulate this controller using loop record functions. The first loop is a special data accumulation loop record. The length of the loop, N, is specified as the closest integer value which satisfies the following relation

$$N \ \Delta t = \tau_D \tag{3.44}$$

The remaining loop records are identical to those described previously for the static feedforward controller, except that the second loop, which calculates the deviation of the feed flow disturbance, obtains its measurement from the oldest historical value in the data accumulation loop record, rather than directly from the multiplexer. This arrangement of loop records is illustrated in Figure 3.4.

3.7.4 Gain plus Time Lag Feedforward Control

The transfer function describing this controller is expressed as

$$G_{ff} = \frac{K_{ff}}{\tau s+1} \tag{3.45}$$

Four loop records are required to implement this controller using standard loop record functions. These loop records are identical to those described previously for gain feedforward control except in this case the first loop applies an exponential filter to the measurement it obtains from the multiplexer. The relation between the filter factor (FF) and the time constant (τ) is given by Jacobson and Fisher (54) to be

$$FF = e^{-\frac{\Delta t}{\tau}} \tag{3.46}$$

This expression is valid for filter factors in the range

$$0.75 \leq FF \leq 1.0 \tag{3.47}$$

Figure 3.5 illustrates this arrangement of loop records.

ain plus Time Lag Feedforward Control

transfer function describing this controller is
d as

$$= \frac{K_{ff}}{\tau s+1} \qquad (3.45)$$

p ~~records are~~ required to implement this controller
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the time constant (τ) is given by Jacobson
er (54) to be

$$= e^{-\frac{\Delta t}{\tau}} \qquad (3.46)$$

ression is valid for filter factors in the range

$$5 \leq FF \leq 1.0 \qquad (3.47)$$

.5 illustrates this arrangement of loop records.

3.7.5 Feedforward-Feedback Control

Combined feedforward and feedback control is implemented with a gain and a gain plus time lag feedforward controllers. The loop records required are illustrated in Figures 3.6 and 3.7 respectively. These loop record configurations result from the combination of those required for feedback control and those required for the feedforward control. The only change required is that instead of applying the output of the feedback loop directly to the current output station, it is used to bias the output of the feedforward controller.

3.7.6 Other Feasible Controllers

A first order lead/lag feedforward controller could be employed by combining loop record functions for exponential filtering and a proportional plus derivative controller. The time constant of the lag element is still given by Equation (3.46), while the time constant of the lead is given by

$$\tau_Z = K_D \quad \Delta t \tag{3.48}$$

Some preliminary tests were performed using this controller, but when the results using the much simpler time lag model proved satisfactory, further testing of this control-

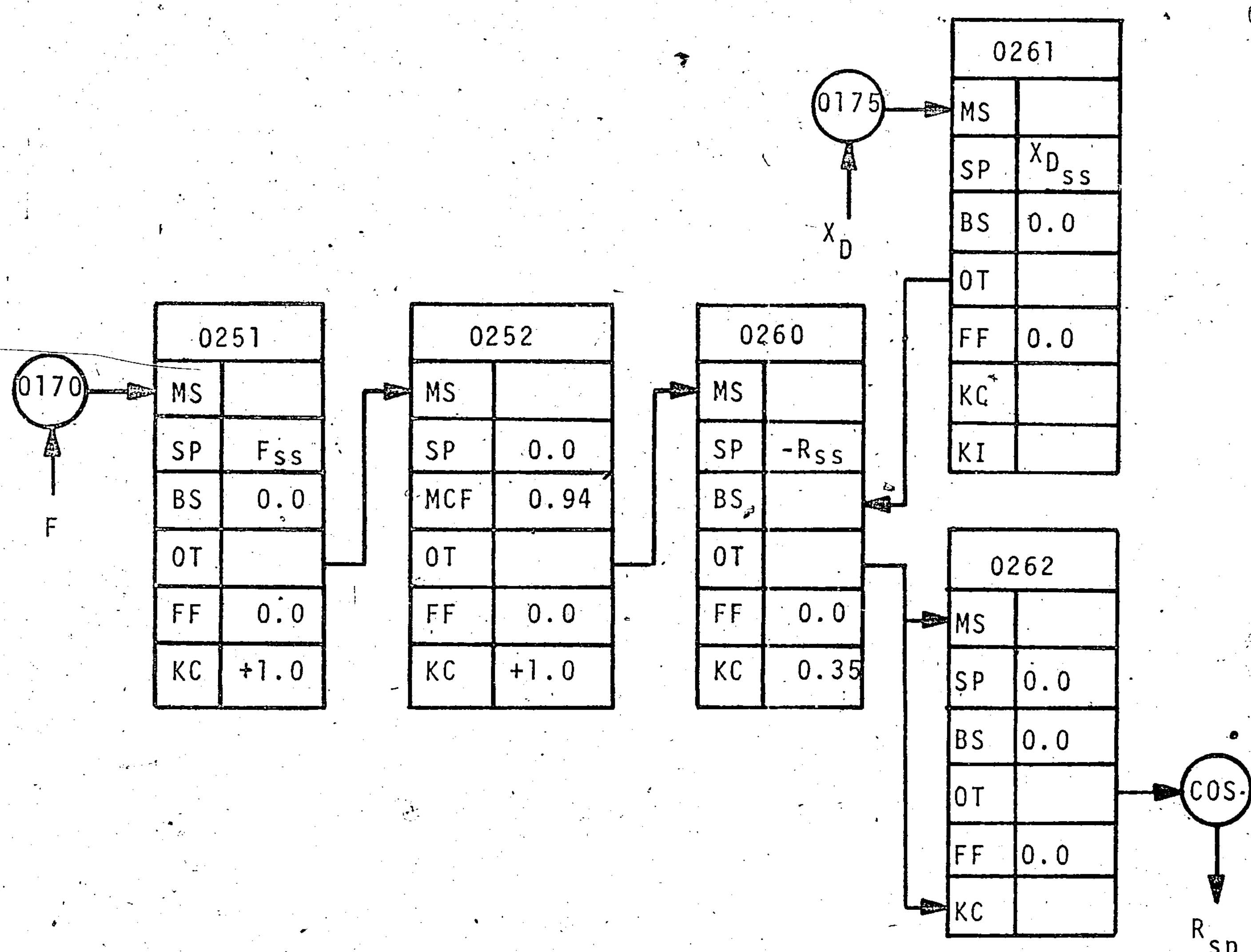

FIGURE 3.6 LOOP RECORD ORGANIZATION FOR GAIN FEEDFORWARD CONTROL WITH FEEDBACK TRIM

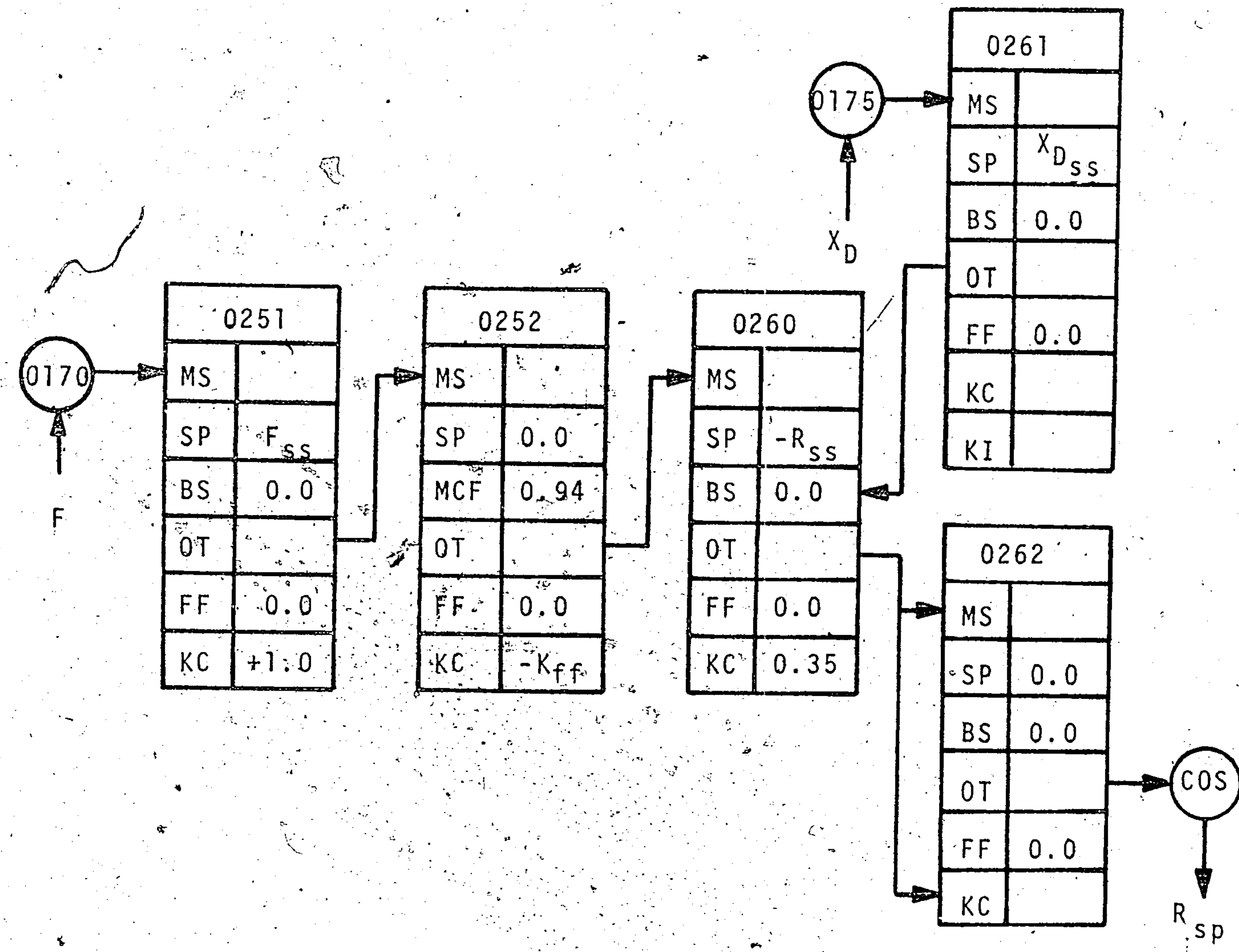

FIGURE 3.7 LOOP RECORD ORGANIZATION FOR GAIN PLUS FIRST ORDER TIME LAG FEEDFORWARD CONTROL WITH FEEDBACK TRIM

ler was discontinued.

It appears that a general transfer function of the form

$$G_{ff} = \frac{\sum_{j=0}^{n} a_j s^j}{\sum_{i=0}^{m} b_i s^i} \tag{3.49}$$

could be obtained by using a series of n proportional plus derivative controllers in series with m exponential filters. This controller would be restricted to those transfer functions in which no complex roots appear.

3.7.7 Discussion

It is important that the loop records be calculated in the proper order, and not be allowed to get out of phase. This can be accomplished by adjusting the phase time parameter of the loop record. The proportional controller algorithm could also have been used in place of the ratio algorithm, which, in some cases, would have saved a loop record. This method was not used because use of the ratio algorithm allowed the storage of the feed-forward gain and the miscalibration constant parameters separately. If the proportional control algorithm were used, it would necessitate lumping these terms into a single parameter.

Although the digital computer is actually a sampled data controller, the analysis which has been performed has assumed that the computer acts as a continuous controller. This assumption is justified since the sample time of 16 seconds is small compared to the dominant time constant of the column, which is of the order of 600 seconds.

CHAPTER 4

COLUMN DESCRIPTION AND OPERATION

4.1 Introduction

This chapter contains a brief description of the distillation column and its ancillary equipment, a tabulation of the process variables that were monitored, a discussion of the data transducing required between the process and the computer and a discussion of the control schemes employed during this investigation. The operating procedures required for column start-up and shut-down and the calibration techniques used are also outlined. The chapter is concluded with a discussion of the initial steady state conditions selected and the material balance errors of closure typical of those obtained throughout the study.

4.2 Distillation Column and Ancillary Equipment

A nine inch diameter, eight tray glass distillation column separating a binary mixture of methanol and water was used to study the effectiveness of the various control schemes studied during this project. Each plate contained four 2 1/4 inch by 1 7/8 inch bubble caps mounted in a square pattern, 1 1/2 inch diameter inlet and outlet weirs, and a one inch diameter downcomer. Each plate was also equipped with a feed location inlet, two thermocouple slots for measuring the plate temperature, a toggle-valve sample line for obtaining a sample of the plate liquid and two sample points used for the continuous plate composition sample (not installed for the present project (128)). The plate spacing was 12 inches.

Four three foot diameter tanks are available to contain the fresh feed and each of the two product streams. These tanks are piped to allow either continuous or batch mode operation. Normally the methanol-rich product is stored in one of the tanks, labelled the top product tank, and the water-rich product is stored in another tank, labelled the bottom product tank. This leaves two tanks available for storage of fresh feed. Two fresh feed tanks are normally required for studies concerning feed composition disturbances. For batch operation, the desired feed composition is prepared in one of the feed tanks by mixing the required amounts of water and methanol from the product tanks. The contents are circulated through the feed pump and back to the feed tank to ensure a uniform composition. While the column is in a batch operation, the feed to the column is pumped from the feed tank, while the products are withdrawn from the column to their respective product tanks. For continuous operation, the feed is withdrawn from the feed tank and the products are returned to that feed tank. At present the feed can be introduced on either the fourth or fifth tray from the bottom through a small shell and tube feed preheater.

The reboiler is a vertical basket type with condensing steam on the shell side. It consists of 38 1/2 inch tubes, two feet in length with a three inch diameter central downcomer to facilitate liquid recirculation. The tube sheet is six inches in diameter. The section above the tube sheet, the glass vapour return line and an extra glass section below the bottom plate, act as the vapour/liquid disengaging volume for the reboiler. The bottom plate of the reboiler contains facilities for installation of a thermocouple, the low level leg of the level differential pressure cell, the bottom product withdrawal, a liquid sample toggle valve, a capacitance probe and its temperature compensation thermistor, (which were not implemented during this project (128)) and the outlet and inlet for a small continuous sample to be withdrawn through an on-line process gas chromatograph. The bottom product, withdrawn from the reboiler base, is returned to the appropriate tank through a series of small coolers.

The vertical nine inch diameter glass condenser contains as many 1/2 inch diameter U-shaped tubes as the volume would allow. The cooling fluid is water on the tube side. The reflux accumulator consists of a nine inch diameter to 1 1/2 inch diameter glass reducer at the base of the condenser. The base plate is equipped with taps for a capacitance probe and its temperature compensating thermistor, a thermocouple, the high and low legs of the level transducer and a liquid sample toggle valve. The overhead product is split with the reflux being pumped back to the top plate through a small shell and tube reflux preheater and the overhead product being pumped back to the appropriate tank through a series of small coolers.

The overhead composition was measured with a small parallel plate capacitance probe connected to a Foxboro Dynalog capacitance recorder/controller. The plates of the probe are approximately 1/4 inch square with a gap of about 1/8 inch. The plates are imbedded in a plug of teflon approximately 1/2 inch diameter. The capacitance analyzer measures the difference in the dielectric constant of the mixture due to changes in the composition. Since the dielectric constant is also sensitive to temperature variations, the recorder is equipped with a thermistor which detects the temperature so that compensation can cancel this effect. The basic theory of this method and detail concerning the design of the capacitance cell are available from Svrcek (128).

The bottom composition is analyzed every 50 seconds using an on-line industrial gas chromatograph. A small stream is circulated from the reboiler bottoms past the chromatograph sample valve and returned to the reboiler. The chromatograph is under complete control of the IBM-1800 computer. The computer generates a control signal causing the chromatograph to take a sample. The computer then monitors the detector current, detects the peaks, calculates the compositions, presents the report, and then commences another analysis cycle by commanding that the next sample be taken at the appropriate time. The chromatograph may also be used to obtain an analysis of the feed composition, when necessary. Operating information and procedures as well as a detailed description of the chromatograph system are available from Berry (9).

All flows are measured with flange tap orifice plates. The reflux, feed overhead and bottoms product flow rates were sufficiently small that quadrant edge (130) orifices could be used. The remaining flows, steam and cooling water were measured using standard square edge orifices. The differential pressure created at the orifice was transduced to a 3-15 psig pneumatic signal using a Foxboro differential pressure cell. The levels in the condenser and reboiler were measured with low range (30 inch) Foxboro differential pressure cells. The column pressure was measured using a Foxboro Pressure Transmitter with a range of -10 to +10 inches of water. All temperatures were measured using iron-constantan thermocouples. A more detailed description of the column and its associated instrumentation is available from Pacey (93). A detailed schematic diagram of the column is presented in Appendix B.

It is worthwhile to note that the column, as used by Svrcek (128), was disassembled and reinstalled in a new laboratory prior to the testing reported in this work. During reinstallation, the column piping was modified so that various possible control schemes could be studied with a minimum of physical effort expended in reconfigurating the control system. Most of the pneumatic instrument lines were installed with a flexible plastic tubing, which can very easily and quickly be changed. Also, all recorders, indicators and controllers were placed in special modules which were installed in a series of cabinets. This modular type of instrumentation permitted the standardization of modules which facilitate the borrowing of a similar instrument from other equipment in the event of an instrument failure.

4.3 Process Variables Monitored

Throughout this project, the data acquisition capability of the IBM-1800 DACS computer facility was used to measure and record the process variables listed in Table 4.1. All variables are available to the computer through the multiplexer, but not all variables were supplied with a loop record. A loop record makes the measurement of a variable available to the direct digital control (DDC) program, which performs the bulk of the data acquisition and control functions. Other special-purpose data acquisition and control functions were performed by user written programs by either direct access to the multiplexers, or by obtaining the measurements from the DDC loop records. In addition, the values of most of the process variables are also available from standard analog indicating and recording instruments. Table 4.1 also indicates that sufficient analog controllers are also available to completely control the column, and that only a few of the digital controllers have been implemented. The column at present can be operated under analog control, but a few extra digital control loops must be implemented before it can be run completely via

TABLE. 4.1 LIST OF PROCESS VARIABLES

Name	Designation	Loop Record Number	Multiplexer Number	Measurement Digital	Measurement Analog	Control Digital	Control Analog
Flows							
feed	FRC-1	0170	[illegible]	yes	yes	yes	yes
reflux	FRC-2	0171	[illegible]	yes	yes	yes	yes
steam	FRC-3	0172	[illegible]	yes	yes	yes	yes
overhead product	FR -4	0174	[illegible]	yes	yes	no +	no
bottom product	FR -5	0173	4104	yes	yes	no +	no
cooling water	FR -6		4105	yes	yes	no +	no
Compositions							
feed	AR -1	0178		no	no	no	no
overhead product	ARC-2	0175	4106	yes	yes	yes	yes
bottom product	AR -3	0176		yes	yes	no	no
Levels							
reflux accumulator	LIC-1		4107	yes	yes	no *	yes
reboiler	LIC-2		4108	yes	yes	no *	yes
Pressures							
column overhead	PIC-1		4109	yes	yes	no *	yes
column differential	PIC-2		4110	yes	yes	no	yes

LIST OF PROCESS VARIABLES

Name	Designation	Loop Record Number	Multiplexer Number	Measurement Digital	Measurement Analog	Control Digital	Control Analog
Temperatures							
reboiler	TR-4.1	0180	00130	yes	yes	no +	no +
plate 1	TR-4.2	0181	00131	yes	yes	no +	no +
plate 2	TR-4.3	0182	00132	yes	yes	no +	no +
plate 3	TR-4.4	0183	00133	yes	yes	no +	no +
plate 4	TR 4.5	0184	00134	yes	yes	no +	no +
plate 5	TR-4.6	0185	00135	yes	yes	no +	no +
plate 6	TR-4.7	0186	00136	yes	yes	no +	no +
plate 7	TR-4.8	0187	00137	yes	yes	no +	no +
plate 8	TR-4.9	0188	00138	yes	yes	no +	no +
reflux accumulator	TR-4.10		00139	yes	yes	no	no
steam	TR-4.11		00140	yes	yes	no	no
steam condensate	TR-4.12		00141	yes	yes	no	no
reflux orifice	TR-4.13		00142	yes	yes	no	no
feed orifice	TR-4.14		00143	yes	yes	no	no
bottom product orifice	TR-4.15		00144	yes	yes	no	no
reboiler overhead	TR-4.1[illegible]		00145	yes	yes	no	no
feed inlet	TR-4.20/TRC-1		00146	yes	yes	no +	yes
reflux inlet	TR-4.21/TRC-2		00147	yes	yes	no +	yes
column overhead vapour	TR-4.22		00148	yes	yes	no	no
cooling water inlet	TR-4.23		00149	yes	yes	no	no
cooling water outlet	TR-4.24		00150	yes	yes	no	no

*These digital controllers were not implemented during this project+, but the necessary digital to analog transducers are available.

+The controllers using intermediate plate temperatures (23) have not been implemented during this project, but the necessary analog controller and digital to analog transducers are available.

computer control.

4.4 Data Conversion Between the Process and the Computer

Before the computer can perform any function, it must have the process information available in a form which it can use. The process variables are available in physical units, ie ^{o}F, psig, in H_2O and lb/min, but the computer requires a discrete digital value, ie 7209. Similarly, to make use of the information generated by the computer, this digital data must be converted into a physical variable in order to drive the final control element. This section will illustrate the methods used during this project to convert the physical process variables into digital values, and the digital computer outputs to physical control variables. Figure 4.1 illustrates the various steps involved with the data transducing process.

4.4.1 Analog to Digital Conversion

The first step in converting the process variables is to generate a signal, either pneumatic or electronic, which is proportional to the measurement. The flows are converted to a pneumatic 3-15 psig signal using

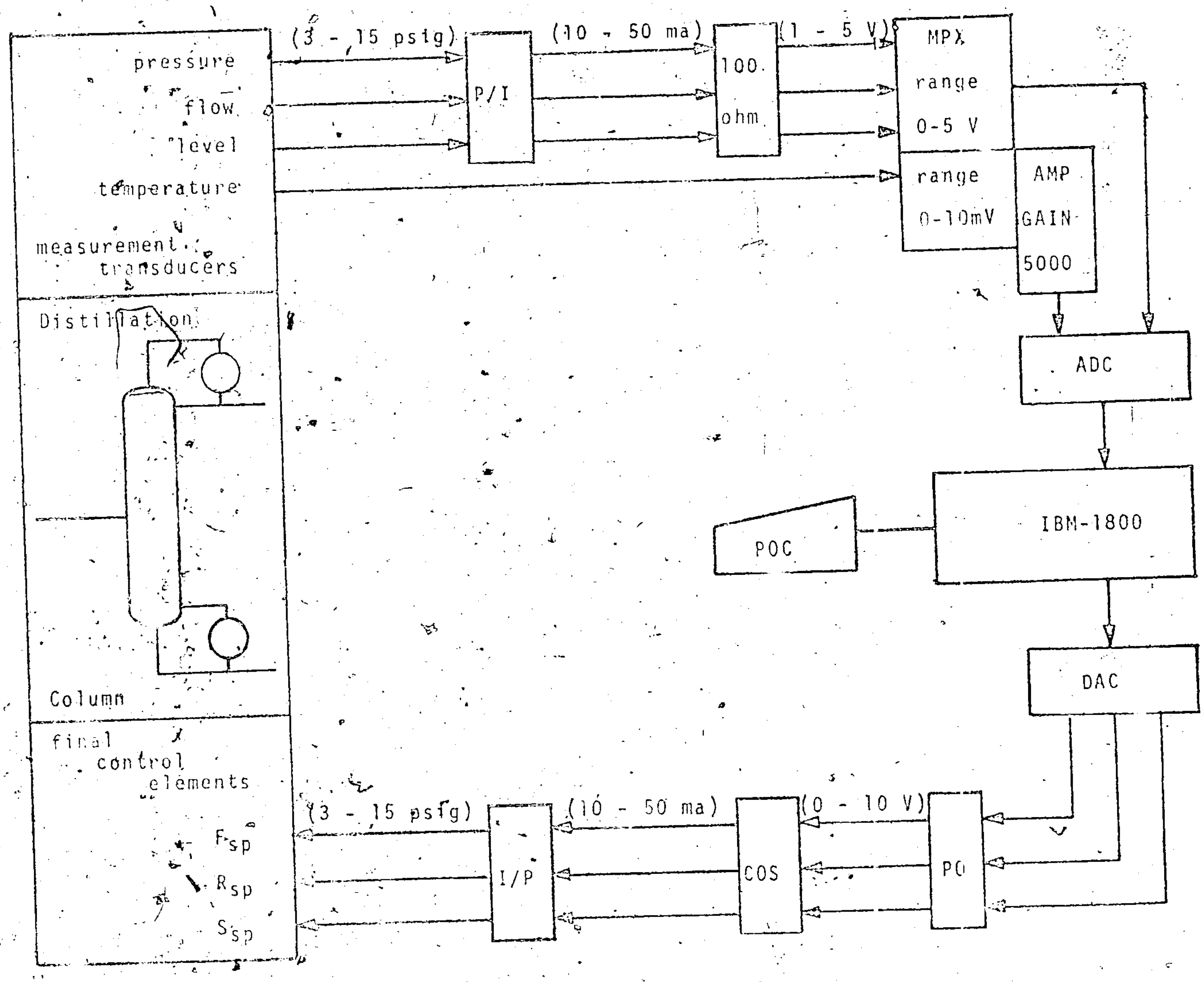

FIGURE 4.1 INFORMATION FLOW BETWEEN THE PROCESS AND THE COMPUTER

an orifice plate and a differential pressure cell. The levels and column differential pressure are also converted to pneumatic 3-15 psig signals using low range differential pressure cells. An absolute pressure transmitter is used to convert the column pressure to a pneumatic 3-15 psig signal. The overhead composition is converted to a pneumatic 3-15 psig signal using a special pen position transducer for use with the Dynalog capacitance recorder. A 10-50 milliampere current proportional to the pneumatic signal is produced from a pressure to [illegible]I) convertor for each pneumatic measurement. The current is further converted to a proportional 1-5 volt electrical signal across a 100 ohm resistor. All temperatures are converted to a 0-10 millivolt electrical signals using iron-constantan thermocouples. These electrical signals are transferred to the computer, usually on shielded twisted pair wires, and each variable is assigned to a unique set of terminals on the multiplexer. The multiplexer is essentially a bank of relay switches. When the computer requires a measurement, the appropriate multiplexer relay is closed, applying the voltage signal to the analog to digital (ADC) converter. The range of the analog to digital converter is normally 0-5 volts. Thus low level electrical signals, ie 0-10 millivolt temperature measurements, must be applied to the analog to digital converter through an appropriate amplifier to produce a proportional 0-5 volt signal. The continuous voltage signal is digitized by the analog to digital converter in the range 0-32767, which is the largest integer contained in the 16 bit word length of the IBM-1800. This measurement is now available to be read into the appropriate core storage location to be used by the computer software.

4.4.2 Digital to Analog Conversion

When the computer has completed the calculation algorithm and is ready to apply the result to the process, it is available in an appropriate core storage location as a binary integer in the range 0-32767. This digital value is applied to a digital to analog converter (DAC), resulting in a proportional 0-10 volt electrical signal. The computer activates the appropriate pulse output (PO) relay which applies the 0-10 volt electrical signal to the current ouput station (COS). The current output station is essentially a zero order hold device which maintains the input signal until changed the next time. The current output station converts the applied voltage to a proportional 10-50 milliampere electrical signal. The 10-50 milliampere electrical signal is further converted to a proportional 3-15 psig pneumatic pressure signal, which is used to drive the final control elements.

The final control elements in this project consisted of the set points of the pneumatic flow control systems. These signals could just as easily have been applied directly to the control valves.

4.4.3 Comments

The major problem associated with signal conversion and transmission is noise pick up, which should be avoided if possible, or at least minimized. The transmission distance of both current and pneumatic signals should be minimized, using high level voltage signals, where possible, to transport process information over longer distances. The high level voltages are preferred for data transmission because of the smaller amount of signal deterioration experienced due to noise pick up than with a low level signal. When low level signals must be used, ie thermocouple voltages, care must be taken not to run the transmission lines past noise generators such as fluorescent lamps, etc.

Pneumatic instrumentation is not normally recommended for computer control applications, since equivalent electronic instrumentation is available which can eliminate some of the required data conversion, ie pneumatic to current. Less opportunities for signal distortion can result from these fewer conversions. However, the pneumatic column instrumentation already existed and the data acquisition system was designed using them.

4.5 Description of the Control Schemes Implemented

4.5.1 Degrees of Freedom

According to Smith (122), the number of degrees of freedom available in defining the steady state operation of a distillation column with a total condenser and a partial reboiler are

$$DF = COMP + 2\ NSTAGE + 9 \tag{4.1}$$

However, Howard (47) has suggested that, for control purposes, this expression is inadequate since it neglects the additional transient degrees of freedom related to the holdups on each stage. This would add an additional NSTAGE + 1 degrees of freedom, so that the total degrees of freedom would become

$$DF = COMP + 3\ NSTAGE + 10 \tag{4.2}$$

Since the control engineer is concerned with an existing column, the following variables are generally fixed by the design of the column:

pressure drop in each stage	NSTAGE
pressure drop in reflux divider	1
heat loss in each stage (except reboiler)	NSTAGE - 1

heat loss in reflux divider		1
holdup in each stage (except reboiler)	NSTAGE	- 1
feed plate location		1
total number of plates		1
Total:	3NSTAGE	+ 2

In addition to these variables, the reflux temperature is generally fixed by the operation of the condenser while the feed conditions are usually determined by the operation of the downstream processes, fixing the following degrees of freedom:

feed conditions	COMP	+ 2
reflux temperature		1
Total:	COMP	+ 3

Of the COMP + 3 NSTAGE + 10 degrees of freedom available to define the column operation, COMP + 3 NSTAGE + 5 degrees of freedom have now been fixed by the column design and operation, leaving only five variables available for specification by the control engineer.

Of these specifications, one is used to define the column pressure, while two are used to define the levels in the condenser and the reboiler. The remaining two degrees of freedom are used to maintain the material and energy balances. The specification of these degrees of freedom, as defined in this study, include:

a) cooling water flow rate (column pressure)

b) overhead product flow rate (condenser level)

c) bottoms product flow rate (reboiler level)

d) reflux flow rate (material balance)

e) steam flow rate (energy balance)

With these specifications, the column operation is completely defined. The control scheme implemented is illustrated schematically in Figure 4.2.

4.5.2 Material and Energy Balance Control

Control of the distillation process requires the maintenance of the material and energy balances. The various control schemes proposed to fulfill these requirements have been reviewed by many authors (16,40,44, 84,89,115). From the many control schemes suggested, only two basic control philosophies emerge, dubbed direct and indirect material balance control. Figures 4.3 and 4.4 illustrate simple examples of indirect and direct material balance control schemes. In the indirect material balance control scheme, the energy balance is manipulated by the steam flow rate and the material balance is adjusted indirectly by the reflux flow rate through the reflux accumulator level controller. Both product flows are then adjusted under level control to maintain the column inventory. In the direct material balance control scheme illustrated, the energy balance is controlled directly by the steam flow rate, while the

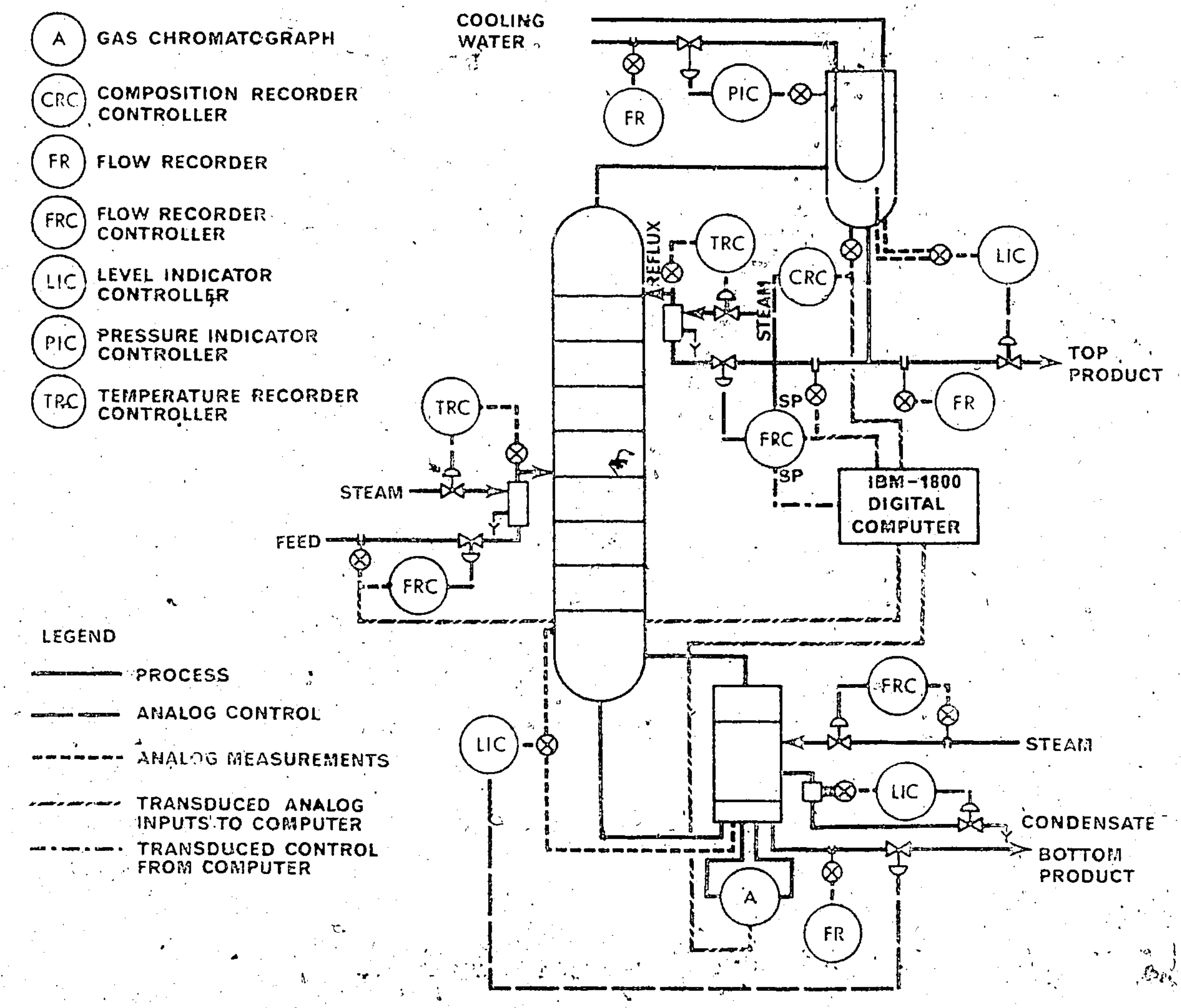

FIGURE 4.2 SCHEMATIC DIAGRAM OF CONTROL SCHEME

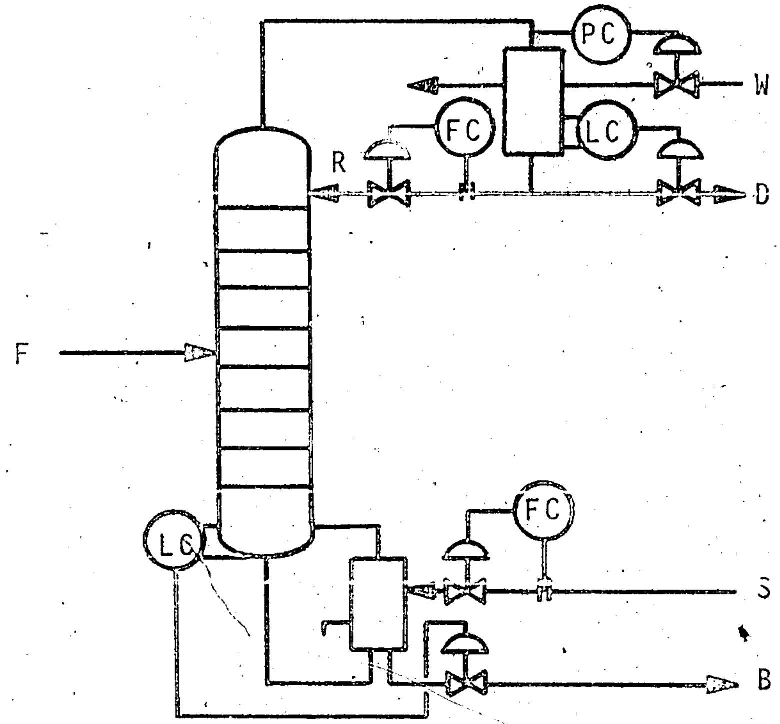

FIGURE 4.3 INDIRECT MATERIAL BALANCE CONTROL

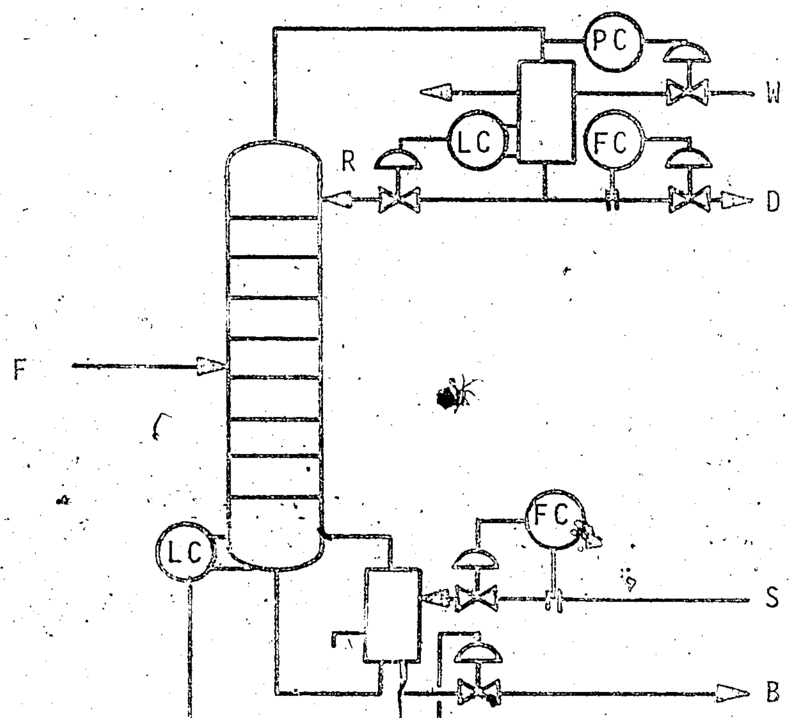

FIGURE 4.4 DIRECT MATERIAL BALANCE CONTROL

material balance is controlled directly by manipulation of the overhead product flow rate. The reflux and bottoms product flow rates are then adjusted under level control to maintain the column inventory. This control scheme could easily be reversed with the reflux flow rate defining the energy balance and the bottoms flow rate directly defining the material balance. The steam and overhead product flow rates would then be adjusted under level control to maintain the column inventory.

Examination of the many articles suggests a controversy exists in the use of the direct and indirect control schemes. The advantages claimed for the direct material balance system (84,89,115) include

- greater degree of self regulation
- greater degree of accuracy
- reduced interactions
- tighter control
- simplicity.

Buckley (16,17) has suggested that the main disadvantage of the direct method is the large reflux cycles which may be set up in the column due to uncontrolled reflux temperature fluctuations, resulting in an oscillatory product quality.

Despite the many advantages claimed for the direct material balance approach, the indirect method of material balance control has remained an extremely popular industrial control scheme. For this reason, the indirect material balance control scheme has been implemented during this project. The fact that such a controversy does exist suggests that an experimental comparison of both control schemes should be made.

In order to complete the definition of the column, the specification of the variables not fixed by the column design is accomplished as follows:

a) the feed and reflux temperatures are maintained a few degrees below their boiling points by a temperature controller which adjusts the steam flow to a small heat exchanger;

b) the feed flow rate, the reflux flow rate and the steam flow rate are maintained under flow control;

c) the feed composition is prepared in the feed tank;

d) the o rhead and bottoms product flow rates are controlled from the levels in the reflux accumulator and the reboiler respectively;

e) the column pressure is maintained by adjustment of the cooling water flow rate to the overhead condenser.

The overhead product flow rate, the bottoms product flow rate and the cooling water flow rate are adjusted by control valves, which receive their pneumatic control signal directly from their respective controller. This method of control is generally considered bad practice (115) due to the hysteresis exhibited by the valve. The installation of either a valve positioner or a flow controller whose set point is cascaded from the primary controller will eliminate any effects caused by these nonlinearities.

4.5.3 Composition Control

Once the operation of the column has been completely defined by the material and energy balances, a composition control scheme may be specified, which will achieve a particular objective. The control objective, chosen during this study, was to maintain the overhead product composition by manipulation of the reflux flow rate. This control strategy is currently implemented using either standard analog controllers or the control capability of the IBM-1800.

Various analog controllers have also been provided, including an intermediate tray temperature controller (23), a differential pressure controller and a ratio relay, although they have not been used during the present study.

A block diagram of the analog feedback control scheme implemented is shown in Figure 4.5. The feedback controller is a special capacitance measuring pneumatic Foxboro Dynalog controller. The output of the composition controller is cascaded to the pneumatic setpoint of the Foxboro Consotrol controller controlling the reflux flow rate. The same control scheme is also illustrated in Figure 4.6, except that the function of the analog composition controller has been replaced by the IBM-1800 digital computer. The only difference between these two control schemes exists in the control algorithm by which each controller calculates its output. The digital computer uses a digital approximation to the ideal feedback controller expression (29), while the analog controller approximates this expression with a series of mechanical linkages and pneumatic resistances and capacitances. An example of a complex expression which has been developed to describe the dynamics of the pneumatic Dynalog controller is available from Svrcek (128). Consequently, a correspondence between the control parameters of the two feedback control schemes would not be expected to exist.

In both cases, the feedback control action is

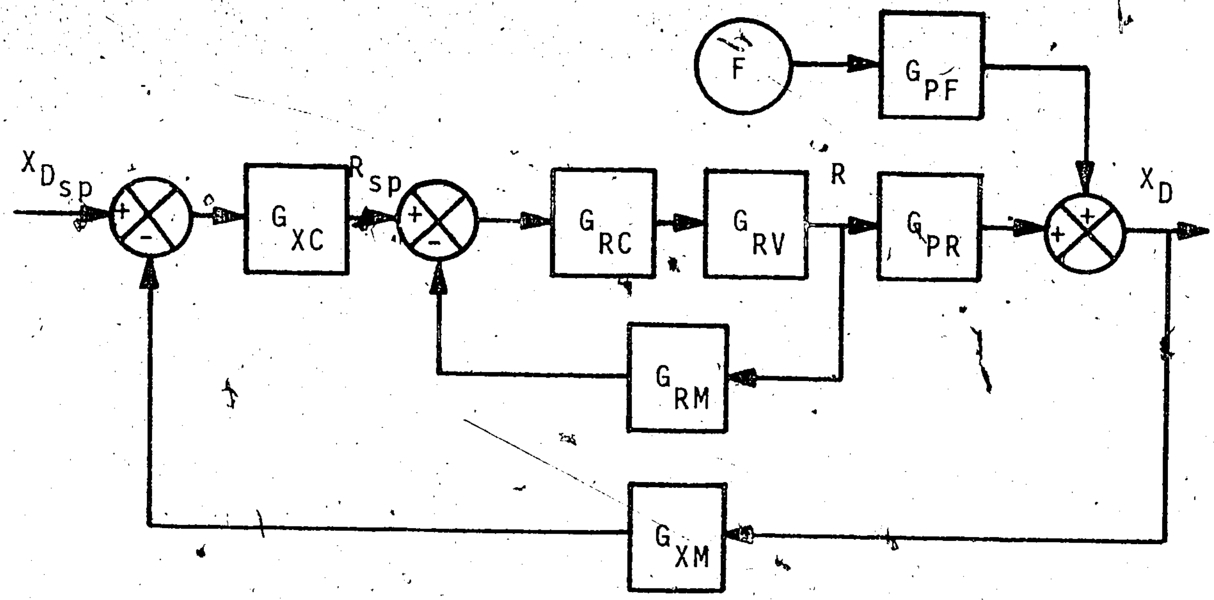

FIGURE 4.5 BLOCK DIAGRAM FOR ANALOG FEEDBACK CONTROL

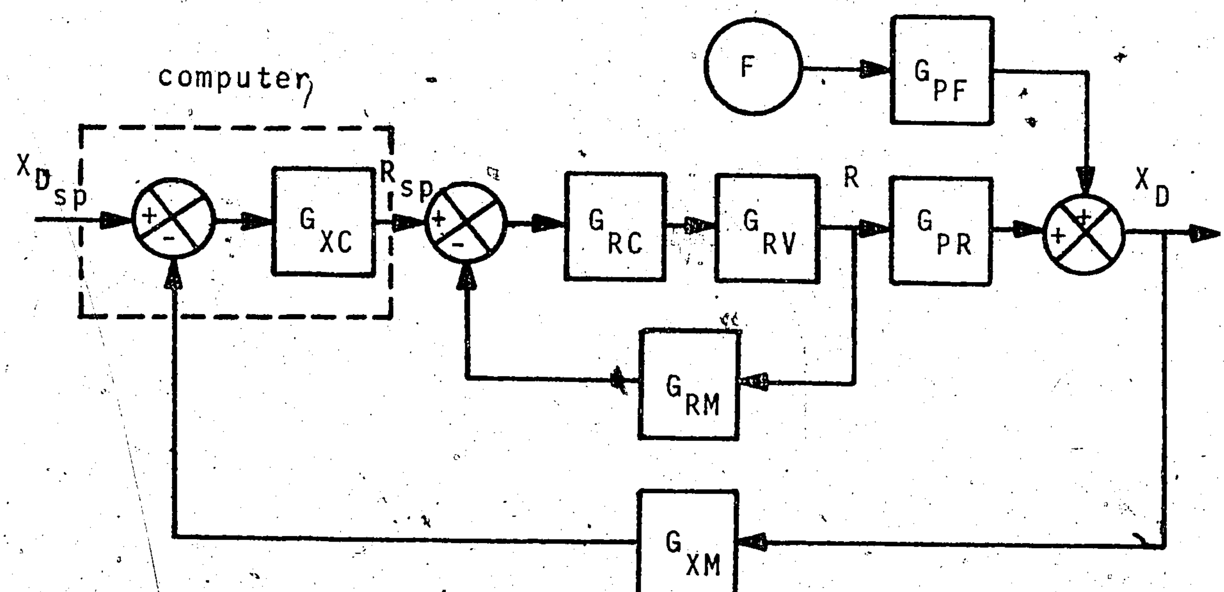

FIGURE 4.6 BLOCK DIAGRAM FOR DIGITAL FEEDBACK CONTROL

activated by disturbances measured in the overhead product quality. The control actions, calculated based on these excursions from the set point, are cascaded to the pneumatic set point of the reflux flow rate controller.

The feedforward control is implemented using the digital computer as shown in Figure 4.7. The feedforward control action is activated by the measured disturbances in the feed flow rate as they enter the column. The control actions, calculated based on these excursions from their set point, are also cascaded to the set point of the reflux flow controller.

Examination of the block diagrams for these two control schemes suggests the next logical step; the combination of the two control schemes, which is illustrated in Figure 4.8. This control scheme has also been implemented using the computer.

The above figures are examples of what is called analog supervisory control, in contrast to direct digital control (33,53,88). Analog supervisory control allows the local analog controllers to control each individual loop, while the computer only supplies the set points for these local controllers. Thus, if the computer fails, the plant can continue to operate by supplying a manual

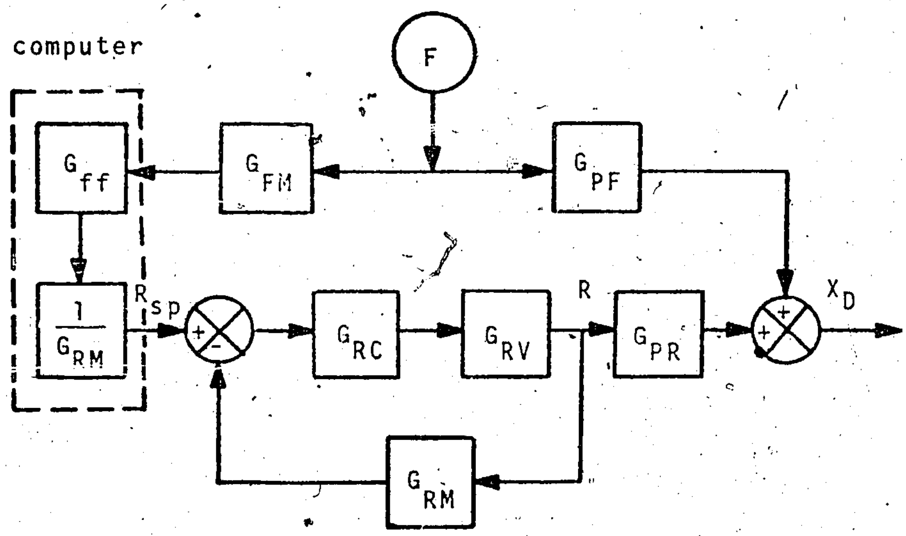

FIGURE 4.7 BLOCK DIAGRAM OF DIGITAL FEEDFORWARD CONTROL.

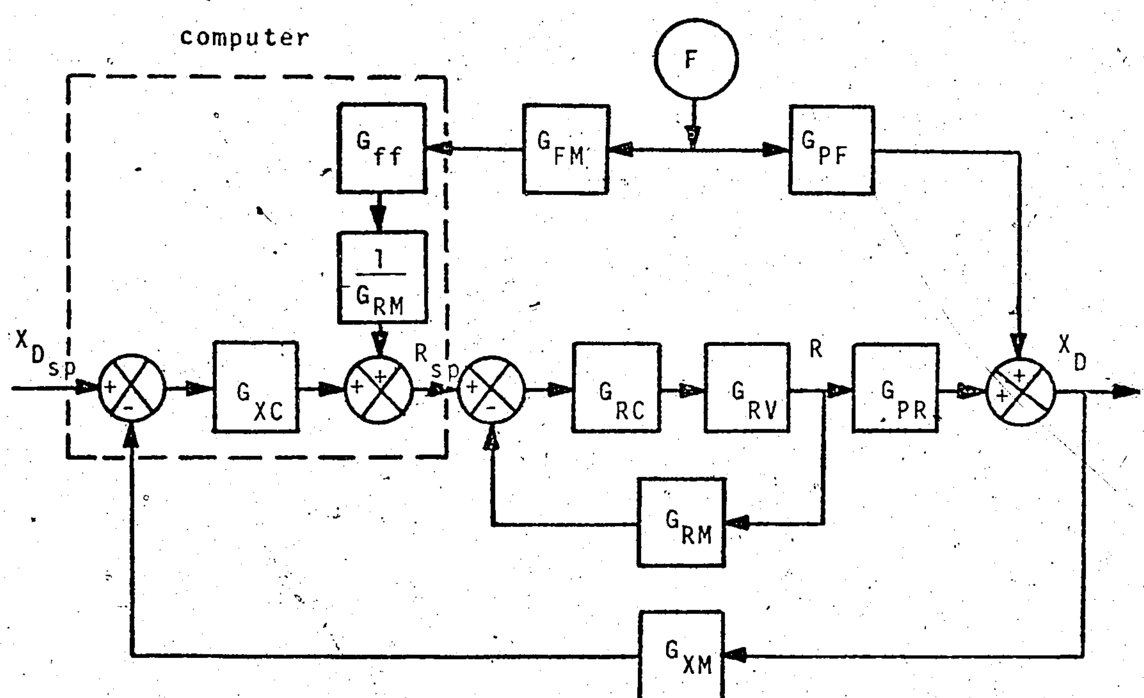

FIGURE 4.8 BLOCK DIAGRAM FOR COMBINED FEEDFORWARD-FEEDBACK CONTROL

set point to each controller. The computer can now spend most of its time optimizing the process operation, rather than controlling the many individual control loops.

4.6 Operating Procedures and Calibration

Since no formal operating procedures existed for the column, the start-up and shutdown procedures (93) were evolved. It will be noted that the column was not started up on total reflux, but rather as soon as the reflux accumulator contained a sufficient level, the value of the desired reflux flow was set on the reflux controller and the column was allowed to come to steady state. Operating in this manner insured that the column could be tested within two to three hours after initial start-up.

All the column measuring instruments were calibrated using the methods outlined in Appendix A.2. The calibration results are given in Figures A.1 to A.7 and Table A.1 of Appendix A.

4.7 Column Operation

The operation of the column was first checked by attempting to reproduce some of the steady state data previously reported by Svrcek (128). Typical results of these steady state comparisons, which involved taking simultaneous samples from all plates, are presented in Table A.3 of Appendix A.

4.8 Selection of the Reference Steady State Conditions

Since the behavior of the column was to be studied for relatively large disturbances (of the order of ± 20%) in the input variables, feed, reflux and steam flow rates, the selection of a suitable initial steady state to allow for normal operation after all such disturbances was necessary. The initial requirements sought included:

a) the disturbance should not cause the column to flood, weep or exceed the condenser or reboiler capacity;

b) the disturbance should not force the column operation through a pinch condition, causing extremely nonlinear responses;

c) the product flow rates should be approximately equal since the feed will nominally be made up of equal quantities by weight of methanol and water.

The typical steady state conditions used by Svrcek (128) were initially employed, but the large increases in the feed, reflux or steam flows caused the column to flood. These steady state conditions were adjusted until the disturbances did not cause flooding. The reference steady state conditions established and the reproducability as given by the standard deviation obtained throughout the course of this study are presented in Table 4.2.

The choice of the steady state operating conditions represented a compromise based on the requirements previously outlined. The responses to the various disturbances exhibit very nonlinear behavior, in that the overhead product composition is driven into or out of the pinched conditions at the top of the column, while the bottoms product composition response is discontinuous at zero percent by weight methanol in addition to similar pinch producing and pinch relieving effects.

The choice of a steady state, exhibiting nonlinear dynamics would, however, be expected to give the proposed feedforward control schemes a much more severe test than would choosing a steady state which exhibits a more linear dynamic response.

4.9 Material Balance Errors of Closure

Material balance calculations were performed using

TABLE 4.2 REFERENCE STEADY STATE

Process Variable	Pulse Test	Transient Response Test	Feedback Control	Feedforward Control	Feedforward Feedback Control	Series	Reference Steady State Overall Project
F	2.46±0.01	2.46±0.01	2.48±0.01	2.48±0.01	2.48±0.01	2.49	2.47±0.01
R	1.94±0.01	1.95±0.02	1.95±0.07	1.96±0.05	1.98±0.05	2.01	1.95±0.04
S	2.01±0.01	1.71±0.01	2.04±0.02	2.04±0.02	2.05±0.02	2.06	1.89±0.20
X_F	46.7 ±0.6	47.0 ±0.9	47.3 ±0.5	47.4 ±0.9	47.1 ±1.0	47.0	47.1 ±0.8
X_D	95.83±0.24	96.13±0.27	96.06±0.04	96.09±0.06	96.07±0.02	96.07	96.00±0.22
X_B	0.53±0.13	0.55±0.17	0.44±0.07	0.44±0.05	0.45±0.05	0.40	0.49±0.12
D	1.17±0.02	1.18±0.05	1.21±0.02	1.22±0.03	1.22±0.04	1.19	1.20±0.04
B	1.27±0.03	1.28±0.04	1.31±0.03	1.31±0.03	1.32±0.04	1.34	1.29±0.06
T_F	167.4 ±1.5	166.5 ±1.0	162.9 ±1.4	162.9 ±0.8	163.0 ±0.8	163.2	164.9 ±2.8
T_R	151.6 ±0.7	152.3 ±0.7	150.5 ±2.5	151.5 ±0.4	151.3 ±0.4	148.7	151.6 ±1.9
N_F	4	4	4	4	4	4	4

the data acquisition capabilities of the IBM-1800, in conjunction with the user written programs DASS, DATAC and BALNC. Table 4.3 illustrates the average material balance error with its standard deviation obtained throughout the course of this study, while Table A.4 of Appendix A gives a specific example of the report produced by the material balance programs.

TABLE 4.3 AVERAGE MATERIAL BALANCE ERRORS OF CLOSURE

Test	T[illegible] Tests	Total Steady States	Errors of Closure - % Overall Balance	Methanol Balance	Water Balance
pulse	18	36	-0.6 ± 1.7	-1.6 ± 2.1	0.3 ± 2.0
transient response	84	84	-0.1 ± 1.5	-1.9 ± 3.6	1.3 ± 2.0
feedback control	18	18	1.6 ± 0.4	0.0 ± 1.0	2.9 ± 1.3
feedforward control	26	26	1.9 ± 0.5	0.3 ± 1.3	3.4 ± 1.1
feedforward-feedback control	11	11	2.1 ± 0.6	0.9 ± 1.5	3.4 ± 0.9
series	2	2	1.6	1.7	4.5
complete project	159	177	0.4 ± 1.7	-1.2 ± 2.9	1.8 ± 2.1

CHAPTER 5

DEVELOPMENT OF AN OPEN LOOP DYNAMIC MODEL OF THE COLUMN

5.1 Introduction

The various methods used in previous studies to describe and evaluate the dynamic behavior of the distillation system have been outlined in Section 2.2. These methods include:

i) deriving the complete set of fundamental equations describing the related phenomena, making as many simplifying assumptions as necessary in order to obtain a solution and applying either analytic, analog or digital computer techniques to obtain a solution

ii) using an approximate correlation method to obtain the parameters of a transfer function representation as a function of the initial steady state operating conditions

iii) determining the parameters of an approximate transfer function representation using standard dynamic testing techniques.

Since obtaining a solution to the general equations characterizing the system is not difficult with the large digital computers currently available, no longer is it necessary to linearize these equations; so consequently the nonlinear behavior of the process may be analyzed. The lack of sufficient accurate information describing such phenomena as plate mixing, plate hydraulics and dynamic plate efficiencies, among others, are the major limitations to this technique. The ease with which the solution can be obtained depends on the number of assumptions made; ie the more assumptions made, the easier the solution. The large digital computer required to obtain the solution is generally unsuitable for controlling the process. Thus, for control purposes, a simplified model would still have to be developed, whose parameters could be updated periodically based on the off-line solution of the large model.

The advantages of obtaining a reasonably good approximate model based on the initial steady state operating conditions are obvious. However, the relationships between the column inputs and outputs, as described previously, are very complex, suggesting that even if a correlation can be found, it too may be equally as complex. Although a recent study by Wahl and Harriott (134) has introduced a correlation which may have merit, in general, the correlation methods developed up to now have not experienced wide spread acceptance.

Approximate models of a number of columns (50,56,85, 98) have been successfully determined using various dynamic testing techniques. The range of the models developed experimentally is, by necessity, limited by the restricted operating range of the column inputs so as not to cause unduly large transients in the column output variables. When the initial steady state conditions of the column change appreciably, a new determination of the approximate model parameters would, ideally, be required. This method could, of course, be used with the large fundamental model to calculate the parameters of a simple model off-line, which could then be used to update the parameters of the on-line model for the changing conditions.

A combination of transient response and pulse testing analyses were employed in this study to obtain the steady state and dynamic relationships describing the column responses. The pulse testing procedure was chosen based on the results presented by Lees (62) and Wildman (135). A computer program, PTAP (pulse testing analysis program) was also available on the IBM-1800, having been developed by Wildman (135), which calculated the Fourier transform of the input and output variable time data, yielding the amplitude ratio and phase angle versus frequency diagram.

Due to the nonlinearity of the system, the steady state process gains could not be accurately determined from the pulse testing data; consequently, a transient response analysis was used to determine them.

In the following sections the data used in the determination of the process gains will be presented first, followed by the data used in determining the dynamic parameters for a simple first order lag plus time delay approximate model. The results of fitting the measured frequency response diagrams to more complex polynomial transfer functions will also be presented.

5.2 Determination of the Process Gain

5.2.1 Procedure

The feed flow rate, reflux flow rate and steam flow rate were the input variables; the overhead product composition and flow rate, and the bottoms product composition and flow rate were the output variables considered during this portion of the study. The column was initially allowed to come to steady state with all the input variables set to their reference steady state values, as listed in Table 4.2. After the column had reached this steady state, the values of all the column variables were documented in a material balance report, similar to that presented in Figure A.4, prepared by programs DASS/DATAC/BALNC. Details of these programs are given in Appendix G. The input variables were then varied, one at a time, over the ranges:

feed flow - 2.47 ± 0.50

reflux flow - 1.95 ± 0.25

steam flow - 2.05 ± 0.[illegible]

When the column has reached the new steady state, a second material balance report is generated, recording all the column measurements. The results obtained using the feed composition as an input variable were also measured, but were reported elsewhere (94).

5.2.2 Results

The experimental operating conditions from which the process gains were calculated are given in Tables C.1, C.2 and C.3 of Appendix C. The gains, initially calculated from the unsmoothed operating data, are presented in Tables C.4, C.5 and C.6 of Appendix C. The dependence of these open loop gains on the magnitude of the disturbance in the various input variables studied is illustrated in Figures 5.1, 5.2 and 5.3. A least squares fit of the process gains, as given by the expressions in Table 5.1, are shown in these figures by the dashed line. Examination

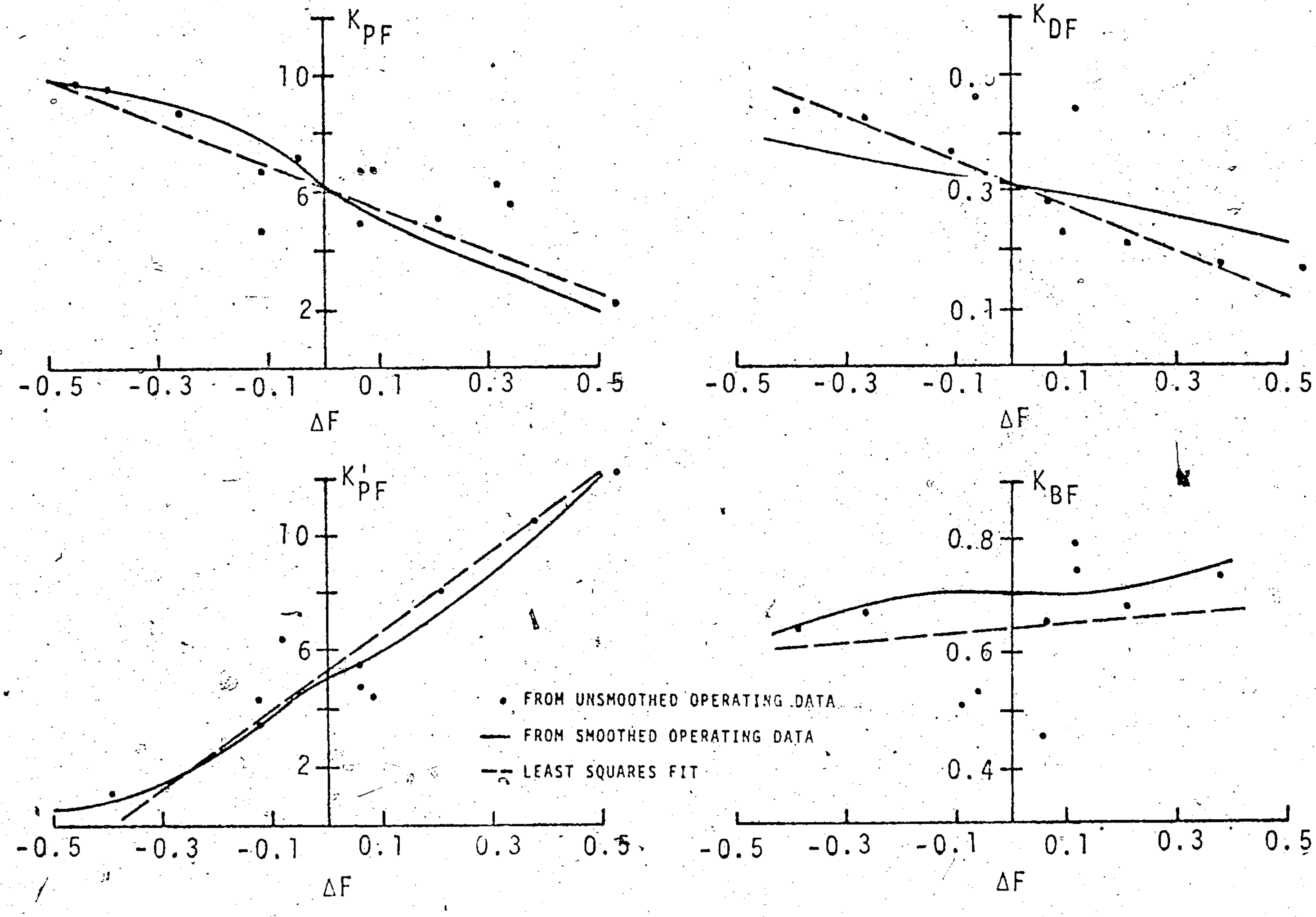

FIGURE 5.1 OPEN LOOP PROCESS GAINS: VARIATION WITH FEED FLOW RATE

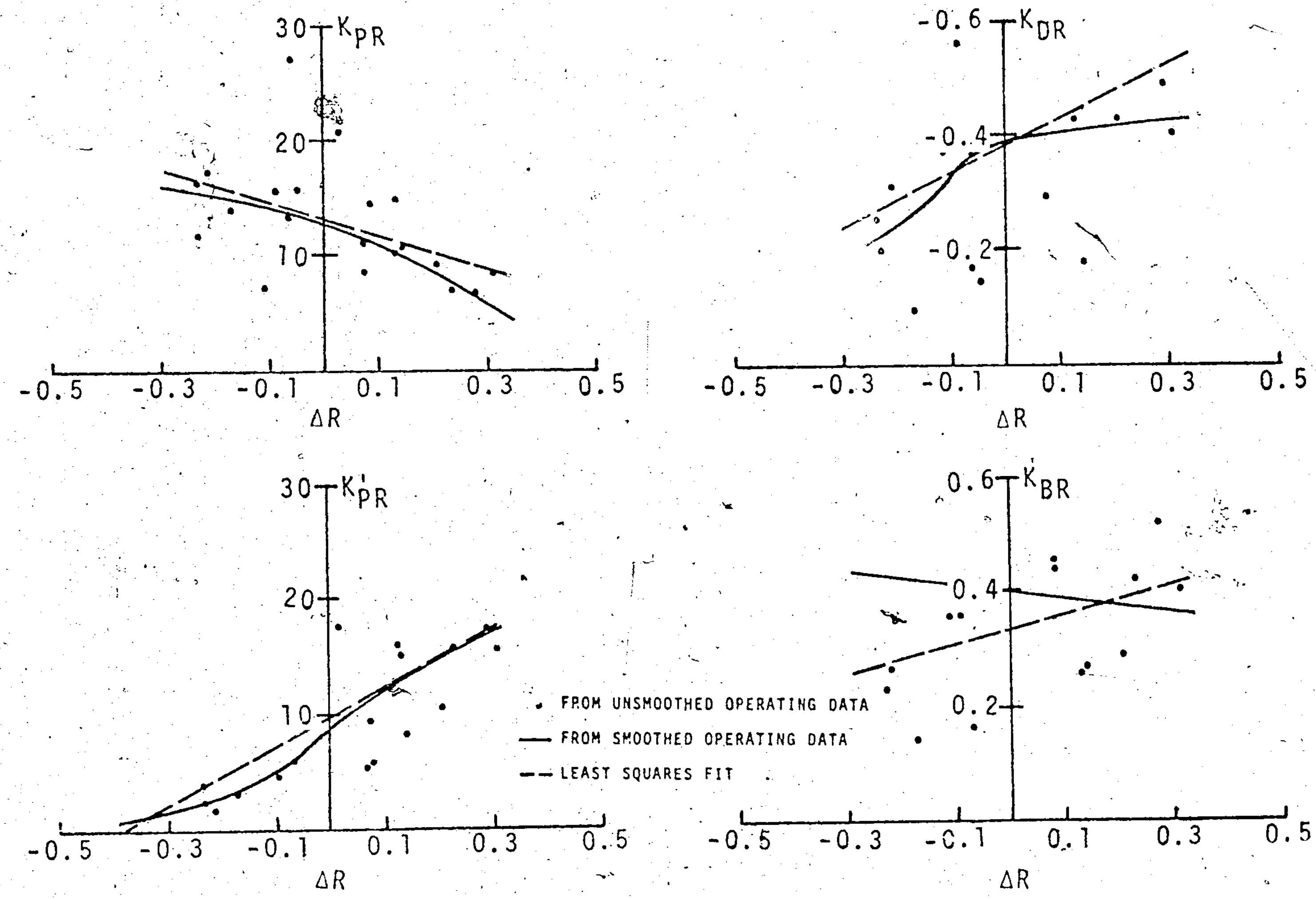

FIGURE 5.2 OPEN LOOP PROCESS GAINS: VARIATION WITH REFLUX FLOW RATE

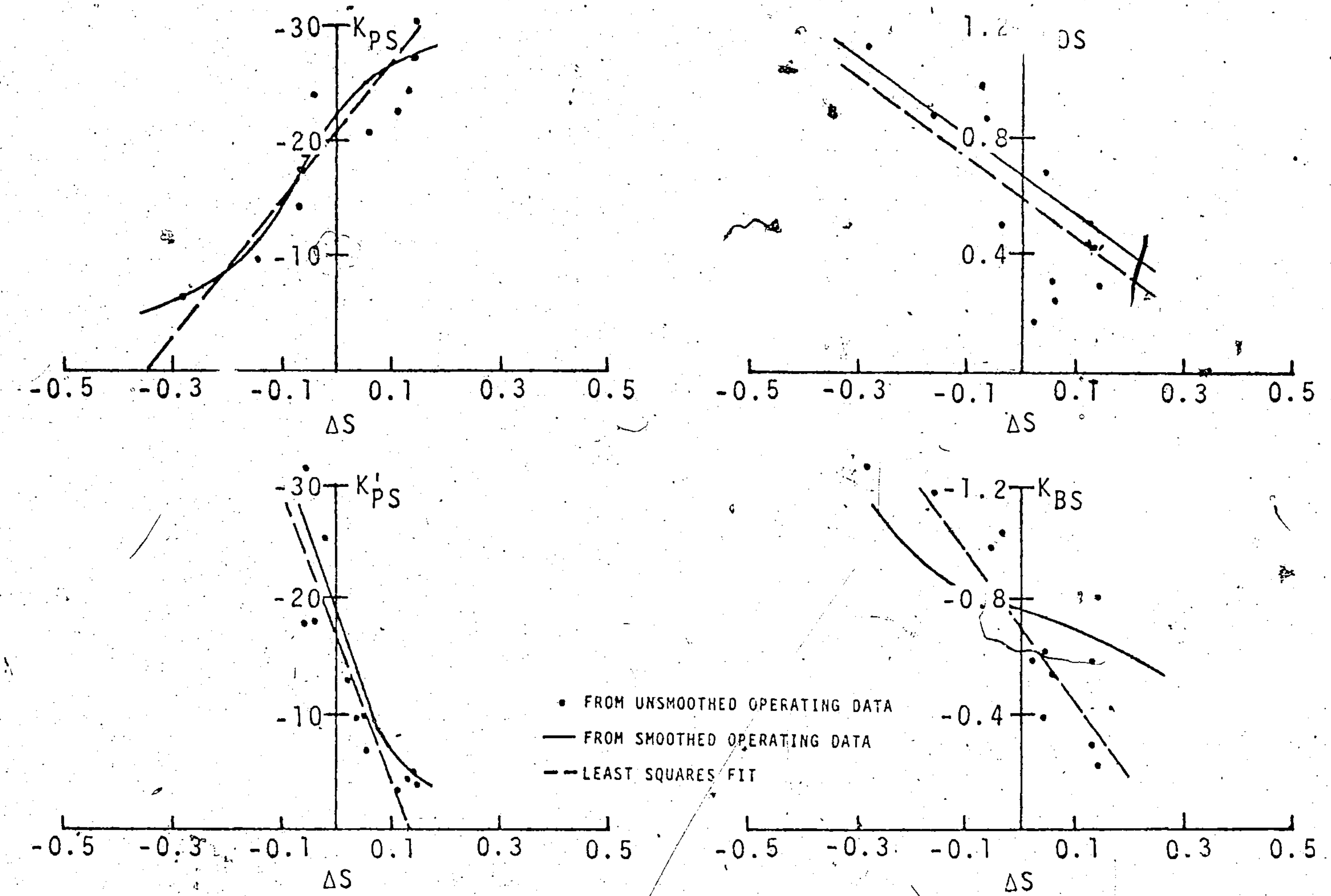

FIGURE 5.3 OPEN LOOP PROCESS GAINS: VARIATION WITH STEAM FLOW RATE

TABLE 5.1 LEAST SQUARES EQUATIONS FIT TO THE EXPERIMENTAL OPEN LOOP PROCESS GAIN DATA

Reflux Flow

range $-0.25 \leq \Delta R \leq 0.40$

$$\frac{X_D}{R} = 13.33 - 14.84\ \Delta R \qquad \frac{X_B}{R} = 9.69 + 25.52\ \Delta R$$

$$\frac{D}{R} = -0.38 - 0.48\ \Delta R \qquad \frac{B}{R} = 0.33 + 0.35\ \Delta R$$

Steam Flow

range $-0.30 \leq \Delta S \leq 0.20$

$$\frac{X_D}{S} = -21.72 - 52.86\ \Delta S \qquad \frac{X_B}{S} = -17.77 + 124.80\ \Delta S$$

$$\frac{D}{S} = 0.62 - 1.41\ \Delta S \qquad \frac{B}{S} = -0.72 + 2.63\ \Delta S$$

Feed Flow

range $-0.40 \leq \Delta F \leq 0.50$

$$\frac{X_D}{F} = 6.21 - 7.40\ \Delta F \qquad \frac{X_B}{F} = 5.28 + 13.38\ \Delta F$$

$$\frac{D}{F} = -0.38 + 0.32\ \Delta F \qquad \frac{B}{F} = 0.64 + 0.07\ \Delta F$$

of these figures indicates that the scatter about these straight lines is appreciable, suggesting that it would be more appropriate to smooth the operating data before calculating the gains. The operating conditions of the column, smoothed by visual examination, are displayed in Figures C.1, C.2 and C.3 of Appendix C. The dependence of the process gains calculated from the smoothed operating data on the magnitude of the input variables is compared to the dependency calculated previously in Figures 5.1, 5.2 and 5.3. The process gains, when calculated from the smoothed operating data, exhibit the 'S'-shaped curve characteristic of nonlinear systems when plotted against the input disturbance. It is interesting to note that, in almost each case, the least squares straight line is a reasonable approximation to the calculated nonlinear relation. This is particularly true in the case of the process gains for the composition variables.

Table 5.2 illustrates the reproducability of the reference steady state as indicated by the standard deviation obtained throughout the transient tests. These errors, although small when compared to the absolute value of the measurement, are important when a calculation requires the use of the difference between two values of the same order of magnitude. Since the process gains are

TABLE 5.2 REPRODUCABILITY OF THE REFERENCE STEADY STATE CONDITIONS OBTAINED DURING TRANSIENT RESPONSE TESTS

Process Variables	Transient Response Test Disturbances			
	Open Loop			Closed Loop
	Feed Flow	Reflux Flow	Steam Flow	Feed Flow
input				
F	2.46±0.01	2.46±0.02	2.46±0.01	2.45±0.01
R	1.94±0.01	1.95±0.01	1.95±0.01	1.92±0.03
S	1.72±0.01	1.71±0.01	1.71±0.01	1.70±0.01
X_F	46.6 ±1.0	46.8 ±0.8	46.7 ±1.1	47.8 ±0.30
output				
X_D	96.15±0.06	96.04±0.34	96.36±0.39	96.0 ±0.25
X_B	0.45±0.07	0.58±0.22	0.54±0.24	0.56±0.13
D	1.16±0.02	1.15±0.06	1.15±0.03	1.23±0.02
B	1.29±0.01	1.29±0.05	1.29±0.04	1.26±0.03

calculated from the ratio of two such differences, small errors can cause quite large errors in the calculated gains.

Certainly some of the scatter, particularly in the case of the overhead and bottoms product flows, was also due to the failure to maintain a constant feed composition. Examination of these results, given in Tables C.1, C.2 and C.3 of Appendix C, indicated that some of the scatter in these process gains appears to correlate with the variation in the feed composition used. These gains appear to be more sensitive to feed composition changes than do the overhead and bottoms product composition gains.

The overhead and bottom product flow rate process gains can be related by the following expressions:

(feed disturbances) $$K_{DF} + K_{BF} = 1 \qquad (5.1a)$$

(reflux disturbances) $$K_{DR} + K_{BR} = 0 \qquad (5.1b)$$

(steam disturbances) $$K_{DS} + K_{BS} = 0 \qquad (5.1c)$$

The consistency of the overhead and bottom product flow rate gains, as calculated from both the original data and a smoothed representation of the same data, have been compared using Equations (5.1) in Tables C.1 - C.6 of Appendix C. Examination of the residuals presented in these tables indicates that despite the good closure on the material balances obtained, generally less than a 3% discrepancy, large deviations occur when the gains calculated from the original operating data are substituted into Equation (5.1). Although the variation in the values calculated from Equation (5.1) is smaller using the gains calculated from the smoothed representation of the operating data, the variations are still larger than would be anticipated for such close material balance closures. These discrepancies reflect some of the difficulties encountered in using the indirect material balance control schemes as discussed by Nisenfeld (89) and Shinskey (115).

5.3 Characterization of the Column Dynamic Behavior

5.3.1 Procedure

The pulse testing procedure, designed using the guidelines presented by Wildman (135), was used to determine the dynamic relationships of the column. The column was allowed to come to steady state with the input variables set to the values listed in Table 4.2. When the steady state conditions were achieved, they were recorded using the on-line material and energy balance programs. In addition, the system data acquisition and buffering to disk programs were initiated to save the data. Collection of the initial steady state data continued for a period of about 10-15 minutes in order to define the steady state conditions for the program PTAP. A rectangular pulse of about 10-15 minutes duration and of sufficient magnitude to cause a measurable response was implemented by adjustment of the set point of the input variable flow controller. The pulse disturbances were applied to each variable in turn, the feed, reflux and steam flow rates. The output variables of interest in this portion of the study included only the overhead and bottoms product compositions. The responses of the plate temperatures were also recorded (94) for use during a subsequent project (23). Each pulse test was repeated a number of times, in both the positive and negative directions, in order to qualitatively compare the reproducability. The magnitude and duration of the pulse were varied slightly during a few tests, but no attempt was made to quantitatively analyze the effects, if any. At the completion of the pulse test, when the response returns to the initial steady state, the steady state conditions were again recorded by the on-line material and energy balance programs.

The complete set of transient response data was then recovered from the system disk files using the program GTDAT. A hard copy of the data was obtained on the system printer in addition to a set of punch cards. The cards, besides acting as a historical record, provided input to the pulse analysis program.

The program PTAP was modified slightly from that described by Wildman (135) in that the program contained an option to punch the calculated frequency response on cards, and the frequency response plotting subroutine employed by Farwell (32) was substituted for the original plotting subroutine.

Based on the guidelines of Wildman (135), the pulse testing procedure was governed by the following characteristics:

a) the approximate time constant determined from the data of Svrcek (128) and by preliminary transient response tests was found to be between 10 and 15 minutes

b) the corner frequency is between 0.067 and 0.1 radians/minute

c) the frequency range of interest is 0.001 to 1.0 radians/minute.

On the basis of these conditions, it followed that:

d) the sample interval should be less than 19 seconds, based on the maximum frequency of interest and the Lees-Dougherty criteria (135) for minimizing the numerical approximation error; consequently the closest sample interval available in the DDC program of 16 seconds was employed.

e) the total time record to be collected should require about 10-15 minutes of steady state data and between 50-70 minutes of transient data

f) the frequency response values calculated below a normalized frequency content value of 0.3 are not considered reliable.

5.3.2 Results

The reference steady state, about which the pulse tests were conducted, along with the pulse characteristics are contained in Table D.1 of Appendix D. The complete transient response data for all the pulse tests are available elsewhere (94). Examples of typical transient responses of the overhead and bottom product compositions to both positive and negative pulses in the input variables are shown in Figures D.1, D.4, D.7 and D.10. Particular note is made of Figure D.7, which illustrates an inverse transient response in the overhead product composition for decreasing disturbances in the steam flow rate. Figures D.2, D.5, D.8 and Figures D.3, D.6, D.9 illustrate the frequency response representation of increasing overhead and bottom product compositions responses respectively, while Figures D.11, D.12, D.13 show the frequency response representation for decreasing overhead product compositions. The frequency response for a decrease in bottoms product composition was not evaluated since the disturbances caused the composition to pass through zero percent by weight methanol, causing a discontinuity in the response.

Comparison of the frequency response results calculated from the repeated pulse test indicates that, by visual examination at least, the agreement obtained is quite good at the lower frequencies, while major discrepancies appear to occur in the slope of the resulting high frequency asymtotes. These discrepancies are undoubtedly a result of the nonlinear behavior of the column response. No special effort had been made to insure that an exact replicate pulse would be formed during the repeated pulse tests; in fact, for a number of tests the pulse duration was deliberately varied.

The frequency response diagrams, for frequencies greater than 0.3 to 0.4 radians/minute, exhibit a great deal of scatter characteristic of the pulse testing procedure. The reasons for this phenomena have been discussed in detail by Wildman (135). These unreliable frequency response values generally occurred for normalized frequency content values in the range 0.3 to 0.5, but in some cases are as high as 0.7 to 0.9 (94), compared to the value of 0.3 suggested by Wildman (135). The termination of the data collection before the process had returned to its steady state and the rounding of the corners of the rectangular pulse , imparted by the response of the flow controller to its set point change, probably contributed to the excessively large values of the normalized frequency content cut off point obtained.

5.4 Approximate Column Model

5.4.1 Procedure

The parameters of a first order lag plus time delay approximate model were obtained using the following methods:

a) the transient response was fit using the Rosenbrock (110) gradient search technique as outlined by Berry (9)

b) the frequency response was fit using the complex curve fitting technique of Levy (63) modified such that a best fit time delay could be included

c) the parameters were determined based on the frequency response characteristics exhibited by a first order lag and time delay functions:

- the time delay makes no contribution to the amplitude ratio
- the first order time constant is defined as the reciprocal of the frequency at which the value of the amplitude ratio is ±0.707 (±3 decibels)
- the time delay is calculated from the difference between the measured phase angle (ϕ) and 45 degrees as

$$\tau_D = (\phi - 45) \frac{\tau}{57.3} \qquad (5.2)$$

d) the process gains, based on the magnitude of the pulse test, were determined from the gain expressions established by transient response testing.

Several of the open loop frequency responses were also fit by a series of more complex polynomial transfer functions. No attempt was made, however, to determine which of these more complex models gave a better approximation to the transient response than does the simple first order lag plus time delay model.

5.4.2 Results

A first order lag plus time delay transfer function of the form

$$G = K \frac{e^{-\tau_D s}}{\tau s+1} \tag{5.3}$$

has been assumed to describe the dynamic response of the distillation column. This form of approximate model has been used successfully in a number of previous control studies (9,14,112). The parameters of this model, determined for the various pulse tests performed, are given in Tables 5.3, 5.4 and 5.5. Average transfer functions are presented in Table 5.6, describing the response of the composition output variables to each of the flow input variables studied based on the values determined from the time domain curve fit. The transient response calculated from these simple models are compared with some actual transient responses measured in Figures D.1, D.4, D.7, D.10 of Appendix D. In addition, the frequency response of these approximate models is also compared with the frequency response calculated from the appropriate transfer function model presented in Table 5.6. This comparison is shown in Figures D.2, D.5, D.8, D.3, D.6, D.9, D.11, D.12 and D.13 of Appendix D. The values of the parameters determined for the more complex polynomial

TABLE 5.3 COMPARISON OF PARAMETERS OF APPROXIMATE MODELS DETERMINED FOR POSITIVE OVERHEAD PRODUCT COMPOSITION RESPONSES

Test	Pulse Magnitude (lb/min)	Pulse Duration (min)	Method 1 K	Method 1 τ_D	Method 1 τ	Method 2 K	Method 2 τ_D	Method 2 τ	Method 3 K	Method 3 τ_D	Method 3 τ	Method 4 K
Feed20	0.54	15.0	2.7	6.1	13.9	2.9	5.3	15.0	2.9	5.1	15.7	1.7
Feed23	0.52	14.8	3.4	8.3	28.2	2.4	2.0	17.5	2.4	10.8	17.4	1.9
Feed24	0.47	9.9	4.9	4.3	27.1	5.1	4.3	11.2	5.8	9.3	16.5	2.1
Reflux20	0.29	14.7	10.1	0.5	16.3	14.8	5.3	23.5	14.8	3.0	20.4	6.0
Reflux22	0.26	14.4	11.1	1.1	17.6	13.3	3.5	18.2	13.3	2.9	18.6	6.8
Reflux23	0.28	10.8	14.5	1.6	16.8	10.6	3.6	11.4	10.6			6.3
Steam20	-0.27	11.0	-17.6	2.4	23.6	-19.7	6.3	19.1	-19.7	5.5	19.8	-6.5
Steam22	-0.27	11.2	-13.4	3.4	20.0	-13.4	7.0	16.9	-13.4	6.0	17.8	-6.5
Steam24	-0.22	9.5	-15.7	2.7	22.0	-17.5	7.1	23.5	-19.4	4.3	12.7	-7.8
Steam27	-0.30	5.8	-19.4	2.4	21.3	-17.2	6.8	12.8	-17.2	6.3	12.0	-6.0

Method 1 time domain curve fit using method of Rosenbrock (110)
2 inspection of frequency response
3 frequency response curve fit using method of Levy (63)
4 gains from expressions determined from transient response testing

TABLE 5.4 COMPARISON OF PARAMETERS OF APPROXIMATE MODEL DETERMINED FOR POSITIVE BOTTOMS PRODUCT COMPOSITION RESPONSES

Test	Pulse Magnitude (lb/min)	Pulse Duration (min)	Method 1 K	Method 1 τ_D	Method 1 τ	Method 2 K	Method 2 τ_D	Method 2 τ	Method 4 K
Feed21	0.54	15.0	4.0	3.8	13.2	4.6	1.8	18.3	12.6
Feed23	0.52	14.8	2.8	4.8	11.6	2.0	0.7	20.6	12.0
Feed24	0.47	9.9	3.0	3.0	13.2	5.8	1.2	16.2	11.3
Reflux20	0.29	14.7	6.7	9.9	12.0	5.6	6.9	15.2	16.5
Reflux22	0.26	14.4	6.4	7.4	11.2	7.1	7.0	14.3	15.8
Reflux23	0.28	10.8	6.8	7.3	9.3	6.3	6.5	8.3	16.3
Steam20	-0.27	11.0	-20.8	3.8	15.5	-19.1	3.5	12.7	-62.0
Steam22	-0.27	11.2	-27.1	5.1	11.2	-28.7	4.8	12.5	-62.0
Steam24	-0.22	9.5	-21.1	4.7	13.6	-21.1	4.7	13.4	-57.0
Steam27	-0.30	5.8	-16.2	2.8	14.1	-14.2	4.8	11.0	-67.0

Method 1 time domain curve fit using method of Rosenbrock (110)
2 inspection of frequency response data
4 gains from expressions determined from transient response testing

TABLE 5.5 COMPARISON OF PARAMETERS OF APPROXIMATE MODELS DETERMINED FOR NEGATIVE OVERHEAD PRODUCT COMPOSITION RESPONSES

Test	Pulse Magnitude (lb/min)	Pulse Duration (min)	Method 1			Method 2			Method 3			Method 4
			K	τ_D	τ	K	τ_D	τ	K	τ_D	τ	K
Feed21	-0.23	16.2	6.0	10.1	10.7	6.0	8.9	13.7	6.0	8.4	14.1	8.2
Feed22	-0.49	18.0	6.8	7.7	9.6	7.9	6.2	13.2	7.9	6.5	12.5	9.8
Reflux21	-0.24	16.2	15.4	1.1	13.5	15.4	3.2	13.0	15.4	3.2	13.0	15.8
Reflux24	-0.25	8.1	14.5	1.6	16.8	18.9	3.7	19.2	18.9	2.8	19.9	15.8
Steam21	0.20	14.4	-27.4	4.5	11.7	-30.3	4.3	13.2	-30.2	4.3	12.8	-28.5
Steam23	0.19	9.0	-33.0	3.2	17.0	-25.2	3.2	12.8	-25.2	3.4	12.9	-28.0
Steam25	0.21	10.0	-22.9	3.2	17.1	-20.0	4.4	13.0	-17.5	5.0	19.6	-28.7
Steam26	0.17	4.8	-28.3	1.6	22.6	-23.1	5.8	13.2	-23.1	5.5	11.9	-27.0

Method 1 time domain curve fit using method of Rosenbrock (110)
2 inspection of frequency response
3 frequency response curve fit using method of Levy (63)
4 gains from expressions determined from transient response testing

TABLE 5.6 NORMALIZED TRANSFER FUNCTION MODELS

Test Designation	Model
Positive Composition Responses	
Feed +	$\frac{X_D}{F} = \frac{e^{-6.0s}}{27.0s + 1}$
Feed B	$\frac{X_B}{F} = \frac{e^{-4.0s}}{12.5s + 1}$
Reflux +	$\frac{X_D}{R} = \frac{e^{-0.8s}}{17.5s + 1}$
Reflux B	$\frac{X_B}{R} = \frac{e^{-8.0s}}{11.0s + 1}$
Steam +	$\frac{X_D}{S} = \frac{e^{-2.5s}}{21.0s + 1}$
Steam B	$\frac{X_B}{S} = \frac{e^{-4.0s}}{12.5s + 1}$
Negative Composition Response	
Feed -	$\frac{X_D}{F} = \frac{e^{-8.0s}}{10.0s + 1}$
Reflux -	$\frac{X_D}{R} = \frac{e^{-1.3s}}{15.0s + 1}$
Steam -	$\frac{X_D}{S} = \frac{e^{-3.3s}}{17.0s + 1}$

transfer function representations of the column's dynamic response, using the method of Levy (63), are given in Tables D.11, D.12, D.13 of Appendix D.

5.5 Discussion

The process gains and the parameters of a first order lag plus time delay transfer function, describing the column dynamic responses, were determined from independent tests. The nonlinear behavior of the column responses necessitated the determination of the process gains as a function of the magnitude of the input disturbances, and the determination of the model parameters as a function of the input disturbance direction. Bornard et al (14) have also experienced the similar nonlinear behavior of the process gains. They also considered that despite the nonlinear behavior of the column, the model parameters of their process model would not vary significantly with the magnitude of the disturbances.

The dynamic parameters determined directly from the time domain data, were used to determine an average model describing the column dynamics, rather than those calculated from the frequency response data. These dynamic parameters were considered the best available, since they were determined directly from the time domain data rather than from a transformed data set (ie frequency response data). The difference between the time delays determined from the two frequency response methods is probably due to the insensitivity of the simple search technique used to find the 'best' time delay incorporated into the method of Levy. The quantitative comparison of time domain and frequency domain curve fitting procedures has been left for a future project.

In general, the gains calculated from the pulse testing procedure do not agree with those determined from the transient response analysis. This is because the pulse does not excite the nonlinearities to the same degree as a step of the same magnitude. Figure 5.4 represents the response of a hypothetical system to both a step and a pulse of the same magnitude. If the process illustrated is nonlinear, the response to the pulse cannot be expected to reflect the nonlinearities in the process that it did not experience (ie those nonlinearities in the output which were excited by the step disturbance at times greater than the pulse duration). Thus the gains calculated from the pulse test would be expected to be biased.

A comparison of the process gains and the dynamic

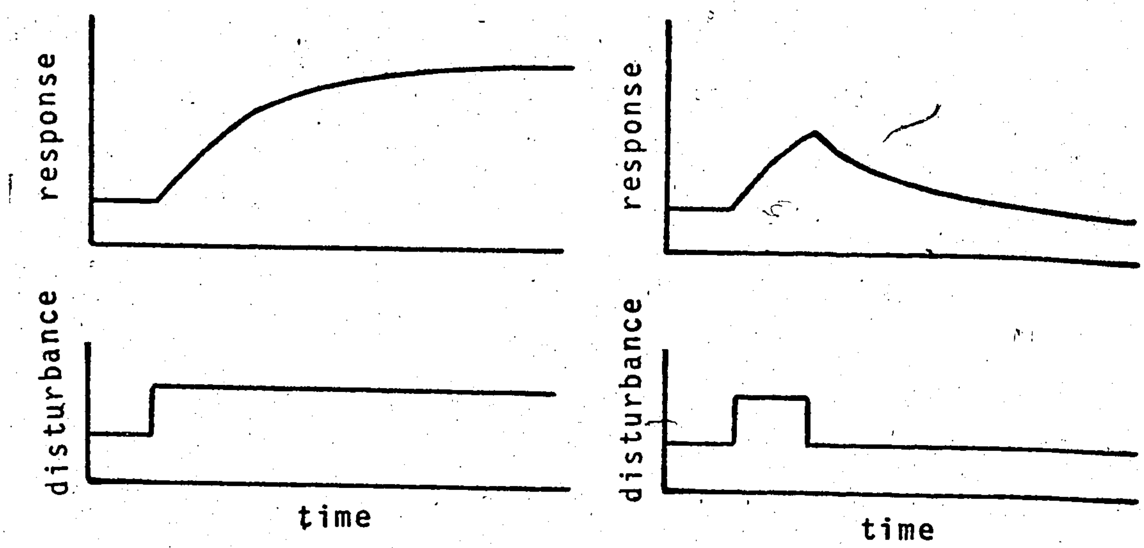

FIGURE 5.4 RESPONSE OF SYSTEM TO A STEP AND PULSE OF SAME MAGNITUDE

parameters of the approximate models indicates that both product composition responses are much more sensitive to the steam flow rate than to either the reflux or feed flow rates. The product composition responses are least sensitive to the feed flow rate.

Since both feed and reflux flow rate disturbances essentially alter the internal reflux distribution in the column, it would be expected that their relative effects would be of the same order of magnitude, while those for steam flow rate disturbances would be expected to be much greater, due to the changes in the vapour boil up rate. The response of the column to vapour flow rate changes is much more rapid than that due to liquid flow rate changes.

The dynamic response of the overhead product composition to feed flow rate resembles that of a feed composition change. Table 5.7 compares the parameters of the approximate dynamic model for each of these disturbances. The values determined for the time delay and time constant are of the same order of magnitude. This suggests that the overhead product composition responds at a rate relative to composition changes, while the bottom product composition responds at a rate relative to liquid flow rate changes. This would indicate why the bottom product composition response is much more sensitive to the feed flow rate disturbances than is the overhead product composition response. These results tend to agree with the discussion presented by Wahl and Harriott (134). The relative speed of response, slowest for composition disturbances, fastest for vapour flow rate disturbances, is also reflected in the order of magnitude of the time delays associated with the response of the overhead and bottom product flow rates given in Table 5.8.

Using the complex curve fitting technique of Levy (63), the fit of the frequency response data can be improved using a more complex polynomial transfer function representation. A first order lead/second order lag

TABLE 5.7 COMPARISON OF APPROXIMATE MODEL PARAMETERS DETERMINED FOR FEED FLOW RATE AND FEED COMPOSITION DISTURBANCES

	Model Parameter	Increasing Composition Responses		Decreasing Composition Responses	
		ΔF	ΔX_F	ΔF	ΔX_F
X_D	τ_D	6.0	9.0	8.0	6.5
	τ	27.0	26.0	10.0	19.0
X_B	τ_D	4.0	9.0		
	τ	12.5	12.0		

TABLE 5.8 APPROXIMATE TIME DELAYS IN THE RESPONSE OF THE OVERHEAD AND BOTTOMS PRODUCT FLOW RATES

Disturbance	Response D	Response B
F	10.0	1.0
R	0.5	4.0
S	1.0	0.5

transfer function representation appears to give the most significant improvement in the fit above that obtained with the simple first order lag plus time delay model. The use of additional terms gives only a marginal improvement in the fit obtained.

CHAPTER 6

FEEDFORWARD CONTROLLER MODEL

6.1 Introduction

The feedforward controller model has been calculated using the method outlined by Bollinger and Lamb (12), with the plant transfer matrix described by the open loop process transfer function models, determined from the transient response and pulse testing program. These models were also used to estimate the parameters required for the experimental feedforward controllers.

6.2 Determination of the Feedforward Gain

6.2.1 Procedure

The feedforward controller gain has been estimated using the following two methods:

a) substitution of the open loop process gains, illustrated in Figures 5.1 and 5.2, into Equation (3.23)

$$K_{ff} = - \frac{K_{XR}}{K_{XF}} \tag{6.1}$$

b) calculation of the slope at the reference steady state of the smoothed feedforward operating conditions (reflux flow rate as a function of the feed flow rate at constant overhead product composition).

$$K_{ff} = \left.\frac{dR}{dF}\right|_{X_D} \tag{6.2}$$

The values calculated from the above methods were also compared with the gains determined during the actual on-line experimental control studies.

6.2.2 Results

The feedforward controller gains calculated from the various methods are displayed in Figure 6.1. The large amount of scatter in the gain calculated from the unsmoothed feedforward operating data amplifies the conclusions reached in the previous chapter, that the gains should be calculated from the smoothed operating data rather than the untreated experimental data. The feedforward operating data, illustrated in Figure E.1 of Appendix E, appear to be best fit by a straight line of the form

$$R = 3.18 - 0.50\ F \tag{6.3}$$

from which the feedforward gain is a constant

$$K_{ff} = -0.50 \tag{6.4}$$

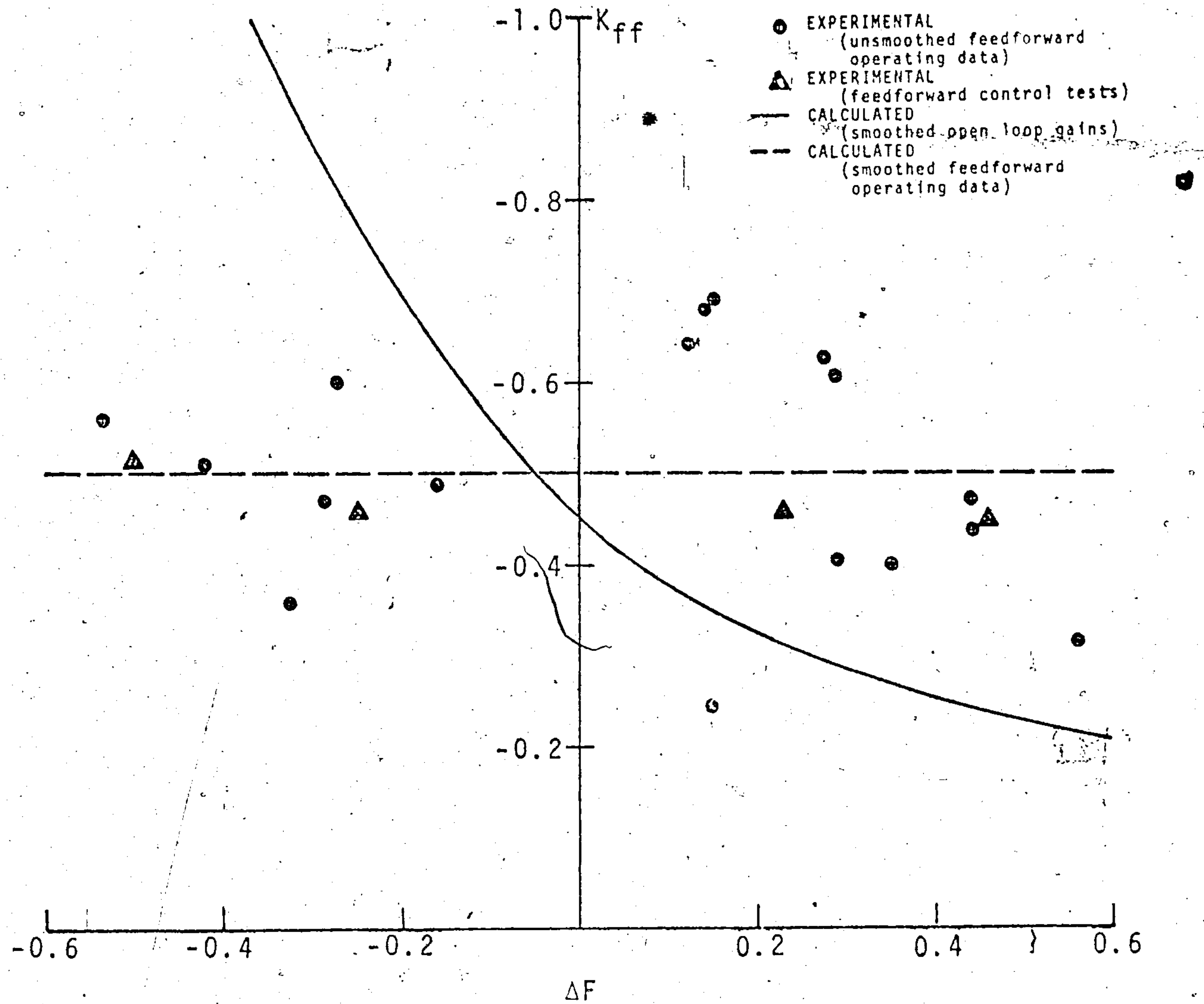

FIGURE 6.1 FEEDFORWARD CONTROLLER GAIN: VARIATION WITH FEED FLOW RATE

This value has also been displayed on Figure 6.1. The feedforward controller gains, calculated according to Equation (6.1), using the open loop process gains, K_{XF} and K_{XR}, have also been included in Figure 6.1.

The two point feedforward controller gains required to maintain both the overhead and bottoms product composition constant by manipulation of the reflux and steam flow rates, were ~~also evaluated by~~ substituting the values of the open loop process gains at the reference steady state into Equation (3.24) as

$$K_{ff} = \begin{bmatrix} 2.27 \\ -2.64 \end{bmatrix} \tag{6.5}$$

This controller, unfortunately, has not been evaluated.

6.3 Characterization of the Controller Dynamic Behavior

6.3.1 Procedure

The dynamic feedforward controller is calculated by substitution of the frequency response representation of the open loop process models, displayed in Figures D.2, D.5, D.12 and D.13 of Appendix D into Equation (3.23). These calculations were performed using the computer program FFC1, which is outlined in Appendix G. The dynamic feedforward controller model was also calculated using the frequency response representations of the approximate open loop transfer functions listed in Table 5.6.

Estimates of the dynamic parameters of simple time delay and first order time lag transfer functions were made, based on these calculated feedforward control frequency response representations. The parameters chosen during the later experimental evaluation were based on these estimates.

The time delay was determined directly from the phase angle of the frequency response representation as

$$\tau_D = \frac{\phi}{57.3\ \omega} \tag{6.6}$$

The time constant was determined u..ing the complex curve fitting procedure of Levy.

6.3.2 Results

The frequency response representations, calculated according to Equation (3.23), using the open loop frequency representations, are displayed in Figures E.2 and E.3 of Appendix E. Also shown on these figures are the frequency response representations of the feedforward controller, calculated using the frequency response representations of the approximate models. These frequency responses illustrate, qualitatively, the characteristics exhibited by the feedforward controller. The values of the time delay and the time constant, presented in Tables 6.1, 6.2 and 6.3, both appear to be less than 12 minutes for the existing column.

6.4 Approximate Feedforward Controller Models

The parameters of other higher order polynomial transfer function representations of the feedforward controller of the form

$$G_{ff} = \frac{\sum_{i=0}^{2} a_i s^i}{b_o + \sum_{j=1}^{3} b_j s^j} \qquad (6.7)$$

are presented in Tables E.4 and E.5 of Appendix E.

6.5 Discussion

The results presented in Figure 6.1 provide additional evidence of the nonlinear behavior of the column. If the column actually responded in a linear manner, the feedforward controller gains should be those calculated using the open loop process gains, K_{XF} and K_{XR}. However, the superposition theorem does not appear to be applicable in this case, as witnessed by the lack of agreement between the gains calculated from the open loop process gains and

TABLE 6.1 TIME DELAY DETERMINED FROM THE FEEDFORWARD CONTROLLER FREQUENCY RESPONSE DATA FOR INCREASING DISTURBANCES IN THE FEED FLOW RATE

Frequency	Test Designation				
	Reflux21 / Feed20	Reflux21 / Feed23	Reflux21 / Feed24	Reflux24 / Feed23	Reflux24 / Feed24
0.0020	4.4	10.5	7.9	6.1	2.6
0.0406	4.8	10.8	8.3	6.1	3.5
0.0308	4.4	10.8	8.2	6.2	3.8
0.0659	2.8	10.3	8.5	7.0	5.2
0.1640	3.3	10.6	4.7	11.5	5.5

TABLE 6.2 TIME DELAY DETERMINED FROM THE FEEDFORWARD CONTROLLER FREQUENCY RESPONSE DATA FOR DECREASING DISTURBANCES IN THE FEED FLOW RATE

Frequency	Test Designation			
	Reflux20 / Feed21	Reflux22 / Feed22	Reflux23 / Feed21	Reflux23 / Feed22
0.0020	0.0	1.7	7.0	4.4
0.0106	0.7	1.8	7.0	4.6
0.0308	0.9	2.0	7.4	4.5
0.0659	1.9	3.0	7.0	4.0
0.1640	5.9	5.5	4.8	3.6

TABLE 6.3 TIME CONSTANT DETERMINED FROM THE FEED-FORWARD CONTROLLER FREQUENCY RESPONSE DATA FOR BOTH INCREASING AND DECREASING DISTURBANCES IN THE FEED FLOW RATE

Increasing		Decreasing	
Test	τ	Test	τ
$\frac{\text{Reflux24}}{\text{Feed24}}$	6.5	$\frac{\text{Reflux22}}{\text{Feed22}}$	2.0
$\frac{\text{Reflux21}}{\text{Feed23}}$	12.2	$\frac{\text{Reflux20}}{\text{Feed21}}$	3.9
$\frac{\text{Reflux21}}{\text{Feed20}}$	3.5	$\frac{\text{Reflux20}}{\text{Feed22}}$	2.1
$\frac{\text{Reflux21}}{\text{Feed24}}$	9.3	$\frac{\text{Reflux22}}{\text{Feed21}}$	4.0

that determined from the smoothed feedforward operating conditions. Reasonable agreement was only obtained in the region of the reference steady state. This would be expected, since it is generally accepted that most nonlinear systems can be approximated in a linear manner for small deviations from the reference steady state. It is this fact which allows the determination of the two point feedforward controller given in Equation (6.5) to be made.

Figure 6.1 has illustrated a rather unexpected result. Despite the nonlinearities exhibited by the open loop responses, the feedforward operating conditions are linear about the reference steady state, resulting in a constant feedforward gain, which is applicable over the range of feed flow rate disturbances studied. These results have been verified by the tuned feedforward gains measured during the evaluation of the on-line closed loop control of the column reported in the next chapter. These gains, shown in Figure 6.1, varied only slightly from this constant value. This result is unexpected, since Luyben (69) has pointed out that the feedforward gain would only be constant at different feed flow rates, if both overhead and bottoms compositions were maintained constant. Clearly, additional work is required to explore this type of behavior that results, no doubt, because of the inherent nonlinearities in the system.

In light of taking into account the nonlinearity of the column in calculating the feedforward gain, it would appear contradictory to now calculate the feedforward dynamics by neglecting these nonlinearities. This was assumed possible by taking into account the following three considerations. First, as described by Bornard (15), that although the process gains exhibit a strong dependency on the operating conditions, the process dynamics would not be expected to possess such a strong dependency. Secondly, it is of prime importance to obtain the best value possible for the feedforward gain, while larger inaccuracies in the dynamic model parameters are tolerable. Thirdly, it is hoped that, despite the nonlinear effects, the same general form of an approximate model would prove adequate. The nonlinearity of the column would only require that the parameters evaluated be fine tuned to the process.

Despite the large amount of scatter evident in Figures E.2 and E.3, the feedforward controllers, calculated from the open loop process models, all tend to exhibit similar characteristics, while those calculated using the approximate transfer function models exhibit markedly different properties, especially at the higher frequencies. This indicates that, even though the simple first order plus time delay transfer function models adequately predict the open loop transient response of the column, they do not appear sufficiently complex to use in the calculation of the feedforward controller.

The intersection of the high frequency asymptote of the calculated feedforward models with the zero decibel low frequency asymptote occurs in the neighbourhood of 0.05 to 0.15 radians/minute. Since the open loop frequency response values were valid only up to frequencies of 0.2 to 0.3 radians/minute, the feedforward controller frequency response values would be expected to have a similar maximum cut off frequency. Thus, the majority of the dynamic information concerning the feedforward controller has been obscured. In order to recover this information, the open loop frequency response of the column must be extended to higher frequencies than were obtained during this study.

Examination of Figures E.2 and E.3 reveals that the main characteristic exhibited by most of the feedforward models is the lead function required to fit the resonant peaks occurring in the amplitude ratio. In fact, within the narrow range of valid amplitude ratio values obtained, the feedforward controller required to compensate for decreasing feed flow rate disturbances appears to be completely described with only a lead function. This observation, though, seems to contradict the observed phase angle, which decreases over the frequency range. These seemingly contradictory observations could be explained if it is assumed that the feedforward controller can be approximated with models of the form

$$G_{ff} = e^{-\tau_D s} (\tau_Z s+1) \tag{6.8}$$

$$G_{ff} = \frac{\tau_Z s+1}{\tau s+1} \qquad \tau_Z > \tau \tag{6.9}$$

$$G_{ff} = \frac{1}{\tau^2 s^2 + 2\tau\zeta s+1} \qquad \zeta < 1.0 \tag{6.10}$$

Since the magnitude of the resonant peaks exhibited by the amplitude ratio are relatively small, and the phase angle generally decreases over the range of frequencies studied, it is also possible to use very simple transfer functions of the form

$$G_{ff} = e^{-\tau_D s} \tag{6.11}$$

$$G_{ff} = \frac{1}{\tau s+1} \tag{6.12}$$

as approximations to the feedforward controller model. The range of the dynamic parameters presented in Tables 6.1, 6.2, E.4 and E.5 are probably caused by neglecting the nonlinearity of the column, and treating it as if it were a linear system.

Since no specific criteria describing the "goodness of fit" are readily available, the choice of which model to use is still a subjective decision, based on considerations such as:

a) is the model realistic?

b) is the model easily implemented?

c) is it worthwhile implementing a more complex model when a much simpler model will perform almost as well?

CHAPTER 7

EXPERIMENTAL CONTROL STUDIES

7.1 Introduction.

In order to make a valid assessment of the effectiveness of feedforward control, the performance of a system using conventional feedback control was initially determined as a base case. A number of previous workers (69, 91,132) have compared the feedforward response to the open loop response of the column. This comparison will only indicate if feedforward control is preferable over no control at all. No evaluation was made as to whether feedforward control can improve the column performance beyond that achieved at present using conventional control schemes, such as feedback control.

During this project therefore, the effectiveness of the various feedforward controllers will be compared to that of a conventional feedback controller, tuned to give the best response based on a minimum total deviation from the set point, as measured by the integral of the absolute error (IAE), a minimum overshoot and a reasonable settling time.

The feedforward models implemented by the computer were purposely kept simple in nature, so that they could easily be implemented without a computer, should the results justify their usefulness. The three models chosen illustrate the simplest possible approximations to the frequency response results presented in Chapter 6.

$$G_{ff} = K_{ff} \tag{7.1}$$

$$G_{ff} = K_{ff}e^{-\tau_D s} \tag{7.2}$$

$$G_{ff} = \frac{K_{ff}}{\tau s+1} \tag{7.3}$$

These simple models are easily tuned to the process, since at most, only one dynamic parameter must be adjusted, in addition to the feedforward gain.

The combined feedforward-feedback control schemes were implemented using the parameters which produced the most effective control behavior, as determined for each feedforward and feedback control implemented independently. The best feedforward-feedback controller was evaluated further to demonstrate its improved performance to a prolonged series of disturbances.

The transient response for each individual test conducted during this project are included in Appendix F. Portions of these figures have been reproduced as composite figures comparing the effectiveness of the various control schemes.

7.2 Procedure

Proportional plus integral feedback control was implemented using both a conventional pneumatic analog controller and the DDC algorithm available on the IBM-1800 digital computer. The controller settings were determined using the ultimate sensitivity method of Ziegler and Nichols (27). These controller constants were also compared with those calculated based on the process reaction curve of Cohen and Coon (27). The constants determined using the Ziegler-Nichols method were initially applied to the feedback controller and subsequently tuned to give an acceptable response.

The gain feedforward controller was initially implemented using the loop record structure, outlined in Figure 3.3, with the gain, defined from Figure 6.2, set to

$$K_{ff} = -0.5 \tag{7.4}$$

This value was then subsequently tuned to eliminate as much offset as possible, in the response of the overhead product composition due to both increasing and decreasing disturbances in the feed flow rate. Once the gain was determined, the other dynamically compensated feedforward controllers, defined by Equations (7.2) and (7.3), were implemented using the loop record structure outlined in Figures 3.4 and 3.5. The values of the dynamic parameters, τ_D and τ, were varied over the range

$$0 \leq (\tau_D, \tau) \leq 12 \qquad (7.5)$$

defined in Tables 6.1 to 6.3 in order to determine the best value for each term. The feedforward-feedback control schemes were implemented using the loop record structure defined in Figures 3.6 and 3.7. The feedforward gain and feedback control parameters were chosen as those 'best' values, based on the previous feedforward control tests. The value of the time constant was purposely chosen not to be the 'best' determined in the previous tests, in order to show the dramatic improvement in the control effectiveness by the addition of the feedback trim. Once the evaluation of the various control schemes was completed, it was decided to compare the effectiveness of the best gain plus time lag feedforward-feedback control with that of the feedback control alone. This comparison used a series of frequently changing step disturbances in the feed flow rate in various directions over a period of approximately six hours. The feedback controller parameters were set to the tuned values determined previously, while the feedforward controller parameters were set to the average of the 'best' values that were determined for both the increasing and decreasing feed flow rate disturbances.

The experimental testing method follows closely that described previously during the open loop process model determination. When the desired controller had been implemented, the proper controller constants entered and the column was operating at its reference steady state, the data acquisition was begun. Data was initially collected over a period of 10-15 minutes in order to establish the initial steady state conditions. These conditions were also recorded using the material and energy balance programs DASS/DATAC/BALNC. After sufficient steady state data had been collected, a single step disturbance in the feed flow rate was introduced. The transient response of the system was accumulated until the new steady state was attained, (generally within about two hours), after which the new steady state material balance was reported. The transient response data were transferred from the system disk files and punched onto cards and listed by the printer using the program GTDAT. When the data from the previous test had been completely removed from the system files, a new test could be initiated, which would return the feed flow rate to its reference steady state value. The documentation of the steady state material balances and the collection of the transient response data proceeded as outlined previously. The transient response data available on cards, could now be plotted offline using the plotting facilities available on the IBM-1800, to examine the effectiveness of the control scheme just tested.

The evaluation of the effectiveness of the various control schemes was based on the calculated integral of the absolute error (IAE) of the transient response of the controlled variable (overhead product composition) using the following expression:

$$IAE = \sum_{i=1}^{175} (X_D - X_{Dsp}) \Delta t \qquad (7.4)$$

This expression calculates the IAE for the first 93.3 minutes of the transient response after the initial entrance of the disturbance into the column.

The magnitude of the step disturbances considered during the remainder of this chapter is either +0.45 or -0.59 lb/min from the reference value of 2.46 lb/min.

7.3 Feedback Control

Table 7.1 gives the feedback controller parameters, (proportional and integral constants) determined using the Ziegler-Nichols method (27), Cohen-Coon method (27) and the final tuned values. The Ziegler-Nichols parameters were determined based on the gain at which the process started to exhibit an oscillating behavior, while the

TABLE 7.1 FEEDBACK CONTROLLER SETTINGS DETERMINED BY VARIOUS METHODS

Method	Analog		Digital		Digital in same units as Analog	
	PB (%)	τ_I(min)	KC	KI	PB (%)	τ_I(min)
Ziegler-Nichols	-100.	5.0	-1.25	0.091	- 80.	6.0
Cohen-Coon	-112.	3.5	-0.90	0.156	-111.	3.5
Tuned	-200.	9.0	-0.75	0.031	-133.	17.0

Cohen-Coon parameters were determined from the parameters of the approximate open loop model of the column given in Figure 5.6. These methods were used to determine the controller parameters for both the conventional analog controller and the digitally implemented controller. Since the controller constants used for both controllers are generally quoted in different units, Table 7.1 also contains the controller constants for the digital controller converted to the same units as the analog controller.

The transient response of the column under analog feedback control, using the Ziegler-Nichols controller constants, is illustrated in Figure 7.1, while the transient response of the digital feedback controller is illustrated in Figure 7.2. The use of the Ziegler-Nichols controller settings produced a very good controlled response with

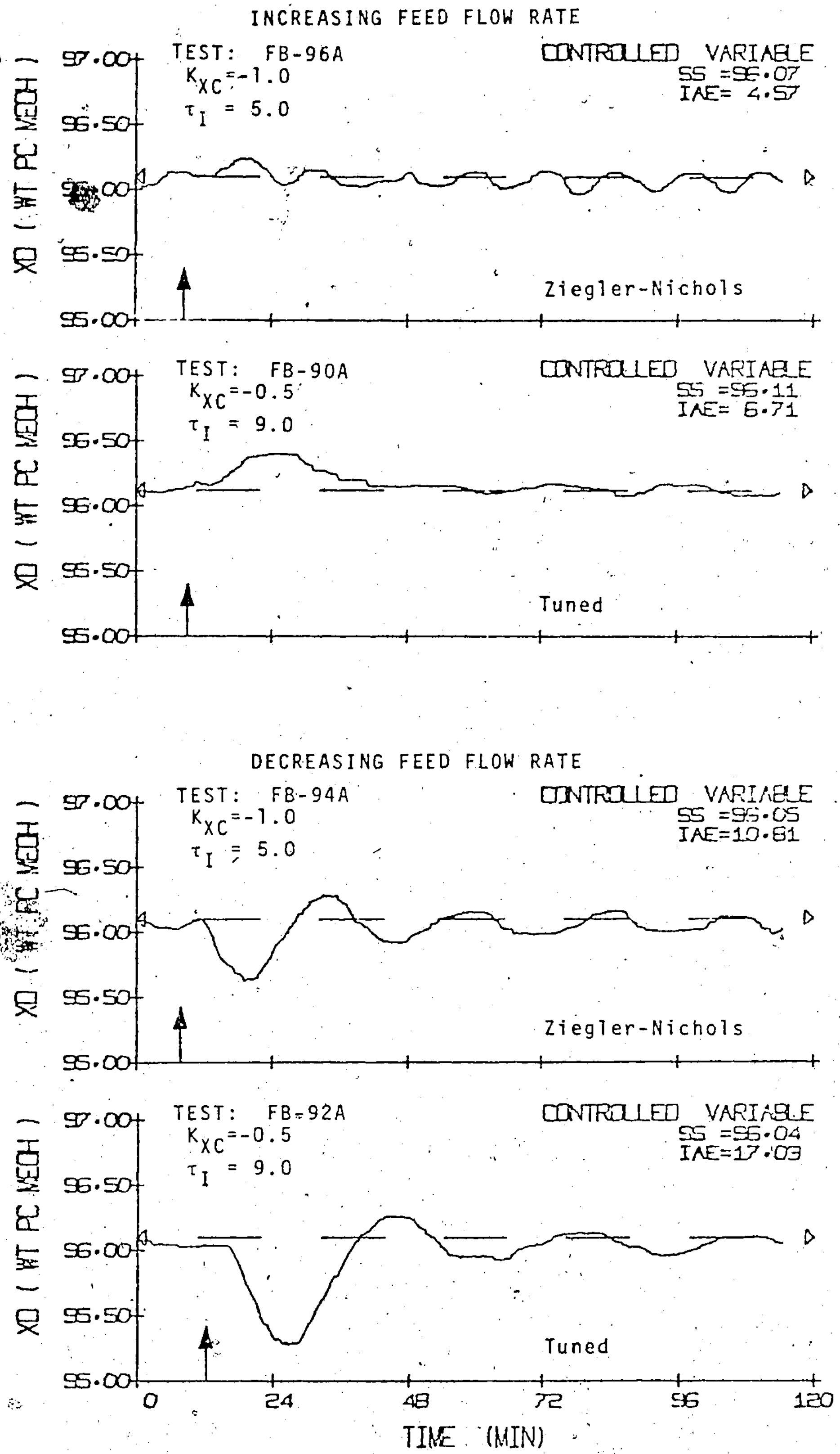

FIGURE 7.1 ANALOG FEEDBACK CONTROL: VARIATION OF CONTROL EFFECTIVENESS USING ZIEGLER-NICHOLS AND TUNED CONTROLLER PARAMETERS

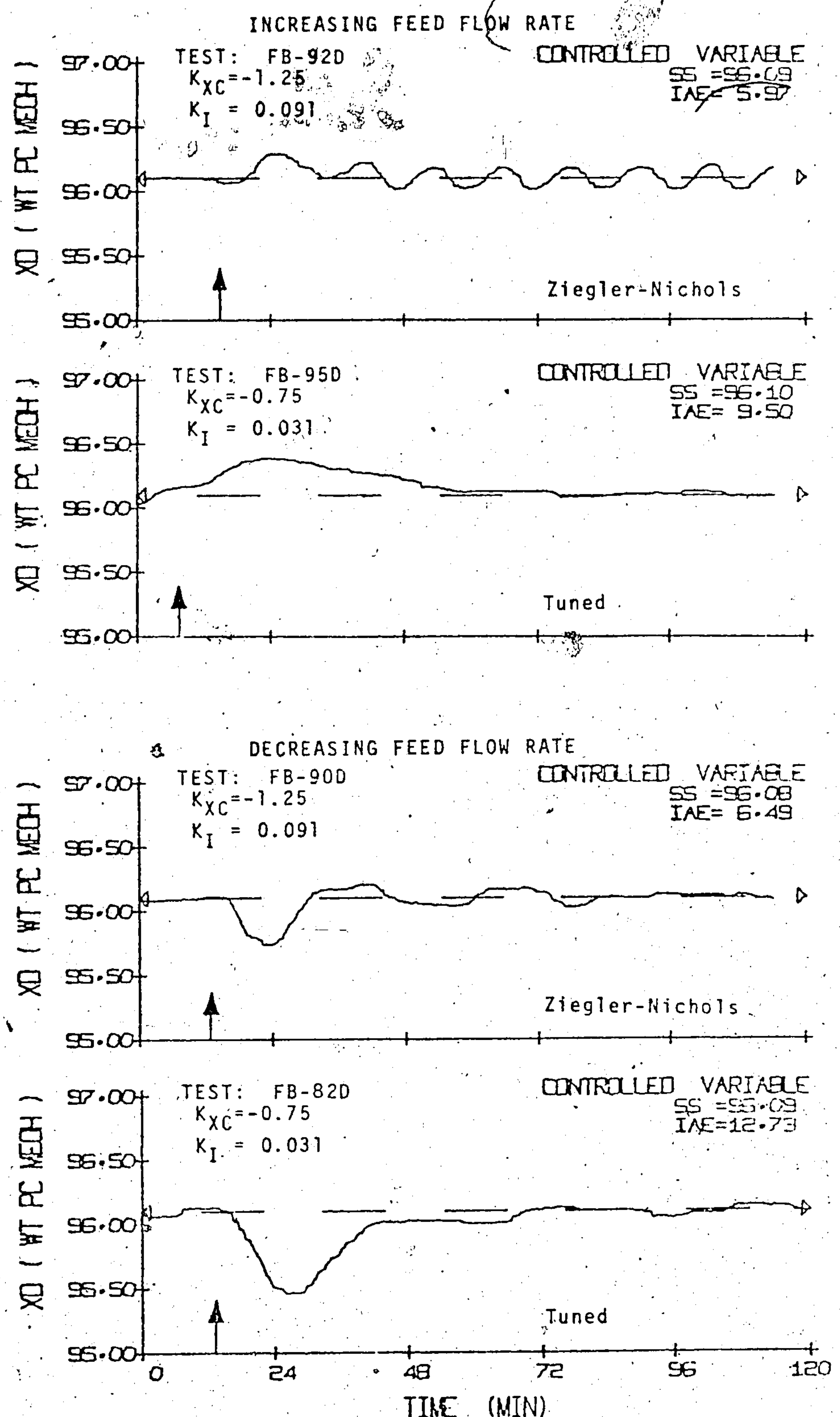

FIGURE 7.2 DIGITAL FEEDBACK CONTROL: VARIATION OF CONTROL EFFECTIVENESS USING ZIEGLER-NICHOLS AND TUNED CONTROLLER PARAMETERS

reference to only the small values of the IAE obtained. However, the response appears to contain very lightly damped oscillations. Although the amplitude of these oscillations was very small, ie of the order of 0.1% by weight methanol, this type of response would likely be unacceptable for industrial control purposes. The control parameters were tuned by relaxing the gain and integral constants somewhat, in order to produce a response in which the oscillations appear to be damped to a larger extent. The transient response of the column under analog feedback control using these tuned controller settings are also shown in Figure 7.1, while those under digital feedback control are shown in Figure 7.2.

The response of the controlled column using these tuned controller settings generally exhibited larger total deviations from the set point, a larger maximum deviation and a greater amount of oscillation damping than does the response obtained using the Ziegler-Nichols settings.

Due to the nonlinear behavior of the overhead product composition to reflux flow rate disturbances, compromise controller settings were selected. The controller parameter settings shown in Table 7.1 tend to give tighter control for input disturbances, causing increasing rather than decreasing deviations in the overhead product composition.

The effectiveness of the feedback control scheme, as measured by the IAE during these tests, is taken as the basis of comparison for evaluating the effectiveness of the various feedforward control configurations.

7.4 Gain Feedforward Control

Initially, the feedforward controller gain was set to 0.5, as suggested in Figure 6.2. The gain was then tuned in order to eliminate any offset resulting from both increasing and decreasing step disturbances in the feed flow rate. The tuned gains were determined to be 0.45 and 0.51 for step disturbances of +0.45 and -0.54 lb/min in the feed flow rate respectively. It was found that slight changes in these values were necessary to compensate for small variations in the feed composition over long periods of continuous operation.

Despite the attempts made to maintain as little offset as possible, many runs do contain an appreciable offset due to unknown causes. This illustrates the major weakness of a feedforward controller; its inability to correct for disturbances other than the ones to which it has been related. This contrasts with a feedback controller which can eventually correct for any disturbance as soon as the disturbance affects the controlled variable.

Figure 7.3 illustrates the response of the column under gain feedforward control for both increasing and decreasing disturbances in the feed flow rate away from the reference steady state. The overhead product composition response can be seen to deviate from its set point in a direction opposite to that expected to be caused by the disturbance alone, and returns to the set point with no overshoot or oscillations. This indicates that the reflux corrective action has been introduced into the column too early, demonstrating the need for dynamic compensation.

It should be noted that if the only two alternatives available for controlling the column were gain feedforward control and conventional feedback control, these tests would suggest that feedback control would be the preferred configuration. The response of the column under feedback control exhibits the smaller IAE value, the smaller maximum deviation and the smaller offset in the face of immeasurable disturbances of the two possible alternatives. If the gain feedforward controller is compared only to the open loop response, then it would be an acceptable control

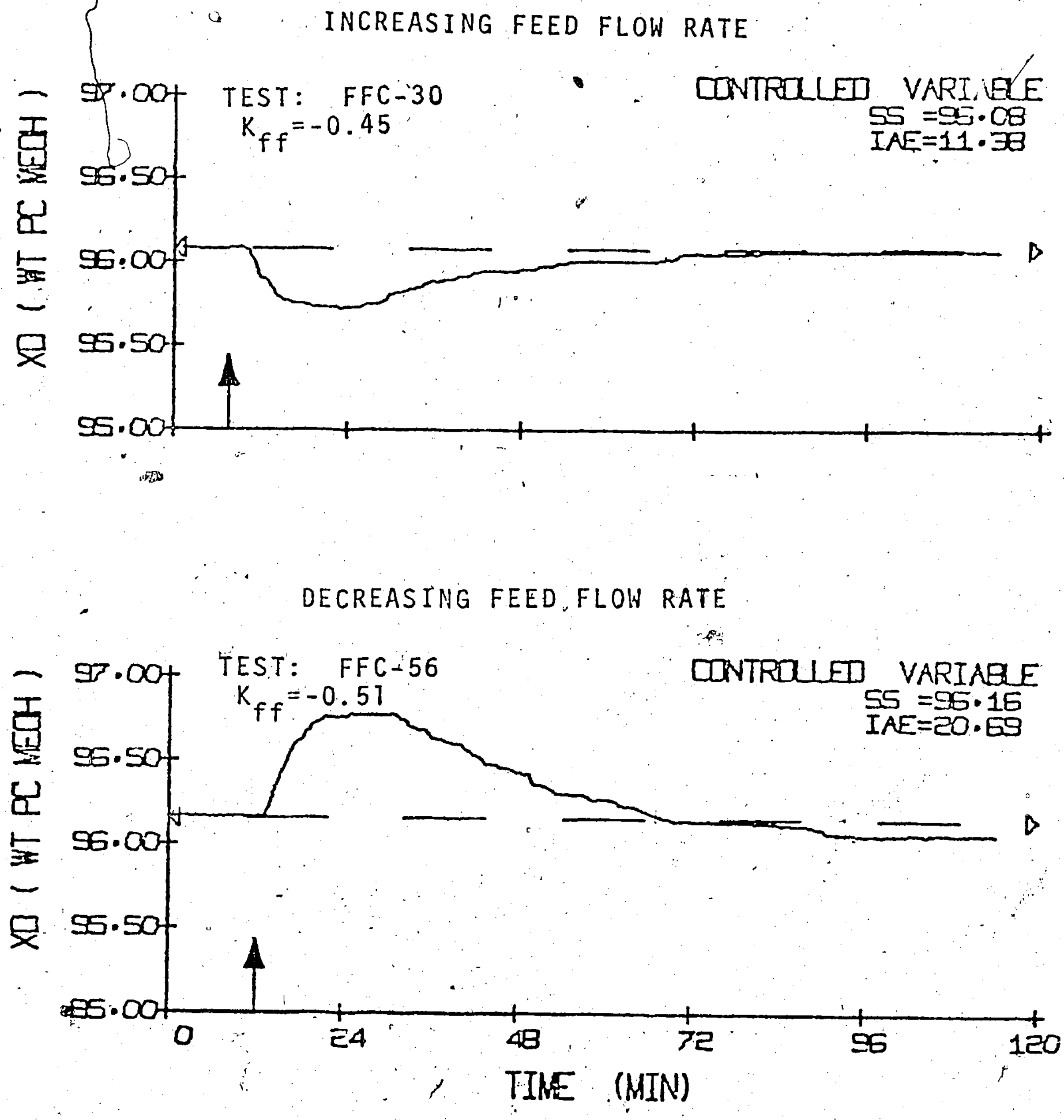

FIGURE 7.3 GAIN FEEDFORWARD CONTROL: VARIATION OF THE GAIN WITH THE DISTURBANCE DIRECTION

configuration.

7.5 Gain plus Time Delay Feedforward Control

Tables 6.1 and 6.2 present values of the time delays, calculated over a range of frequencies, from the feedforward controller frequency response data in Tables E.2 and E.3. These results suggest that for increasing feed flow rate disturbances, a time delay of between three to nine minutes is required, while for decreasing disturbances, a time delay of the order of zero to six minutes would be desirable. The fact that the time delay values calculated for increasing disturbances appear reasonably constant over the frequency interval of interest, suggests that a time delay is a fairly good representation of the data, whereas the time delays calculated for decreasing disturbances exhibit a considerable variation over the same frequency interval, suggesting that it is a less representative model in this instance.

Figure 7.4 compares the transient responses of the column using a gain plus time delay feedforward controller as a function of the time delay. Examination of this figure indicates that the best value of the time delay is of the order of five minutes. Examination of Figures F.34, F.35 and F.36 reveals that this value appears equally valid for

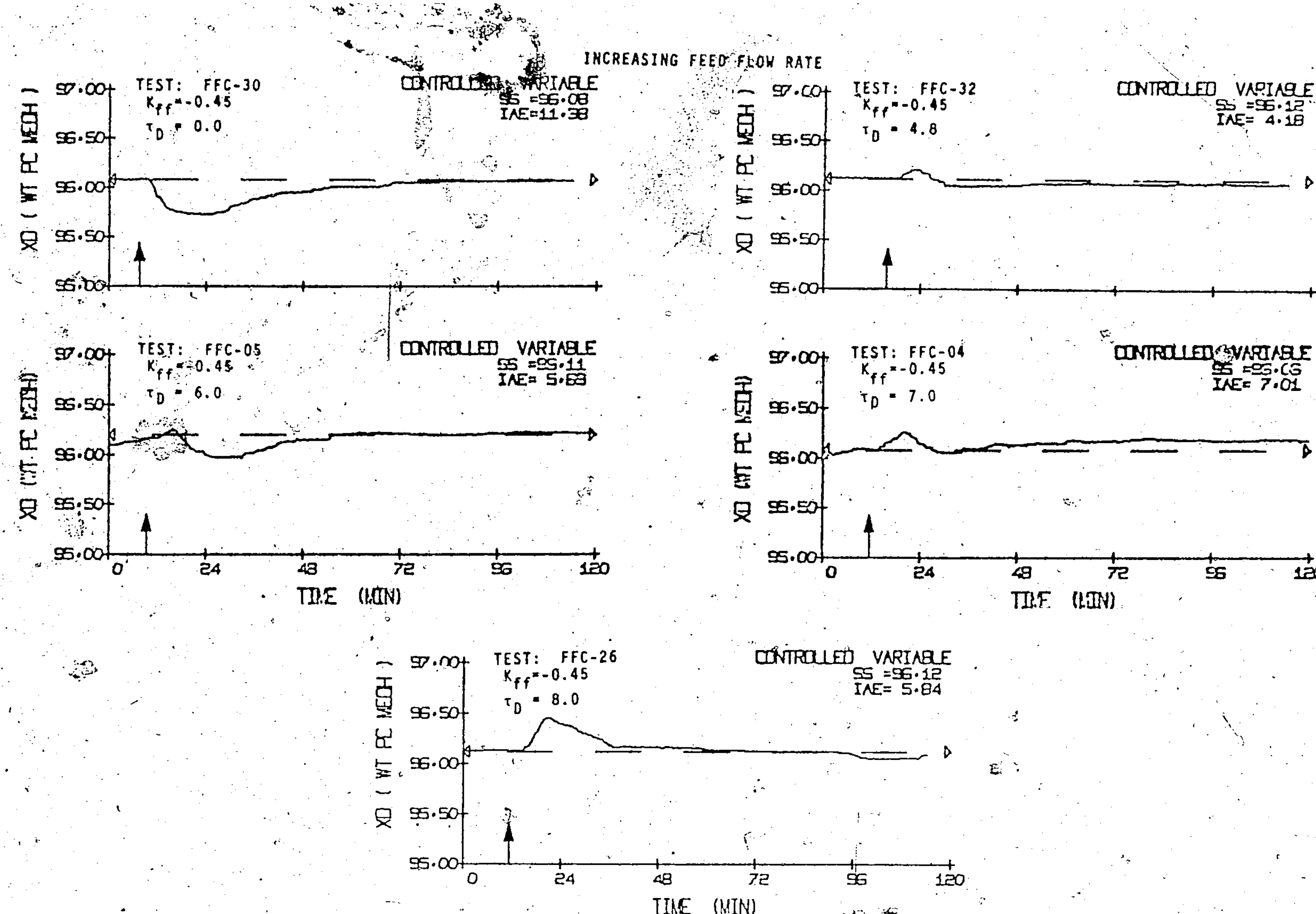

FIGURE 7.4 GAIN PLUS TIME DELAY FEEDFORWARD CONTROL: VARIATION OF THE CONTROL EFFECTIVENESS AS A FUNCTION OF THE TIME DELAY

both increasing and decreasing disturbances away from and returning to the reference steady state conditions.

The use of a time delay dynamic compensation cannot be expected to totally eliminate the transient response in the overhead product composition, unless the feed and reflux dynamics differ only by a time delay. The best that can be expected is to adjust the time delay such that the area under the transient response is approximately equally divided on either side of the set point cross over, as shown in Figure 7.4 for a time delay of 4.8 minutes.

The addition of time delay dynamic compensation to a gain feedforward controller has reduced the maximum deviation by about 75% and the IAE by approximately 75 to 90% from that achieved using a gain feedforward controller alone. The time delay feedforward controller has also reduced the IAE below that achieved using only a feedback controller.

7.6 Gain plus Time Lag Feedforward Control

Table 6.3 presents the time constant obtained by fitting the feedforward controller frequency response values given in Tables E.2 and E.3 to a first order time lag representation using the method of Levy (63). The results suggest a time constant ranging between four and ten minutes for increasing disturbances and between two and four minutes for decreasing disturbances in the feed flow rate away from the reference steady state.

Figure 7.5 compares the overhead product composition response of the column as a function of the time constant of the time lag feedforward controller, for an increasing step disturbance in the feed flow rate away from the reference steady state. Figures 7.6 and 7.7 present similar comparisons for a decreasing step disturbance in the feed flow rate away from and an increasing step disturbance in the feed flow rate returning to the reference steady state.

Examination of Figure 7.5 indicates that for an increasing disturbance away from the reference steady state, the best value of the time constant appears to be of the order of 6.6 minutes. In fact, the feedforward controller using this value of the time constant maintained the overhead product composition essentially constant, despite the approximately +20% feed flow rate disturbance. Examination of Figures 7.6 and 7.7 indicate that for a decreasing feed flow rate disturbance away from the reference steady state and for an increasing feed flow rate disturbance returning

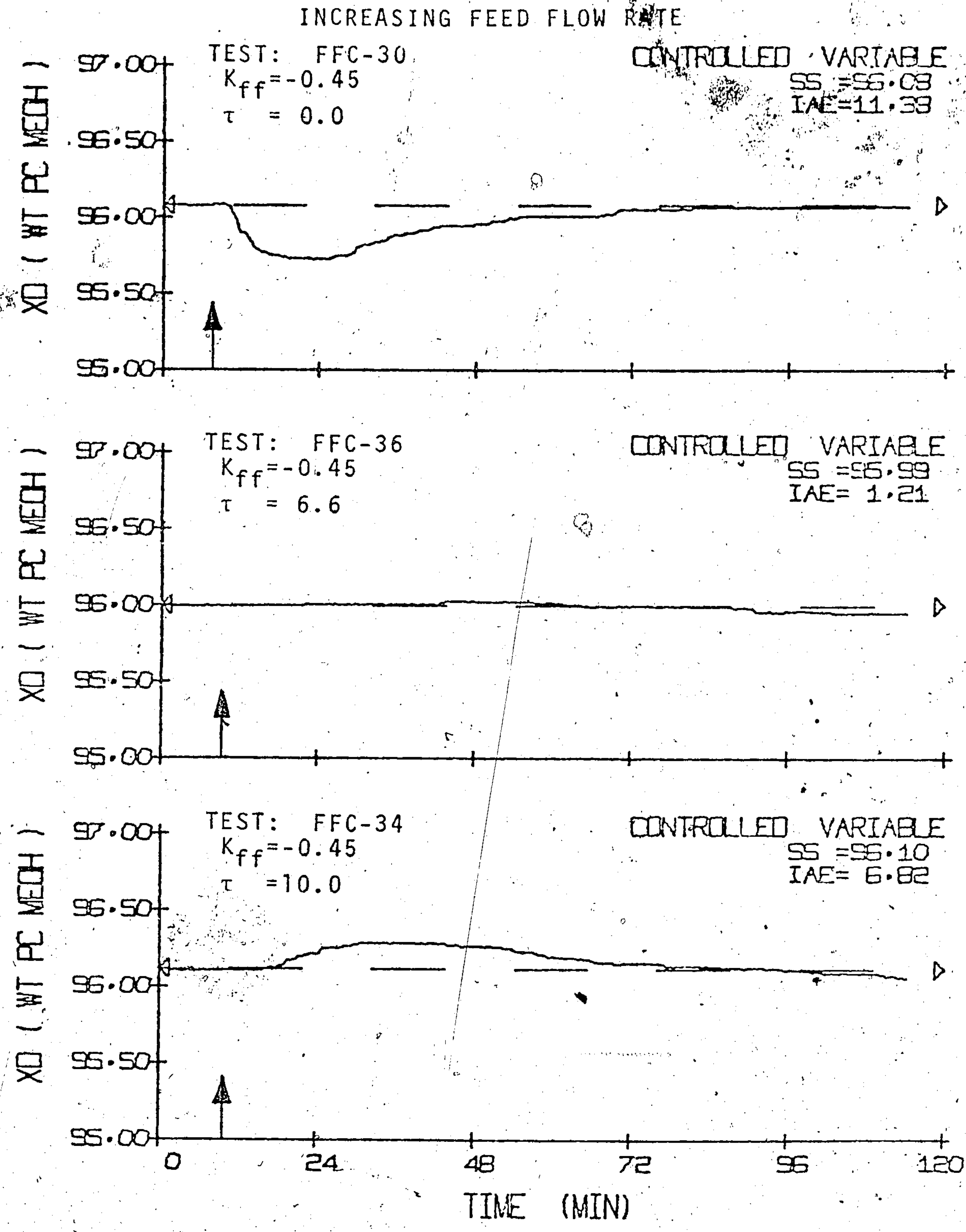

FIGURE 7.5 GAIN PLUS TIME LAG FEEDFORWARD CONTROL: VARIATION OF THE CONTROL EFFECTIVENESS AS A FUNCTION OF THE TIME CONSTANT

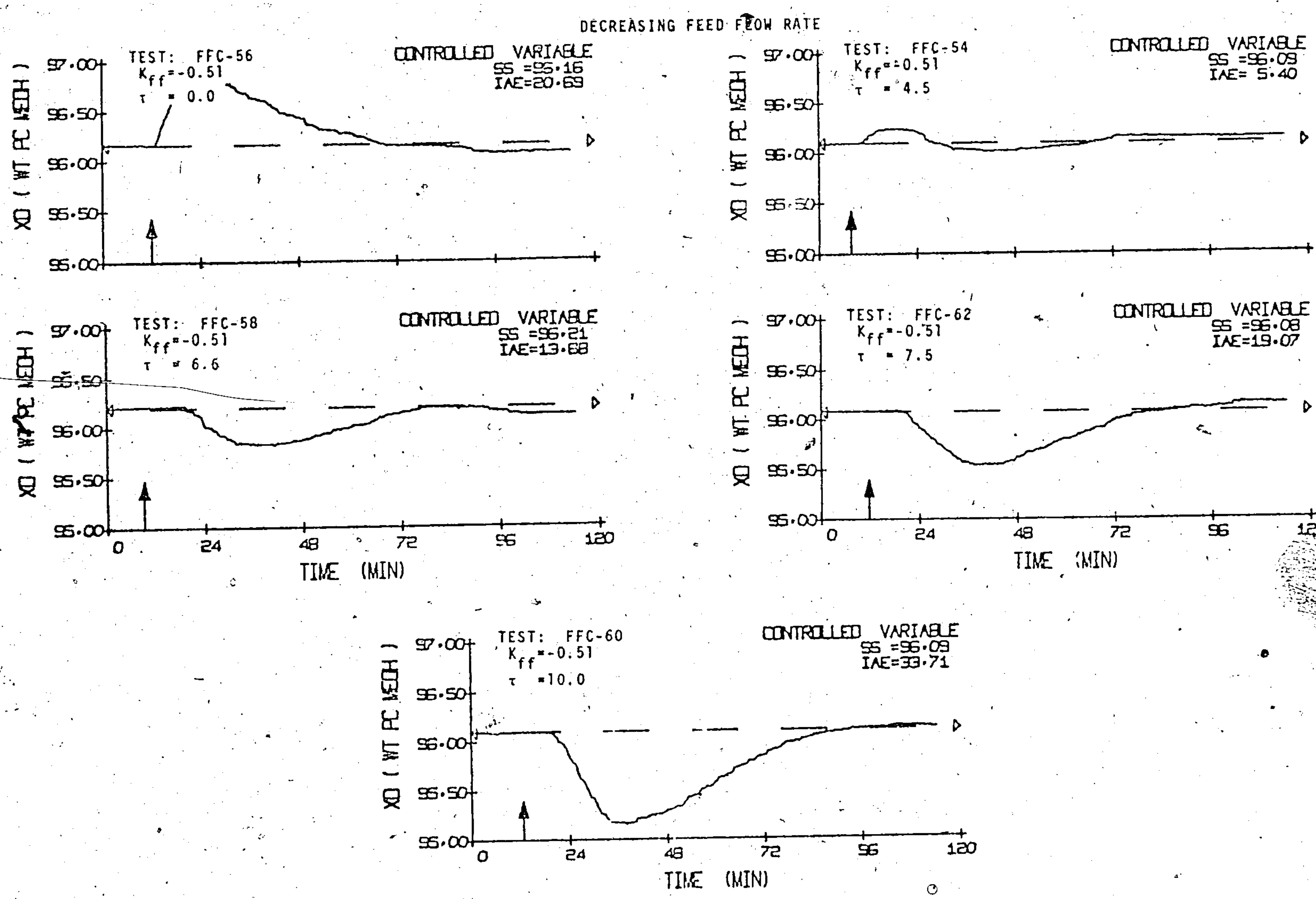

FIGURE 7.6 GAIN PLUS TIME LAG FEEDFORWARD CONTROL: VARIATION OF THE CONTROL EFFECTIVENESS AS A FUNCTION OF THE TIME CONSTANT

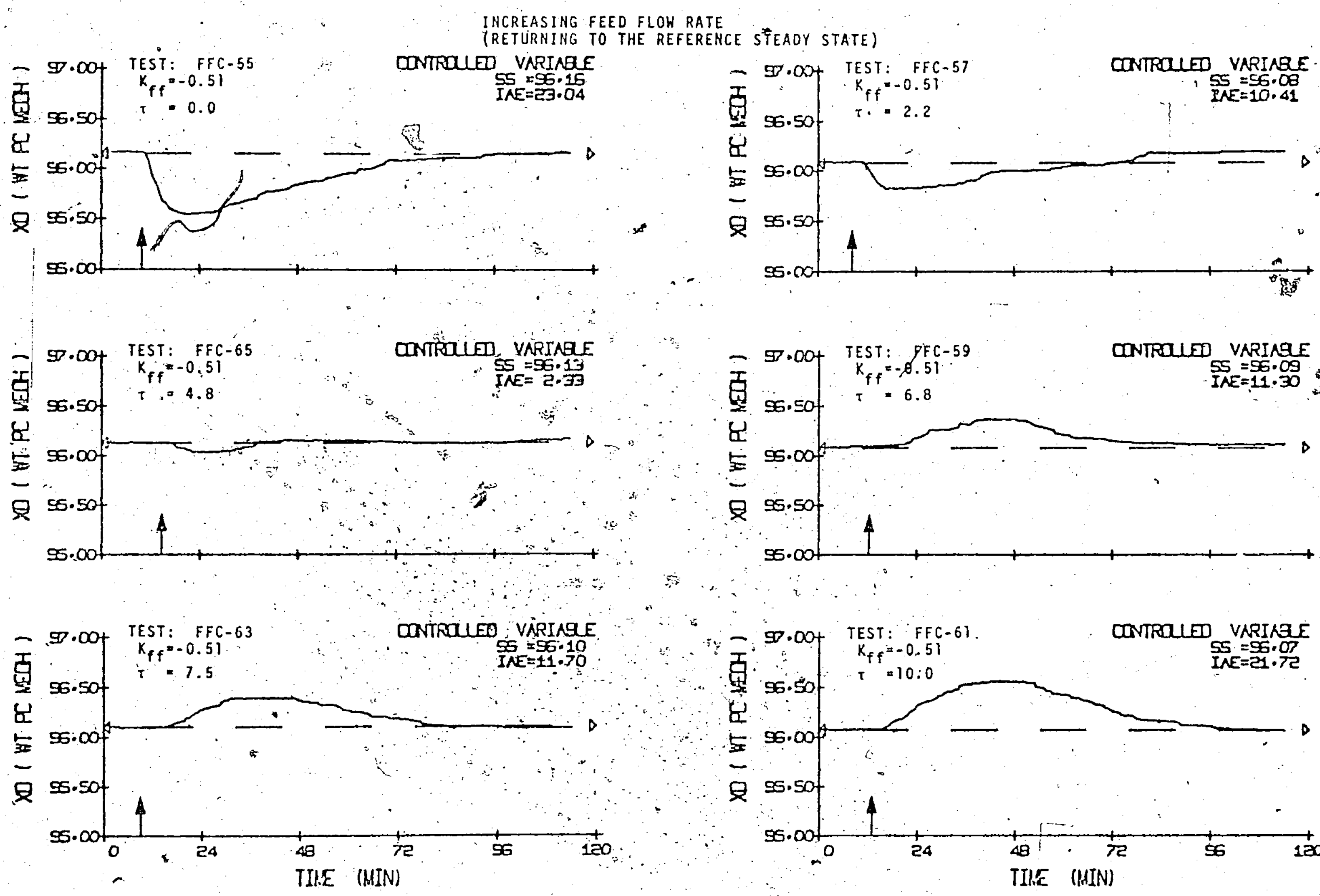

FIGURE 7.7 GAIN PLUS TIME LAG FEEDFORWARD CONTROL: VARIATION OF THE CONTROL EFFECTIVENESS AS A FUNCTION OF THE TIME CONSTANT

to the reference steady state, the best value of the time constant appears to be of the order of 4.5 minutes.

Tables 7.2 and 7.3 compare the controlled responses achieved in Figures 7.6 and 7.7 respectively. The IAE values have been corrected to remove the bias caused by the final offset observed in the response prior to the run termination. This would tend to make the comparison between the various time constants more dramatic. These results illustrate that, if care is not exercised in the selection of the time constant, the effectiveness of a time lag feedforward controller can be degraded to a level below that obtainable using only a gain feedforward controller. Undoubtedly, a similar result would be evident when using a time delay feedforward controller, had a larger range of time delay values been studied. Obviously, too much dynamic compensation can be much more harmful than not enough.

A well tuned gain plus time lag feedforward controller appears capable of reducing the IAE and the maximum deviation of the controlled response by at least 70% below those obtainable using a gain feedforward controller, and by at least 50% below those obtainable using a conventional feedback controller. A gain plus time lag feedforward controller and a gain plus time delay feedforward control-

TABLE 7.2 EFFECTIVENESS OF TIME LAG FEEDFORWARD CONTROL AS A FUNCTION OF THE TIME CONSTANT FOR DECREASING DISTURBANCES IN THE FEED FLOW RATE AWAY FROM THE REFERENCE STEADY STATE

Test	τ (min)	IAE*	Maximum Deviation wt % MeOH
FFC-56	0.0	20.7	+0.63
FFC-64	4.5	5.4	+0.13/-0.08
FFC-58	6.6	13.9	-0.36
FFC-62	7.5	19.1	-0.54
FFC-60	10.0	33.7	-0.92

*IAE - corrected for offset

TABLE 7.3 EFFECTIVENESS OF TIME LAG FEEDFORWARD CONTROL AS A FUNCTION OF THE TIME CONSTANT FOR INCREASING DISTURBANCES IN THE FEED FLOW RATE RETURNING TO THE REFERENCE STEADY STATE

Test	τ (min)	IAE*	Maximum Deviation wt % MeOH
FFC-55	0.0	23.0	-0.30
FFC-57	2.1	8.4	-0.25
FFC-65	4.5	2.3	-0.12
FFC-59	6.6	10.6	+0.30
FFC-63	7.5	11.7	+0.30
FFC-61	10.0	21.7	+0.50

*IAE - corrected for offset

ler, when both are properly tuned, appear to be equally effective in reducing both the IAE and the maximum deviation.

7.7 Feedforward-Feedback Control

Figures 7.8 and 7.9 compare the transient response of the column obtained by adding feedback trim to the feedforward controller with that obtained using the feedforward controller alone. The best time constants of the gain plus time lag feedforward controller, determined in the previous section, were 6.6 minutes for increasing and 4.5 minutes for decreasing feed flow rate disturbances away from the reference steady state. Despite this fact, the values of the time constants were chosen as 10.0 and 6.6 respectively, in order to illustrate the rather dramatic increase in the effectiveness of even a poorly tuned gain plus time lag feedforward controller with feedback trim, over that obtained using a gain feedforward controller.

The addition of feedback trim to a gain feedforward controller has markedly reduced the effect of the disturbance on the overhead product composition response, as measured by the IAE; by at least 60% over that obtainable

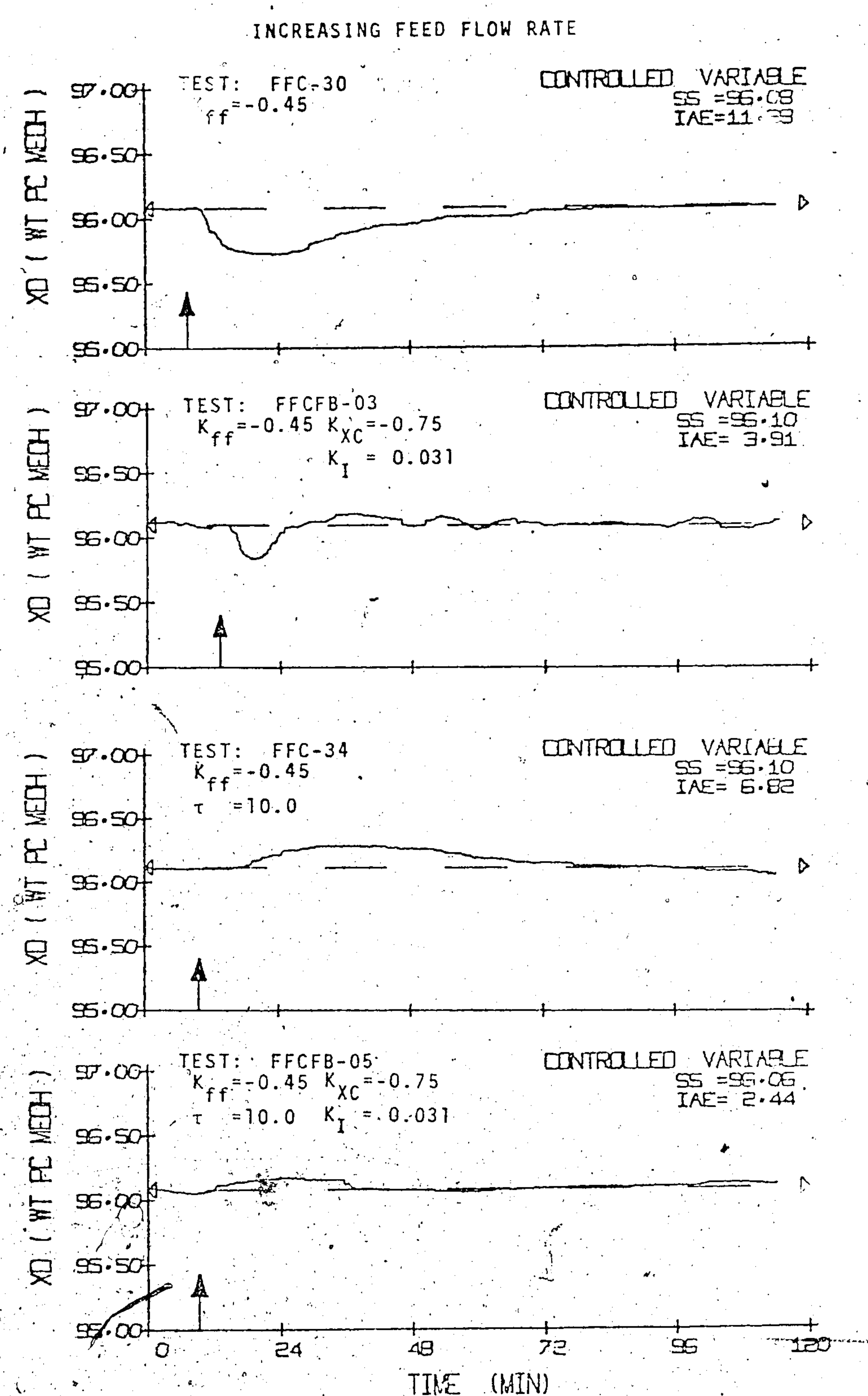

FIGURE 7.8 FEEDFORWARD-FEEDBACK CONTROL: COMPARISON OF THE CONTROL EFFECTIVENESS WITH THAT OBTAINED USING FEEDFORWARD CONTROL

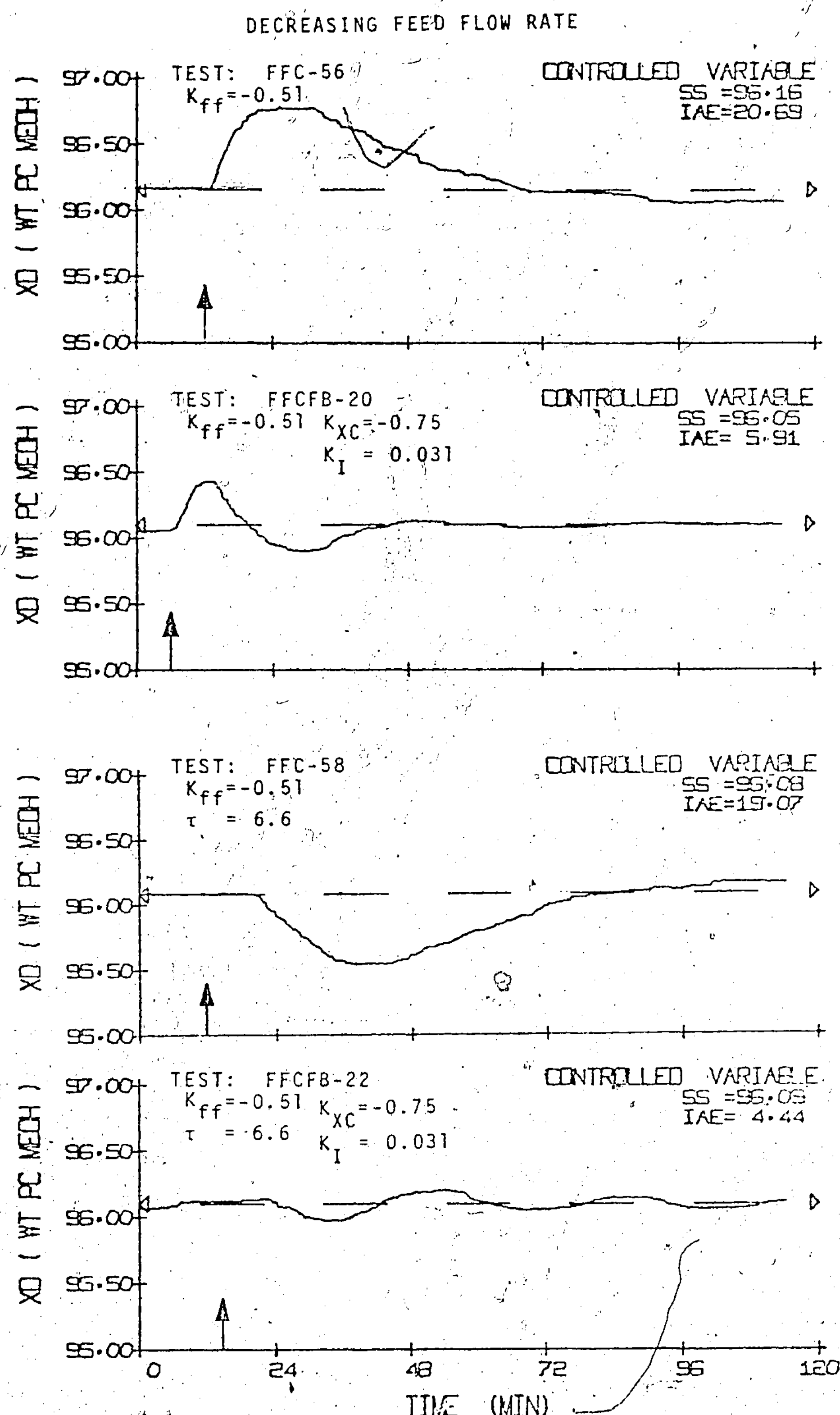

FIGURE 7.9 FEEDFORWARD-FEEDBACK CONTROL: COMPARISON OF THE CONTROL EFFECTIVENESS WITH THAT OBTAINED USING FEEDFORWARD CONTROL

using a gain feedforward controller; and by at least 20% over that obtainable using a conventional feedback controller. When the feedback trim is added to the gain plus time lag feedforward, the effect of the disturbance was reduced by at least 80% from that obtained using a poorly tuned gain plus time lag feedforward controller alone; by at least 20% from that obtained using a well tuned gain plus time lag feedforward controller alone; and by at least 50% over that obtained using a conventional feedback controller alone. The addition of feedback trim has also eliminated the majority of the offset due to either model errors or immeasurable disturbances.

This property has been dramatically illustrated in Figure 7.10. During the evaluation of the gain feedforward controller, the feed temperature controller failed, allowing the feed temperature to decrease from 163°F to 153°F, creating a large offset in the overhead product composition. A similar test was also performed when evaluating the gain feedforward controller with feedback trim. A step decrease in the feed temperature of -10°F was introduced into the column, in addition to the step increase in the feed flow rate. The resulting response contained no offset and a marked decrease in the effect of the combined disturbance. In fact, the IAE for the gain feedforward controller with feedback trim was even reduced well below that obtained

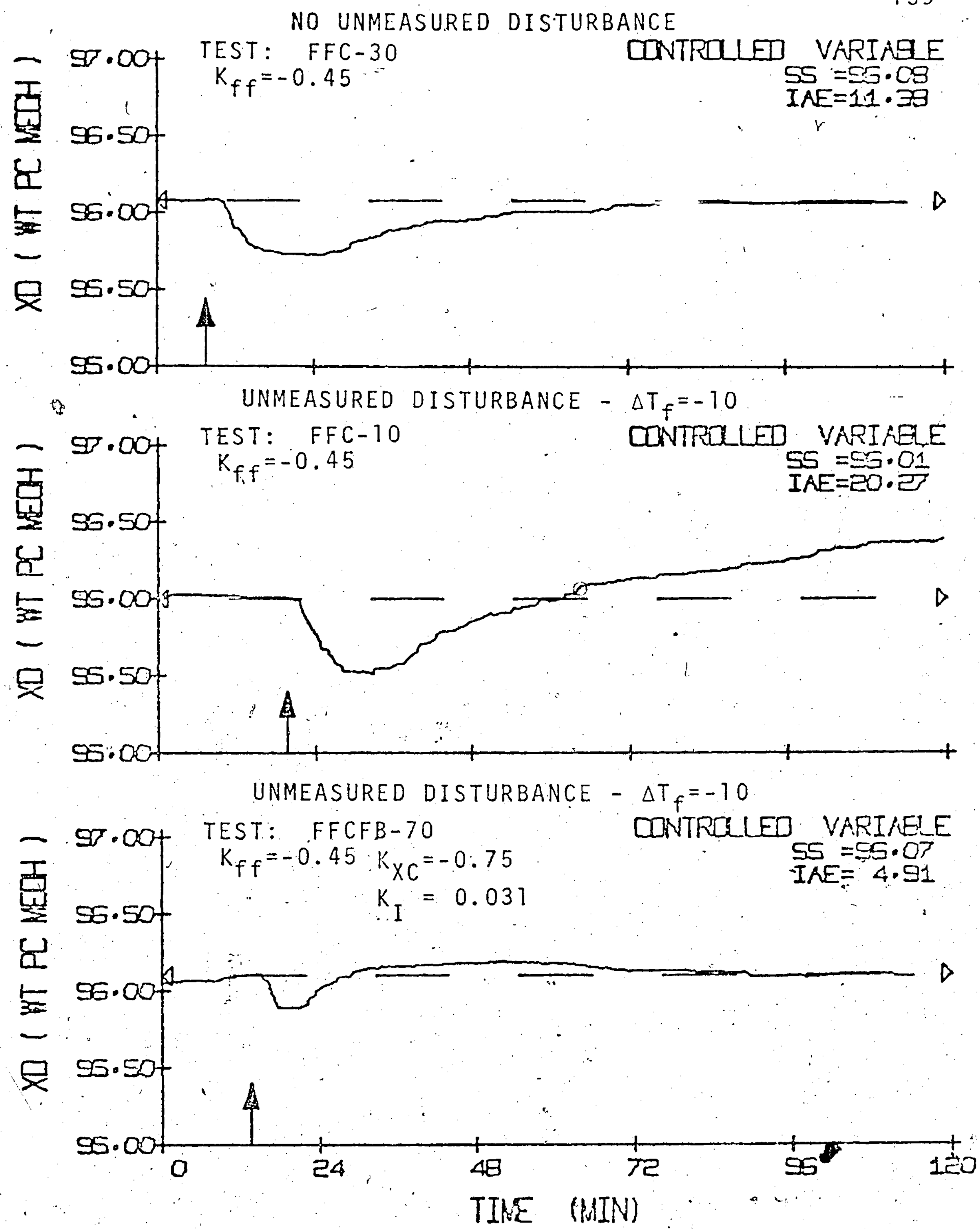

FIGURE 7.10 COMPARISON OF THE EFFECT OF AN UNMEASURED DISTURBANCE ON THE CONTROL EFFECTIVENESS OF FEEDFORWARD AND FEEDFORWARD-FEEDBACK CONTROLLERS

with the gain feedforward controller when no immeasurable disturbance was present.

7.8 Comparison of Controller Responses to a Series of Step Disturbances

Since the feedforward-feedback controller produced such a significant increase in the effectiveness of the control, it was decided to compare the effectiveness of this control scheme to that obtained using a conventional feedback controller, using a prolonged series of increasing and decreasing step disturbances in the feed flow rate about the reference steady state. Examination of Figure 7.11 indicates a reduction of approximately 70% in both the IAE and the maximum deviation about the set point has been achieved using a gain plus time lag feedforward controller with feedback trim in place of the feedback controller. The feedforward controller parameters used were the averages of those values reported previously as the best values for the increasing and decreasing disturbances in the feed flow rate.

$$G_{ff} = -\frac{0.48}{5.7s+1} \qquad (7.5)$$

The tuned values of the feedback controller parameters reported previously in Table 7.1 were also used.

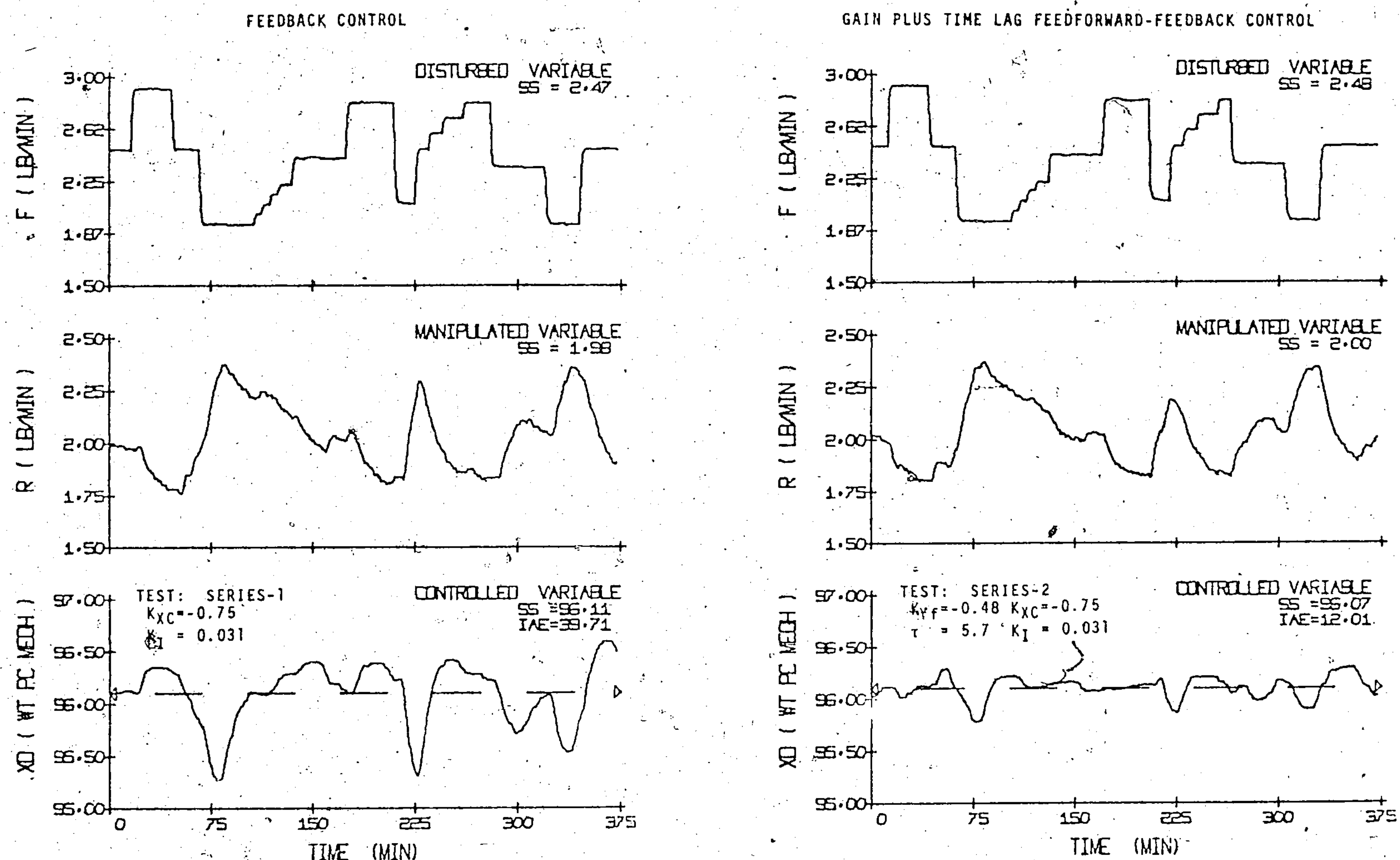

FIGURE 7.11 COMPARISON OF THE EFFECT OF A SERIES OF DISTURBANCES ON THE CONTROL EFFECTIVENESS OF FEEDFORWARD-FEEDBACK AND FEEDBACK CONTROL

7.9 Discussion

Figures 7.12 and 7.13 compare the effectiveness of the various control schemes evaluated during this study. Examination of these figures suggests that the order of preference of the control schemes discussed throughout this project would be

a) dynamic compensated feedforward control with feedback trim
b) gain feedforward control with feedback trim
c) dynamic compensated feedforward control
d) feedback control
e) gain feedforward control
f) open loop response.

Feedforward-feedback control schemes, in addition to improving the effectiveness of the control, also corrects for the effect of any immeasurable disturbances or of any errors in the feedforward control model. These errors may include either errors in the feedforward gain or in the use of poorly tuned dynamic parameters. Feedforward control alone does not exhibit this regulation capability. The addition of feedback trim to a feedforward controller actually reduces the transient response below that obtainable using feedforward control alone, in addition to the

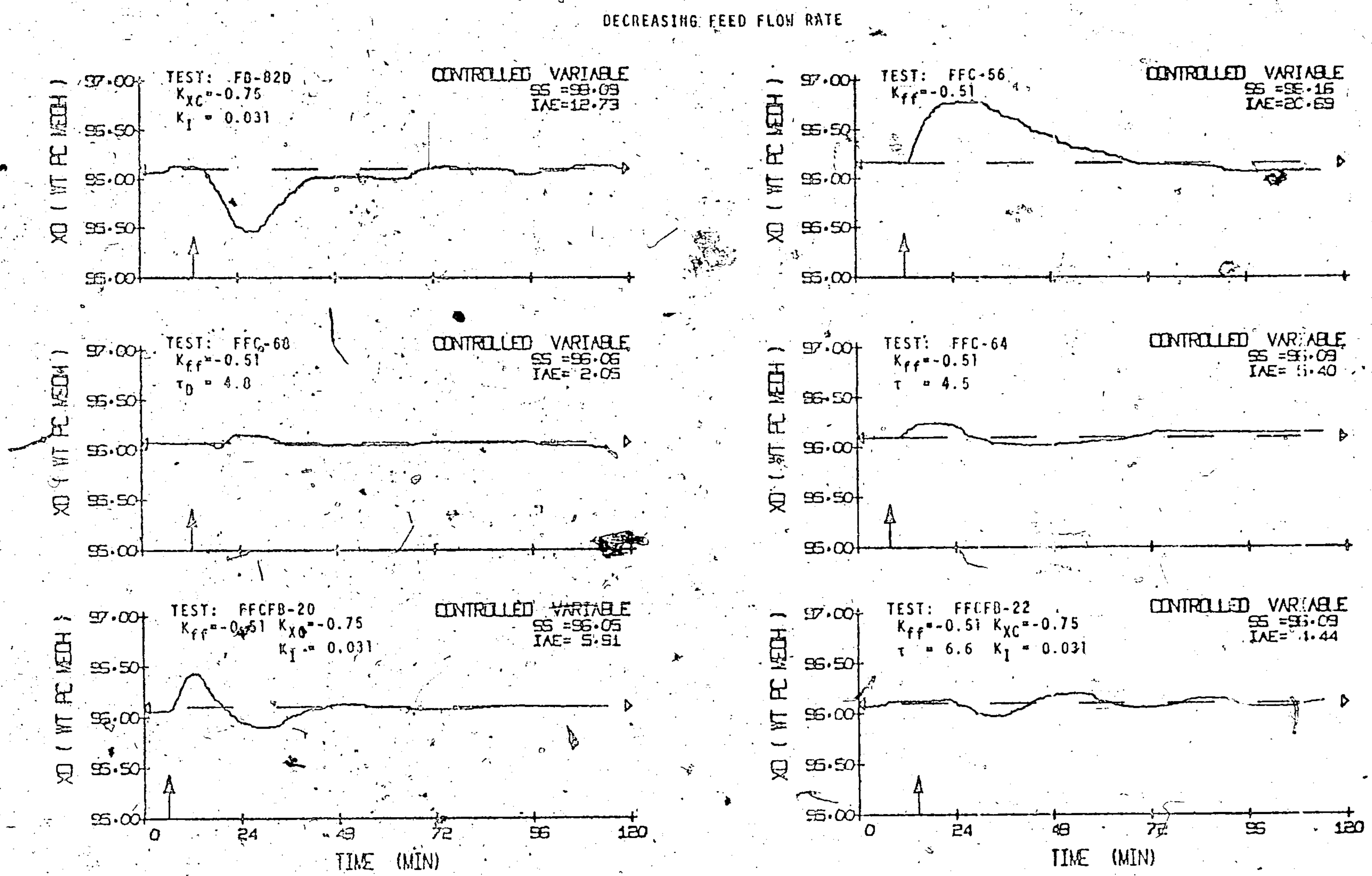

FIGURE 7.12 SUMMARY OF RESULTS:
COMPARISON OF THE CONTROL EFFECTIVENESS OF FEEDFORWARD AND FEEDFORWARD-FEEDBACK CONTROL WITH THAT OF FEEDBACK CONTROL

INCREASING FEED FLOW RATE

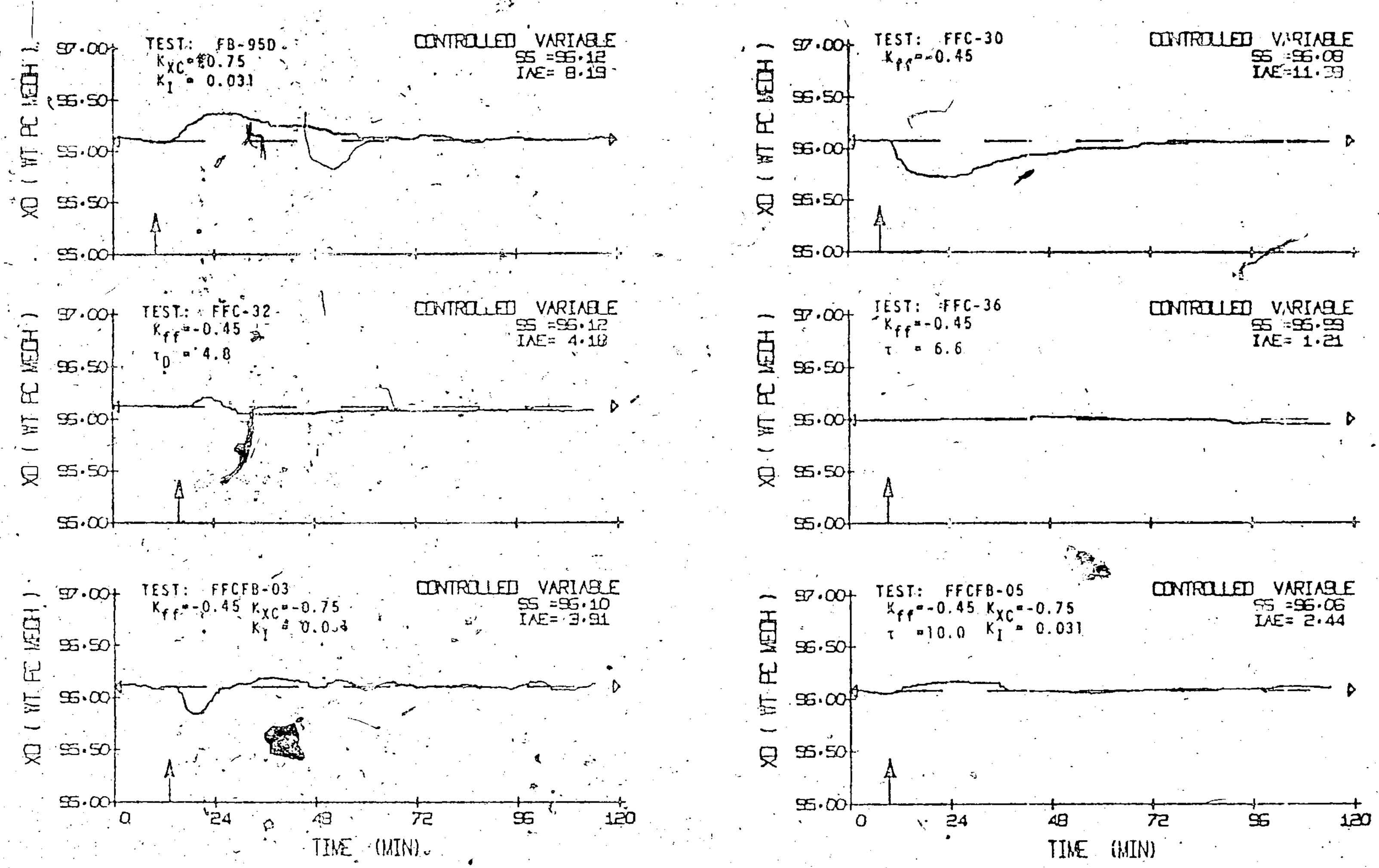

FIGURE 7.13 SUMMARY OF RESULTS:
COMPARISON OF THE CONTROL EFFECTIVENESS OF FEEDFORWARD AND FEEDFORWARD-FEEDBACK CONTROL WITH THAT OF FEEDBACK CONTROL

long term elimination of offsets.

Both time delay and time lag dynamic compensation appear of about equal ability in improving the effectiveness of the control. However, without access to a digital computer, a pure time delay function is quite difficult to implement. For this reason, the time lag transfer function was used in all the subsequent feedforward-feedback control schemes. It is interesting to note that for the present column, the value of the time delay and the time constant are both of the order of five minutes. These values are contained within the ranges presented in Tables 7.2 and 7.3.

CHAPTER 8

CONCLUSIONS AND RECOMMENDATIONS

8.1 Conclusions

1) A series of approximate transfer function models, which adequately describe the transient response of the column, have been successfully developed using a combination of transient response and pulse testing techniques.

2) The process gains should be calculated from the smoothed operating conditions, rather than from the original experimental data. This suggests that any attempt to adaptively tune the gain of a controller on-line will require a smoothing or filtering algorithm in the procedure.

3) The calculated process gains exhibit a strong nonlinear dependence on the magnitude of the disturbance. This nonlinear behavior is typical of columns producing relatively pure products, due to the operation of the column within the pinched regions. The nonlinear process gains could be adequately represented by a least squares straight line over the range of interest.

4) The open loop dynamic response of the column was successfully represented by a simple transfer function model of the form

$$G(s) = \frac{K \cdot e^{-\tau_D s}}{\tau s+1} \qquad (8.1)$$

Due to the nonlinear behavior of the column, two sets of parameters were determined, one set each for increasing and decreasing disturbances about the reference steady state. No attempt was made to determine a functional dependence between the magnitude of the time delay or time constant of this model and the magnitude of the disturbance. It should be noted that a similar model has been successfully used to describe the dynamics of large industrial columns (14,112).

5) The response of the column was most sensitive to the steam flow rate, followed by reflux flow rate, feed flow rate and least sensitive to the feed composition. This indicates that the column responds most rapidly to vapour flow rates, followed by liquid flow rates, but responds very slowly for composition changes.

6) The feedforward gain could not be calculated directly from the open loop process gains due to the nonlinear behavior of the column. The feedforward gain

was measured from the slope of the feedforward operating conditions, according to

$$K_{ff} = \left.\frac{dR}{dF}\right|_{X_D} \tag{8.2}$$

7) Despite the nonlinear behavior of the column, the feedforward operating conditions turned out to be linear, resulting in a constant value of the feedforward gain. The feedforward gain determined during the control system evaluation tests varied only slightly from this constant value.

8) The feedforward controller gain can, however, be calculated from the open loop process gains valid at the reference steady state.

9) The frequency response representation of the feedforward controllers calculated from the experimentally measured open loop models could not be predicted, within satisfactory limits, using the approximate open loop transfer function models. This situation resulted, despite the fact that the approximate models adequately represented the open loop transient response of the column. These simple models do not appear sufficiently complex to be used to determine the feedforward controller.

10) Since the range of the frequency response values did not extend to sufficiently high values of frequency, most of the dynamic information describing the feedforward controller has not been obtained. However, ranges of dynamic parameters of very simple transfer functions of the form

$$G_{ff} = e^{-\tau_D s} \tag{8.3}$$

$$G_{ff} = \frac{1}{\tau s+1} \tag{8.4}$$

were determined.

11) The implementation of feedforward controllers incorporating these models yielded a significant improvement in the effectiveness of the control of the column. The experimentally tuned values of the parameters of these models were included within the range of values determined from the frequency response representations. These results agree with those of Luyben (69,70) that a first order dynam compensation should be adequate.

12) Despite the nonlinear behavior exhibited by the column, simple and effective feedforward and feedforward-feedback control schemes were successfully implemented.

13) Based on the results presented in this project, the order of preference for the implementation of the various control schemes evaluated during this study would include:

a) dynamic feedforward control with feedback trim

b) gain feedforward control with feedback trim

c) dynamic feedforward control,

d) feedback control

e) gain feedforward control

14) Besides eliminating offsets caused by either the effect of immeasurable disturbances or errors in the feedforward model, due to either the approximate nature of the model or the varying column conditions, the audition of a feedback trim to a feedforward controller actually increases the effectiveness of the control.

15) Although the gain feedforward controller is not as effective as a conventional feedback control, the combination of the two greatly enhances the effectiveness of the overall composition control.

16) Only a limited knowledge of the column dynamics need be known in order to implement a feedforward controller

with feedback trim.

8.2 Recommendations

8.2.1 Distillation Column

Some of the flow recorders used during this project, including reflux, steam, overhead and bottoms product flows, were operated below approximately 30% of the transmitter span. This reduces the accuracy with which these flows can be measured and reset. Since the column operates very close to the flooding constraint, the span should be reduced to improve the accuracy and repeatability of metering these flow rates.

The cause of the flooding appears to be undersized downcomers. If they could be made larger, so that the internal reflux flow could be increased, the external flows could also be increased, resulting in shorter time constants for the dynamic response of the column, thus decreasing the time required for the transient response tests.

During the present study, the column was operated for long periods of time, often unattended. Any failure of any of the utilities, such as steam, cooling water, instrument air, or electricity, or a pump failure, could cause levels to be lost, pressures to rise, pumps to be operated dry, or many other potentially hazardous conditions, which could result in damage to the equipment. Since the computer has access to all the process measurements, it would be advantageous to have an operations monitor type of program , similar to that available on the evaporator (54), which could automatically shut down the column when symptoms of any hazardous conditions are detected.

8.2.2 Distillation Column Model

Sufficient experimental data has been presented to form the basis of a fundamental model which could be developed. This model could be used in future studies to aid in the study and development of other advanced control schemes. A satisfactory model should permit determination of the column response about any other column operating condition, rather than require the collection of new experimental data, since an experimental program is generally expensive, time consuming and inconvenient when applied to an industrial column.

8.2.3 Distillation Column Approximate Model

Further pulse tests are required to extend the frequency response of the open loop responses, so that the feedforward controller frequency response can be calculated to frequencies above 0.2 radian/minute. The frequency response of the feedforward controller was only valid up to this frequency, due to the pulse testing limitation imposed by the measured open loop frequency response. Since the time constant of the feedforward controller is of the order of five minutes, the break frequency of the model is in the vicinity of 0.2 radians/minute. Thus most of the high frequency characteristics of the model have been lost.

Further study should be required to determine the effect on the pulse test results of the nonlinear behavior of the column, and the causes of the abnormally high values of the normalized frequency content at the limit of the valid frequency response data. It is also possible that other stochastic testing techniques, such as Kalman filtering, pseudo-random binary inputs or an auto and cross correlation analysis of process noise, may provide better estimates of the dynamic model of the column than pulse testing. The existing column, with its data acquisition and control capabilities, is an ideal vehicle for the evaluation of these alternate methods.

8.2.4 Transient Response versus Frequency Response

The determination of the parameters of an approximate model using data in the time or frequency domain requires further study. It is necessary to determine criteria to decide which of the many methods of fitting the data should be used, and in the case of complex models, to determine when a sufficient number of terms have been included.

8.2.5 Feedforward Control

The feedforward controllers designed during this study are based on conventional continuous methods. The sample time chosen when implementing the controllers on the computer was small relative to the time constants associated with the process. This, however, does not make the best use of the computer. A digital feedforward model, based on sampled data techniques, should be designed, evaluated experimentally and compared with the control obtained using the continuous model. Further study would then be required to find the best value of the sampling interval for this model.

Dynamic feedforward control, incorporating only simple models such as a time delay or a time lag, has reduced the amount of deviation in the overhead product composition response in the order of 75% of that obtained using conventional feedback control alone. Further study could determine if the use of more complex models, such as a first order lead/lag (115), first order lag plus time delay (14), a second order lag (66), a second order lag plus time delay (66) or a first order lead/second order lag (82) would give sufficiently increased effectiveness of control to justify their implementation.

Under certain circumstances, it may be advantageous to control both overhead and bottoms product composition Various suggestions have been made (1,9,73,102,103,108 to implement non-interacting control, including a few based on feedforward concepts (1,73). Further experimental study is required to determine the feasibility and parameters required to implement the two-point feedforward control system.

8.2.6 Material Balance Control Scheme

The column material balance is currently controlled indirectly, with both overhead and bottoms product flow rates on level control. An alternate control scheme, recommended by Shinskey (115) and Nisenfeld (89), controls the material balance directly by adjusting either the overhead or bottoms flow rate directly, rather than either reflux or steam flow rates. Since each control philosophy appears to have obtained about equal acceptance, based on literature, references, it would seem worthwhile to experimentally study the direct method of material balance control and compare the results with those presented during the current project. This comparison could include the column dynamics, the feedforward controllers and the effectiveness of the feedforward control schemes. A study of this nature may help resolve the controversy concerning which material balance control scheme is more desirable, or under what circumstances each control scheme might be preferred.

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NOMENCLATURE

i) Upper Case

AA	-	defined by Equation (3.5a)
ADC	-	analog to digital converter
AMP	-	amplifier
AR	-	amplitude ratio, decibels
B	-	bottoms product flow rate, lb/min
BB	-	defined by Equation (3.5b)
BS	-	bias in the DDC loop record
CC	-	defined in Equation (3.5c)
COEF	-	vector of unknown transfer function coefficients
COS	-	current output station
D	-	overhead product flow rate, lb/min
DD	-	defined by Equation (3.5d)
DEN(iω)	-	complex denominator defined by Equation (3.9)
DF	-	degrees of freedom
E	-	error function defined by Equation (3.13)
E(iω)	-	error function in frequency domain defined by Equation (3.9)
F	-	feed flow rate, lb/min
FC	-	flow controller
FF	-	exponential filter constant in the DDC algorithm
F_{sp}	-	set point of feed flow rate controller

F_{ss} - steady state value of feed flow rate, lb/min

$F(i\omega)$ - function defined to fit the experimental frequency response data exactly, as defined by Equation (3.8)

FR - flow recorder

FRC - flow recorder/controller

$G(i\omega)$ - general transfer function in the frequency domain

G - general transfer function in the Laplace domain

$G_c(t)$ - general feedback controller equation in the time domain

G_{FM} - transfer function of the feed flow rate measurement transmitter

G_{pj} - transfer function relating the overhead product composition to the 'jth' input variable

G'_{pj} - transfer function relating the bottom product composition to the 'jth' input variable

G_{ij} - transfer function relating the 'ith' output variable to the 'jth' input variable
where the 'ith' output variable refers to

B - bottom product flow rate

D - overhead product flow rate

O - any other output variable of interest

and the 'jth' input variable refers to

F - feed flow rate

L - any other input variable of interest

R - reflux flow rate

- S - steam flow rate
- T_F - feed temperature
- V - boilup flow rate
- X_F - feed composition

G_R - closed loop transfer function of the reflux flow rate to a set point change

G_{RC} - transfer function of the reflux flow rate controller

G_{RM} - transfer function of the reflux flow rate measurement transmitter

G_{RV} - transfer function of the reflux flow rate control valve

G_{XC} - transfer function of the overhead product composition controller

G_{XM} - transfer function of the overhead product composition measurement transmitter

G_f - transfer function of the feedforward controller

H_2O - water

IAE - integral of the absolute error

$I(i\omega)$ - complex function defined by Equation (3.8)

I/P - current to pressure transducer

K - gain of a general transfer function

K_c - feedback controller gain

KC - feedback controller gain in DDC loop record

K_{FM} - gain of the feed flow rate measurement transmitter

K_I - integral controller constant for digital feedback controller

KI	-	integral controller constant in DDC loop record
K_{Nj}	-	gain of the transfer function relating the composition of plate N to the 'jth' input variable
K_{pj}	-	gain of the transfer function relating the overhead product composition to the 'jth' input variable
K'_{pj}	-	gain of the transfer function relating the bottoms product composition to the 'jth' input variable
K_{ij}	-	gain of the transfer function relating the 'ith' output variable to the 'jth' input variable where the 'ith' output variable refers to B - bottoms flow rate D - overhead flow rate O - any other output variable and the 'jth' input variable refers to F - feed flow rate L - any other input variable R - reflux flow rate S - steam flow rate T_F - feed temperature V - boilup flow rate X_F - feed composition
L	-	any other input variable not otherwise defined, ie T_R, T_A, T_C, T_S
$\mathcal{L}(\ \ )$	-	Laplace transform operator
LC	-	level controller
LIC	-	level indicating controller

MeOH	-	methanol
MPX	-	multiplexer
MS	-	measurement in DDC loop record
N	-	general plate number
NPTS	-	number of data points collected
NSTAGE	-	total number of stages including the reboiler
NUM($i\omega$)	-	complex numerator defined by Equation (3.7)
O	-	any other output variable not otherwise defined, ie X_N, P
OT	-	output in DDC loop record
P	-	pressure, inch H_2O
PB	-	proportional band of feedback controller, %
PC	-	pressure controller
P/I	-	pressure to current transducer
PIC	-	pressure indicating controller
PO	-	pulse output
POC	-	process operators console
R	-	reflux flow rate, lb/min
R_{sp}	-	set point of reflux flow controller
R_{ss}	-	steady state value of the reflux flow rate, lb/min
R($i\omega$)	-	complex function defined by Equation (3.8)
S	-	steam flow rate
SP	-	set point in the DDC loop record
SS	-	steady state value

t	-	finite integration limit introduced into Equation (3.5)
T_A	-	atmospheric temperature, deg F
T_C	-	cooling water inlet temperature, deg F
T_F	-	feed temperature, deg F
T_R	-	reflux temperature, deg F
TR	-	temperature recorder
TRC	-	temperature recorder/controller
V	-	vapour boilup rate, lb/min
VV	-	vector defined by Equation (3.15)
WW	-	matrix defined by Equation (3.15)
X_B	-	bottom product composition, weight % methanol
$X_{B_{ss}}$	-	steady state value of the bottom product composition, weight % methanol
X_D	-	overhead product composition, weight % methanol
$X_{D_{sp}}$	-	set point of the overhead product composition controller
$X_{D_{ss}}$	-	steady state value of the overhead product composition, weight % methanol
X_F	-	feed composition, weight % methanol
X_N	-	composition on plate N, weight % methanol

ii) Lower Case

$a_o \ldots a_n$	-	coefficients of numerator in polynomial transfer function defined in Equation (6.8)
$b_o \ldots b_m$	-	coefficients of denominator in polynomial transfer function defined in Equation (6.8)

e	-	exponential operator, 2.7183...
i	-	$\sqrt{-1}$
m	-	degree of denominator in polynomial transfer function defined in Equation (6.8)
n	-	degree of numerator in polynomial transfer function defined in Equation (6.8)
$r(\omega)$	-	real part of complex function defined by Equation (3.11)
s	-	Laplace transform operator, min^{-1}
$s(\omega)$	-	imaginary part of complex function defined by Equation (3.11)
t	-	time, minutes
$x(t)$	-	transient response of an input variable
$y(t)$	-	transient response of an output variable

iii) Greek

α	-	defined in Equation (3.7a)
β	-	defined by Equation (3.7b)
γ	-	defined by Equation (3.7d)
ΔB	-	variation of the bottoms flow rate about the reference steady state value, lb/min
ΔD	-	variation of the overhead flow rate about the reference steady state value, lb/min
ΔF	-	variation of the feed flow rate about the reference steady state value, lb/min
ΔR	-	variation of the reflux flow rate about the reference steady state value, lb/min

ΔS	-	variation of the steam flow rate about the reference steady state value, lb/min
ΔT_f	-	variation of the feed temperature about the reference steady state value
ΔX_B	-	variation of the bottoms product composition about the reference steady state value, weight % methanol
ΔX_D	-	variation of the overhead product composition about the reference steady state value, weight % methanol
Δt	-	sampling interval, minutes
ε_F	-	error in the process gain determinations for feed flow rate disturbances, defined by Equation (5.1a)
ε_{H_2O}	-	component material balance error of closure for water
ε_{MeOH}	-	component material balance error of closure for methanol
ε_{OV}	-	overall material balance error of closure for flow rates
ε_R	-	error in process gain determinations for reflux flow rate disturbances, defined by Equation (5.1b)
ε_S	-	error in process gain determinations for steam flow rate disturbances, defined by Equation (5.1c)
ζ	-	damping coefficient
π	-	constant, 3.1416...
ρ	-	density, gm/ml
σ	-	defined by Equation (5.1c)
$\tau_1, \tau_2, \tau_H, \tau_N$	-	time constants defined in Equations (2.1) to (2.6)
τ	-	time constant of first order lag transfer function, minutes
τ_D	-	time delay, minutes

τ_I	-	reset time, minutes
τ_Z	-	time constant of a first order lead transfer function, minute
ϕ	-	phase angle, degrees
ω	-	frequency, radians/minute
ω_{max}	-	maximum frequency, radians/minute

APPENDIX A

PHYSICAL PROPERTIES, EQUIPMENT DESCRIPTION AND CALIBRATION PROCEDURES

A.1 Introduction

This section contains a compilation of data required throughout the project, including the pertinent physical property data for water, methanol and water methanol solutions, the methods used to calibrate the column instrumentation and select the controller constants, a tabulation of the results comparing the present column operation with that obtained by Svrcek (128), examples of typical material balances obtained and listings of the loop record formats.

A.2 Physical Property Data

The pertinent physical property data for methanol, water and methanol-water mixtures were subjected to a regression analysis in order to obtain a form which could easily be used in the digital computer.

a) density of liquid water (96)

$$\text{DENSITY} = 1.005 - 2.14\times10^{-4}T - 2.51\times10^{-6}T^2 \qquad \text{(A.2)}$$

Range: $40 \leq T \leq 120$

Units: DENSITY - gm/ml
T - °C

b) density of liquid methanol (35)

$$DENSITY = 0.808 - 7.74\times10^{-4}T - 1.98\times10^{-6}T^2 \quad (A.2)$$

Range: $0 \le T \le 120$

Units: DENSITY - gm/ml
T - ^oC

c) density of methanol-water mixture (22,96)

$$DENSITY = 1.020 - 5.12\times10^{-4}T - 1.51\times10^{-6}T^2 - 1.52\times10^{-3}X - 8.11\times10^{-6}X^2 \quad (A.3)$$

Range: $40 \le T \le 120$
$0 \le X \le 100$

Units: DENSITY - gm/ml
T - ^oC
X - wt % methanol

d) saturated density of methanol-water solutions (128)

$$DENSITY = 0.960 - 1.36\times10^{-3}X - 7.46\times10^{-6}X^2 \quad (A.4)$$

Range: $0 \le X \le 100$

Units: DENSITY - gm/ml
X - wt % methanol

e) density of steam (96)

$$DENSITY = -0.113 + 1.[illegible]6\times10^{-3}T - 1.07\times10^{-5}T^2 + [illegible].50\times10^{-8}T^3 \quad (A.5)$$

Range: $170 \le T \le 320$

Units: DENSITY - lb/ft^3
T - ^oF

f) heat of vapourization of water (96)

$$HEATV = 1080 - 3.99\times10^{-1}T - 5.62\times10^{-4}T^2 \quad (A.6)$$

Range: $150 \le T \le 300$

Units: HEATV - BTU/lb
T - ^oF

g) heat of vapourization of methanol (35)

$$HEATV = 291.2 - 3.31\times10^{-1}T - 1.53\times10^{-3}T^2 \quad (A.7)$$

Range: $0 \leq T \leq 160$

Units: HEATV - cal/gm
T - ^oC

h) heat capacity of steam (43)

$$HEATC = 7.80 + 3.2\times10^{-3}T - 4.83\times10^{-3}T^2 - \quad (A.8)$$

Range: $0 \leq T \leq 3500$

Units: HEATC - cal/gm mole ^oC
T - ^oC

i) heat capacity of liquid methanol (35)

$$HEATC = 0.557 + 1.87\times10^{-3}T + 9.01\times10^{-6}T^2 \quad (A.9)$$

Range: $0 \leq T \leq 120$

Units: HEATC - cal/gm ^oC
T - ^oC

j) heat capacity of liquid water (43)

$$HEATC = 1.001 - 1.36\times10^{-4}T + 2.03\times10^{-6}T^2 \quad (A.10)$$

Range: $35 \leq T \leq 100$

Units: HEATC - cal/gm ^oC
T - ^oC

k) heat capacity of methanol vapour (43

$$HEATC = 10.26 + 1.98\times10^{-2}T - 1.44\times10^{-6}T^2 - 1.92\times10^{-9}T^3 \quad (A.11)$$

Range: $0 \leq T \leq 500$

Units: HEATC - cal/gm mole ^oC
T - ^oC

l) saturated liquid temperature of methanol-water mixtures (128)

$$TEMP = 211.29 - 1.55\,X + 1.85\times10^{-2}X^{2} - 9.33\times10^{-5}X^{3} \qquad (A.12)$$

Range: $0 \le X \le 100$
$212 \le TEMP \le 148.5$

Units: TEMP - ^{o}F
X - wt % methanol

m) Saturated vapour temperature of methanol-water mixtures (128)

$$TEMP = 211.84 - 2.25\times10^{-1}T - 1.79\times10^{-3}T^{2} - 2.33\times10^{-5}T^{3} \qquad (A.13)$$

Range: $0 \le X \le 100$
$212 \le TEMP \le 148.5$

Units: TEMP - ^{o}F
X - wt % methanol

n) Enthalpy of saturated methanol-water vapour (128)

$$ENTHV = 1143 - 5.29\,X \qquad (A.14)$$

Range: $0 \le X \le 100$

Units: ENTHV - BTU/lb
X - wt % methanol

o) Enthalpy of saturated methanol-water liquid (128)

$$ENTHL = 188.6 - 2.59\,X + 1.42\times10^{-2}X^{2} \qquad (A.15)$$

Range: $0 \le X \le 100$

Units: ENTHL - BTU/lb
X - wt % methanol

p) Enthalpy of superheated methanol-water vapour (128)

$$ENTHV = 1151.4 - 5.97\,X + 2.08\times10^{-3}X^{2} - 1.55\,T + 1.53\times10^{-2}T^{2} \qquad (A.16)$$

Range: $0 \le X \le 100$
$70 \le T \le 100$

Units: ENTHV - BTU/lb
X - wt % methanol
T - ^{o}C

q) vapour pressure of liquid methanol (61)

$$\text{Log}_{10}VP = 7.87863 - \frac{1473.11}{(T + 230)} \qquad (A.17)$$

Range: $0 \leq T \leq 120$

Units: VP - mm Hg
T - ^oC

r) vapour pressure of liquid water (61)

$$\text{Log}_{10}VP = 7.96680 - \frac{1668.2}{(T + 228)} \qquad (A.18)$$

Range: $60 \leq T \leq 100$

Units: VP - mm Hg
T - ^oC

Regression equations could not be obtained which adequately predicted the following physical properties (94):

s) equilibrium vapour composition of methanol-water system (128) as a function of equilibrium liquid composition

t) equilibrium liquid composition of methanol-water system (128) as a function of the equilibrium vapour composition

u) composition of liquid methanol-water mixtures (22,96) as a function of both density and temperature of mixture

v) equilibrium liquid composition of methanol-water system (25) as a function of temperature and pressure of the system.

A.3 Calibration and Instrument Adjustment Techniques

This section will outline the procedures that were employed to calibrate and/or adjust the control settings on the following instruments:

i) overhead product composition analyzer

ii) all flow transducers and their controllers

iii) overhead pressure transmitter

iv) both level transducers and their controllers.

A.3.1 Overhead Product Composition

The overhead product composition analyzer was calibrated using the pycnometer method. The weight and volume of a set of pycnometers were determined from replicate measurements. The volume was determined by weighing the pycnometers full of water at a known temperature. Volume corrections were determined for the level of the sample in the capillary tube. The composition of the methanol-water samples was determined by the density of the solution measured at room temperature. The density is calculated from the known volume of the pycnometer, and the weight of the pycnometer filled with sample. The sample was initially allowed to come to equilibrium with the room temperature in a sealed flask. The filled pycnometer was also allowed a few minutes to equilibrate with the room temperature before weighing it. The composition was then obtained from standard composition-density-temperature data available in the literature (22,35,96).

The calibration of the overhead product analyzer was performed on-line. The column was operated to produce the desired range of methanol concentrations. The calibration technique followed that outlined in the Foxboro manual available for the Dynalog capacitance analyzer (94). The analyzer was initially operated with the temperature compensation disconnected in order to select the desired span. After this was accomplished, the temperature compensation was reconnected and the column was operated to produce a mid-span composition measurement. The temperature of the condensate was decreased and the amount of temperature compensation was adjusted, returning the composition to its previous mid-span measurement. The temperature of the condensate was changed by recycling the cooler reflux pump output back to the reflux accumulator. This procedure was repeated until a temperature change in the reflux accumulator did not cause any deviations in the measurement of the composition.

After the analyzer span and temperature compensation have been set up as desired, the column was operated to produce compositions covering the complete recorder span. Samples of these overhead products were taken at various chart readings and the compositions determined using the pycnometers. The results are shown in Figure A.1, complete with the best fit least squares straight line through the

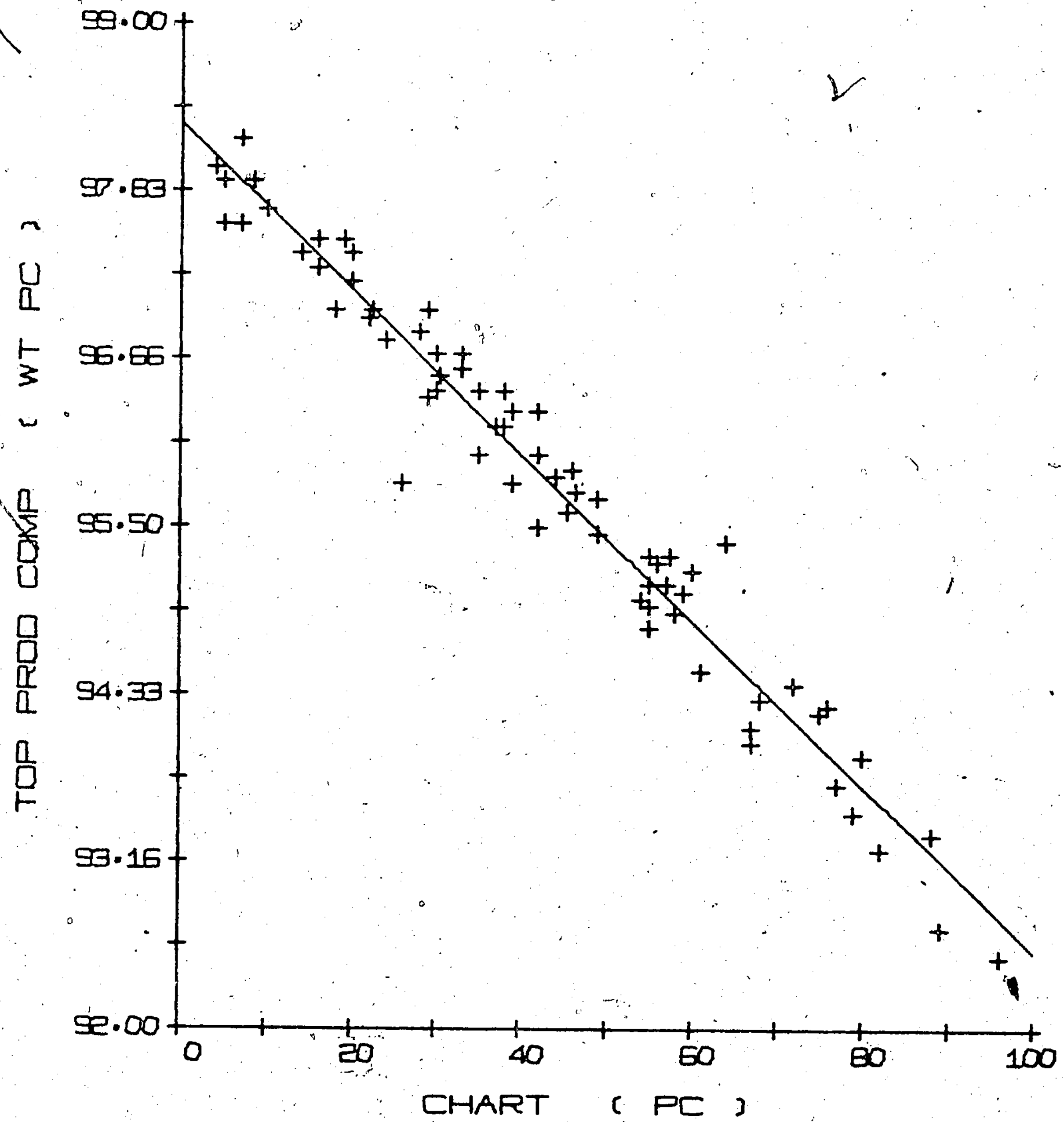

FIGURE A.1 OVERHEAD PRODUCT COMPOSITION CALIBRATION data. The parameters of the least squares line are given in Table A.1.

The calibration procedure for the bottoms product analyzer, an on-line gas chromatograph, has been outlined by Berry (9).

A.3.2 Flow Rates

The flows are measured using orifice plates with flange taps. The pressure drop is transduced to a 3-15 psig pneumatic signal using standard Foxboro d/p cells. A number of flows, feed, reflux, overhead and bottoms product, use quadrant-edge orifices (130) since the discharge coefficients are reported to be constant at the lower values of Reynolds number associated with the smaller flow rates. Sharp-edge orifices are used for the remaining flows, steam and cooling water, since the discharge coefficient is constant for the larger values of Reynolds numbers associated with these flow rates. The flow calibration data are expected to follow the characteristic orifice equation (130)

$$\frac{FLOW}{(sg)^{\frac{1}{2}}} = K\,(\Delta P)^{\frac{1}{2}} \tag{A.19}$$

where FLOW is the flow rate measured in lb/min, ΔP is the pressure drop as measured in percent of the recorder chart, sg is the specific gravity of the fluid flowing and K is the proportionality constant.

All liquid flows were initially calibrated using water. The weight of water at a measured temperature passing through the orifice creating a pressure drop indicated by the chart reading was measured over a known time interval. The calculated flow rates, specific gravities and chart positions were fit to Equation (A.19) using a least squares procedure. For all flows, the intercept calculated was not significantly different from zero, so the least squares fit was repeated forcing the straight line through the origin.

The flow rates of the methanol-water solutions, namely feed, reflux, overhead and bottom product, can now be determined from Equation (A.19) using the specific gravity of these solutions determined from Equation (A.3) and the measured composition and temperature of the stream.

The steam flow orifice was calibrated by introducing steam to the reboiler at various chart measurements and weighing the condensate collected in a previously weighted ice bath in a known time interval. These flow rates were fit to Equation (A.19), modified such that the specific gravity term was combined in the slope, using the least squares procedure. The intercept obtained for this flow equation was significantly different than zero and was thus retained.

The calibration data and best fit flow equations for the different orifices are shown in Figures A.2 to A.7. The constants of these equations are displayed in Table A.1.

A.3.3 Column Pressure

The column pressure transmitter was initially calibrated off-line to determine that the scale readings were correct. It was then installed above the vent line so that any methanol which condensed in the measurement line would return to the process.

A.3.4 Condenser and Reboiler Levels

Since a knowledge of the absolute values of the levels was not required during this project, the span of these d/p cells were adjusted to give a reasonable operating range for the levels in the column. Any levels were then measured relative to the 3-15 psig output produced by the full range d/p cells. The parameters related to these measurement transducers are summarized in Table A.1.

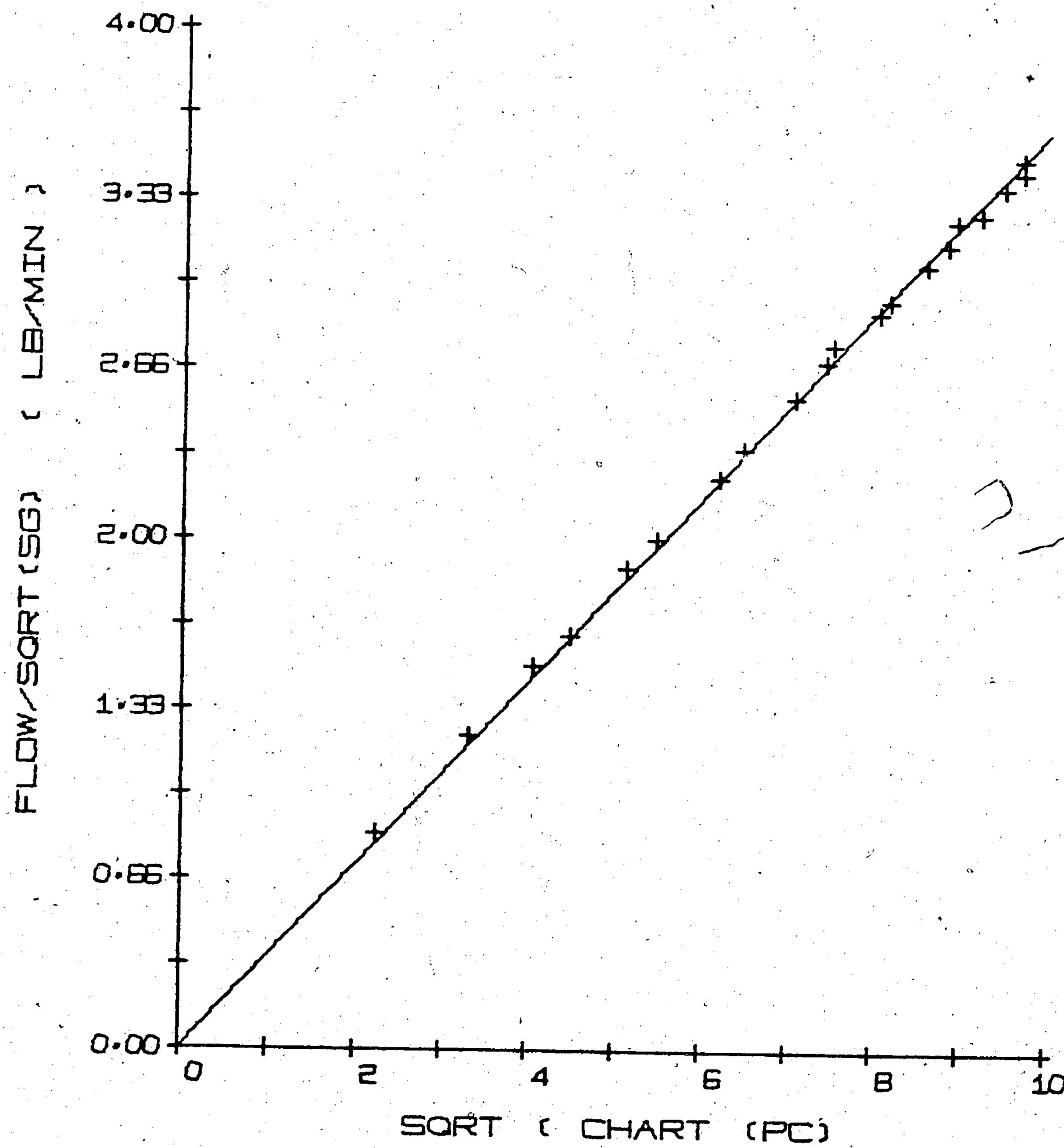

FIGURE A.2 FEED FLOW RATE CALIBRATION

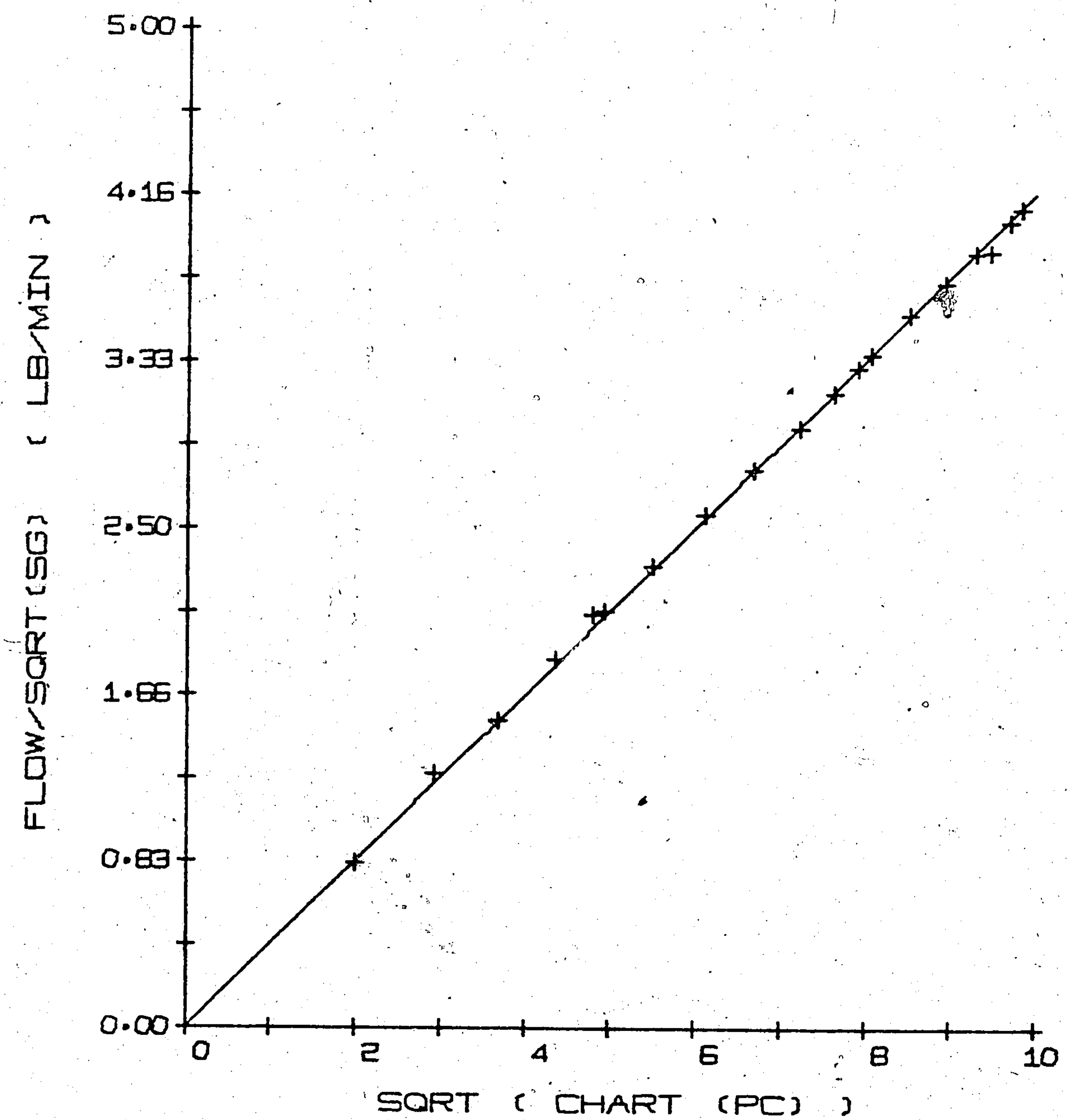

FIGURE A.3 REFLUX FLOW RATE CALIBRATION

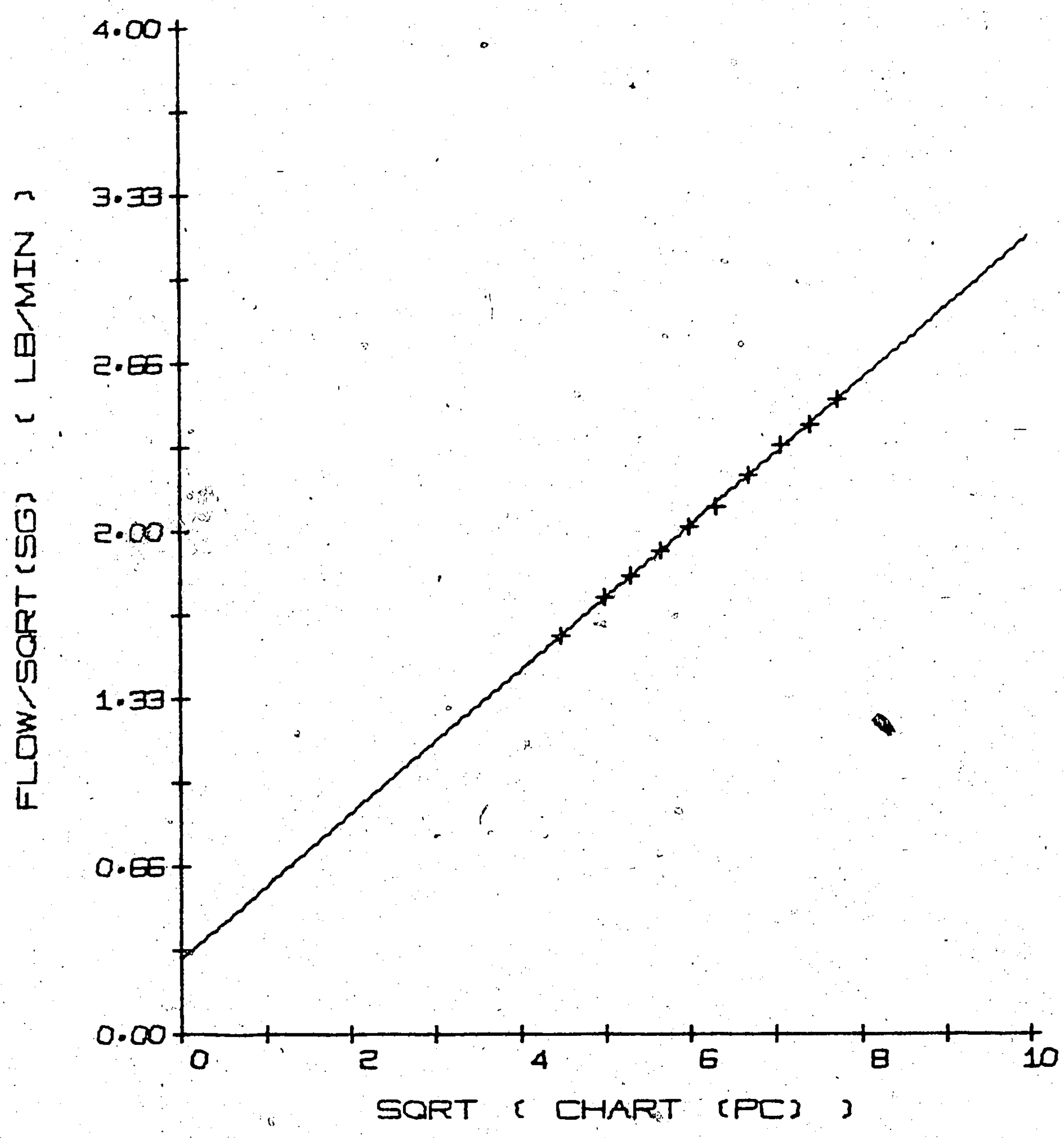

FIGURE A.4 STEAM FLOW RATE CALIBRATION

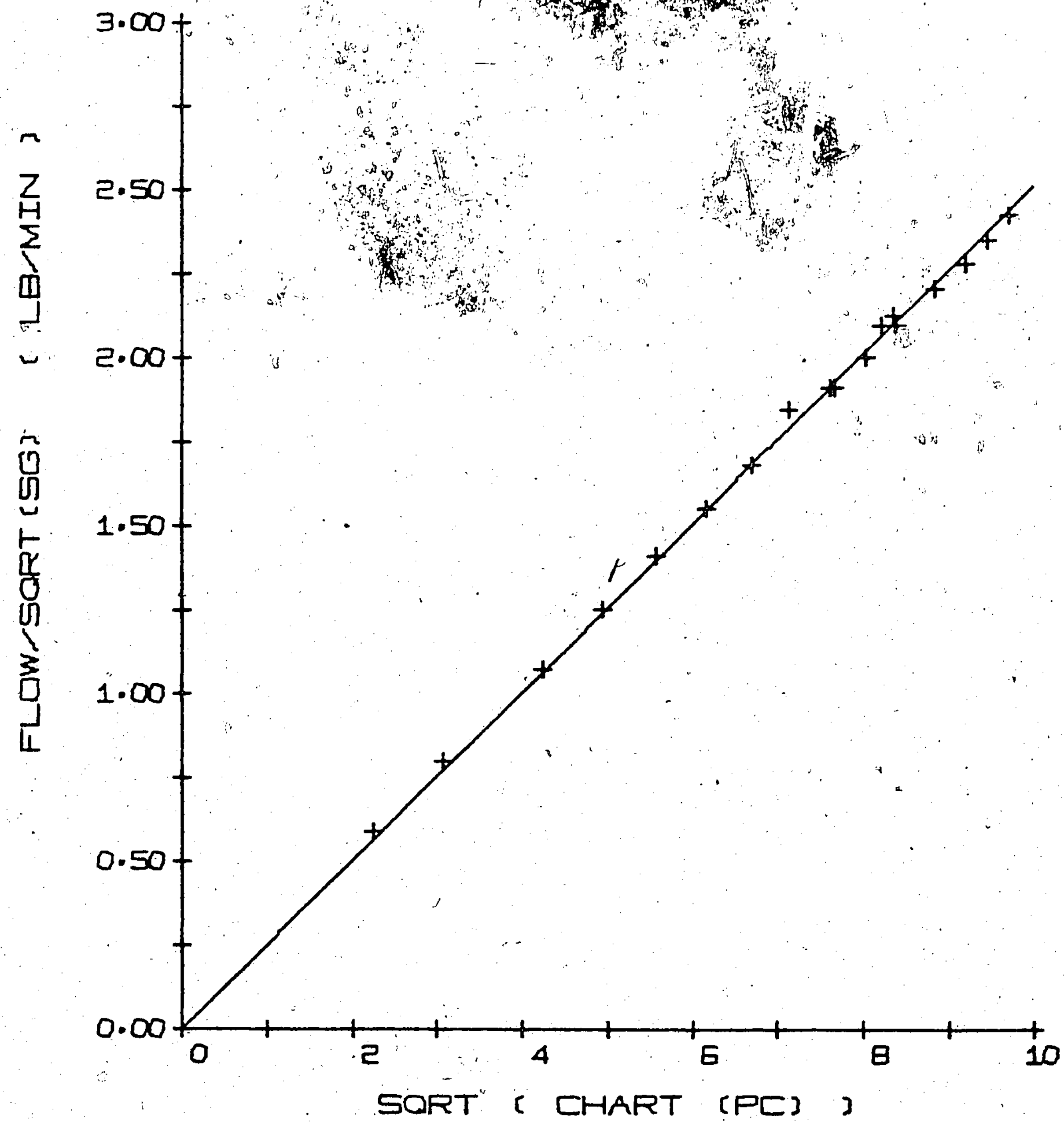

FIGURE A.5 BOTTOM PRODUCT FLOW RATE CALIBRATION

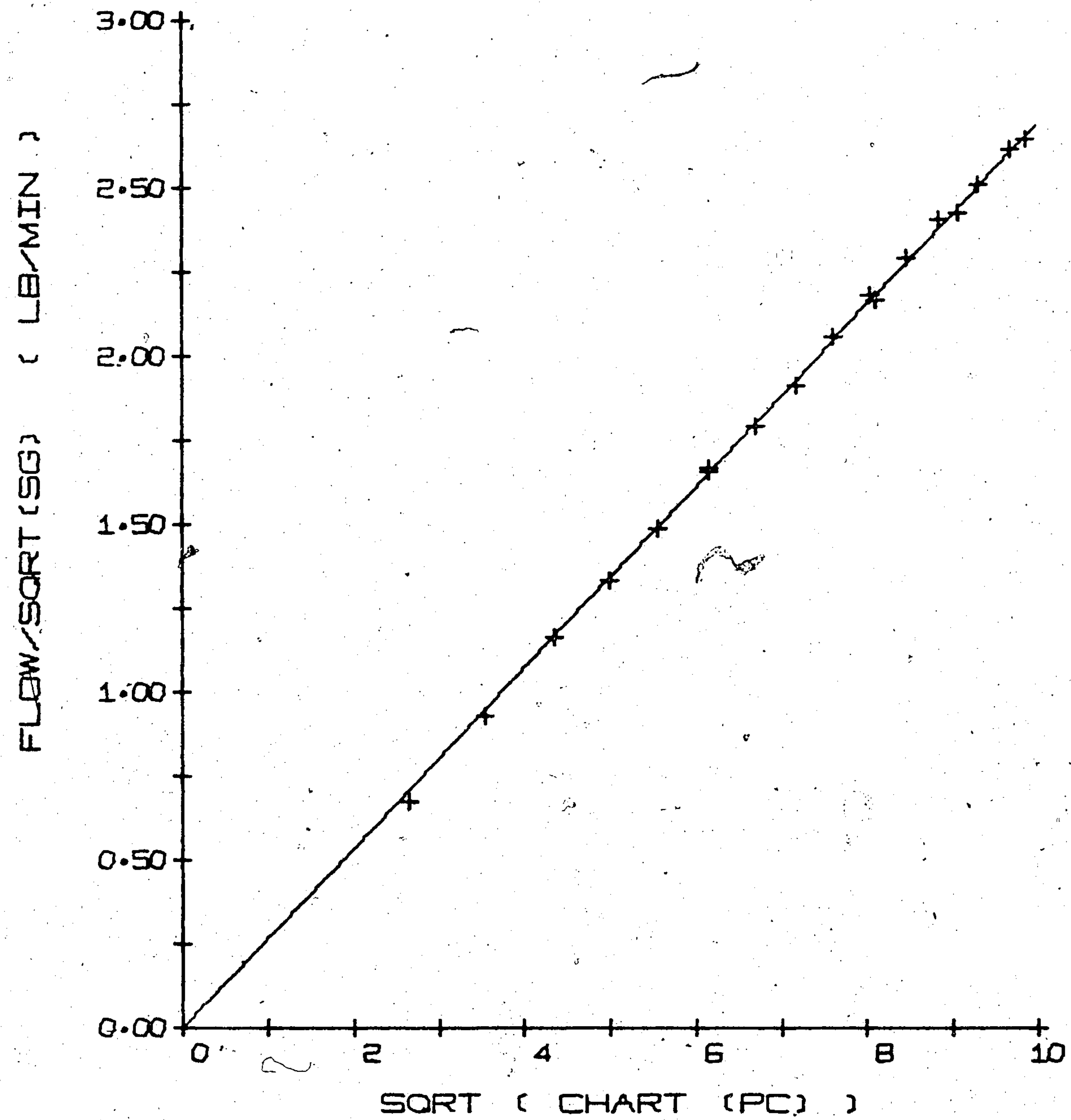

FIGURE A.6 OVERHEAD PRODUCT FLOW RATE CALIBRATION

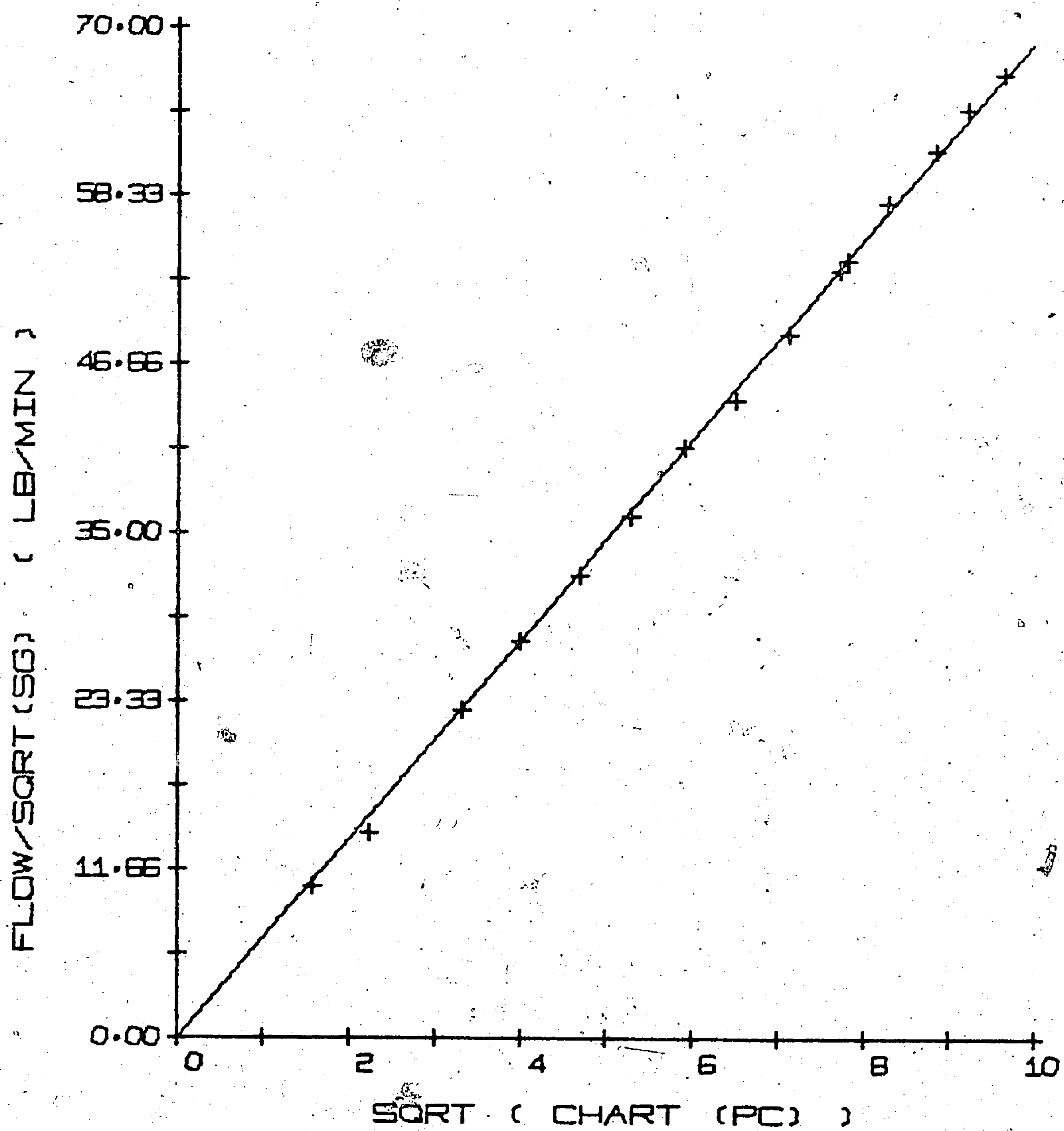

FIGURE A.7 COOLING WATER FLOW RATE CALIBRATION

TABLE A.1 SUMMARY OF MEASUREMENT DEVICE PARAMETERS

Process Variable	Orifice Size (Inch)	Orifice Type	Span Inch H_2O	A	B	Eq'n Type	Units
feed flow	0.10	quadrant	80	0.36	0.0	1	lb/min
reflux flow	0.09	quadrant	254	0.42	0.0	1	lb/min
steam flow	0.30	sharp	156	0.29	0.3	1	lb/min
overhead product flow	0.10	quadrant	51	0.27	0.0	1	lb/min
bottoms product flow	0.10	quadrant	46	0.25	0.0	1	lb/min
cooling water flow	0.35	sharp	250	6.90	0.0	1	lb/min
overhead product composition				-0.058	98.30	2	wt %
column pressure				0.20	-10.0	2	in H_2O
condenser level				0.12	3.0	2	psig
reboiler level				0.12	3.0	2	psig

Equation Types

1 $Y = A\sqrt{X} + B$ Y - measurement

2 $Y = A \cdot X + B$ X - chart per cent

A.3.5 Column Temperatures

All temperatures on the column were measured using iron-constantan thermocouples. Before installation, all thermocouples were checked against published thermocouple tables at various temperatures between 0 and 100°C. The temperatures measured by the thermocouples generally agreed to within 1°C of the calibration temperature.

A.3.6 Analog Controller Constants

Table A.2 summarizes the controller constants used for the various analog controllers. The settings for the feed, reflux and steam controllers were chosen to give the most rapid response with minimum overshoot for a set point change. The level controller settings were chosen to give

TABLE A.2 TYPICAL ANALOG CONTROLLER CONSTANTS

Controller	K_C	τ_I
feed flow	200	0.2
reflux flow	100	0.25
steam flow	175	0.15
overhead composition	-200	7.
condenser level	50	8.
reboiler level	300	10.
column pressure	100	5.
feed temperature	-150	1.8
reflux temperature	-20	2.

averaging level control, adjusting the product flows smoothly while allowing the level to vary between limits. The settings of the overhead pressure controller were chosen to maintain as tight a control as possible, without causing oscillations to occur in either the overhead pressure or cooling water flow. The reflux and feed pre-heat temperature controllers were tuned to give the most rapid possible response with minimum overshoot to the flow rate disturbances.

A.4 Comparison of Column Operation with that of Svrcek (128)

Typical results are presented in Table A.3 comparing the operation of the column attained during this study with that reported by Svrcek (128) before the column was dis-assembled and moved. Sufficiently good agreement was obtained between the runs to conclude that the present operation of the column is consistent. The small differences in the temperature and composition profiles may be attributed to the inaccuracies and insensitivity in manually adjusting the pneumatic set points of the pneumatic controllers controlling the input flow rates and the difficulty in reproducing the feed composition.

TABLE A.3 COMPARISON OF THE COLUMN OPERATING DATA WITH THOSE OBTAINED BY SVRCEK (128)

Process Variable	F1 Initial Svrcek	F1 Initial this study	F1 final Svrcek	F1 final this study	R1 Initial Svrcek	R1 Initial this study	R1 final Svrcek	R1 final this study
flow								
feed	2.20	2.24	2.51	2.54	2.51	2.54	2.51	2.54
reflux	2.25	2.29	2.25	2.29	2.09	2.12	2.24	2.14
steam	1.93	1.98	1.93	1.98	1.90	2.00	1.90	1.96
overhead product	1.16	1.14	1.26	1.26	1.33	1.37	1.24	1.24
bottoms product	1.05	1.10	1.24	1.26	1.18	1.17	1.23	1.26
feed plate	3	3	3	3	3	3	3	3
composition								
feed	50.6	48.5	50.6	48.5	51.3	51.8	51.3	51.8
overhead product	95.6	96.2	97.8	97.9	96.5	96.5	98.1	98.5
bottoms product	0.8	0.6	2.5	2.1	0.5	1.2	2.9	4.0
plate 1	2.8	3.3			7.4	6.5	16.8	13.5
plate 2	8.7	7.2			15.9	20.0	34.5	34.0
plate 3	23.5	20.8			31.7	38.0	49.4	50.0
plate 4	32.8	34.5			47.1	51.3	67.6	68.0
plate 5	52.4	52.8			63.8	68.0	79.6	80.0
plate 6	70.4	71.0			76.7	79.0	87.6	88.3
plate 7	83.1	84.0			86.6	88.0	92.8	92.5
plate 8	90.3	91.3			90.2	92.8	95.6	95.0
temperature								
reboiler	207.0	207.5	204.0	204.5	206.0	209.0	201.5	201.0
plate 1	201.0	202.5	190.0	192.0	197.0	199.0	184.0	187.0
plate 2	184.5	189.0	170.0	178.0	178.0	185.0	168.0	174.0
plate 3	180.0	179.0	167.5	167.5	173.0	172.0	165.0	166.0
plate 4	170.5	173.5	157.0	162.5	164.0	168.5	156.5	161.0
plate 5	162.0	163.0	152.5	154.5	158.0	160.5	151.5	154.5
plate 6	154.0	155.0	148.0	149.5	152.0	154.5	148.0	150.0
plate 7	149.0	150.0	145.0	147.0	148.0	150.5	145.0	148.0
plate 8	147.0	145.5	144.0	144.0	146.0	147.5	144.0	146.0

A.5 Material Balance Errors of Closure

Table A.4 gives, as an example, a typical report issued by the program BALNC documenting the operation of the distillation column. A statistical analysis of the errors of closure obtained for each run gives the following limits on the average closure obtained:

$$\varepsilon_{OV} = 0.4 \pm 1.7$$

$$\varepsilon_{MeOH} = -1.2 \pm 2.9$$

$$\varepsilon_{H_2O} = 1.8 \pm 2.1$$

This indicates that for 95% of the runs, the overall balance closed within ± 3.5% and the component balances closed within ± 5%.

A.6 Loop Record Listings

Tables A.5 to A.9 contain the listings of all DDC loop records utilized during the project.

TABLE A.4 TYPICAL MATERIAL BALANCE REPORT

STEADY STATE DATA
RUN NO FB-90D
APR 15

FEED FLOW	1.947 LB/MIN	BOTTOM PROD	1.016 LB/MIN
REFLUX FLOW	2.278 LB/MIN	TOP PROD	0.939 LB/MIN
STEAM FLOW	2.051 LB/MIN	COOL WATER	50.939 LB/MIN
FEED PLATE	4	FEED COMP	47.00 WT P C
TOP PROD	96.06 WT P C	BOTTOMS COMP	0.00 WT P C
FEED INLET	162.6 DEG F	REFLUX INLET	148.8 DEG F
STEAM TEMP	234.6 DEG F	PRESSURE	-0.1 IN H2O

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

	FLOW (LB/MIN)	COMP (WT PCT)	METHANOL (LB/MIN)	WATER (LB/MIN)
FEED	1.947	47.000	0.915	1.032
BOTTOM PRODUCT	1.016	0.000	0.000	1.016
TOP PRODUCT	0.939	96.068	0.902	0.036
CLOSURE ERROR-PC	0.4		-1.3	2.0

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

	ENTHALPY IN (BTU/MIN)	ENTHALPY OUT (BTU/MIN)
COOLING WATER	2923.7	4491.3
REFLUX	246.7	238.6
TOP PRODUCT		98.4
FEED	278.5	
STEAM	2457.4	482.5
BOTTOM PRODUCT		216.2
TOTAL	5906.5	5527.3
HEAT LOSS		379.1

TABLE A.5 DATA ACQUISITION LOOP RECORDS

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				

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 F[illegible]W

01	6010	205F+04104	3340	84[illegible]0+05032	000[illegible]	0000	0000	7FFF
02	7FFF	2110	7FFF	0000	21[illegible]			

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	641E	404C+04106	0000	9100-01310+09879	0000	0000	7FFF		
02	7FFF	2990	7FFF	0000	2990	7FFF	0000	2990	116A	0000
03	00C0	0000	0000	0000	001A	003A	0000	0000	0400

DATA ACQUISITION LOOP RECORDS

ID = 0176 BOTTOM COMPOSITION

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 = 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

01	0010	207F	0179	0000	9180+20000+00000	0000	0000	7FFF
02	7FFF	2110+10000+00000	2110					

ID = 0179 FEED COMPOSITION(DUMMY)

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 = 0184 PLATE 4 TEMPERATURE

01	7010	21BF+00134	0000	A280+20000	0000	0000	0000	7FFF
02	7FFF	2110	7FFF	0000	2110			

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			

TABLE A.6 RING BUFFER LOOP RECORDS FOR PULSE TESTS

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 BUFFER LOOP RECORDS FOR PULSE TESTS

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

ID = 0228 PLATE 8 TEMPERATURE

01 0615 4208 0188 0000 0000 0200

TABLE A.8 FEEDFORWARD CONTROLLER LOOP RECORDS

ID = 0250 TIME DELAY RING BUFFER

01	0614	510C	0170	0000	0000	0000

ID = 0251 DYNAMIC COMPENSATION

01	241E	5274	0170	3B10	8400+07214	0000	0000	0000	7FFF
02	7FFF	2110	7FFF-32767	2110	7FFF-32767	2110	0252	0000	
03	10A0	0000	0000	0000	0000	0080	0000	FF00	0000

ID = 0252 GAIN COMPENSATION

01	0A1B	5364	0251	0000	8400+07214	0000	0000	0000	7FFF
02	7FFF	2110	7FFF-32767	2110	7FFF-32767	2110	0259	0000	
03	10E0	0000	0000	0000	0000	0733			

ID = 0259 ADDITION OF REFLUX CHANGES

01	021B	5364	0252	0000	8400+08364	0000	0000	0000	7FFF
02	7FFF	2110	7FFF-32767	2110	7FFF-32767	2110	0260	0000	
03	13A0	0000	0000	0000	0000	0080			

ID = 0260 REFLUX SET POINT ADDITION

01	021B	5364	0259	0000	8400+08364	0000	0000	0000	7FFF
01	7FFF	2110	7FFF-32767	2110	7FFF-32767	2110	0262	0000	
03	13A3	0000	0000	0000	0000	0008			

ID = 0261 FEEDBACK COMPENSATION

01	641E	534C+04106	0000	9100-01310+09879	0000	0000	7FFF		
02	7FFF	2550	7FFF	0000	2550	7FFF	0000	2550	0262	0000
03	00A2	0000	0000	0000	0020	0060	0000	0000	0000	0000

FEEDFORWARD CONTROLLER LOOP RECORDS

ID = 0262 OUTPUT TO COS

01	021B	5364	0260	0000	8400+08364	0000	0000	0000	7FFF	
02	7FFF	2110	7FFF-32767	2110	7FFF-32767	2110	116A	0000		
03	1380	0000	0000	0000	0000	0000				

TABLE A.7 GAS CHROMATOGRAPH LOOP RECORDS

ID = 0201 GAS CHROMATOGRAPH

01	4010	0004	004D	0000	8F00	7FFF	0000	0000	0000	0000
02	0000	0000	7FFF	0000	0000					

ID = 0202 GAS CHROMATOGRAPH

01	067F	0000	0201	0000	0000	0000

TABLE A.9 RING BUFFER LOOP RECORDS FOR CONTROL STUDIES

ID = 0210 REFLUX (DISPLAY)

01	0665	6508	0171	0000	0000	0200

ID = 0211 FEED (DISPLAY)

01	0665	6508	0170	0000	0000	0200

ID = 0212 TOP COMPOSITION (DISPLAY)

01	0665	6508	0261	0000	0000	0200

ID = 0213 STEAM FLOW

01	0611	5408	0172	0000	0000	0200

ID = 0214 BOTTOM PRODUCT FLOW

01	061B	5408	0173	0000	0000	0200

ID = 0215 TOP PRODUCT FLOW

01	061B	5408	0174	0000	0000	0200

ID = 0216 BOTTOM COMPOSITION

01	0611	5408	0176	0000	0000	0200

ID = 0217 FEED FLOW

01	0610	5408	0170	0000	0000	0200

RING BUFFER LOOP RECORDS FOR CONTROL STUDIES

ID = 0218 REFLUX FLOW

01 0610 5408 0171 0000 0000 0200

ID = 0219 TOP COMPOSITION

01 0610 5408 0261 0000 0000 0200

01 0610 5408 0261 0000 0000 0200

APPENDIX B

SCHEMATIC DIAGRAM OF DISTILLATION COLUMN

This section contains the detailed schematic diagrams of the various sections of the distillation column system.

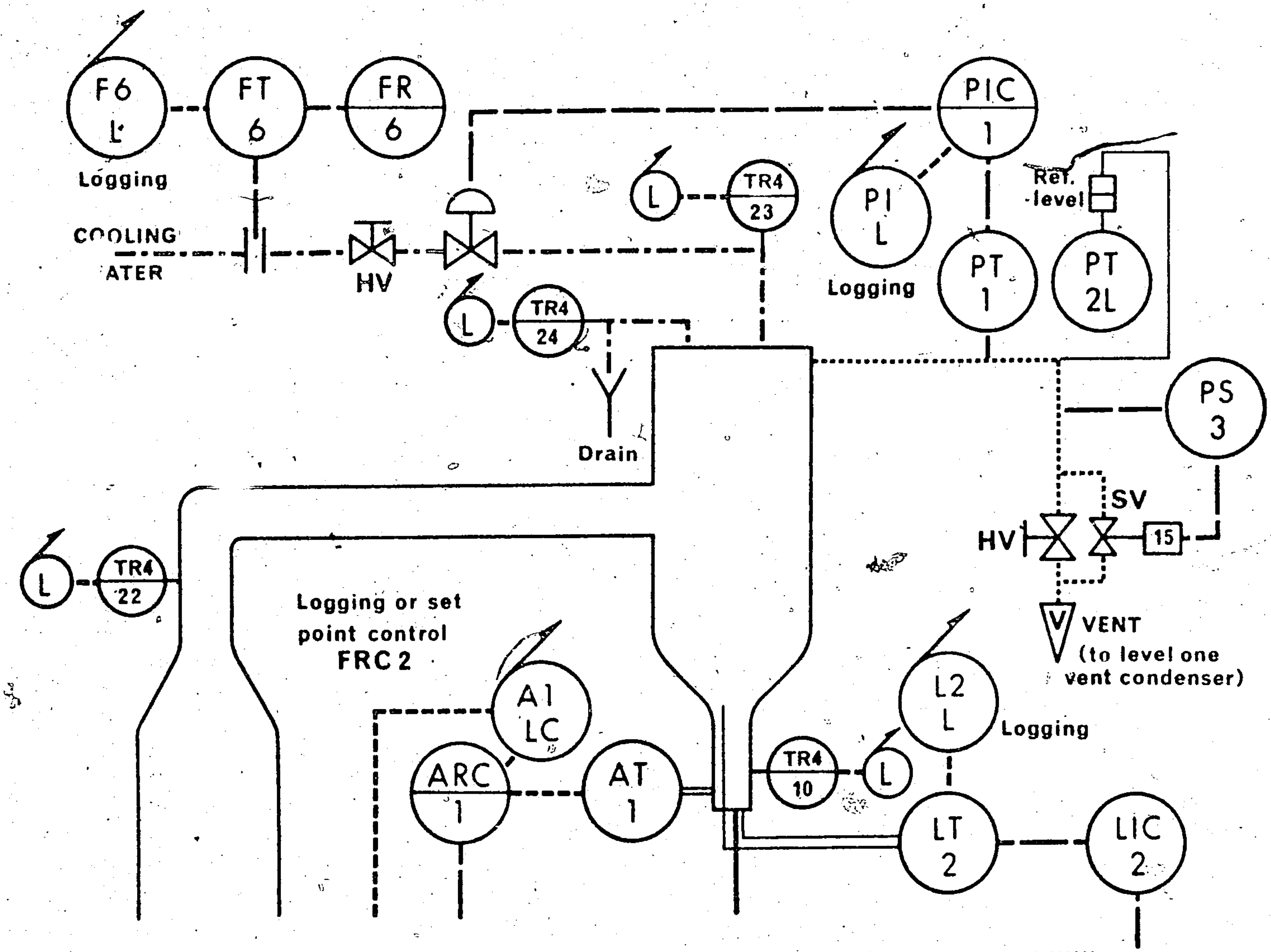

FIGURE B.1 SCHEMATIC DIAGRAM OF DISTILLATION COLUMN: OVERHEAD SECTION

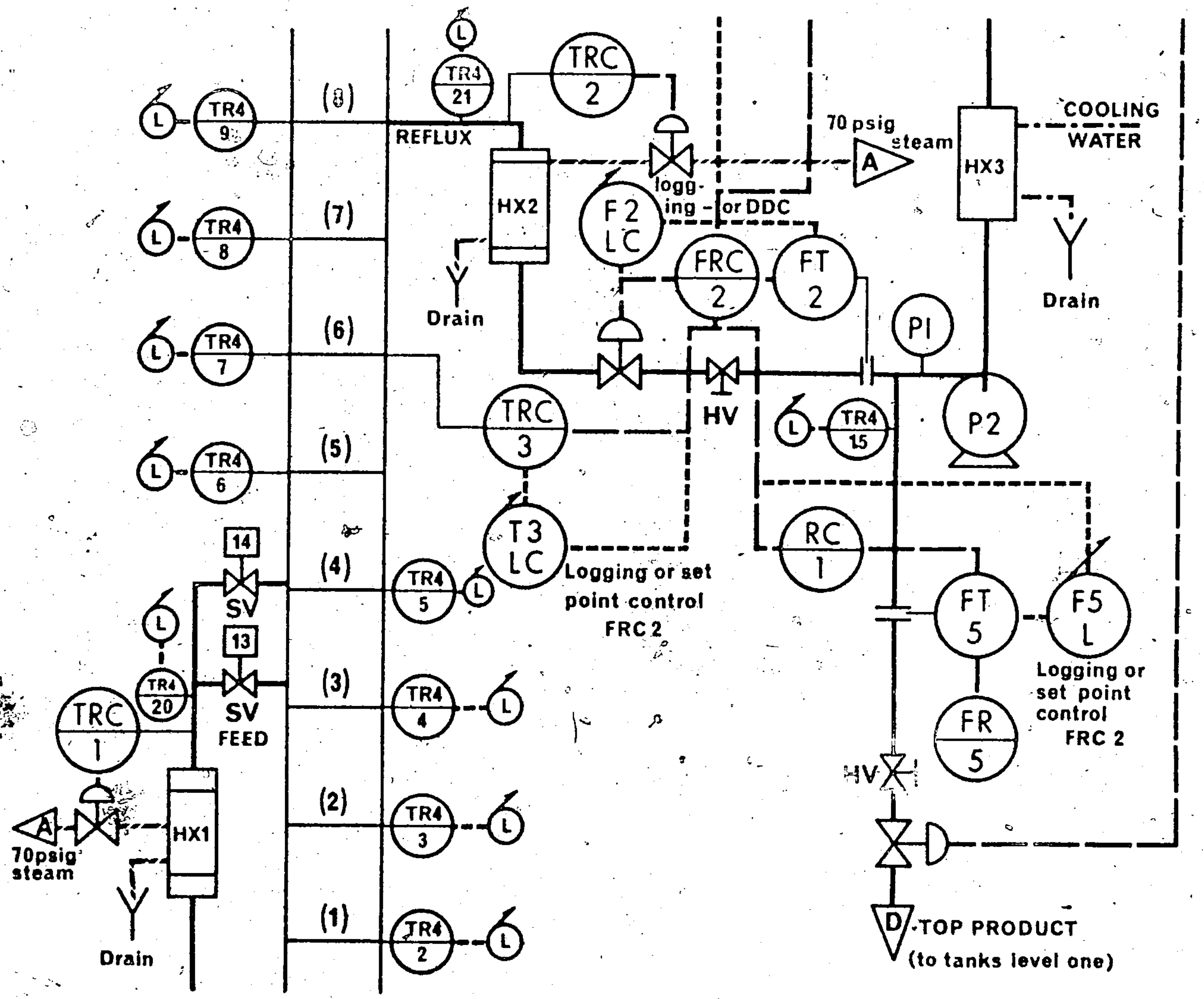

FIGURE B.2 SCHEMATIC DIAGRAM OF DISTILLATION COLUMN: MID SECTION

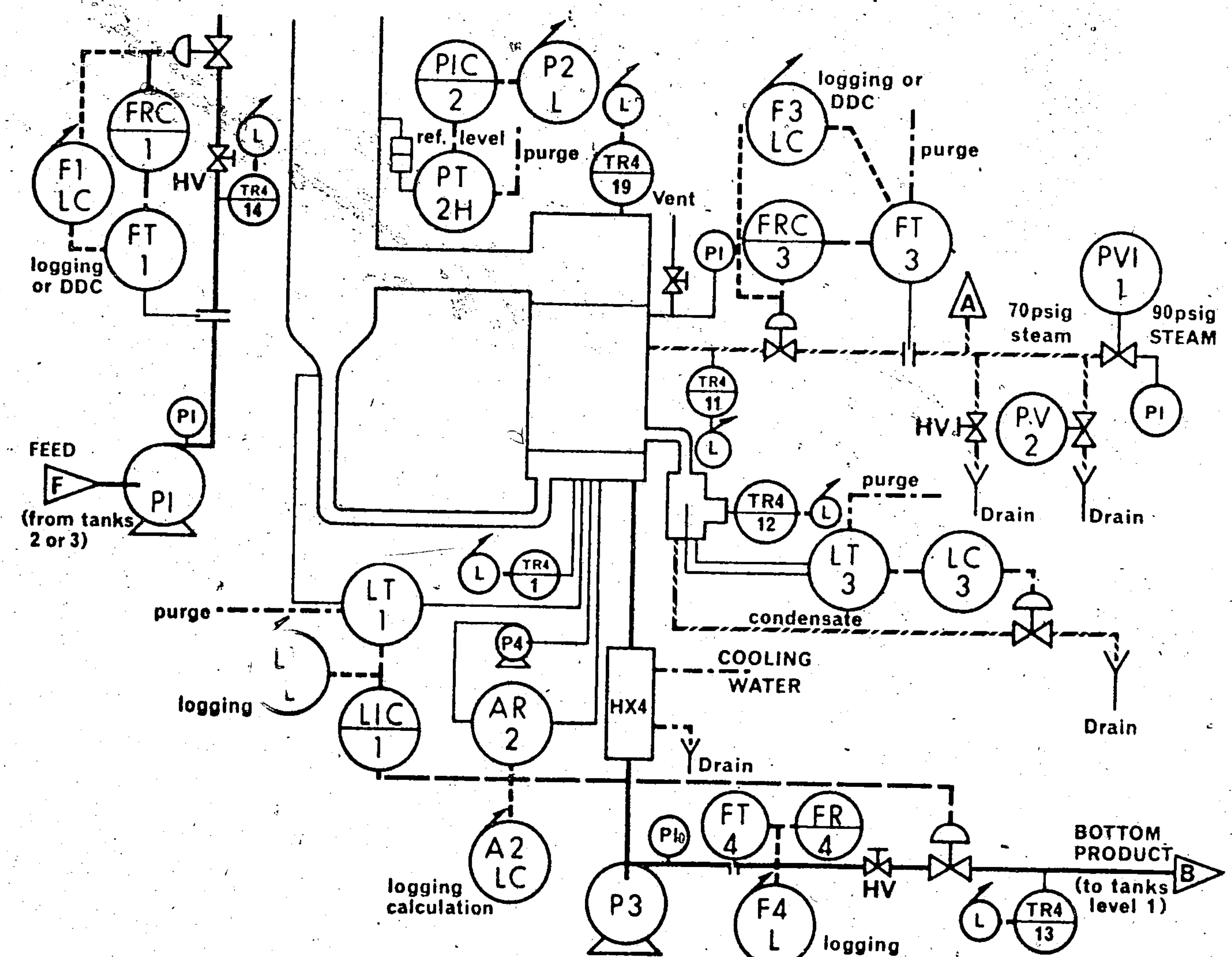

FIGURE B.3 SCHEMATIC DIAGRAM OF DISTILLATION COLUMN: BOTTOM SECTION

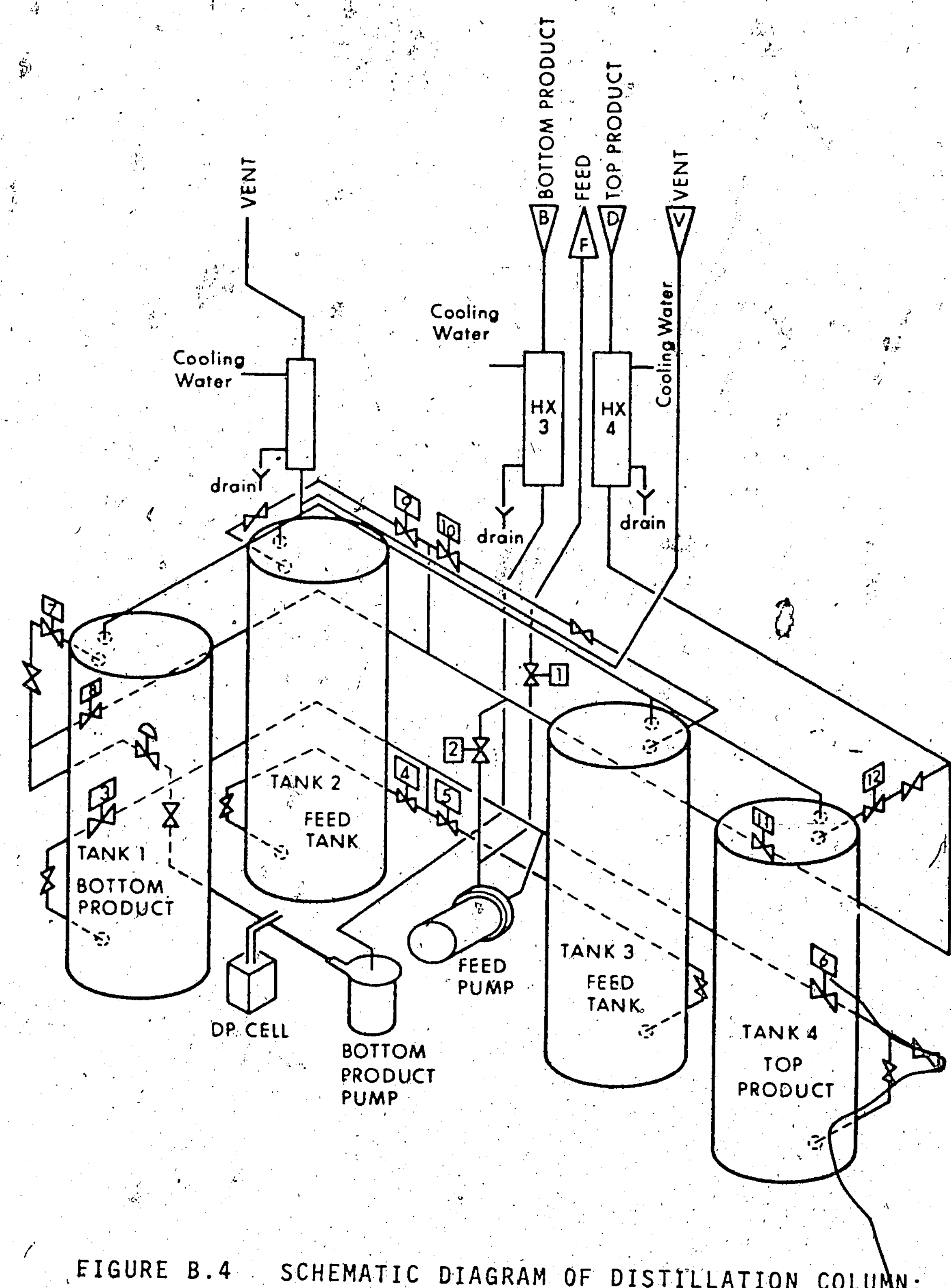

FIGURE B.4 SCHEMATIC DIAGRAM OF DISTILLATION COLUMN: STORAGE TANKS

AR/n	COMPOSITION RECORDER (Chromatograph)	PIC/n	PRESSURE INDICATOR/CONTROLLER
ARC/n	COMPOSITION RECORDER/CONTROLLER (capacitance)	PRV	PRESSURE REGULATING VALVE
AT/n	COMPOSITION TRANSDUCER (Capacitance probe)	Ps	PRESSURE SWITCH
FR/n	FLOW RECORDER	PT/n	PRESSURE TRANSDUCER
FRC/n	FLOW RECORDER/CONTROLLER	PT/nH PT/nL	PRESSURE TRANSDUCER High Low
FT/n	FLOW TRANSDUCER (d/p cell)	TRn/m	MULTIPOINT TEMPERATURE RECORDER
HV	HAND VALVE	TRC/n	TEMPERATURE RECORDER/CONTROLLER
LIC/n	LEVEL INDICATOR/CONTROLLER	SV	SOLENOID VALVE
LT/n	LEVEL TRANSDUCER	nn/nL	LOGGING (VARIABLE nn-n) BY COMPUTER
P-n	PUMP	nn/nLC	LOGGING AND CONTROL (VARIABLE nn-n) BY COMPUTER

PROCESS LINES

PNEUMATIC INSTRUMENT LINES	METHANOL/WATER PROCESS LINES
PROCESS INSTRUMENT LINES	VENT
COOLING WATER PROCESS LINES	70 PSIG STREAM
	DATA TRANSMISSION LINES TO COMPUTER

FIGURE B.5 SCHEMATIC DIAGRAM OF DISTILLATION COLUMN: LEGEND

APPENDIX C

PROCESS GAIN DATA

This section contains a summary of the experimental data used to determine the process gain portion of the open loop column model.

TABLE C.1 COLUMN OPEN LOOP OPERATING DATA: VARIATION WITH FEED FLOW RATE

Test No		F	R	S	X_F	X_D	X_B	D	B	ε_{OV}	ε_{MeOH}	ε_{H_2O}
F-30S	SS	2.46	1.93	1.73	47.	96.11	0.50	1.18	1.29	-0.1	-3.6	3.0
F-31S		2.99	1.93	1.73	47.	97.27	7.00	1.26	1.71	-0.7	-5.7	3.8
F-32S		2.84	1.93	1.72	47.	97.25	4.50	1.24	1.57	-1.2	-5.4	2.5
F-33S		2.67	1.93	1.73	47.	97.07	2.20	1.22	1.43	-1.0	-4.4	2.0
F-34S		2.20	1.94	1.73	47.	93.86	0.00	1.06	1.11	-1.2	-4.7	1.9
F-35S		2.07	1.95	1.73	47.	92.50	0.00	1.00	1.04	-1.5	-5.7	2.2
F-36S		2.35	1.93	1.73	47.	95.57	0.00	1.13	1.20	-0.8	-3.1	1.1
F-37S		2.55	1.94	1.73	47.	96.72	0.90	1.19	1.33	-1.0	-3.7	1.4
F-40S	SS	2.46	1.94	1.72	46.	96.19	0.40	1.14	1.30	-0.6	-2.3	0.7
F-41S		2.40	1.95	1.71	46.	95.75	0.00	1.11	1.27	-0.6	-3.1	1.5
F-42S		2.52	1.95	1.71	46.	96.50	0.75	1.16	1.33	-1.2	-2.5	0.1
F-43S		2.34	1.95	1.71	46.	95.32	0.00	1.09	1.21	-1.5	-3.3	0.0
F-45S		2.52	1.94	1.71	46.	96.61	0.70	1.15	1.34	-1.0	-3.1	0.6

TABLE C.2 COLUMN OPEN LOOP OPERATING DATA: VARIATION WITH REFLUX FLOW RATE

Test No		F	R	S	X_F	X_D	X_B	D	B	ε_{OV}	ε_{MeOH}	ε_{H_2O}
R-30S	SS	2.47	1.96	1.72	47.	95.60	0.50	1.16	1.26	-2.2	-3.8	-0.8
R-31S		2.46	2.27	1.72	47.	98.13	5.30	1.03	1.38	-1.9	-6.1	1.7
R-32S		2.46	2.09	1.71	47.	97.55	2.60	1.10	1.29	-2.7	-4.1	-1.5
R-34S		2.47	1.85	1.71	47.	94.80	0.00	1.20	1.22	-2.0	-2.6	-1.5
R-35S		2.46	1.74	1.71	47.	93.03	[illegible].00	1.20	1.20	-2.4	-3.7	-1.1
R-40S	SS	2.45	1.94	1.72	47.	95.98	0.50	1.14	1.29	-0.9	-5.8	3.4
R-41S		2.45	2.02	1.72	47.	97.19	1.00	1.09	1.33	-1.1	-6.7	3.7
R-42S		2.46	2.14	1.73	47.	97.87	2.60	1.05	1.36	-2.3	-7.5	2.1
R-43S		2.46	1.88	1.72	47.	94.50	0.00	1.16	1.21	-3.7	-4.8	-2.6
R-50S	SS	2.46	1.95	1.71	46.	96.18	0.40	1.09	1.36	-0.6	-6.9	4.7
R-51S		2.46	2.03	1.71	46.	97.02	0.80	1.07	1.39	-1.0	-6.7	3.7
R-52S		2.46	2.10	1.71	46.	97.74	1.60	1.07	1.40	-0.8	-5.7	3.3
R-54S		2.46	1.86	1.71	46.	94.75	0.00	1.14	1.33	0.4	-3.8	4.0
R-44S		2.46	1.74	1.71	46.	92.49	0.00	1.15	1.29	0.9	-5.3	2.8
R-46S		2.46	1.91	1.71	46.	95.52	0.00	1.15	1.29	-0.6	-2.2	0.7
R-60S	SS	2.46	1.95	1.71	48.	96.40	0.90	1.22	1.26	0.6	1.7	-0.3
R-61S		2.46	2.03	1.71	48.	97.03	1.60	1.17	1.32	1.2	-0.8	3.2
R-62S		2.46	2.24	1.71	48.	98.28	6.00	1.08	1.42	1.5	-0.1	3.1
R-63S		2.46	2.18	1.71	48.	98.01	4.50	1.07	1.35	-1.5	-5.5	1.9
R-64S		2.47	2.09	1.71	48.	97.71	2.90	1.17	1.33	1.1	0.5	1.5
R-65S		2.47	1.98	1.70	48.	96.94	1.35	1.21	1.30	1.8	1.6	1.9
R-66S		2.46	1.89	1.71	48.	95.53	0.50	1.24	1.25	0.6	1.3	0.0
R-67S		2.46	1.78	1.70	48.	94.03	0.40	1.24	1.23	0.5	0.2	0.7
R-68S		2.46	1.72	1.71	48.	92.66	0.00	1.28	1.20	0.8	1.4	0.2
R-69S		2.46	1.94	1.70	48.	96.10	0.65	1.20	1.27	0.2	-0.3	0.8

TABLE C.3 COLUMN OPEN LOOP OPERATING DATA: VARIATION WITH STEAM FLOW RATE

Test No		F	R	S	X_F	X_D	X_B	D	B	ε_{OV}	ε_{MeOH}	ε_{H_2O}
S-30S	SS	2.46	1.94	1.72	47.	96.37	0.60	1.18	1.23	-2.0	-1.9	-2.1
S-31S		2.46	1.93	1.44	47.	98.11	15.50	0.86	1.59	-0.5	[illegible]	4.7
S-32S		2.46	1.94	1.58	47.	97.76	6.90	1.05	1.41	-0.1	[illegible]	3.0
S-33S		2.46	1.94	1.66	47.	97.30	2.70	1.10	1.33	-1.1	[illegible]	2.0
S-34S		2.47	1.95	1.85	47.	93.26	0.00	1.23	1.19	-1.7	-1.6	-1.8
S-35S		2.46	1.95	1.79	47.	95.07	0.00	1.19	1.25	-0.1	-2.1	-0.5
S-36S		2.46	1.95	1.87	47.	92.41	0.00	1.22	1.20	-1.6	-3.0	-0.3
S-40S	SS	2.46	1.96	1.72	46.	96.25	0.40	1.12	1.29	-1.8	-5.5	1.3
S-41S		2.46	1.96	1.67	46.	97.29	1.40	1.08	1.35	-1.3	-6.5	3.1
S-43		2.46	1.96	1.69	46.	97.13	1.00	1.11	1.33	[illegible]	-3.8	1.3
S-44S		2.46	1.97	1.76	46.	95.06	0.00	1.15	1.28	-1.4	-3.5	0.2
S-45S		2.46	1.97	1.84	46.	92.64	0.00	1.19	1.22	-2.1	-2.7	-1.6
S-50S	SS	2.46	1.95	1.71	45.	96.20	0.40	1.12	1.33	-0.2	-3.9	2.8
S-51S		2.46	1.94	1.65	45.	97.32	1.75	1.06	1.36	-1.6	-6.7	2.6
S-52S		2.46	1.95	1.77	45.	94.23	0.00	1.10	1.30	-2.2	[illegible]	1.9
S-53S		2.46	1.95	1.74	45.	95.40	0.00	1.12	1.32	-0.6	[illegible]	2.3
S-60S	SS	2.46	1.97	1.71	48.	96.63	0.90	1.17	1.32	1.4	-3.3	5.8
S-61S		2.46	1.97	1.85	48.	92.20	0.00	1.24	1.20	-0.9	-4.0	2.0
S-62S		2.46	1.97	1.76	48.	95.01	0.40	1.24	1.25	1.1	-0.5	2.8
S-63S		2.46	1.97	1.71	48.	95.97	0.55	1.20	1.26	0.4	-1.8	2.[illegible]

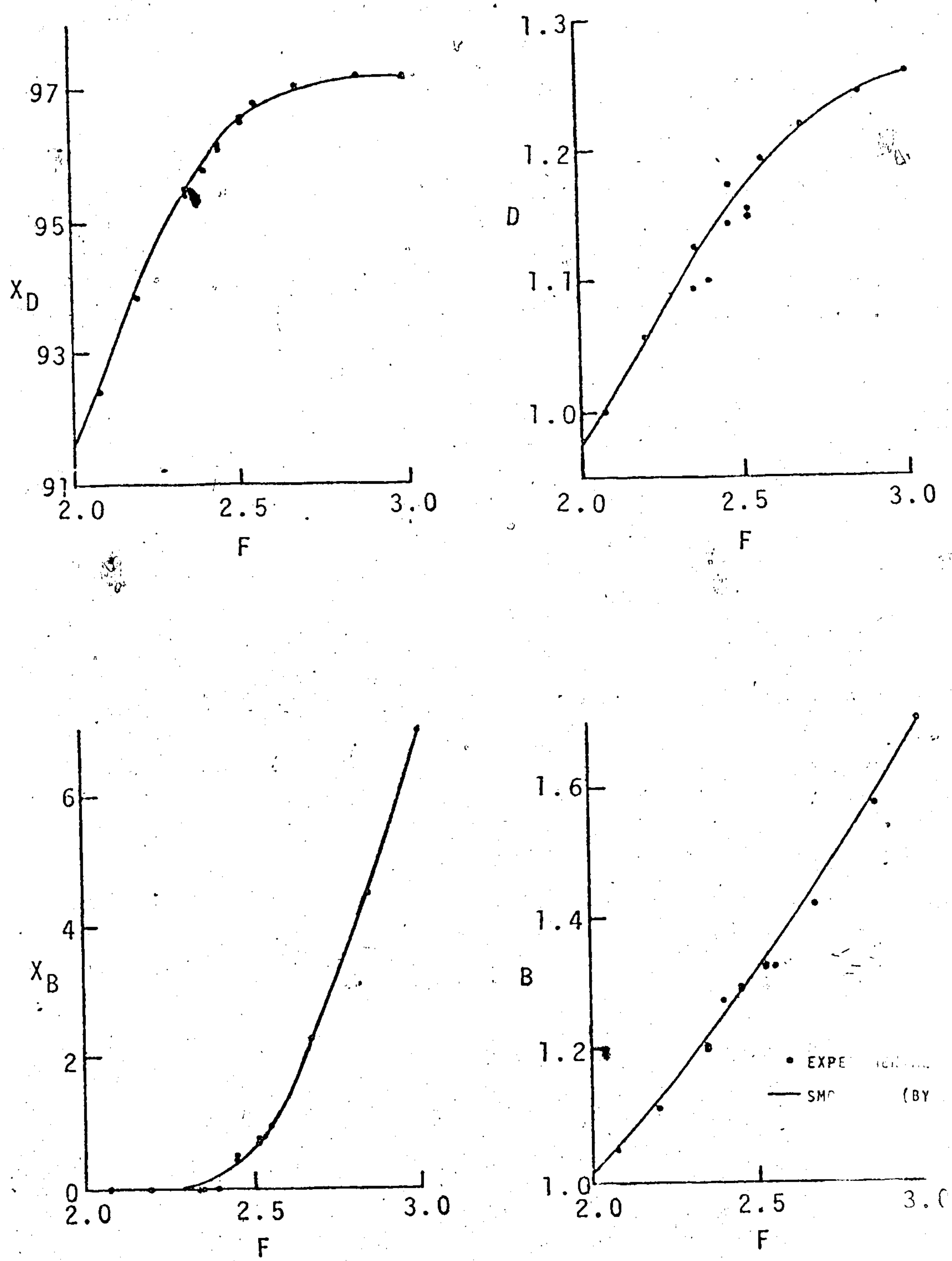

FIGURE C.1 COLUMN OPEN LOOP OPERATING DATA: VARIATION WITH FEED FLOW RATE

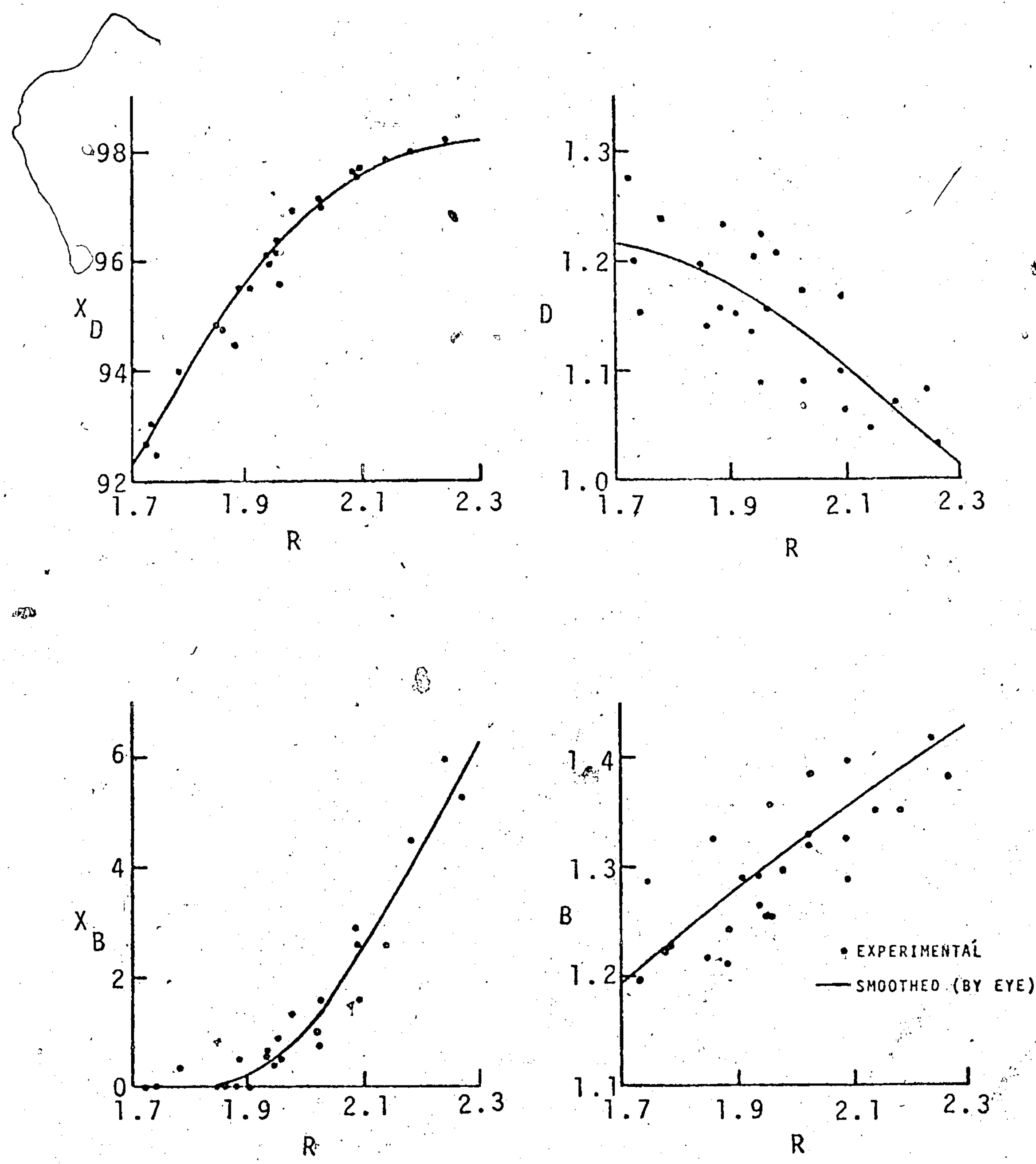

FIGURE C.2 COLUMN OPEN LOOP OPERATING DATA: VARIATION WITH REFLUX FLOW RATE

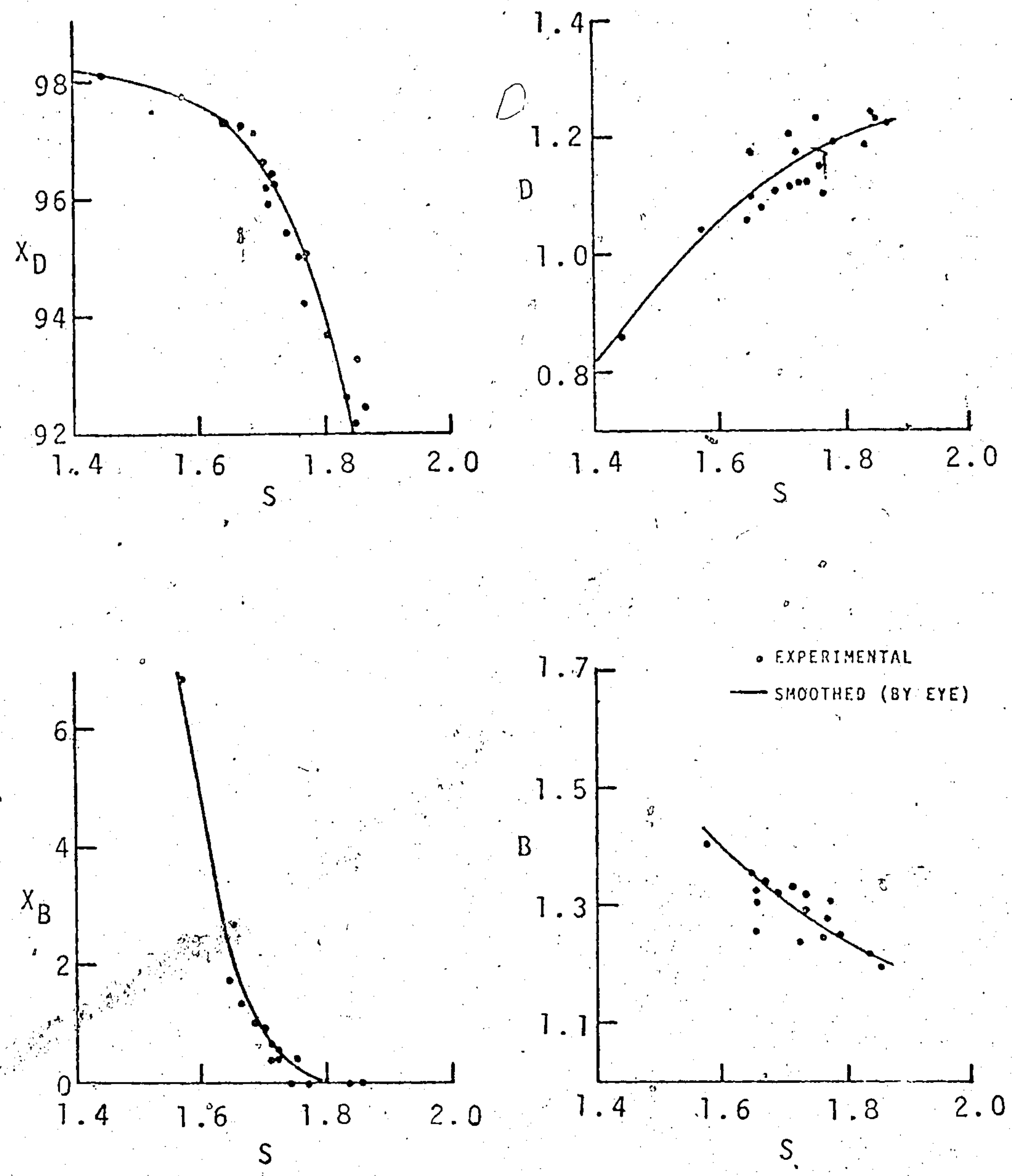

FIGURE C.3 COLUMN OPEN LOOP OPERATING DATA: VARIATION WITH STEAM FLOW RATE

TABLE C.4 SMOOTHED OPEN LOOP OPERATING DATA: VARIATION WITH FEED FLOW RATE

F	X_D	X_B	D	B
2.01	91.80	0.00	0.96	1.01
2.16	93.40	0.00	1.04	1.10
2.31	95.00	0.00	1.10	1.19
2.46	96.20	0.45	1.15	1.29
2.61	96.90	1.40	1.19	1.40
2.76	97.15	3.10	1.22	1.51
2.91	97.30	5.40	1.25	1.64

TABLE C.5 SMOOTHED OPEN LOOP OPERATING DATA: VARIATION WITH REFLUX FLOW RATE

R	X_D	X_B	D	B
1.74	92.80	0.00	1.21	1.21
1.81	94.00	0.00	1.20	1.25
1.89	95.25	0.10	1.18	1.28
1.96	96.30	0.55	1.16	1.31
2.04	97.15	1.45	1.13	1.34
2.11	97.70	2.55	1.09	1.36
2.19	98.05	4.00	1.06	1.39

TABLE C.6 SMOOTHED OPEN LOOP OPERATING DATA: VARIATION WITH STEAM FLOW RATE

S	X_D	X_B	D	B
1.49	98.00	12.60	0.92	1.53
1.56	97.80	7.60	1.01	1.43
1.64	97.45	2.80	1.08	1.36
1.71	96.25	0.60	1.14	1.30
1.79	94.30	0.00	1.19	1.25
1.86	91.80	0.00	1.21	1.21
1.94	89.90	0.00	1.23	1.17

TABLE G.7 CALCULATION OF PROCESS GAINS FROM COLUMN OPERATING DATA: VARIATION WITH MAGNITUDE OF FEED FLOW RATE DISTURBANCES

Test	ΔF	ΔX_D	ΔX_B	ΔD	ΔB	K_{PF}	K'_{PF}	K_{DF}	K_{BF}	ε_F
F-31S	0.53	1.16	6.50	0.09	0.42	2.2	12.2	0.17	0.79	-0.04
F-32S	0.38	1.14	4.00	0.07	0.28	3.0	10.5	0.18	0.73	-0.09
F-33S	0.21	1.04	1.70	0.04	0.14	5.0	8.1	0.21	0.68	-0.11
F-34S	-0.26	-2.25	-0.50	-0.11	-0.17	8.7	1.9	0.43	0.67	0.10
F-35S	-0.39	-3.58	-0.50	-0.17	-0.24	9.5	1.3	0.44	0.63	0.08
F-36S	-0.12	-0.54	-0.50	0.02	0.05	4.7	4.3	0.23	0.51	-0.26
F-37S	0.09	0.61	0.40	-0.04	-0.09	6.8	4.4	0.37	0.79	0.16
F-41S	-0.06	-0.44	-0.40	-0.03	-0.03	7.2	6.4	0.46	0.53	-0.01
F-42S	0.06	0.31	0.35	0.02	0.03	4.9	5.5	0.28	0.46	-0.26
F-43S	-0.12	-0.77	-0.40	-0.05	-0.09	6.6	3.5	0.44	0.74	0.18
F-45S	0.06	0.42	0.30	0.01	0.04	6.8	4.8	0.16	0.65	-0.19

TABLE C.8 CALCULATION OF PROCESS GAINS FROM COLUMN OPERATING DATA: VARIATION WITH MAGNITUDE OF REFLUX FLOW RATE DISTURBANCES

Test	ΔR	ΔX_D	ΔX_B	ΔD	ΔB	K_{PR}	K'_{PR}	K_{DR}	K_{BR}	ϵ_R
R-31S	0.31	2.53	4.80	-0.13	0.13	8.2	15.5	-0.40	0.41	~~0.00~~
R-32S	0.13	1.95	2.10	-0.06	0.03	14.9	16.0	-0.43	0.26	-0.17
R-34S	-0.11	-0.80	-0.50	0.04	-0.04	7.2	4.5	-0.37	0.35	-0.02
R-35S	-0.22	-2.57	-0.50	0.05	-0.06	11.5	2.2	-0.20	0.27	0.07
R-41S	0.09	1.21	0.50	-0.05	0.04	14.2	5.9	-0.53	0.45	-0.08
R-42S	0.20	1.89	2.10	-0.09	0.06	9.3	10.5	-0.43	0.29	-0.14
R-43S	-0.06	-1.48	-0.50	0.02	-0.04	27.0	9.1	-0.36	0.73	0.37
R-51S	0.08	0.84	0.40	-0.02	0.03	11.1	5.3	-0.29	0.43	0.14
R-52S	0.14	1.56	1.20	-0.03	0.04	10.8	8.3	-0.17	0.27	0.10
R-54S	-0.09	-1.43	-0.40	0.05	-0.03	15.7	4.4	-0.56	0.35	-0.21
R-44S	-0.21	-3.69	-0.40	0.06	-0.07	17.2	1.9	-0.31	0.35	0.04
R-46S	-0.04	-0.66	-0.40	0.06	-0.07	15.7	9.5	-1.45	1.59	0.14
R-61S	0.73	0.63	0.70	~~-0.05~~	0.08	8.6	9.6	-0.69	1.03	0.34
R-62S	0.29	1.88	5.10	-0.14	0.15	6.6	17.3	-0.49	0.53	0.04
R-63S	0.23	1.61	3.60	-0.15	0.10	7.0	15.6	-0.65	0.42	-0.23
R-64S	0.13	1.31	2.00	-0.06	0.07	9.9	15.0	-0.43	0.54	0.11
R-65S	0.03	0.54	0.45	-0.02	0.04	20.8	17.3	-0.65	1.61	0.96
R-66S	-0.07	-0.87	-0.40	0.01	-0.01	13.2	6.1	-0.17	0.17	0.00
R-67S	-0.17	-2.37	-0.52	0.02	-0.03	14.0	3.1	-0.09	0.15	0.06
R-68S	-0.23	-3.74	-0.90	0.06	-0.05	16.2	3.9	-0.25	0.23	-0.02
R-69S	-0.03	-0.30	-0.25	0.02	-0.01	11.1	9.6	-0.80	0.35	-0.45

TABLE C.9 CALCULATION OF PROCESS GAINS FROM COLUMN OPERATING DATA: VARIATION WITH MAGNITUDE OF STEAM FLOW RATE DISTURBANCES

Test	ΔS	ΔX_D	ΔX_B	ΔD	ΔB	K_{PS}	K'_{PS}	K_{DS}	K_{BS}	ε_S
S-31S	-0.28	1.74	14.90	-0.32	0.35	- 6.3	-53.4	1.14	-1.28	-0.14
S-32S	-0.15	1.39	6.30	-0.13	0.18	- 9.5	-43.0	0.89	-1.20	-0.31
S-33S	-0.07	0.93	2.10	-0.07	0.09	-14.1	-31.8	1.11	-1.43	-0.32
S-34S	0.13	-3.11	- 0.60	0.06	-0.04	-24.2	- 4.6	0.43	-0.31	0.12
S-35S	0.06	-1.30	- 0.60	0.02	-0.02	-20.6	- 9.5	0.25	-0.32	-0.07
S-36S	0.15	-3.96	- 0.60	0.05	-0.04	-27.3	- 4.2	0.31	-0.24	0.07
S-41S	-0.06	1.04	0.98	-0.05	0.06	-18.9	-17.8	0.85	-0.98	-0.13
S-42S	-0.03	0.88	0.60	-0.02	0.04	-24.2	-18.1	0.52	-1.04	-0.52
S-44S	0.04	-1.19	-0.40	0.03	-0.02	-29.7	-10.0	0.68	-0.40	0.28
S-45S	0.11	-2.56	-0.40	0.06	-0.07	-22.9	- 3.6	0.52	-0.59	-0.07
S-51S	-0.07	1.12	1.35	-0.06	0.03	-17.2	-20.8	0.89	-0.40	0.49
S-52S	0.06	-1.97	-0.40	0.02	-0.03	-34.6	- 7.0	0.32	-0.55	-0.23
S-53S	0.03	-0.80	-0.40	0.01	-0.02	-29.6	-14.8	0.18	-0.59	-0.41
S-61S	0.14	-4.43	-0.90	0.06	-0.12	-31.2	- 6.3	0.44	-0.81	-0.37
S-62S	0.05	-1.62	0.50	0.06	-0.07	-31.2	- 9.6	1.19	-1.29	-0.10
S-63S	0.01	-0.66	0.35	0.02	-0.06	-94.5	-50.0	4.14	-8.00	-4.14

TABLE C.10 CALCULATION OF PROCESS GAINS FROM SMOOTHED OPERATING DATA: VARIATION WITH MAGNITUDE OF FEED FLOW RATE DISTURBANCES

ΔF	ΔX_D	ΔX_B	ΔD	ΔB	K_{PF}	K'_{PF}	K_{DF}	K_{BF}	ε_F
-0.45	-4.40	-0.45	-0.18	-0.29	9.8	1.0	0.39	0.63	0.02
-0.30	-2.80	-0.45	-0.12	-0.20	9.3	1.5	0.37	0.67	0.04
-0.15	-1.20	-0.45	-0.05	-0.11	8.0	3.0	0.33	0.70	0.03
0.00	0.00	0.00	0.00	0.00	6.0	5.7	0.30	0.70	0.00
0.15	0.70	0.95	0.40	0.10	4.7	6.3	0.27	0.70	-0.03
0.30	0.95	2.65	0.75	0.22	3.2	8.8	0.25	0.72	-0.03
0.45	1.10	4.95	0.10	0.35	2.4	11.0	0.23	0.77	0.00

TABLE C.11 CALCULATION OF PROCESS GAINS FROM SMOOTHED OPERATING DATA: VARIATION WITH MAGNITUDE OF REFLUX FLOW RATE DISTURBANCES

ΔR	ΔX_D	ΔX_B	ΔD	ΔB	K_{PR}	K'_{PR}	K_{DR}	K_{BR}	ϵ_R
-0.23	-3.50	-0.55	0.05	-0.10	15.6	2.4	-0.24	0.42	0.18
-0.15	-2.30	-0.55	0.04	-0.06	15.3	3.7	-0.27	0.40	0.13
-0.08	-1.05	-0.45	0.03	-0.03	14.0	6.0	-0.37	0.40	0.03
0.00	0.00	0.00	0.00	0.00	12.5	8.3	-0.39	0.40	0.01
0.08	0.85	0.90	-0.03	0.03	11.3	12.0	-0.40	0.40	0.00
0.15	1.40	2.00	-0.06	0.06	9.3	13.3	-0.41	0.39	-0.02
0.23	1.75	3.45	-0.10	0.08	7.8	15.3	-0.42	0.37	-0.05

TABLE C.12 CALCULATION OF PROCESS GAINS FROM SMOOTHED OPERATING DATA: VARIATION WITH MAGNITUDE OF STEAM FLOW RATE DISTURBANCES

ΔS	ΔX_D	ΔX_B	ΔD	ΔB	K_{PS}	K'_{PS}	K_{DS}	K_{BS}	ε_S
-0.23	1.75	12.00	-0.22	10.23	- 7.8	-53.3	0.98	-1.00	-0.02
-0.15	1.55	7.00	-0.14	0.13	-10.3	-46.7	0.90	-0.87	0.03
-0.08	1.15	2.20	-0.60	0.06	-15.3	-29.3	0.80	-0.80	0.00
0.00	0.00	0.00	0.00	0.00	-22.0	-15.2	0.65	-0.78	-0.11
0.08	-1.95	-0.60	0.50	-0.06	-26.0	- 8.0	0.60	-0.73	-0.13
0.15	-4.45	-0.60	0.73	-0.10	-29.7	- 4.0	0.49	-0.63	-0.24
0.23	-6.35	-0.60	0.85	-0.13	-28.2	- 2.7	0.38	-0.58	-0.20

APPENDIX D

OPEN LOOP DYNAMIC MODEL DATA

D.1 Pulse Test Results

This section contains a summary of the experimental data used to obtain the dynamic parameters of a first order plus time delay model used to describe the open loop response of the distillation column. The frequency response diagrams displayed in Figures D.2, D.3, D.5, D.6, D.8, D.9, D.11, D.12 and D.13 were prepared from the measured transient response of the column to rectangular pulses in the various input variables using the Pulse Testing Analysis Program (PTAP) (135). The dynamic parameters of a first order plus time delay model were determined using both the 'Levy' procedure and the properties exhibited by first order and time delay transfer functions. These parameters were compared to those calculated from the transient response data using the Rosenbrock procedure (110).

The parameters of higher order polynomial transfer function models have been calculated using the 'Levy' procedure. The parameters are presented in Tables D.11, D.12 and D.13.

TABLE D.1 STEADY STATE OPERATING DATA FOR PULSE TESTS

		Inputs				Outputs					Material Balance		
Test	*	F	R	S	X_F	X_D	X_B	D	B		ε_{OV}	ε_{MeOH}	ε_{H_2O}
F-20	I	2.45	1.94	2.02	46.6	95.68	0.51	1.20	1.27		0.4	0.9	0.0
	F	2.45	1.94	2.02	46.6	95.63	0.52	1.19	1.29		1.2	0.3	2.0
F-21	I	2.46	1.95	2.00	46.6	95.87	0.75	1.13	1.29		-1.6	-4.7	0.9
	F	2.45	1.94	2.00	46.6	96.02	0.50	1.18	1.25		-0.7	-0.1	-1.2
F-22	I	2.45	1.95	2.00	46.6	96.02	0.50	1.18	1.27		0.0	-0.4	0.2
	F	2.45	1.94	2.00	46.6	95.92	0.50	1.18	1.29		0.5	-0.6	1.6
F-23	I	2.45	1.94	1.99	46.3	96.15	0.60	1.17	1.32		1.6	-0.3	3.4
	F	2.46	1.94	1.99	46.3	96.06	0.60	1.21	1.30		2.1	2.3	2.0
F-24	I	2.45	1.93	2.03	47.6	95.77	0.40	1.17	1.18		-3.8	-3.1	-4.5
	F	2.45	1.93	2.00	47.6	95.71	0.50	1.17	1.22		-2.4	-3.4	-1.4
R-20	I	2.45	1.91	2.02	46.6	95.49	0.50	1.18	1.28		0.6	-0.5	1.6
	F	2.45	1.92	2.02	46.6	95.75	0.50	1.18	1.28		0.2	0.1	0.4
R-21	I	2.45	1.94	2.00	46.6	95.99	0.50	1.18	1.27		-0.3	-0.7	0.0
	F	2.46	1.95	2.01	46.6	95.98	0.45	1.19	1.27		0.1	0.0	0.2
R-22	I	2.46	1.94	2.01	46.6	95.99	0.55	1.17	1.30		0.5	-1.2	2.1
	F	2.46	1.95	2.00	46.6	96.12	0.50	1.17	1.29		0.1	-0.9	1.2
R-23	I	2.46	1.96	1.99	45.5	95.56	0.50	1.12	1.30		-1.4	-3.5	0.3
	F	2.47	1.96	1.99	45.5	95.43	0.50	1.12	1.24		-4.4	-4.1	-4.6
R-24	I	2.45	1.93	2.02	47.5	95.71	0.50	1.17	1.22		-2.4	-3.4	-1.4
	F	2.45	1.93	2.02	47.5	95.77	0.40	1.17	1.24		-1.9	-3.9	-0.1

* I - initial steady state before pulse entered
F - final steady state after pulse entered

STEADY STATE OPERATING DATA FOR PULSE TESTS

Test*		Inputs				Outputs				Material Balance		
		F	R	S	X_F	X_D	X_B	D	B	ε_{OV}	ε_{MeOH}	ε_{H_2O}
S-20	I	2.45	1.92	2.02	46.6	95.66	0.50	1.20	1.30	1.9	0.9	2.7
	F	2.46	1.91	2.02	46.6	95.61	0.50	1.22	1.30	2.4	2.3	2.4
S-21	I	2.45	1.94	2.00	46.6	96.01	0.60	1.17	1.28	0.2	-0.7	1.1
	F	2.46	1.94	2.00	46.6	95.90	0.50	1.19	1.28	0.6	0.3	0.8
S-22	I	2.46	1.95	1.99	46.2	95.95	0.70	1.20	1.28	0.9	2.1	-0.1
	F	2.46	1.95	2.00	46.2	96.02	0.80	1.18	1.28	0.2	-0.2	0.5
S-23	I	2.47	1.96	2.00	45.5	95.13	0.50	1.13	1.24	-3.8	-3.6	-3.8
	F	2.46	1.95	1.99	45.5	95.56	0.55	1.12	1.32	-0.8	-4.1	1.8
S-24	I	2.47	1.96	2.01	47.0	95.90	0.60	1.15	1.31	-0.5	-4.3	2.9
	F	2.46	1.96	2.01	47.0	96.02	0.50	1.16	1.27	-1.1	-3.1	0 5
S-25	I	2.46	1.94	2.01	47.0	96.04	0.60	1.17	1.26	-1.5	-2.5	-[illegible].7
	F	2.46	1.95	2.02	47.0	96.26	0.60	1.17	1.27	-0.8	-1.7	-0.0
S-26	I	2.46	1.93	2.02	47.5	95.74	0.40	1.16	1.21	-3.4	-4.0	-2.7
	F	2.46	1.93	2.02	47.5	95.83	0.45	1.16	1.24	-2.2	-3.8	-0.7
S-27	I	2.45	1.93	2.02	47.5	95.83	0.45	1.16	1.23	-2.4	-[illegible].9	-1.0
	F	2.46	1.94	2.02	47.5	95.73	0.45	1.16	1.28	-0.6	-4.3	2.7

I - initial steady state before pulse entered
F - final steady state after pulse entered

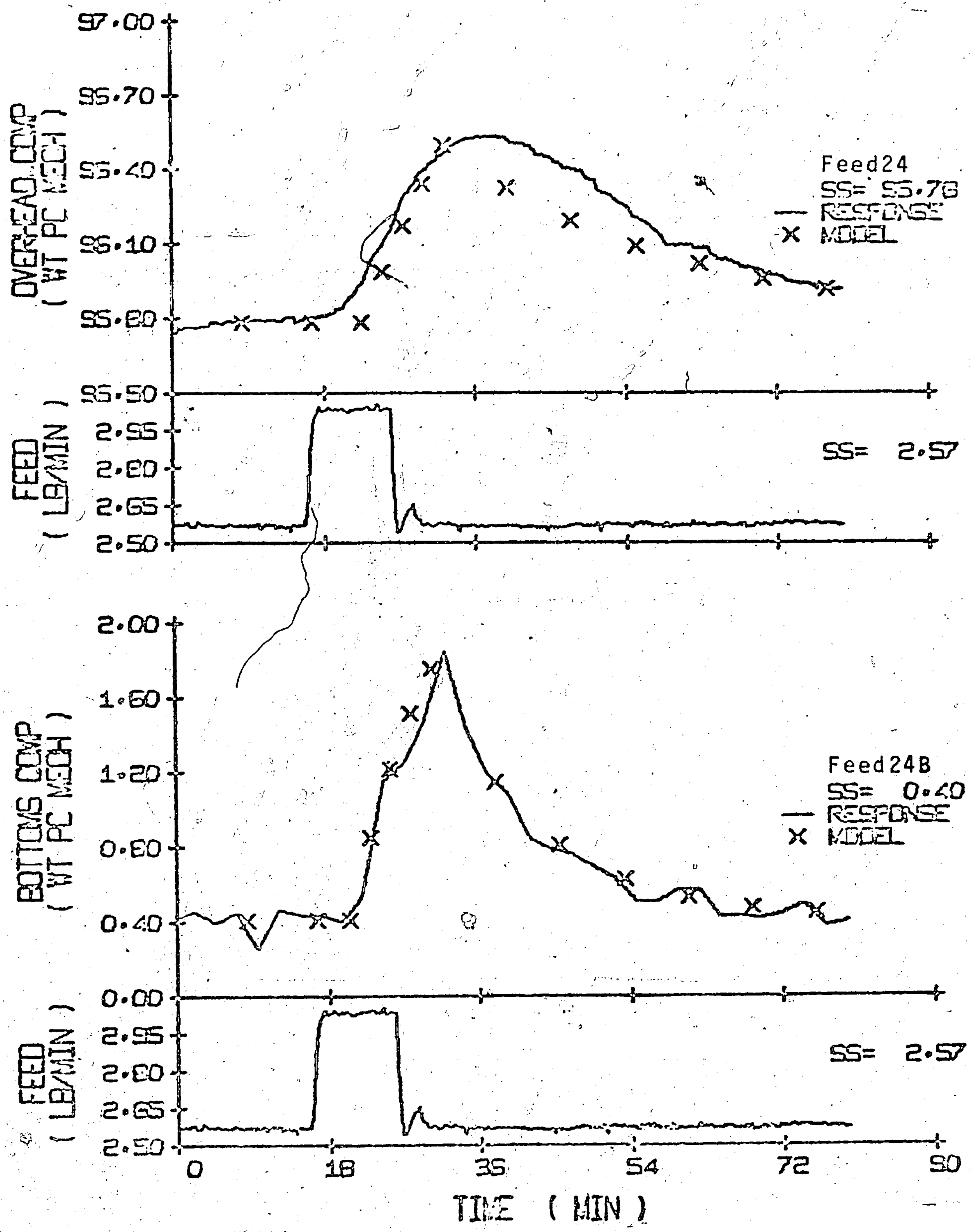

FIGURE D.1 TRANSIENT RESPONSE OF THE OVERHEAD AND BOTTOMS PRODUCT COMPOSITION TO A POSITIVE PULSE IN THE FEED FLOW RATE

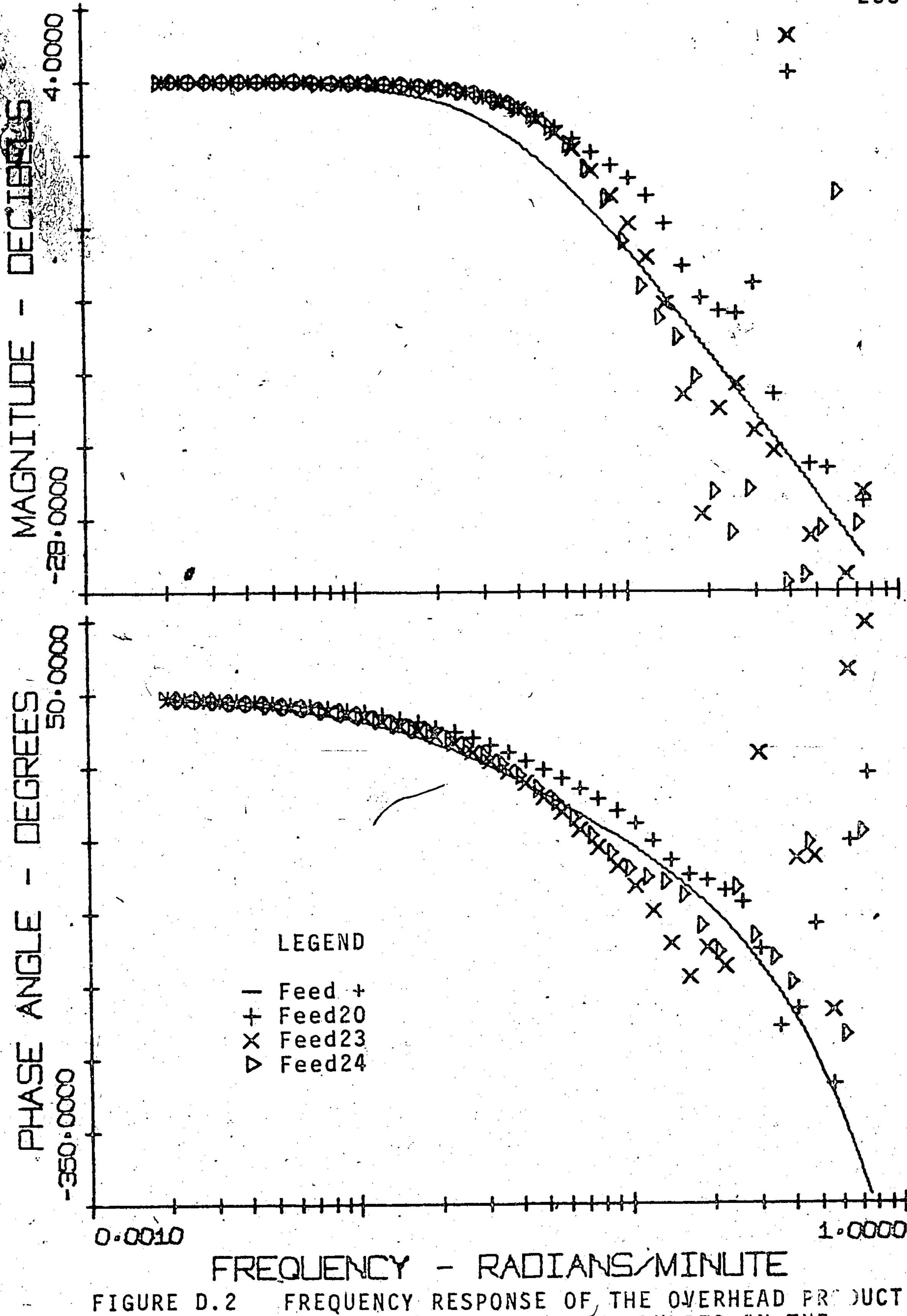

FIGURE D.2 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO POSITIVE PULSES IN THE FEED FLOW RATE

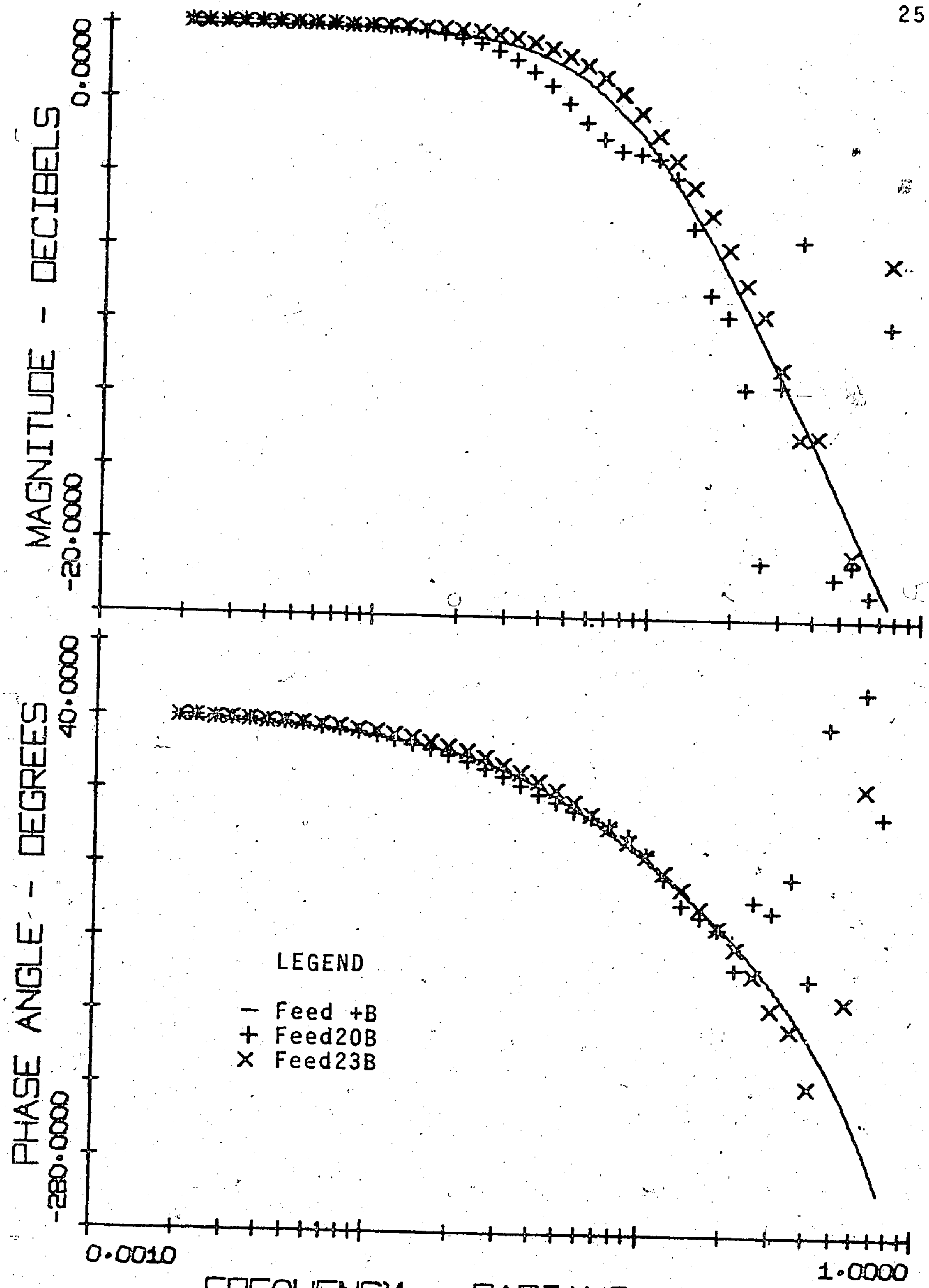

FIGURE D.3 FREQUENCY RESPONSE OF THE BOTTOMS PRODUCT COMPOSITION TO POSITIVE PULSES IN THE FEED FLOW RATE

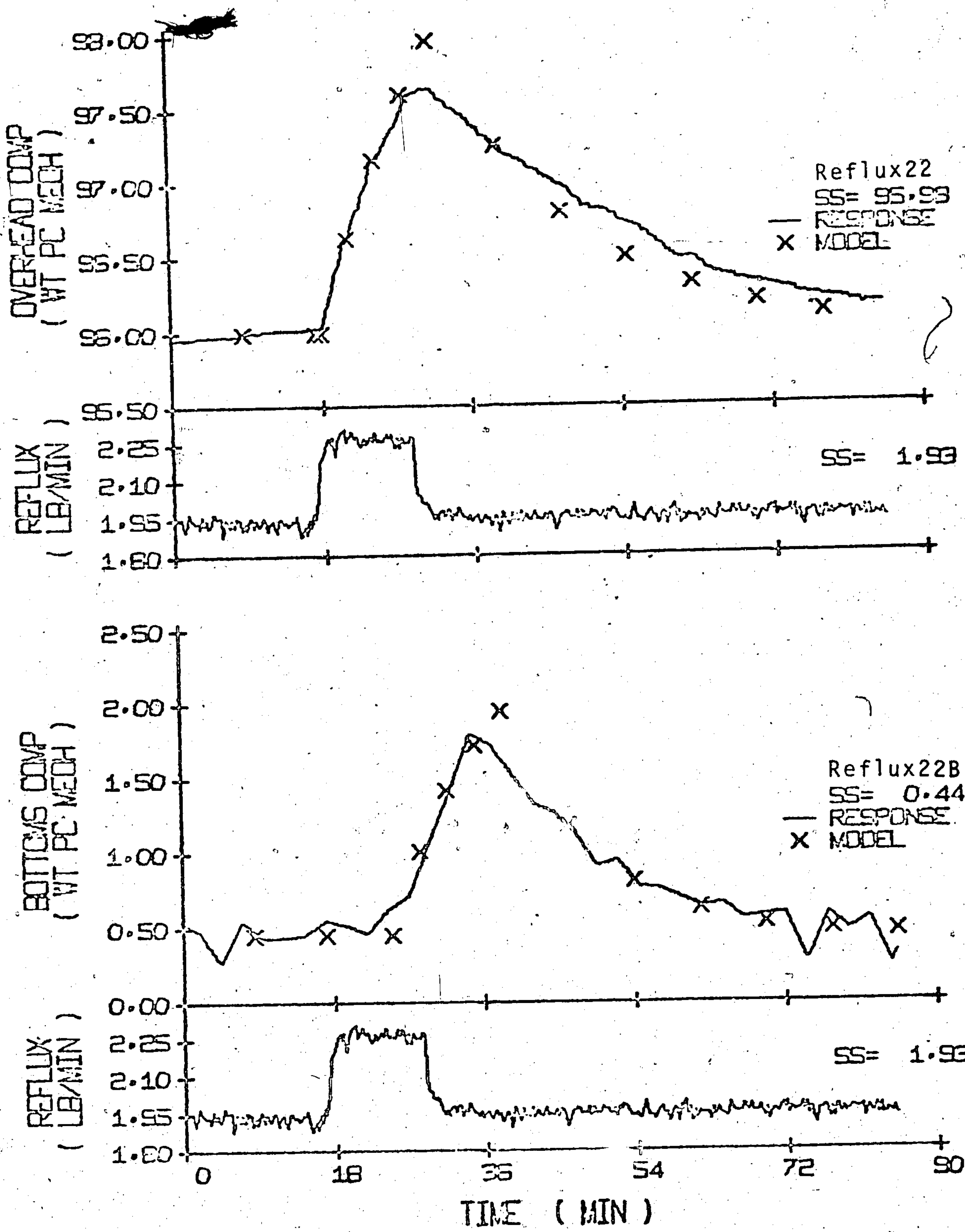

FIGURE D.4 TRANSIENT RESPONSE OF THE OVERHEAD AND BOTTOMS PRODUCT COMPOSITION TO A POSITIVE PULSE IN THE REFLUX FLOW RATE

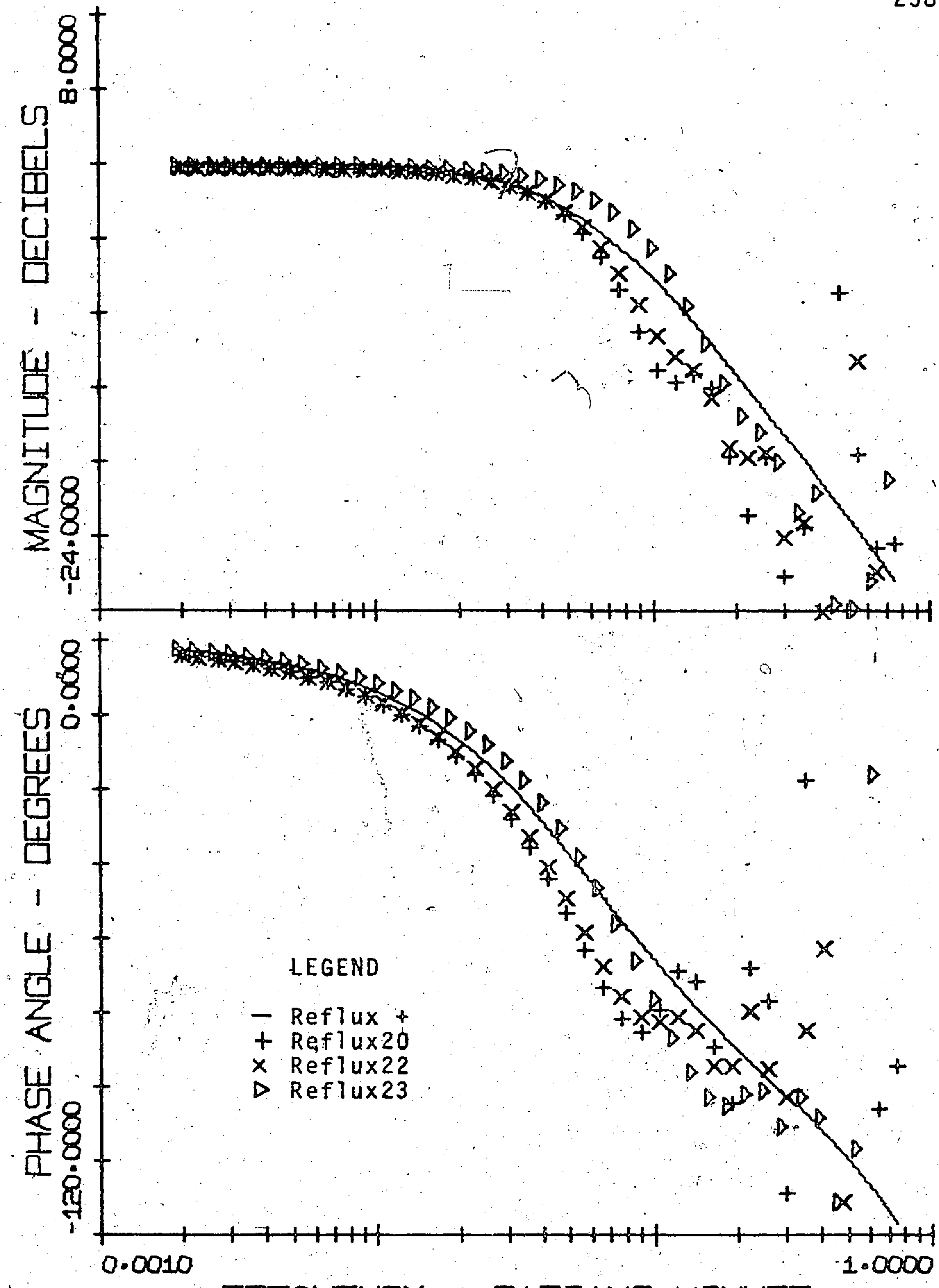

FIGURE D.5 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO POSITIVE PULSES IN THE REFLUX FLOW RATE

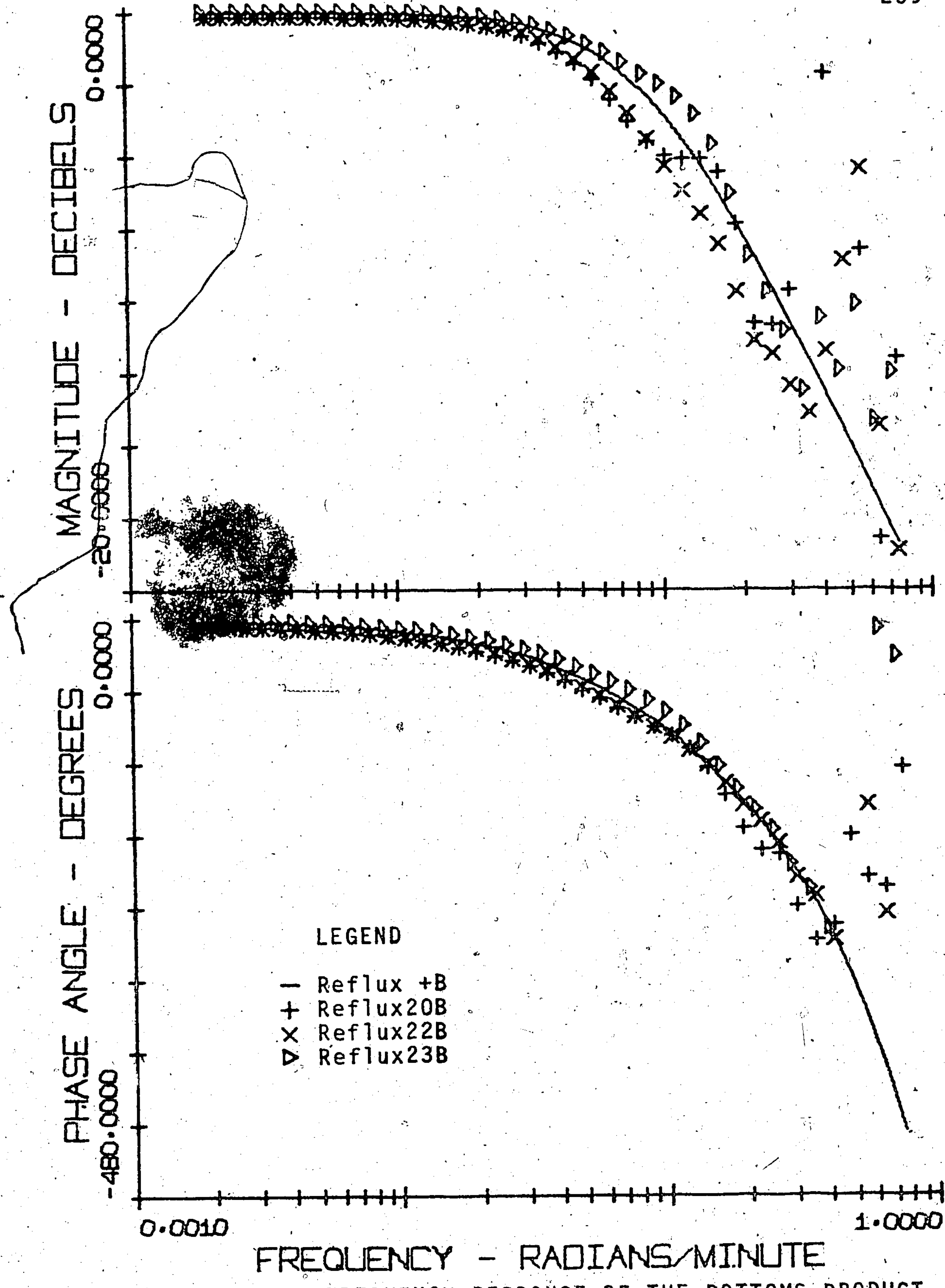

FIGURE D.6 - FREQUENCY RESPONSE OF THE BOTTOMS PRODUCT COMPOSITION TO POSITIVE PULSES IN THE REFLUX FLOW RATE

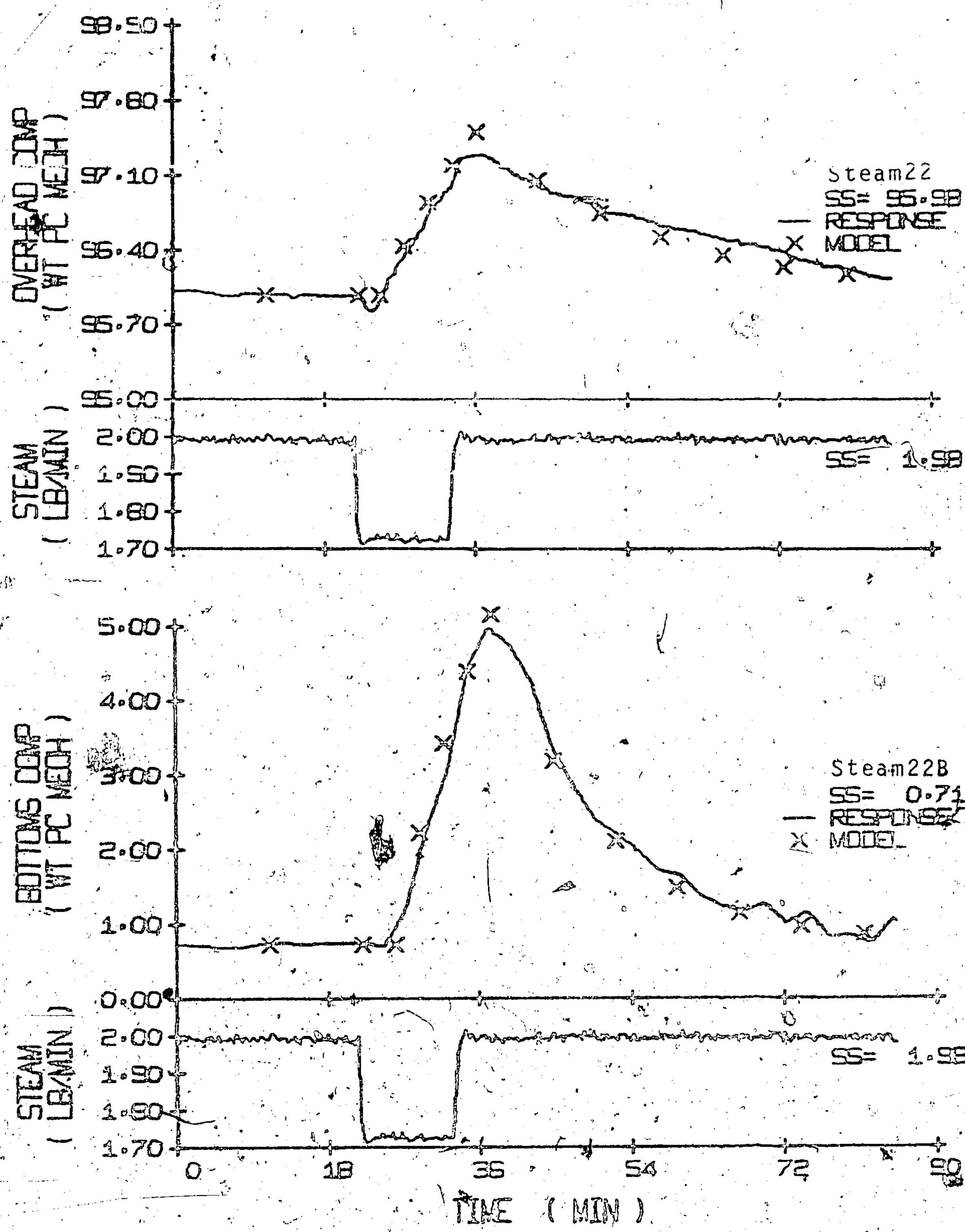

FIGURE D.7 TRANSIENT RESPONSE OF THE OVERHEAD AND BOTTOMS PRODUCT COMPOSITION TO A NEGATIVE PULSE IN THE STEAM FLOW RATE

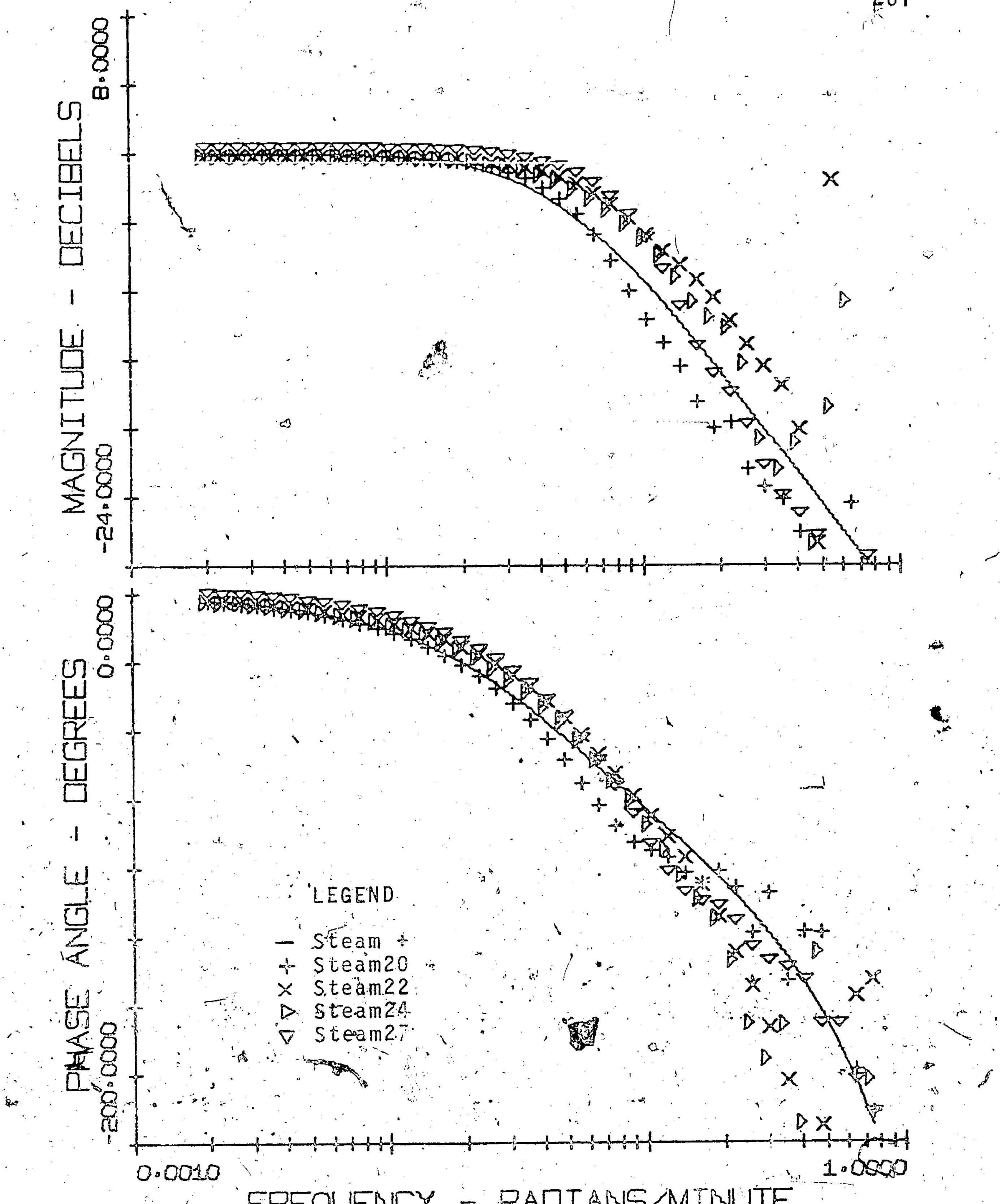

FIGURE D.8 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO NEGATIVE PULSES IN THE STEAM FLOW RATE

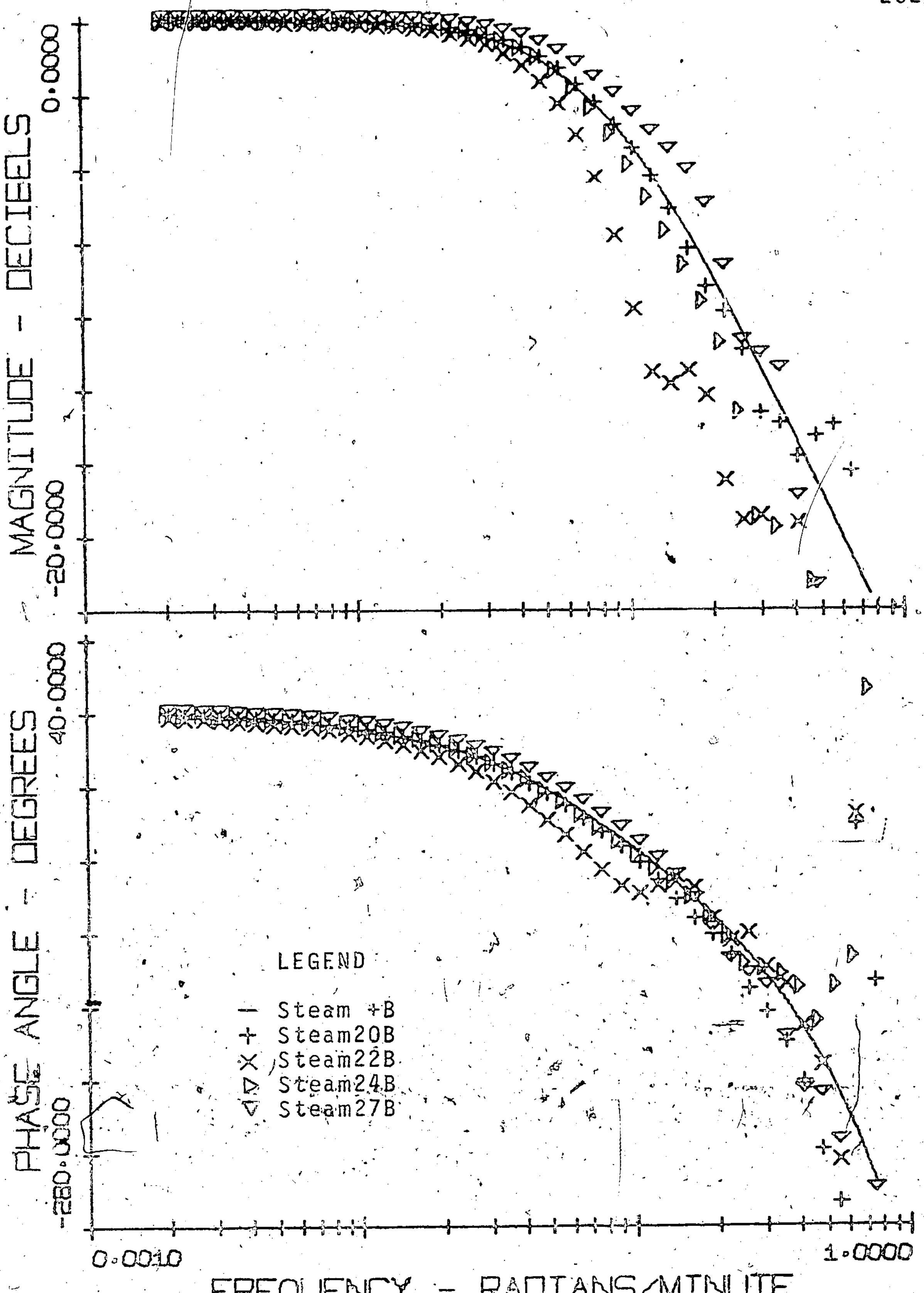

FIGURE D.9 FREQUENCY RESPONSE OF THE BOTTOMS PRODUCT COMPOSITION TO NEGATIVE PULSES IN THE STEAM FLOW RATE

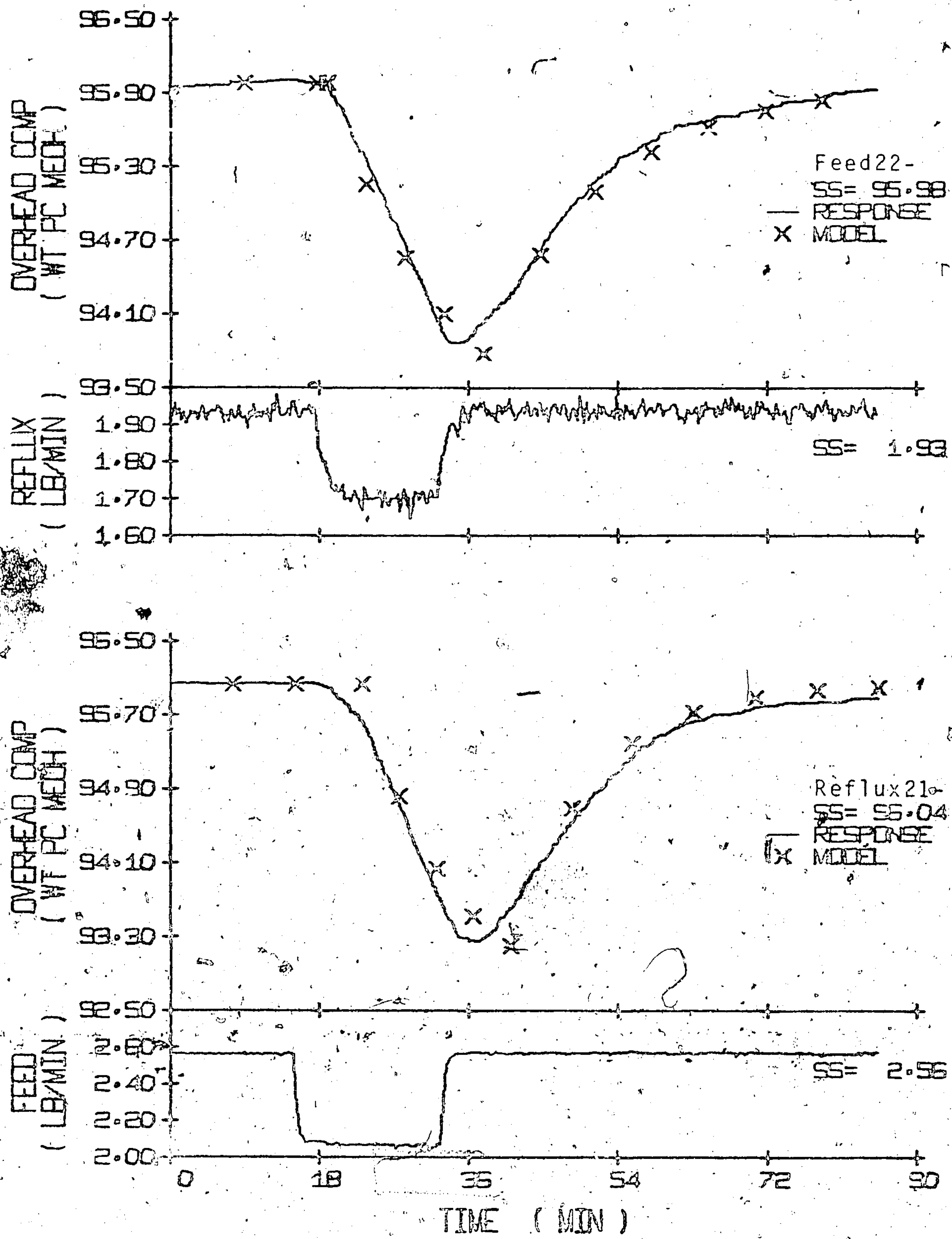

FIGURE D.10 TRANSIENT RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO A NEGATIVE PULSE IN THE FEED AND REFLUX FLOW RATES

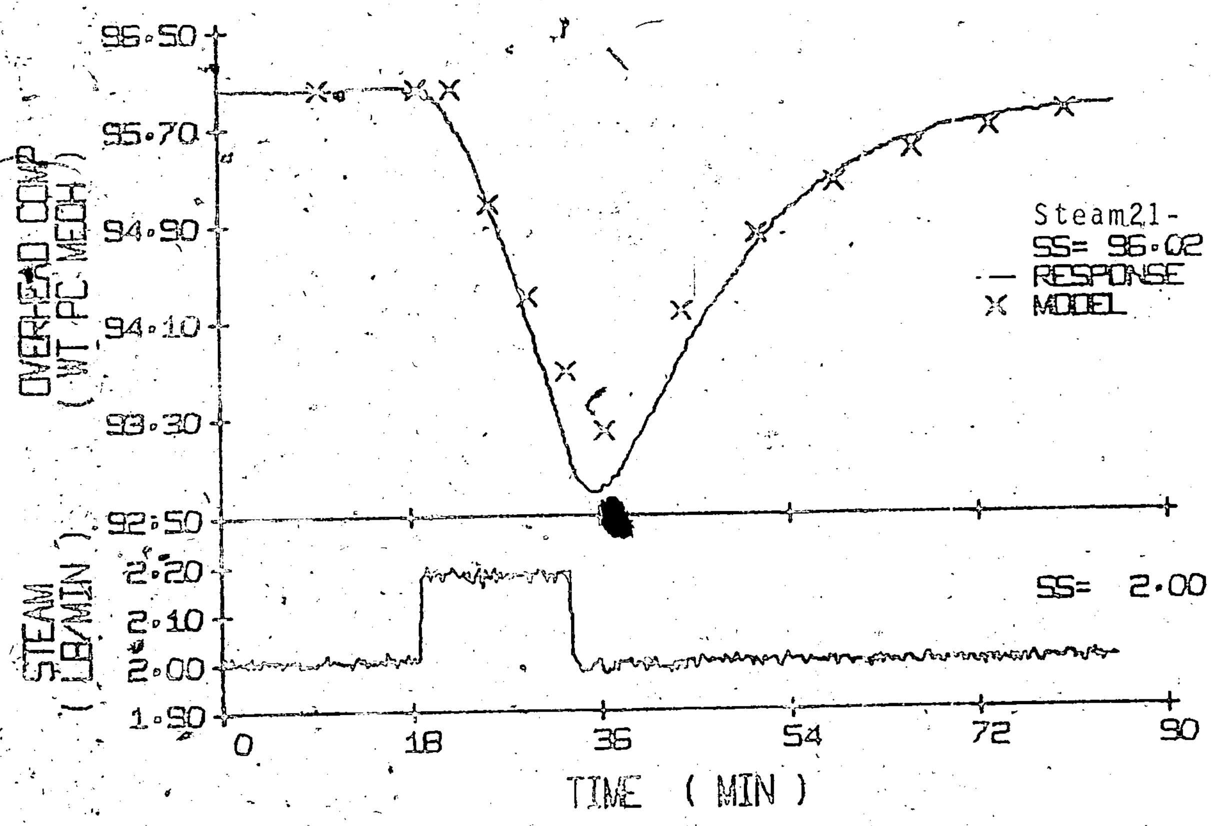

FIGURE D.11 TRANSIENT RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO A POSITIVE PULSE IN THE STEAM FLOW RATE

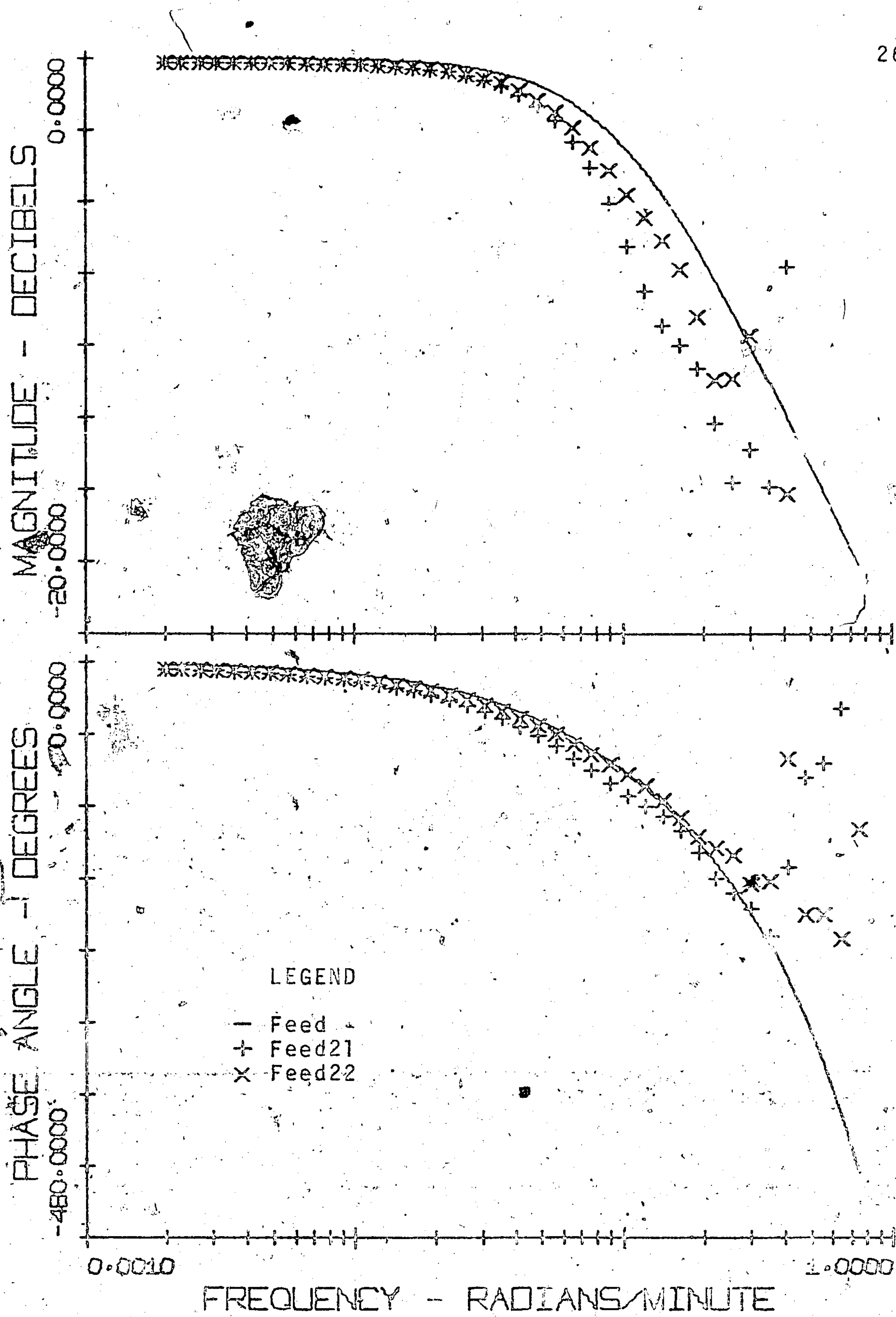

FIGURE D.12 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO NEGATIVE PULSES IN THE FEED FLOW RATE

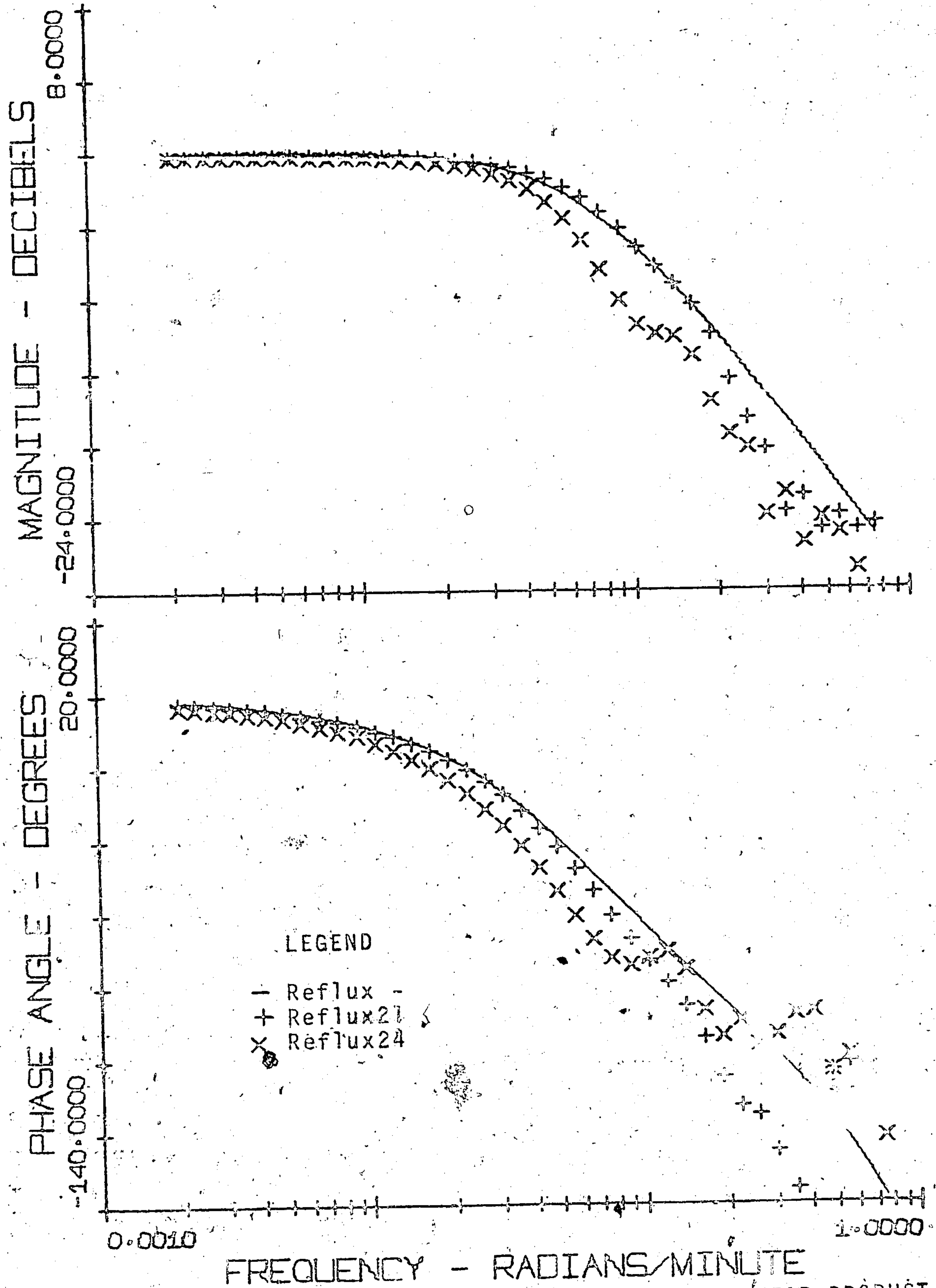

FIGURE D.13 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO NEGATIVE PULSES IN THE REFLUX FLOW RATE

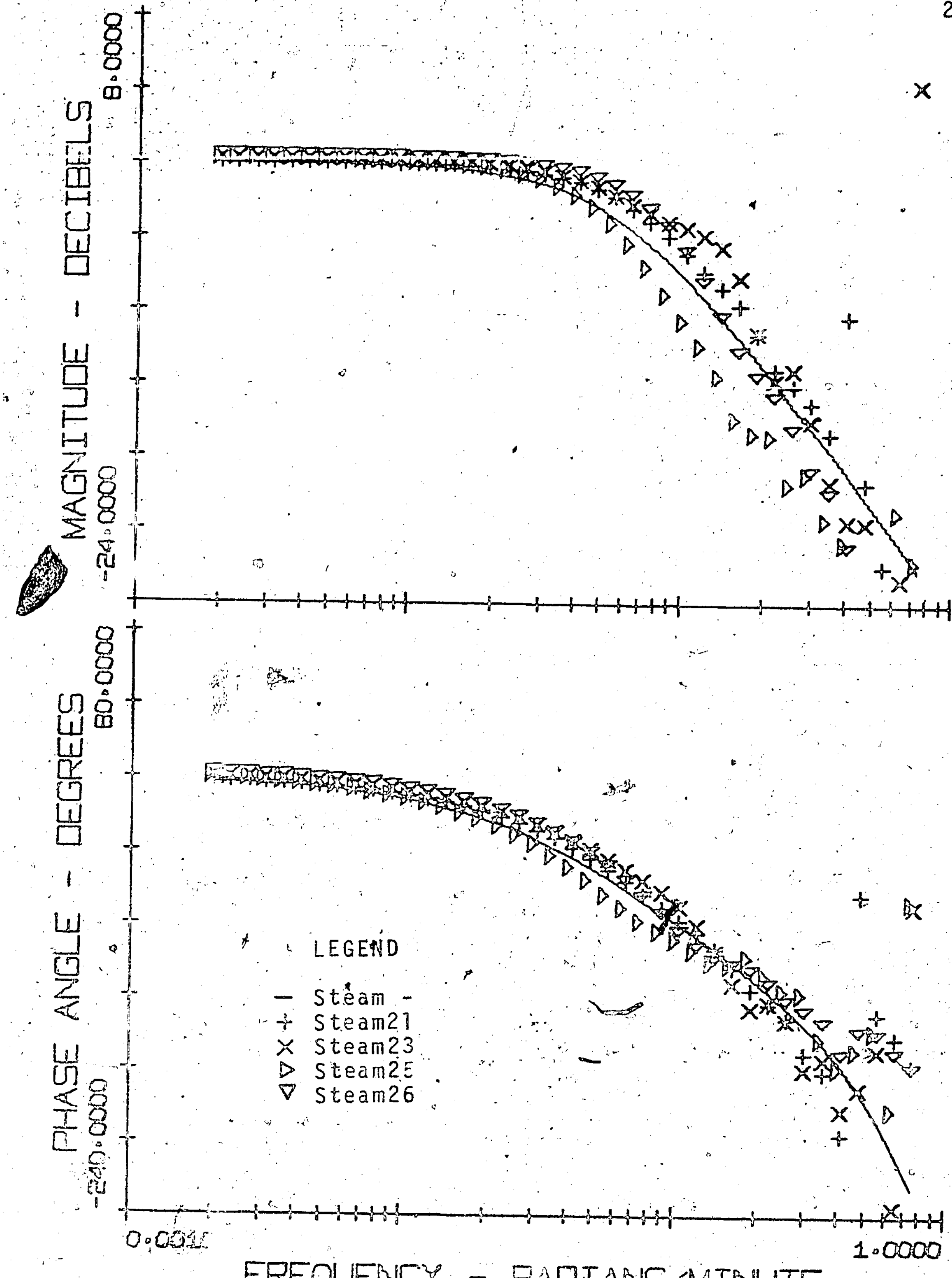

FIGURE D.14 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO POSITIVE PULSES IN THE STEAM FLOW RATE

TABLE D.2 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO POSITIVE PULSES IN THE FEED FLOW RATE

	Feed +		Feed20		Feed23		Feed24	
GAIN	0.99		2.93		2.41		5.82	
FREQ RAD/ MIN	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE
0.0020	1.000	-3.[illegible]	1.000	-2.3	1.000	-3.0	1.000	-2.7
0.0023	0.999	-4.3	1.000	-2.7	0.999	-3.5	0.999	-3.1
0.0027	0.998	-5.1	1.987	-3.6	0.999	-4.1	0.999	-3.1
0.0031	0.997	-5.9	0.999	-3.6	0.999	-4.7	0.999	-4.2
0.0036	0.996	-6.9	0.998	-4.2	0.998	-5.5	0.999	-4.9
0.0042	0.994	-8.0	0.998	-4.9	0.998	-6.4	0.998	-5.8
0.0049	0.992	-9.3	0.997	-5.8	0.997	-7.5	0.998	-6.7
0.0057	0.989	-10.8	0.996	-6.7	0.996	-8.7	0.997	-7.8
0.0067	0.985	-12.6	0.995	-7.8	0.995	-10.2	0.996	-9.1
0.0078	0.979	-14.6	0.993	-9.1	0.993	-11.8	0.994	-10.6
0.0091	0.972	-17.0	0.991	-10.6	0.991	-13.8	0.992	-12.4
0.0106	0.962	-19.6	0.988	-12.3	0.988	-16.0	0.990	-14.4
0.0123	0.949	-22.7	0.984	-14.4	0.984	-18.7	0.986	-16.8
0.0144	0.933	-26.2	0.978	-16.7	0.978	-21.7	0.981	-19.5
0.0167	0.912	-30.1	0.970	-19.4	0.970	-25.2	0.975	-22.7
0.0195	0.885	-34.5	0.960	-22.5	0.960	-29.3	0.966	-26.4
0.0227	0.853	-39.3	0.947	-26.0	0.946	-34.1	0.954	-30.7
0.0264	0.814	-44.6	0.929	-30.1	0.928	-39.6	0.938	-35.7
0.0308	0.769	-50.3	0.906	-34.7	0.905	-45.9	0.917	-41.5
0.0358	0.719	-56.4	0.877	-39.9	0.873	-53.1	0.888	-48.2
0.0417	0.664	-62.8	0.841	-45.6	0.833	-61.4	0.851	-55.8
0.0486	0.606	-69.4	0.796	-51.8	0.783	-70.7	0.803	-64.5
0.0566	0.548	-76.2	0.745	-58.3	0.721	-81.1	0.742	-74.4
0.0659	0.490	-83.3	0.698	-65.1	0.648	-92.5	0.666	-85.4
0.0767	0.435	-90.6	0.636	-72.1	0.567	-104.8	0.574	-97.2
0.0893	0.383	-98.1	0.586	-79.8	0.483	-117.7	0.470	-10[illegible].1
0.1040	0.335	-106.1	0.540	-89.1	0.403	-131.9	0.363	-119.5
0.1210	0.292	-114.6	0.484	-101.0	0.327	-148.8	0.273	-125.7
0.1409	0.254	-123.7	0.404	-114.4	0.243	-170.5	0.222	-128.5
0.1640	0.220	-133.6	0.310	-124.4	0.138	-194.0	0.198	-137.8
0.1910	0.190	-144.6	0.253	-127.5	0.065	-174.2	0.154	-158.7
0.2223	0.164	-156.9	0.232	-134.9	0.126	-187.3	0.074	-176.5
0.2588	0.141	-170.8	0.228	-142.6	0.109	-41.1	0.057	-133.0
0.3013	0.122	-186.5	0.278	-175.0	0.147	87.9	0.075	-166.2
0.3507	0.105	-204.5	0.137	-228.2	0.096	74.9	0.025	-181.4
0.4083	0.090	-225.2	1.048	-215.9	1.321	-112.9	0.042	-198.3
0.4753	0.077	-248.9	0.089	-157.6	0.056	-112.3	0.044	-101.9
0.5534	0.066	-276.4	0.086	-268.2	0.019	-7.2	0.059	88.9
0.6442	0.057	-308.1	0.030	-101.0	0.043	16.3	0.490	-234.1
0.7499	0.049	-344.9	0.069	-54.9	0.074	47.9	0.060	-94.5

TABLE D.3 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO POSITIVE PULSES IN THE REFLUX FLOW RATE

	Reflux +		Reflux20		Reflux [illegible]		Reflux23	
GAIN	0.99		14.80		13.29		10.59	
FREQ RAD/ MIN	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE
0.0020	1.000	-2.0	1.000	-2.3	1.000	-2.2	1.000	-1.5
0.0023	0.999	-2.4	0.999	-2.7	0.999	-2.5	0.999	-1.8
0.0027	0.999	-2.8	0.999	-3.1	0.999	-3.0	0.999	-2.1
0.0031	0.999	-3.2	0.999	-3.7	0.999	-3.5	0.999	-2.4
0.0036	0.998	-3.8	0.998	-4.3	0.998	-4.0	0.999	-2.9
0.0042	0.997	-4.4	0.998	-5.0	0.998	-4.7	0.999	-3.3
0.0049	0.996	-5.1	0.997	-5.8	0.997	-5.5	0.998	-3.9
0.0057	0.995	-6.0	0.996	-6.8	0.996	-6.4	0.998	-4.5
0.0067	0.993	-7.0	0.995	-7.9	0.995	-7.5	0.997	-5.3
0.0078	0.991	-8.1	0.993	-9.2	0.993	-8.7	0.997	-6.2
0.0091	0.988	-9.4	0.990	-10.7	0.991	-10.1	0.996	-7.2
0.0106	0.983	-11.0	0.986	-12.4	0.987	-11.8	0.994	-8.4
0.0123	0.977	-12.7	0.982	-14.5	0.983	-13.7	0.992	-9.7
0.0144	0.970	-14.7	0.975	-16.8	0.977	-15.9	0.989	-11.3
0.0167	0.960	-17.0	0.966	-19.6	0.969	-18.5	0.986	-13.2
0.0195	0.946	-19.7	0.955	-22.7	0.958	-21.5	0.981	-15.3
0.0227	0.929	-22.6	0.939	-26.4	0.943	-25.0	0.974	-17.8
0.0264	0.907	-26.0	0.918	-30.6	0.924	-29.0	0.965	-20.7
0.0308	0.880	-29.6	0.890	-35.5	0.898	-33.5	0.953	-24.1
0.0358	0.847	-33.6	0.853	-41.0	0.865	-38.7	0.937	-27.9
0.0417	0.807	-37.9	0.805	-47.3	0.821	-44.5	0.916	-32.3
0.0486	0.761	-42.4	0.743	-54.2	0.766	-50.9	0.888	-37.4
0.0566	0.710	-47.1	0.666	-61.6	0.697	-57.8	0.852	-43.1
0.0659	0.655	-51.9	0.573	-69.1	0.615	-64.7	0.807	-49.5
0.0767	0.597	-56.6	0.468	-75.5	0.523	-70.9	0.750	-56.5
0.0893	0.539	-61.2	0.362	-78.2	0.431	-75.0	0.681	-64.1
0.1040	0.481	-65.6	0.284	-73.6	0.357	-75.7	0.601	-72.0
0.1210	0.427	-69.9	0.265	-65.8	0.313	-74.9	0.513	-79.7
0.1409	0.375	-73.9	0.277	-68.0	0.287	-77.8	0.422	-86.7
0.1640	0.329	-77.8	0.254	-81.1	0.243	-84.7	[illegible]	[illegible]
0.1910	0.286	-81.5	0.167	-92.7	0.178	-84.9	0.260	-93.6
0.2223	0.248	-85.1	0.115	-65.3	0.166	-73.8	0.212	[illegible]
0.2588	0.215	-88.6	0.165	-72.0	0.171	-85.5	0.192	-90.7
0.3013	0.186	-92.2	0.079	-110.9	0.101	-91.1	[illegible]	[illegible]
0.3507	0.160	-95.8	0.107	-27.7	0.111	-77.6	[illegible]	[illegible]
0.4083	0.138	-99.5	0.061	-224.5	0.064	-61.0	0.131	[illegible]
0.4753	0.119	-103.5	0.462	-179.9	0.062	-112.5	0.066	-113.1
0.5534	0.102	-107.8	0.169	-140.8	0.304	-149.2	0.064	-102.3
0.6442	0.088	-112.6	0.094	-93.8	0.082	-157.2	0.076	-26.8
0.7499	0.075	-117.8	0.097	-85.3	0.061	-121.9	0.142	53.1

TABLE D.4 FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO NEGATIVE PULSES IN THE STEAM FLOW RATE

	Steam +		Steam20		Steam22	
GAIN	0.99		19.70		28.74	
FREQ RAD/MIN	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE
0.0020	1.000	-2.6	1.000	-2.6	1.000	-1.8
0.0023	0.999	-3.1	0.999	-3.0	0.999	-2.1
0.0027	0.999	-3.6	0.999	-3.5	0.999	-2.5
0.0031	0.998	-4.2	0.999	-4.1	0.999	-2.9
0.0036	0.997	-4.9	0.998	-4.8	0.999	-3.4
0.0042	0.996	-5.7	0.998	-5.6	0.999	-4.0
0.0049	0.995	-6.6	0.997	-6.5	0.998	-4.6
0.0057	0.993	-7.7	0.996	-7.5	0.998	-5.4
0.0067	0.990	-9.0	0.994	-8.8	0.997	-6.3
0.0078	0.987	-10.4	0.992	-10.2	0.996	-7.2
0.0091	0.982	-12.1	0.990	-11.9	0.995	-8.5
0.0106	0.976	-14.1	0.986	-13.9	0.993	-9.9
0.0123	0.968	-16.3	0.981	-16.2	0.991	-11.6
0.0144	0.957	-18.9	0.974	-18.8	0.987	-13.5
0.0167	0.943	-21.8	0.966	-21.8	0.983	-15.7
0.0195	0.925	-25.1	0.954	-25.4	0.977	-18.2
0.0227	0.903	-28.7	0.938	-29.5	0.969	-21.2
0.0264	0.874	-32.8	0.916	-34.2	0.959	-24.6
0.0308	0.840	-37.3	0.888	-39.5	0.945	-28.5
0.0358	0.799	-42.1	0.852	-45.6	0.926	-33.0
0.0417	0.752	-47.2	0.805	-52.5	0.902	-38.1
0.0486	0.700	-52.5	0.745	-60.1	0.871	-43.8
0.0566	0.644	-58.0	0.672	-68.2	0.831	-50.2
0.0659	0.586	-63.6	0.587	-76.5	0.782	-57.2
0.0767	0.527	-69.1	0.494	-84.2	0.724	-64.7
0.0893	0.470	-74.7	0.403	-89.9	0.659	-72.3
0.1040	0.416	-80.2	0.330	-93.0	0.593	-79.9
0.1210	0.366	-85.8	0.282	-95.5	0.531	-87.2
0.1409	0.320	-91.5	0.241	-100.7	0.481	-95.1
0.1640	0.278	-97.3	0.189	-105.0	0.437	-104.7
0.1910	0.242	-103.3	0.159	-100.2	0.388	-116.6
0.2223	0.209	-109.7	0.165	-106.4	0.332	-129.2
0.2588	0.181	-116.6	0.121	-123.0	0.283	-141.8
0.3013	0.156	-124.1	0.108	-108.5	0.244	-156.9
0.3507	0.134	-132.5	0.098	-140.4	0.215	-176.7
0.4083	0.115	-141.8	0.079	-122.7	0.159	-216.2
0.4753	0.099	-152.3	0.036	-123.0	0.074	-192.8
0.5534	0.085	-164.3	0.039	78.3	0.848	40.8
0.6442	0.073	-178.0	0.096	-173.2	0.044	-145.3
0.7499	0.063	-193.7	0.051	-188.4	0.053	-139.3

FREQUENCY RESPONSE OF THE OVERHEAD PRODUCT COMPOSITION TO NEGATIVE PULSES IN THE STEAM FLOW RATE

	Steam24		Steam27	
GAIN	21.11		17.18	
FREQ RAD/MIN	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE
0.0020	1.000	-2.0	1.000	-1.9
0.0023	0.999	-2.3	0.999	-2.3
0.0027	0.999	-2.7	0.999	-2.6
0.0031	0.999	-3.1	0.999	-3.1
0.0036	0.999	-3.7	0.999	-3.6
0.0042	0.998	-4.3	0.999	-4.2
0.0049	0.998	-5.0	0.998	-4.9
0.0057	0.997	-5.8	0.998	-5.7
0.0067	0.996	-6.8	0.997	-6.6
0.0078	0.995	-7.9	0.996	-7.7
0.0091	0.993	-9.2	0.995	-9.0
0.0106	0.991	-10.7	0.993	-10.5
0.0123	0.988	-12.5	0.991	-12.2
0.0144	0.984	-14.5	0.988	-14.2
0.0167	0.978	-16.8	0.984	-16.6
0.0195	0.971	-19.5	0.979	-19.3
0.0227	0.961	-22.7	0.971	-22.4
0.0264	0.948	-26.3	0.962	-26.1
0.0308	0.931	-30.4	0.948	-30.3
0.0358	0.908	-35.0	0.931	-35.2
0.0417	0.880	-40.3	0.907	-40.8
0.0486	0.844	-46.1	0.876	-47.3
0.0566	0.801	-52.5	0.836	-54.7
0.0659	0.751	-59.4	0.784	-63.1
0.0767	0.695	-66.7	0.718	-72.4
0.0893	0.637	-74.5	0.638	-82.6
0.1040	0.577	-82.9	0.544	-93.2
0.1210	0.513	-92.1	0.441	-103.3
0.1409	0.444	-101.7	0.341	-111.0
0.1640	0.377	-109.8	0.263	-114.5
0.1910	0.339	-117.4	0.218	-115.8
0.2223	0.317	-132.4	0.191	-121.5
0.2588	0.249	-155.3	0.155	-130.9
0.3013	0.149	-168.5	0.117	-136.2
0.3507	0.121	-155.9	0.096	-138.3
0.4083	0.145	-191.9	0.085	-143.1
0.4753	0.074	-129.4	0.073	-159.2
0.5534	0.186	-212.9	0.055	-158.9
0.6442	0.377	69.0	0.056	-179.4
0.7499	0.025	-175.8	0.064	-191.5

TABLE D.5 FREQUENCY RESPONSE OF THE BOTTOM PRODUCT COMPOSITION TO POSITIVE PULSES IN THE FEED FLOW RATE

	Feed +B		Feed20B		Feed24B	
GAIN	0.99		4.63		5.12	
FREQ RAD/MIN	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE
0.0020	1.000	-1.8	1.000	-2.3	1.000	-1.7
0.0023	0.999	-2.2	0.999	-2.7	0.999	-1.9
0.0027	0.999	-2.5	0.999	-3.1	0.999	-2.3
0.0031	0.999	-2.9	0.999	-3.6	0.999	-2.7
0.0036	0.999	-3.4	0.998	-4.3	0.999	-3.1
0.0042	0.998	-4.0	0.997	-5.0	0.999	-3.6
0.0049	0.998	-4.7	0.996	-5.8	0.998	[illegible]
0.0057	0.997	-5.4	0.995	-6.7	0.998	[illegible]
0.0067	0.996	-6.3	0.993	-7.8	0.997	-5[illegible]
0.0078	0.995	-7.4	0.991	-9.1	0.997	6.[illegible]
0.0091	0.993	-8.6	0.988	-10.6	0.995	7.8
0.0106	0.991	-10.0	0.983	-12.3	0.994	[illegible]
0.0123	0.988	-11.6	0.977	-14.3	0.992	-10.5
0.0144	0.984	-13.5	0.969	-16.6	0.989	-12.3
0.0167	0.978	-15.7	0.959	-19.3	0.986	-14.3
0.0195	0.971	-18.2	0.945	-22.3	0.981	-16.6
0.0227	0.962	-21.0	0.926	-25.7	0.974	-19.3
0.0264	0.949	-24.3	0.902	-29.5	0.966	-22.4
0.0308	0.933	-28.1	0.871	-33.7	0.954	-26.0
0.0358	0.912	-32.3	0.833	-38.2	0.939	-30.1
0.0417	0.886	-37.1	0.788	-42.9	0.919	-34.8
0.0486	0.854	-42.4	0.737	-47.4	0.893	-40.1
0.0566	0.816	-48.2	0.685	-51.6	0.860	-46.0
0.0659	0.772	-54.5	0.641	-55.2	0.820	-52.6
0.0767	0.721	-61.3	0.614	-59.0	0.771	-59.8
0.0893	0.667	-68.6	0.606	-64.8	0.716	-67.4
0.1040	0.609	-76.2	0.596	-74.6	0.656	-75.4
0.1210	0.551	-84.2	0.552	-88.3	0.596	-83.8
0.1409	0.493	-92.7	0.455	-102.5	0.538	-92.9
0.1640	0.438	-101.6	0.351	-109.1	0.480	-103.0
0.1910	0.386	-111.0	0.321	-114.8	0.420	-113.8
0.2223	0.338	-121.1	0.242	-136.4	0.367	-124.8
0.2588	0.295	-132.1	0.122	-99.7	0.325	-138.8
0.3013	0.256	-144.1	0.245	-105.3	0.264	-157.7
0.3507	0.222	-157.5	0.433	-87.2	0.201	-168.6
0.4083	0.192	-172.5	3.963	-142.4	0.203	-199.6
0.4753	0.165	-189.3	0.116	-4.5	0.073	43.1
0.5534	0.143	-208.6	0.122	53.9	0.127	-153.7
0.6442	0.123	-230.5	0.108	14.0	2.269	-37.3
0.7499	0.106	-255.7	0.311	-53.3	0.401	64.8

TABLE D.6 FREQUENCY RESPONSE OF THE BOTTOM PRODUCT COMPOSITION TO POSITIVE PULSES IN THE REFLUX FLOW RATE

	Reflux +B		Reflux20B		Reflux22B		Reflux23B	
GAIN	0.99		5.62		7.13		6.33	
FREQ RAD/ MIN	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE	MAG RATIO	PHASE ANGLE DEGREE
0.0020	1.000	-2.1	1.000	-2.4	1.000	-2.2	1.000	-1.7
0.0023	0.999	-2.5	0.999	-2.8	0.999	-2.6	1.000	-2.0
0.0027	0.999	-2.9	0.999	-3.2	0.999	-3.0	0.999	-2.3
0.0031	0.999	-3.4	0.999	-3.8	0.999	-3.5	0.999	-2.7
0.0036	0.999	-3.9	0.999	-4.4	0.999	-4.1	0.999	-3.2
0.0042	0.999	-4.6	0.998	-5.1	0.998	-4.8	0.999	-3.7
0.0049	0.998	-5.4	0.998	-6.0	0.998	-5.6	0.998	-4.3
0.0057	0.998	-6.3	0.997	-7.0	0.997	-6.5	0.998	-5.0
0.0067	0.997	-7.3	0.996	-8.1	0.996	-7.6	0.997	-5.8
0.0078	0.996	-8.5	0.995	-9.5	0.995	-8.9	0.997	-6.8
0.0091	0.995	-9.9	0.993	-11.0	0.993	-10.3	0.996	-7.9
0.0106	0.993	-11.5	0.990	-12.8	0.991	-12.0	0.994	-9.2
0.0123	0.991	-13.4	0.987	-14.9	0.988	-14.0	0.992	-10.7
0.0144	0.987	-15.6	0.982	-17.3	0.984	-16.3	0.990	-12.5
0.0167	0.983	-18.1	0.976	-20.2	0.979	-18.9	0.986	-14.5
0.0195	0.977	-21.0	0.968	-23.4	0.971	-22.0	0.981	-16.9
0.0227	0.970	-24.4	0.957	-27.2	0.962	-25.5	0.975	-19.6
0.0264	0.960	-28.3	0.942	-31.5	0.949	-29.6	0.967	-22.7
0.0308	0.947	-32.8	0.923	-36.4	0.932	-34.3	0.956	-26.3
0.0358	0.930	-38.0	0.898	-42.1	[illegible]	-39.7	0.942	-30.5
0.0417	0.908	-43.8	0.865	-48.4	0.880	-45.8	0.924	-35.1
0.0486	0.881	-50.4	0.824	-55.3	0.843	-52.7	0.901	-40.4
0.0566	0.848	-57.8	0.775	-62.9	0.796	-60.3	0.875	-46.3
0.0659	0.809	-66.1	0.717	-70.7	0.741	-68.5	0.844	-52.8
0.0767	0.764	-75.3	0.658	-78.5	0.678	-77.2	0.810	-60.1
0.0893	0.713	-85.4	0.605	-85.9	0.612	-86.1	0.776	[illegible]
0.1040	0.658	-96.5	0.572	-93.6	0.551	-95.2	0.743	-78.0
0.1210	0.600	-108.5	0.565	-104.0	0.499	-105.2	0.708	-90.0
0.1409	0.542	-121.7	0.567	-120.0	0.454	-127.7	0.661	-104.9
0.1640	0.484	-136.2	0.536	-143.0	0.402	-133.6	0.536	-123.0
0.1910	0.429	-152.0	0.436	-170.1	0.332	-150.9	0.430	-142.5
0.2223	0.378	-169.6	0.293	-188.3	0.273	-164.7	0.376	-159.2
0.2588	0.331	-189.2	0.291	-192.1	0.260	-182.2	0.326	-176.6
0.3013	0.288	-211.3	0.335	-235.4	0.229	-211.6	0.279	-205.1
0.3507	0.250	-236.2	0.084	-263.7	0.205	-226.2	0.219	-225.2
0.4083	0.217	-264.6	0.793	-250.3	0.262	-263.0	0.294	-258.3
0.4753	0.187	-297.0	1.435	-176.6	0.377	73.9	0.239	49.8
0.5534	0.162	-334.3	0.392	-211.2	0.543	-151.4	0.309	15.4
0.6442	0.139	-377.2	0.124	-219.1	0.195	-240.8	0.195	-10.1
0.7499	0.120	-426.8	0.255	-119.6	0.118	41.3	0.235	-32.1