STEADY-STATE EVALUATION OF CONTROL SCHEMES FOR BINARY DISTILLATION

by

Alan Lik Chan

A thesis submitted to the

Department of Chemical Engineering

Of

The University of Ottawa

in partial fulfillment of the requirements

for the degree of Master of Science

August, 1967

Research Director

M.Sc. Candidate

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

ABSTRACT

Utilising the Smoker equation, an analytical method equivalent to the McCabe-Thiele diagram method was developed to evaluate some simple control schemes for a binary distillation tower. The basis of evaluation was the steady-state deviations of product purities when the tower was subject to a sustained disturbance in feed composition, rate or enthalpy. The schemes proposed in this work were a simple feedforward and a feedback heat-balance control scheme. These control schemes were simulated on an IBM-360 and IBM-1620 digital computer. The feedback heat-balance control was further verified on a laboratory pilot-scale binary distillation tower separating a methanol-water mixture.

These two control schemes handled feed composition and enthalpy disturbances more successfully than the rate disturbance. In general, for one end control, the feedback heat-balance control was better than the simple feedforward one. For feed composition disturbance and two end control the simple feedforward control was the better one. The feedback heat-balance control was more suitable for system with high relative volatility and top composition control. Also, the closer the control tray to the end of the column, the better the product purity.

ACKNOWLEDGEMENT

The author is greatly indebted to his director Dr. F.D.F. Talbot for his guidance, patience and encouragement throughout the course of this work.

He also wishes to express his sincere gratitude to the Chairman of the Department Dr. B. C.-Y. Lu for providing the facilities for this research work.

He is grateful to Messrs. F. Giacobbi and G. Gasperetti for their helps in setting up the equipment; also to Mr. H.K. Lau for carrying out part of the experimental work.

Financial assistance from the National Research Council of Canada under Research Grant No. A-2925 during the year 1966-67 and from the Department of Chemical Engineering for a teaching assistantship during the 1966-67 academic year are gratefully acknowledged.

Last, but not the least, he wants to extend his thankfulness to Mr. S.D. Chang for his valuable suggestions in the computer program.

I Introduction

Continuous distillation is one of the most common and important unit operations used in the chemical and petroleum industry to-day. Efficient continuous operation of the distillation column requires a suitable automatic control scheme. In the past, the application of control schemes to distillation columns was practiced as an art. These schemes evolved empirically utilized available control equipment . However, to-day, aided by analog and digital computers, the design engineer has more powerful tools that can be used to investigate the consequence of various alternate control schemes. This ability is necessary in an age where various sophisticated alternate control schemes are now available which need evaluation. By utilizing analyzing equipment such as the infrared analyzer and the gas chromatograph with analog and digital computers, distillation column control has entered a new era. An example of a new control technique is the "feedforward control" scheme. This scheme has been successfully applied to columns where the conventional feedback control failed. By the combination of computers and analysing equipment , the combined feedforward and feedback control scheme is now possible. This control scheme is theoretically, by far the best.

This thesis consists of the analytical evaluation of the effectiveness of alternate control schemes for binary distillation where the column was subjected to feed rate, feed composition and feed enthalpy disturbances. The evaluation was made on a steady-state basis using a specially written digital computer program. The results were compared on a qualitative basis and some general conclusions were drawn.

An attempt was made to check the validity of the computer program against the actual operation of a pilot-plant distillation column. The experimental results were found to agree qualitatively with those calculated.

II Literature review

Most of the published works on distillation column control can be divided into two main parts; column dynamics and column control. In 1961, an excellent survey of the published studies on dynamics and control of distillation columns and other mass-transfer equipment was made by Archer and Rothfus (1). An extensive bibliography of both dynamics and control of distillation columns can also be found in a historical and critical review article by Rosenbrock (26) in 1962. One year later, a survey on the application of automatic control in the chemical industry by Rijnsdorp et. al., (25) listed several other important references on the dynamics and control of distillation column . Goulds (15) reviewed and analyzed the research activities on the characterization and control of distillation column, but also pointed out the shortcomings of various reported studies as well as many promising aspects. Very little information about multicomponent distillation has been reported and since it is beyond the scope of this investigation the following discussion will refer to a binary system unless otherwise specified. Further, this work was restricted on the control aspect of the effectiveness of the control schemes which were evaluated on a steady-state basis. No attempt was made to include the column dynamics.

1) Control Schemes General

Throughout the past twenty years, a great number of satisfactory and practical control schemes for distillation column control have been reported in the literature, yet, only a few typical and illustrative references will be cited in this work.

Rademaker and Rijnsdorp (25) introduced the principles of the classification for the most common types of columns and control systems. Harriott (14) in his book discussed various simple schemes and possible combinations of these schemes while Buckley (3) in his text deals with the more practical aspects of distillation control systems. Various other more sophisticated control schemes can be found in the literature (9,25).

2) Feedback Control

A fairly complete review was done by Fu (13) in his thesis on the general feedback control, the control point location and the interaction between control loops. Besides, an article by Forman (12) has given excellent consideration to every aspect of feedback distillation control.

3) Feedforward Control

In 1961, Calvert and Coulman (7) gave one of the first qualitative discussions on the possibility of the application of feedforward control to a chemical process. A year later, Lupfer and Oglesby (32) reported the design of a special purpose pneumatic version of a feed enthalpy computer which was the first actual application of feedforward control to distillation column. Later, Lupfer and Parsons (19) described the synthesis of feedforward controller of a distillation column from an empirically determined transient response of this column. Lamb, Pigford and Rippin (18) synthesized a feedforward controller using a linear model for a seven-plate column. Luyben and Gerster (20) modified the approach of Lamb, Pigford and Rippin and did the analog simulation of a ten-plate and a forty plate column, as well as by experimental tests with a ten-plate, 2-foot diameter pilot plant column. They also pointed out the need for the investigation of combined feedforward and feedback control schemes. An article by A. Crico in 1963 described the advantages and limits of certain types of combined simple feedforward and feedback control schemes.

MacMullan and Shinskey (22) reported the successful result of applying the feedforward analog computer control to a superfractionator where feedback control was found to be unacceptable. Utilizing the dynamic model developed by May and Franke (17) Distefano (10) programmed it on an IBM 709 digital computer for the feedforward control of a binary distillation tower. For a composition disturbance in feed stream, the overhead composition was maintained at $\pm .05$ percent of its initial steady-state value.

Shinskey (28) stated the theories and applications of feedforward control to chemical processes and also stressed the fact that feedforward control was more superior than conventional feedback control scheme. Forman (11) also gave a thorough discussion and comparison on feedback and feedforward control. Fu (13) investigated the performance of these two control schemes and the hybrid control scheme on an analog computer for a ten-plate pilot-plant binary distillation column.

Control scheme was usually justified by the transient performance. Yet for a chemical process, the steady-state is of great importance too. Keating and Townend (21) pointed out the fact that for the proper design and operation of a control system for a distillation column, the quantitative knowledge of steady and transient states is of equal importance. In 1954, Bauer and Orr (4) used McCabe-Thiele diagram method to determine the degree of control attainable with different control schemes. The idea of this paper was to investigate how the control variables affected the steady-state performance. Bertrand and Jones (2) also showed the effect of feed disturbances on the steady-state responses of the product compositions by using a digital computer and also McCabe-Thiele diagram. Lubyen (31) recently studied the dependence of steady-state gains on feed composition disturbance for a feedforward control scheme. Cadman, et al (6) used steady-state method to evaluate the effectiveness of feedforward controllers to compensate for step changes in feed compositions.

III Basis of Study

1) Principle of Column Control

The objects of any column control scheme can be listed as follows; (a) product quality control (b) material balance control and (c) a secondary aspect, constraints.

Product quality control. The column control system must:

i) Maintain either the overhead or bottom composition at a specified value, and

ii) Maintain the composition at the other end of the column as close as possible to a desired composition.

Material balance control. The column control system must:

i) Cause the average sum of the distillate and tail streams to be exactly equal to the average feed rate.

ii) The resulting adjustment in process flow must be smooth and gradual to avoid upsetting either the column or downstream process equipment fed by the column.

iii) Column holdup and overhead and bottoms inventories, should be maintained between maximum and minimum limits.

Constraints. For safe satisfactory operation of the column, certain constraints must be observed. For example:

i) The column should not flood;

ii) Column pressure drop should be high enough to maintain effective column operation;

iii) The temperature difference in the reboiler should not exceed the critical temperature difference.

As the operating conditions change, the above will change too. In order to offset this, a control scheme is necessary. The usual types of operating disturbance of a column are feed rate, feed composition and feed enthalpy changes. Feed enthalpy disturbance can easily be eliminated using auxiliary control equipment. The main disturbances would be the feed rate and feed composition and of these two, feed rate is more important because composition disturbances usually are gradual.

By applying a control scheme which adjusts the control variables to produce the correct product volume split and maintain the required "separating power" of the column, the end product purities are ensured. The term "separating power" is specific to each tower and depends upon such things as reflux rate, actual number of trays, tray efficiency, feed enthalpy and feed tray location. Usually most of these variables are fixed and the ones that can be manipulated are the reflux, reboiler heat input and product withdrawal rates. These manipulated variables can be controlled directly or indirectly according to different types of scheme. There are two basic methods by which a tower can be controlled namely, the heat-balance control and the material-balance control. Both methods have been used successfully in the past, however, the heat-balance method has been the more popular probably because of the lack of development of process stream analysers. Both the above schemes are conventional feedback control scheme. The heat-balance method is simply establishing a heat balance in the tower and accepting whatever product composition and volume results. The material balance method consists of setting the rate of removal of one of the product streams and providing a control scheme which will allow the heat and material balance to adjust automatically.

2) Method of Control or Adjustment of Manipulated Variables

Nowadays, the control schemes of Chemical industry can be divided in two main categories. The first is feedback control scheme and the second is feedforward control scheme. The feedback control scheme can be applied with the minimum amount of knowledge about the internal workings of the process beyond that of the proper variables to accomplish the end result. On the other hand, feedworward implies considerable knowledge of all the process characteristics. Simply speaking, feedforward control needs to have a mathematical model while the feedback does not. This model no matter how simple or complex it is can be generated either theoretically or from operating correlations.

Usually for tower control, if feedforward control system is applied, it is best used with the material balance method because it is easier to calculate the required flow rate of one product stream e.g. the distillate rate than the necessary heat addition or removal required for the heat-balance method. For the feedback control scheme, the heat balance method is more popular because it does not require a stream analyzer. This is especially true for the binary system at constant pressure operation, the sole effect on composition is the temperature. In this heat balance control scheme, the liquid temperature of a tray is maintained constant by manipulating either the reboiler heat input or the reflux rate.

If product compositions are not controlled directly, there will always be a certain amount of offset in the product purity when a disturbance enters and remains in the system. In order to eliminate this offset, product composition control should be introduced. For the conventional feedback control, this can be accomplished by cascading product analysis control loop around the tray temperature control loop. Initial responses to upsets are controlled with the tray liquid temperature and then the final adjustment made with a product analyzer. In this case, the dynamics of the short term are improved and the inaccuracy of the correlation is eliminated in the long term through adjustment of the setpoint of the controlled tray temperature by the analyzer.

The successful use of the feedforward control depends upon the accuracy of the required product flow calculation which is in term depended upon the accuracy of the mathematical model of the tower. Unless a perfect model is used, the tower products will have composition offset. An improvement on this type of control consists of adding a feedback loop containing a product stream analyzer which readjust the set product flow. This introduces a correction for the errors made in setting product flow rates and this makes the method more practical.

Usually, there will be an amount of product purity offset which is tolerable. If in evaluating the effectiveness of a control scheme the deviation is within product purity limits, the control scheme can be considered acceptable.

3) Control Schemes Study

In this work, two basic control schemes were studied, the conventional feedback heat-balance control scheme with first the temperature control tray located in the rectifying section and reflux rate as the manipulated variable, and second with the temperature control tray located in the stripping section and the reboiler heat input as the manipulated variable. In addition to the above, a simple feedforward scheme was studied. In this scheme the distillate withdrawal rate instead of the reflux rate was used as the manipulated variable. The mathematical model used in this scheme was a simple algebraic equation that related the feed stream rate to the distillate withdrawal rate

$$D = KF$$

where K was a function of feed, and top and bottom product compositions. An analytical method for both types of control schemes was derived in this work to study the performance of the control schemes. Besides the analytical study experimental work was carried out on the conventional feedback heat-balance control system in order to verify the analytical method. No experimental work was done on the feedforward control scheme because of the limitation of the equipment. Schematic diagram of the two control scheme are shown on Figure 1 and 2.

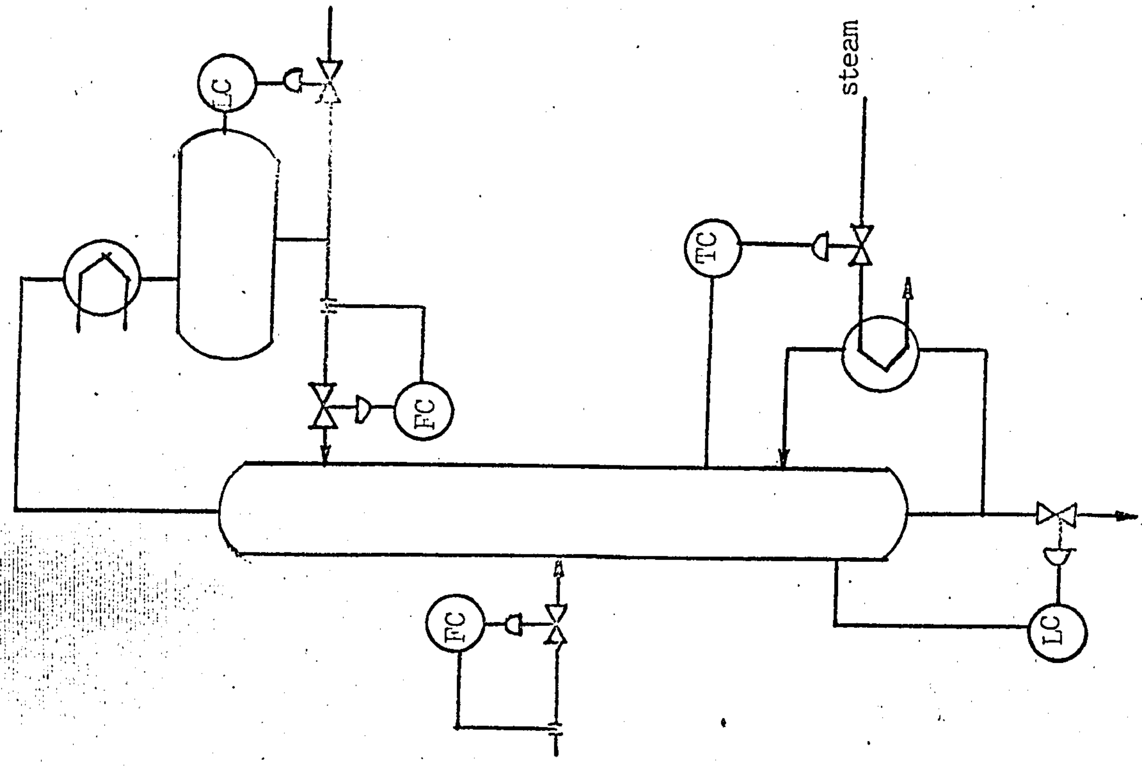
4) Basis for Comparison of Alternate Schemes

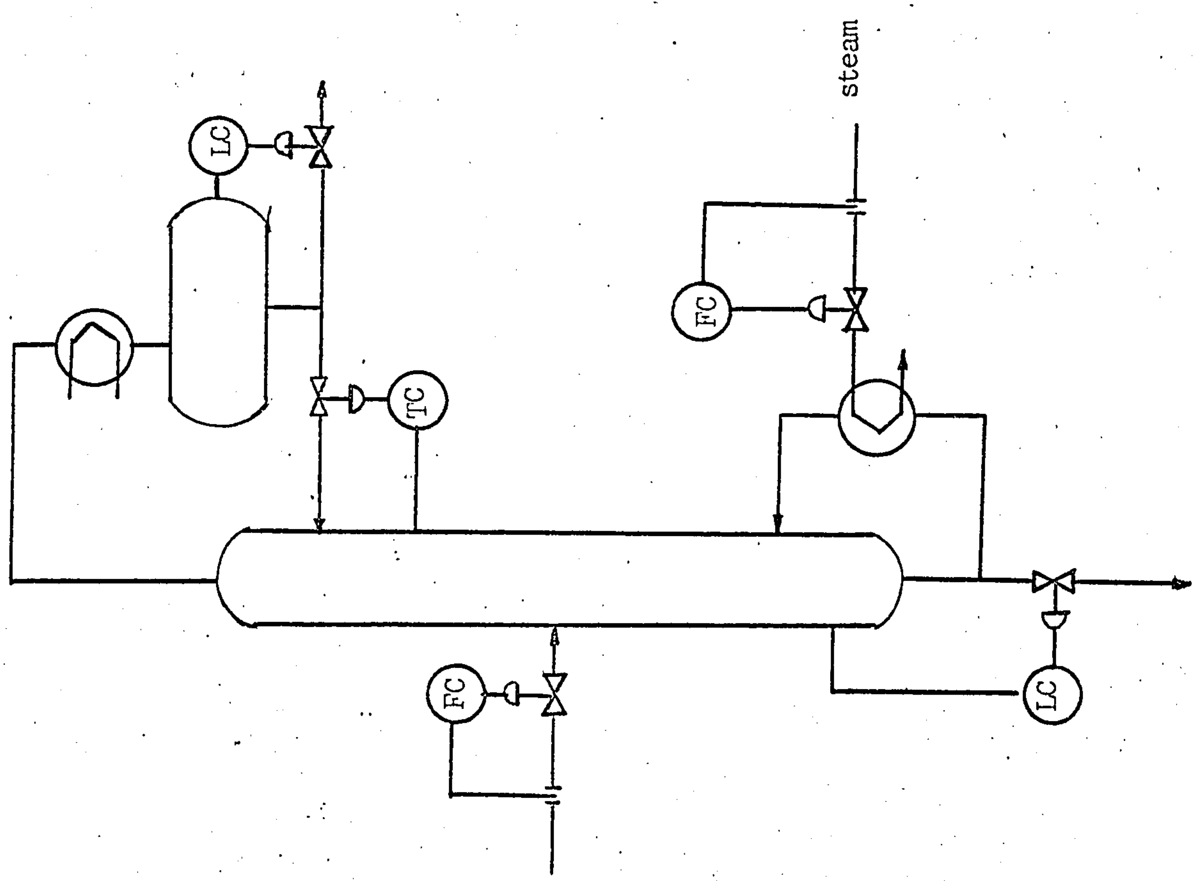
The steady-state product purity offset was used as a criterion to compare different type of control schemes. Of course, the smaller the amount of offset, the better performance the control scheme.

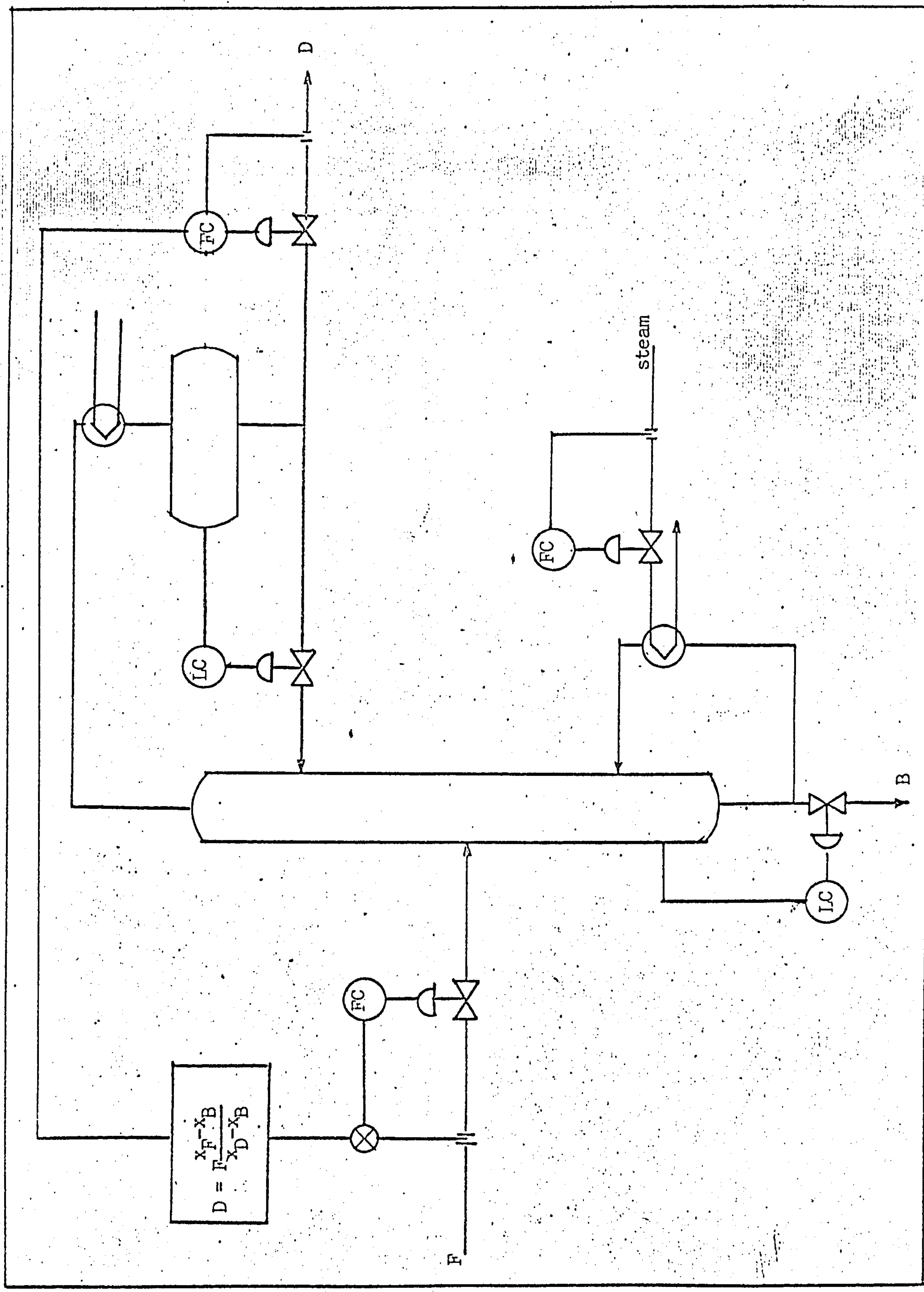
IV Calculation Method

Bauer and Orr (4) used the McCabe-Thiele diagram method to study some basic control schemes for a binary distillation column under feed composition disturbance. However, this method is considered by the author as tedious, time consuming and inaccurate. An analytical method using a high speed digital computer is preferable.

Smoker (30) derived an equation which is equivalent to the McCabe-Thiele diagram. There are some basic assumptions in the Smoker equation


LC
FC
TC
steam
FC
LC


LC
TC
steam
FC
FC
LC


D
FC
LC
steam
FC
B
LC
FC
$D = F\frac{x_F - x_B}{x_D - x_B}$
F

a) Constant molal overflow which includes

i) Constant liquid and vapor rates.

ii) The system components follow the Trouton's rule.

iii) Negligible heat loss.

b) Constant relative volatility

From the first assumption, the equation of any operating line (in either section of the column) can be represented by

$$y = mx + b \tag{IV-1}$$

From the second assumption, the equilibrium curve can be represented by

$$y = \frac{\alpha x}{1 + (\alpha - 1)x} \tag{IV-2}$$

The intersection of the operating and equilibrium lines gives the following equation

$$m(\alpha-1)x^2 + [m + b(\alpha-1) - \alpha]x + b = 0 \tag{IV-3}$$

From the above three equations and the technique of transformation of coordinates, the Smoker equation was derived. (30)

$$x'_n = \frac{m^n c^{2n} x'_o}{\alpha^n - mc(\alpha-1)[(\alpha^n - m^n c^{2n})/(\alpha - mc^2)]x'_o} \tag{IV-4}$$

$$n = \frac{\log \dfrac{x'_o\{1-[mc(\alpha-1)x'_n/(\alpha-mc^2)]\}}{x'_n\{1-[mc(\alpha-1)x'_o/(\alpha-mc^2)]\}}}{\log(\alpha/mc^2)} \tag{IV-5}$$

The Smoker equation must be applied to the rectifying and stripping section separately. For the rectifying section

$$m = \frac{R}{R+1} \qquad R = \frac{L_{N+1}}{D} \qquad x_o = x_D$$

$$x_o' = x_o - k \qquad x_n' = x_n - k \qquad c = 1 + (\alpha-1)k$$

where k is the root of equation (IV-3) between 0.0 and 1.0.

For the stripping section, the expression for m can be obtained from the known reflux ratio, feed thermal condition and feed and product composition as followed:

$$m = \frac{L'}{V'} = \frac{L + qF}{V - (1-q)F} = \frac{R + q + q\,(B/D)}{R + q - (1-q)(B/D)}$$

$$= \frac{Rx_F + qx_D - (R+q)x_B}{(R+1)x_F + (q-1)x_D - (R+q)x_B}$$

$$x_o' = x_o - k \qquad x_n = x_B$$

$$x_n' = x_B - k \qquad c = 1 + (\alpha-1)k$$

In the Smoker equation, the number of trays is very sensitive to the relative volatility of the system. When the value of α approaches one, the n value will become very large. The Smoker equation can be used not only to calculate the top, bottom compositions and the number of trays but also can be used to perform tray to tray calculation so that the composition of each tray can be evaluated.

With the help of the basic column material-balance equations, operating-line equations, q-line equation and the Smoker equation, an analytical method was developed to calculate the steady-state performance of the column control scheme when subjected to an operating disturbance.

V Calculation Procedures

1) Equations Used in the Calculations

a) Smoker equations and all the other equations that appear in the previous section.

b) Basic equations for distillation column

The over-all material balance equation

$$F = D + B \tag{V-1}$$

A material-balance for the lighter component

$$Fx_F = Dx_D + Bx_B \tag{V-2}$$

Substituting equation (V-1) into equation (V-2) gives

$$D = \frac{x_F - x_B}{x_D - x_B} F \tag{V-3}$$

and

$$B = \frac{x_D - x_F}{x_D - x_B} F \tag{V-4}$$

c) Operating line equations

An over-all material-balance for the rectifying section of a column gives

$$V = L + D \qquad \text{(V-5)}$$

for the lighter component

$$V\, y_{n+1} = Lx_n + Dx_D \qquad \text{(V-6)}$$

Dividing (V-6) by V, the following equation is obtained

$$y_{n+1} = \frac{L}{V}\, x_n + \frac{D}{V}\, x_D \qquad \text{(V-7)}$$

This is the operating line equation for the rectifying section. Similarly an over-all material balance for the stripping section of a column gives

$$V' = L' - B \qquad \text{(V-8)}$$

for the lighter component

$$V' = y_{n'+1} = L' x_{n'} - Bx_B \qquad \text{(V-9)}$$

Dividing (9) by V' gives

$$y_{n'+1} = \frac{L'}{V'}\, x_{n'} - \frac{B}{V'}\, x_B \qquad \text{(V-10)}$$

This is the operating line equation for the stripping section.

d) q-line equation

The quantity q is defined by the equation

$$L' = L + qF \qquad \text{(V-11)}$$

The q-line equation, on which the intersection of the two operating lines will fall is

$$y = \frac{q}{q-1}x - \frac{x_F}{q-1} \qquad \text{(V-12)}$$

The coordinates of the point of intersection must then satisfy the equation

$$y_1 = \frac{q}{q-1}x_1 - \frac{x_F}{q-1} \qquad \text{(V-13)}$$

Since x_1y_1 is a point on the two operating lines its coordinates must also satisfy the following equations

$$y_1 = \frac{L}{V}x_1 + \frac{D}{V}x_D \qquad \text{(V-14)}$$

$$y_1 = \frac{L'}{V'}x_1 - \frac{B}{V'}x_B \qquad \text{(V-15)}$$

In addition to the above equations, two other equations are needed, namely

$$\text{Slope of rectifying line} = \frac{L}{V} = \frac{x_D - y_1}{x_D - x_1} \qquad \text{(V-16)}$$

$$\text{Slope of stripping line} = \frac{L'}{V'} = \frac{y_1 - x_B}{x_1 - x_B} \qquad \text{(V-17)}$$

Some basic quantities must be set in order to carry out the calculation. Usually, the following variables are set before hand; the q-value, the feed composition, the top product composition, the bottom product composition, the relative volatility, reflux ratio and for an existing tower the number of theoretical trays and the feed location.

2) Basic Case for Study

A binary system of relative volatility equal to two, with the q-value equal to one was used as the base case. The steady-state top product purity was 95 mole % while the bottom product purity was 5.6 mole %. The feed composition was 50 mole %. The feed location was the sixth tray from the bottom. From the Smoker equation calculation, a column with 12.77 theoretical trays was needed.

3) Description of Calculation Procedures

From the values that have already been determined in the base case, a simple relationship can be established between the variables D, B, V, V', L, and L' with F e.g. from equation (V-3)

$$D = \frac{x_F - x_B}{x_D - x_B} F$$

Knowing the value of x_F, x_D, x_B, D can be related with F as D = aF where a is a numerical constant.

a) Feedback Heat-Balance Control Scheme

When the control tray was in the stripping section quantity control was applied to the reflux rate. The reboiler heat input was the manipulated variable. For feed composition disturbance the calculation was as follows.

i) First the location of the control tray was picked and then the tray composition was calculated using the Smoker equation Eq. (IV-4).

ii) Next x_B was assumed.

iii) The values of x_i, y_i were then assumed. Then, m, the slope of the operating line and the number of trays between x_B and the control tray were calculated using equations (V-17) (IV-3) and (IV-5).

iv) The trays calculated in step (iii) were then checked with the desired value and if they were equal the calculation proceeded to step V, if they were not, the calculation went back to step (iii) and was repeated.

v) Next the value of x_D was calculated using equation (V-18). This equation was obtained by combining equations (V-3), (V-5) and (V-16)

$$x_D = \frac{x_B(y_i - x_i)L - y_i(x_F - x_B)F}{(y_i - x_i)L - (x_F - x_B)F} \qquad \text{(V-18)}$$

vi) The liquid composition of the feed tray was calculated using x_B and the total number of trays in the stripping section and equations (V-17) and (IV-4).

vii) Using the values calculated in steps (v) and (vi) the number of trays in the rectifying section was calculated.

viii) The total number of trays was compared with the desired number. If they were equal the problem was ended, if not the calculation went back to step (ii) and was repeated.

In the case for the feed rate disturbance, the calculating procedure was the same except for the use of equation (V-18) step (V). In this equation the feedrate F was multiplied by one plus the fraction increase or decrease in the feed rate.

When feed enthalpy was the disturbance the relationship between x_1 and y_1 must satisfy equation (V-13); i.e. one of x_1 or y_1 was assumed, and the other calculated. The calculation steps were all the same as the feed composition disturbance except the x_F never changed. The variable that did change was the q-value.

When the control tray was in the rectifying section reboiler vapour rate was quantity controlled. The iteration variable for step (ii) was x_D and equation (V-19)

$$x_B = \frac{x_1 (x_D - x_F) - x_D (y_1 - x_1)V'}{(x_D - x_F) F - (y_1 - x_1)V'} \qquad \text{(V-19)}$$

was used in step (v). The calculation procedure is almost the same except for steps (v), (vi) and (vii) where x_B and the number of trays in the stripping section were calculated.

b) Feedforward Control Scheme

The calculation involved here was simpler than that in the feedback control scheme. The reflux ratio was the manipulated variable. Since the reflux rate was kept constant in this scheme. It turned out that the distillate withdrawal rate was the actual manipulated quantity.

The steps of the calculation are listed as follows

i) $$D = \frac{x_F - x_B}{x_D - x_B} = aF$$

The value of x_D, x_B remained constant no matter what disturbance the column was subjected to. For feed composition disturbance only; F was constant, therefore x_F was the only variable in the above equation.

ii) $$R = \frac{L}{D} \quad \text{where} \quad L = dF$$
$$D = aF$$

L was the quantity that must be maintained constant throughout the calculation so

$$R = \frac{dF}{aF} = \frac{d}{a}$$

iii) Next a value of x_D was assumed.

iv) By combining the equations in step (i) and (ii) an expression for x_B was obtained

$$x_B = \frac{x_F - a\,x_D}{1 - a}$$

v) Utilizing Smoker equation (IV-4) and the number of trays in the rectifying section, the composition of the feed tray was calculated.

vi) With the values from step (iv) and (v) and Smoker equation (IV-5) the number of trays in the stripping section was calculated.

vii) The total numbers of trays was then checked with the desired one. If they were identical then the problem was finished. If not, the calculation went back to step (iii) and was repeated.

Calculation for the feed rate disturbance required some changes in the beginning of the calculation.

1) $$D = aF \qquad R = \frac{dF}{aF}$$

L was kept constant in the calculation, the numerator in the above expression was constant, any change in F would change the denominator only. For this reason, the expression for R was regarded as a constant over a variable quantity which was the disturbance in feed flow rate. The calculation procedures were the same from step (iii) of the feed composition disturbance calculation.

VI Trial and Error Method

The first important item in the trial and error calculation was the choice of iteration variables. For feedback heat-balance control, it depended on the location of the control tray. If the control tray was in the stripping section, the bottom product purity was the iteration variable. When the control tray was in the rectifying section, the top product purity was the iteration variable. These two can be called the primary iteration variables, where the intersection of the q-line and the equilibrium line were regarded as the secondary iteration variable. In case of the feedforward control scheme, the top product composition was the one and only iteration variable and the calculation was much simpler.

Feedback control scheme with the control tray located in the stripping will be used as an example to illustrate the trial and error method.

1) Limits of Iteration Variables

There are two limiting cases for the operation of a distillation column, total reflux and minimum reflux. Total reflux means no withdrawal of the top product, i.e. all the overhead material is returned to the column as reflux. This gives the maximum separation of components in the column. On the other hand, minimum reflux implies an infinite number of stages to obtain a certain separation which in turn implies a minimum of component separation. Usually, a column will be operated between these two limiting conditions and therefore the end product purities must lie between the values of these two extreme cases.

The bottom product purities for these two cases were calculated. This was done by using the Smoker equation the tray number between the control and bottom trays and the conditions for these two limiting cases. When these two values were established, calculation was begun. The secondary iteration variable was the intersecting point of the operating line and the q-line. From the McCabe-Thiele diagram the intersection must be between the equilibrium and the 45° lines. These give the upper and lower limits of this iteration variable. Intersection of the q-line with the 45° line and the equilibrium line were calculated and then the secondary iteration variable was chosen between these two limited values.

2) Method of convergence

The bisectional method was used for both iteration variables. According to the calculation procedure the inner loop must converge first before the problem can be completed. The first convergence was the match between the calculated value of the number of trays between the control tray and the end of the column and the desired one. This match was obtained by setting a value of bottom product composition and using the $x_i y_i$ (intersection of q and operating lines) as the iteration variable. For the basic case under study, the value of x_i was fixed and always equal to the value of the feed composition because the q-value was equal to 1 and q-line was always making a 45^o angle with the 45^o line in the McCabe-Thiele diagram. This simplified the calculation to iterate the y_i value only. If the calculated tray number in this loop was smaller than the desired one, a larger value of y_i was chosen next and vice versa. When the value of y_i reached its upper limiting value and the convergence had not been achieved, a smaller value of the bottom product composition was used and the inner iteration was repeated again until the calculated value matched the desired one. In this convergence step, the acceptable mismatch between the calculated and desired number of trays was very small. Throughout the cases that have been studied, the tolerable error was specified as between 1×10^{-5} to 5×10^{-5}. If not, the error would be amplified in the later calculation and give an unacceptable result.

After the calculation had passed through the first convergence, a value of the top product purity was calculated using equation (V-18). Using this information together with the known number of rectifying trays, the liquid composition of the feed tray was obtained. Then the number of stages of the stripping section was calculated and added to that in the rectifying section to check with the desired total number of stages. If the calculated value exceeded the tolerable error and was smaller than the desired one, a larger value of bottom product composition was needed and vice versa. The tolerance in this final step was larger than in the first one. A value of about 10^{-2} was used. From experience it was found that the results were accurate enough using this value. A smaller acceptable error would improve the final result very little and waste more extra computing time was not justified.

The first converging step was slower than the second because it demanded higher accuracy. Usually, convergence could be obtained after ten to fifteen complete trials. The time of convergence was also affected by values of relative volatility, q-value , total number of trays location of the control tray and the reflux ratio. Throughout the calculation, it seemed that higher reflux ratio was more favorable to the second converging step. When the control tray was nearer the feed tray, the convergence was slower. A feed composition disturbance required more time to converge than the feed rate disturbance.

The trial and error method was by comparison simpler for the feed-forward control scheme. The top product composition was the only iteration variable. There were two ways of choosing the first iteration value. It could be started using a top product composition equal to 1 which was higher than the final value. This composition was then decreased by a certain amount in each iteration until the convergence was obtained. The other way was to choose a value which was smaller than the one approximated from the McCabe-Thiele diagram or from experience. In this work, the initial value of 0.8 was chosen. This value was then increased by a certain amount in every iteration until the desired result was obtained. The time required for convergence was dependent upon the increment of the iteration variable. The larger this value, the faster the convergence. The program never converged if the increment was too large. An optimal value of this increment was chosen by trial and error.

The factor such as type of disturbance, reflux ratio etc. did not affect the convergence of the feedforward control scheme so much as in the feedback control scheme because of the simplicity of this scheme.

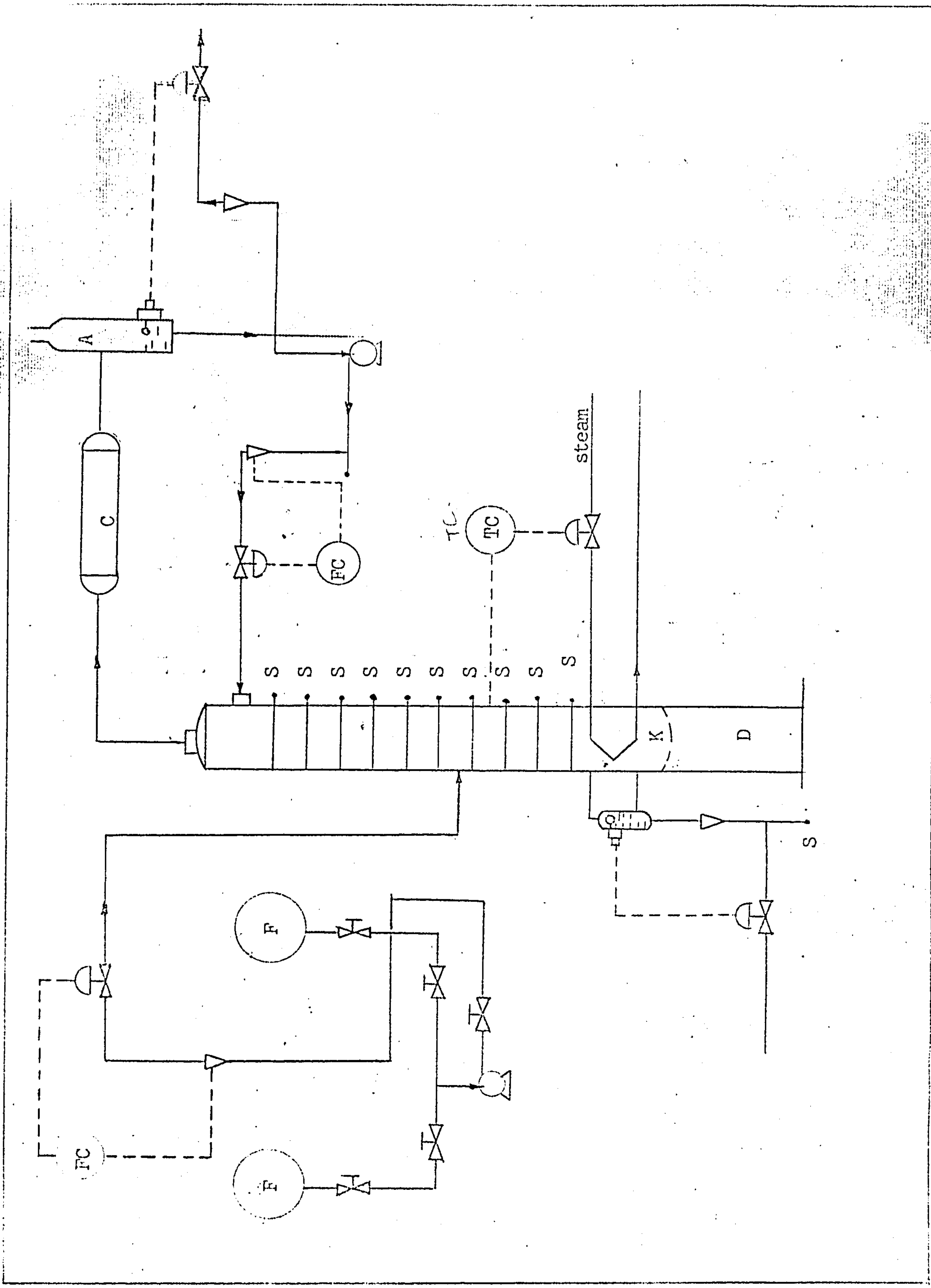
All the calculation in this work were programmed using Fortran II and IV language and solved on IBM-360 and IBM-1620 digital computer. Double precision was used in the calculation for accuracy. A listing of the computer programs and the results of the calculations are included in the appendix for reference.

VII Experimental Work

The object of this experimental work was to check how well the proposed calculation method agreed with actual column operation and to test the performance of the feedback heat-balance control scheme. No attempt was made to run the column under feedforward control because of the limitation of equipment.

1) Description of Apparatus

The apparatus used in this experiment was a pilot-scale laboratory distillation column. A schematic diagram of the equipment is shown in figure (3).


A
C
FC
TC
TC
steam
S
K
D
F
F
FC

The column was constructed of Q.V.F. 9 inch glass pipe with internal diameter of 8.38 inches. The column contained ten bubble-cap trays, each tray with two copper bubble caps. The bubble cap risers were $1 - \frac{7}{8}$ inches in diameter. The circular downcomers were 2 inch copper pipes. The spacing between trays was one foot. There were two stainless steel drums in the feed system for feed storage. A Foxboro model 52A consotrol controller, a Brooks model 5500 indicating pneumatic transmitter and a Brooks full-view rotameter were used to control the feed flow rate. Feed to the tower was introduced into the circular downcomer from the tray above. In this work, the feed was introduced into the column at the fourth tray from the bottom.

The overhead system consisted of a total condenser, and accumulator. The condenser was of copper construction and contained two shell and tube heat exchanges. Water was used as the condensing medium. The rate of cold water was manipulated using a manual valve and a rotameter. No attempt was made to control the temperature of cold water. The accumulator was a copper cylinder with diameter of 18 inches. The distillate withdrawal rate was controlled by the liquid level of the accumulator. A Honeywell liquid level transmitter, A Foxboro model 52A consotrol controller and a Brooks full view rotameter were installed for this control action.

The bottom system consisted of a thermosyphon reboiler which was a shell and tube heat exchanges mounted on the bottom of the column. Steam was used as the heating medium. The steam rate was controlled by a Foxboro model 52A consotrol temperature controller and Foxboro EMF/pneumatic

Table 1

LEGEND FOR EXPERIMENTAL APPARATUS

Letters

A	-	Accumulator
C	-	Condenser
D	-	Distillation column
F	-	Feed tank
FC	-	Flow rate controller
K	-	Reboiler
LC	-	Liquid level controller
S	-	Sample outlet
TC	-	Temperature controller

Symbols

	-	Manual valve
	-	Air operated control valve
	-	Pump
	-	Rotameter
	-	Pipe line
	-	Sample point
	-	Signal flow line

temperature transmitter. The steam rate was changed automatically in order to maintain the particular control tray temperature constant. The bottom product was withdrawn using bottom liquid level control, a Foxboro type 15A d/p cell transmitter, a Foxboro model 52-A controller and a Brooks full view rotameter was used for this purpose.

The reflux was under flow control. If the control tray was in the rectifying section, the temperature of that tray would manipulate the reflux flow rate. If not, the reflux flow rate was maintained constant. A Brooks model 5500 indicating pneumatic transmitter, a Brooks full view rotameter and a Foxboro model 52-A consotrol controller was responsible for the control of the reflux rate.

Two centrifugal pumps were installed for pumping the feed and the reflux into the column. The tray temperatures were recorded using a Honeywell type 153 electronik multipoint recorder. All the thermocouples were made of copper-constantan. An Acromag 330 series thermocouple references was used for the reference junction. A Foxboro model 53 consotrol recorder with three channels was used to record the feed flow rate, reflux flow rate and the temperature of the control tray. All the control valves were bought from Honeywell Controls Limited. The transmitter and the control and recording elements except the multipoint recorder were pneumatically operated.

The system used in this work was methanol-water. The methanol was the grade A-412 Fisher certified A.C.S. from Fisher Scientific Co.

A Westphal balance was used to analyze the end product purities. A direct reading of the specific gravity to the fourth decimal place of methanol could be obtained from this instrument.

2) Experimental Procedures

Methanol-water system was used in all experimental runs. The normal feed composition was 12 wt % (7.2 mole %). The first run was a total reflux run. The object was to obtain an approximate temperature profile of the column.

a) Control Tray in the Stripping Section

Using the second tray from the bottom as the control tray, several runs on feed composition and feed rate disturbance were performed. The control tray was then moved to the third tray from the bottom. Only feed composition disturbances were studied for this case.

The following is an outline of the startup procedure.

i) The equipment was checked out and the condenser cooling water turned on. The feed temperature was then measured.

ii) Feed was then pumped into the column until the reboiler was flooded. Steam was then admitted into the reboiler and the feed pump was shut off. When the liquid inside the column started to boil, the feed pump was started and the feed rate was maintained constant by an automatic flow controller.

iii) The steam flow rate to the reboiler was controlled by the temperature of the control tray - second from bottom - which was set equal to 204°F. This temperature was maintained throughout the entire run.

iv) The reflux pump was started when the liquid in the overhead accumulator was built up to a certain level. The reflux rate was then maintained at a small value and increased gradually to the desired one. During startup, manual control was applied to the reflux flow controller and the level controller which controlled the distillate withdrawal rate.

v) When the level in the overhead accumulator remained fairly constant, the reflux flow rate was increased to the desired value and the level and flow controller switched to automatic control.

vi) When the temperature profile of the tower showed little changes, steady-state was assumed to be obtained. Top and bottom products were withdrawn and analyzed with a Westphal balance.

vii) A disturbance was introduced and again when steady state was reached, top and bottom compositions were withdrawn and analyzed.

When the control tray moved to the third tray from the bottom, all the experimental procedures were the same except the temperature of the control tray was set at 200°F.

b) Control Tray in the Rectifying Section

Reboiler heat input was set constant and the reflux rate was controlled by the temperature of the control tray which was in the rectifying section.

In this work, the third tray from the top was chosen as the control tray and the temperature was maintained at 176°F.

Due to the shortage of equipment, automatic control of steam rate was impossible. In order to maintain a constant steam rate, the upstream pressure of the steam line was kept fairly high at about 45 psi. Then a relative low pressure input at about 6 psi was introduced into the reboiler. In this way, the steam rate could be assumed constant as long as no other disturbance existed in the steam line.

All the experimental procedures were the same as when the control tray was in the stripping section.

VIII Discussion of Results

The base case used in this study was a system with the following specifications

a) relative volatility = 2

b) q = 1

c) Feed composition = 50 mole % of lighter component.

d) Total number of theoretical tray in the column = 12.77 of which 6.18 trays were in the rectifying section and 6.59 trays in the stripping section.

e) Feed enters the column at the sixth tray from the bottom.

f) The steady-state top product purity was 95 mole % where the bottom purity was 5.6 mole %.

All the calculated results are tabulated in appendix some of them are shown graphically in the following figures.

1) Feedback Heat-Balance Control Scheme

A fairly detailed study of the base case was carried out. Every tray except the feed tray was simulated as the control tray in this scheme. Cases for feed rate, composition and enthalpy disturbance were studied separately. Systems with relative volatility equal to 1.4, 1.6, 4, 4.88, 6 were also studied. However, in these cases, not every tray was simulated as the control tray. The control tray was just picked at random.

The reason for simulating a system with α = 4.88 was because this is the estimated average relative volatility of the methanol-water system used in the experimental work. The results were used for comparison of the calculated and experimental results.

Sixteen experimental runs were carried out in the column. All of these runs were under the heat-balance control scheme. No attempt was made to run the column under feedforward control because of limitation of the apparatus. Most of the experimental runs were the cases where the control tray was in the stripping section and of feed composition disturbance. This was because the feed composition disturbance required a more intensive study. In the experimental runs, there was no control of the cooling water and the feed temperatures. Therefore, the experimental base case steady-state end compositions were different from run to run although the feed composition as well as the feed rate did not vary in each case.

As a result of analytical study, the following points were observed:

a) The deviations of both end purities caused by feed rate disturbance were the largest among the three types of disturbance under study.

b) The deviations caused by feed rate and enthalpy disturbances followed a regular trend.

c) The deviations caused by feed composition disturbance acted more randomly. They were a strong function of control tray location and quantity of disturbance. When the control tray was in the rectifying section, the top product composition responded more regularly to the disturbance, but the bottom composition did not. When the control tray was in the stripping section, the bottom product composition responded more regularly than the top. Irregularities of the end product responses were more prominent when the control tray was nearer the feed tray.

d) For top product purity control, the best tray to control should be the first tray from the top if time lag were not taken into consideration. The best tray for controlling the bottom product purity should be the first tray from the bottom.

e) If both end compositions are to be controlled, the control tray should be located in the stripping section rather than in the rectifying section.

2) Feedforward Control Scheme

No experimental work was done on this part of the work. Both feed composition and feed rate disturbance were studied analytically.

This simple feedforward control scheme handled the feed composition disturbance quite satisfactorily. From the calculated result, this scheme was good for either end product purity control. For feed rate disturbance, this control scheme did not give as good a result. The deviations in both ends were much larger especially in the bottom product purity. Also, the two end compositions deviated according to a regular trend whether feed rate or feed composition disturbance was introduced.

3) Comparison of the Two Control Schemes

Feedback heat-balance control scheme could handle the rate disturbance better than the feedforward control no matter where the control tray was located. For control of both end compositions the feedback scheme was still the better one.

For feed composition disturbance with the top product as the control criterion, the simple feedforward control scheme gave a better result than the feedback one provided that the control tray of the latter was in the stripping section. There was not much difference for both schemes when the control tray was in the rectifying section for the feedback control scheme.

For the control of bottom product composition, feedback control with control tray in the stripping section was superior than the simple feedforward one. In the case when the control tray was in the rectifying section, the simple feedforward control was the better one.

For the control of both end compositions, the simple feedforward scheme seemed to be the more acceptable one.

4) Effect of Relative Volatility to the Steady-State Product Purities

System with higher relative volatility usually had larger deviations in the steady-state product purities when subjected to the feed rate or composition disturbances. The site of the control tray had more influence on the end product composition in the higher relative volatility cases. This was due to the fact that in the higher relative volatility cases the composition changes across a theoretical tray was higher than in the low relative volatility cases.

In general, rate disturbance had a greater effect on end product purities than composition disturbance. However, in some cases for high relative volatility and the location of the control tray was located in the stripping section, the deviation of the bottom product purity was greater for composition disturbance than rate disturbance.

5) Comparison of Calculated and Experimental Results

A system with relative volatility of 4.88 was simulated. The value 4.88 was the estimated average relative volatility of the methanol-water system used in the experimental work. Most of the experimental results agreed qualitatively with the calculated ones. The amount of deviation in the experimental work was slightly larger than those predicted from calculations.

Small discrepencies between these two results were expected, because of the simplifying assumptions used in the calculation method. Besides these assumptions, factors such as tray efficiency, the variation of tray efficiency with loading etc. would also attribute to the observed deviations.

Steady-state was difficult to achieve in the experimental work. All tray temperatures on the column oscillated with almost the same period of oscillation. However the amplitude of the oscillation was larger for the trays located in the center portion of the column. This phenomena could be called "quasi-static" as introduced by Rademaker and Rijnsdorp (25). These oscillations might have been caused by a number of factors;

a) Variation of heat input because no automatic control was applied to the reboiler steam rate.

b) The reflux flow control instrument could have required finer tuning.

c) A possible hydraulic instability in the column itself.

The last source is included because the amplitude of the cycling seemed to be a function of the disturbance. The cycling became more serious when the feed flow rate or composition had increased in quantity and vice versa.

IX Conclusions

1) Feedback Heat-Balance Control Scheme

This control scheme studied is generally suitable for one end product composition control. No matter which end product is going to be controlled. The closer the tray to the end of the column, the better the product purities. For top product purity control, the best tray would be the top tray. If both end product purities are to be controlled, the best location

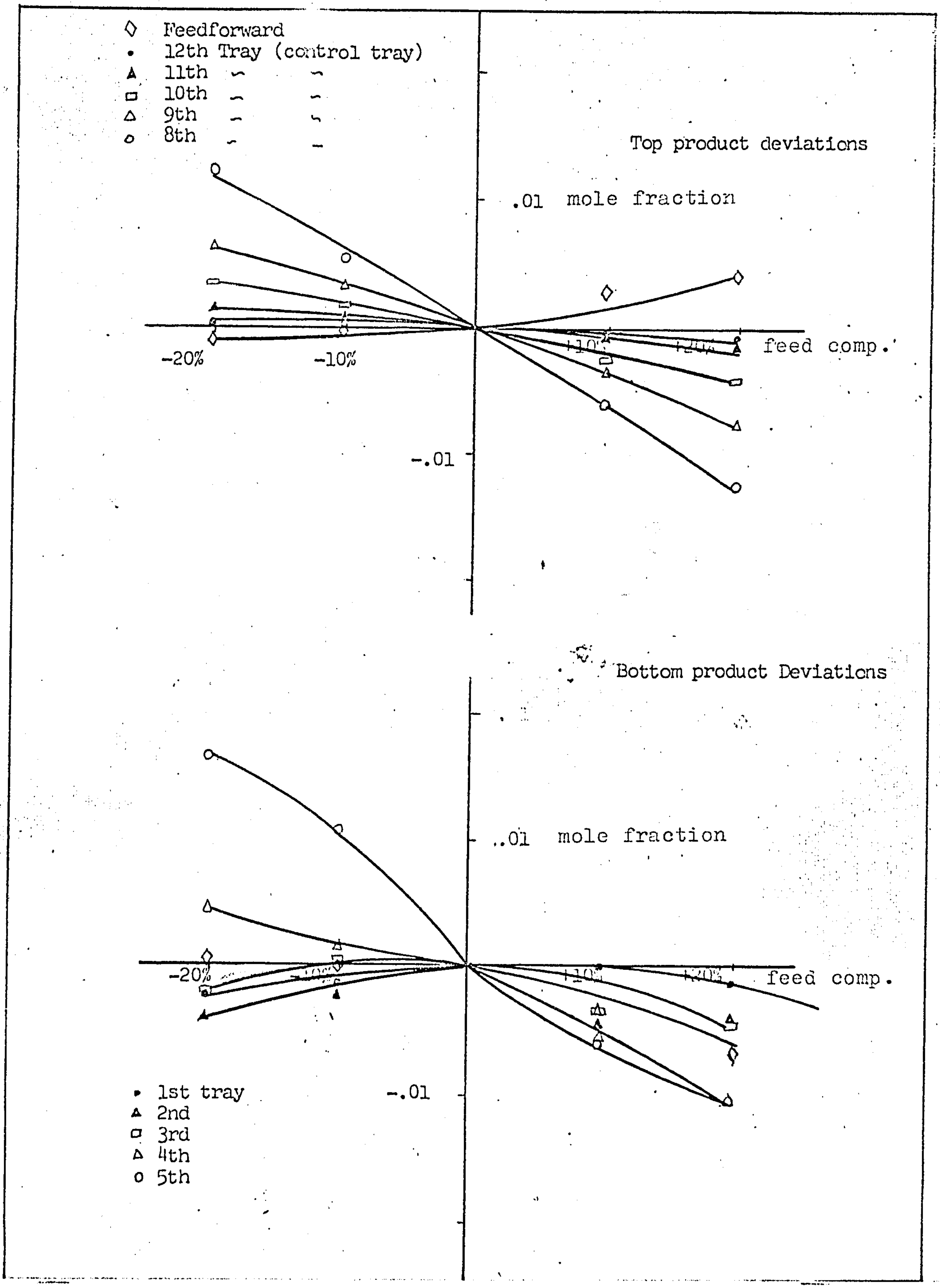


Figure 4. Deviations in x_D and x_B to changes in x_F for α=2 system with control tray in rectifying section and feedforward control

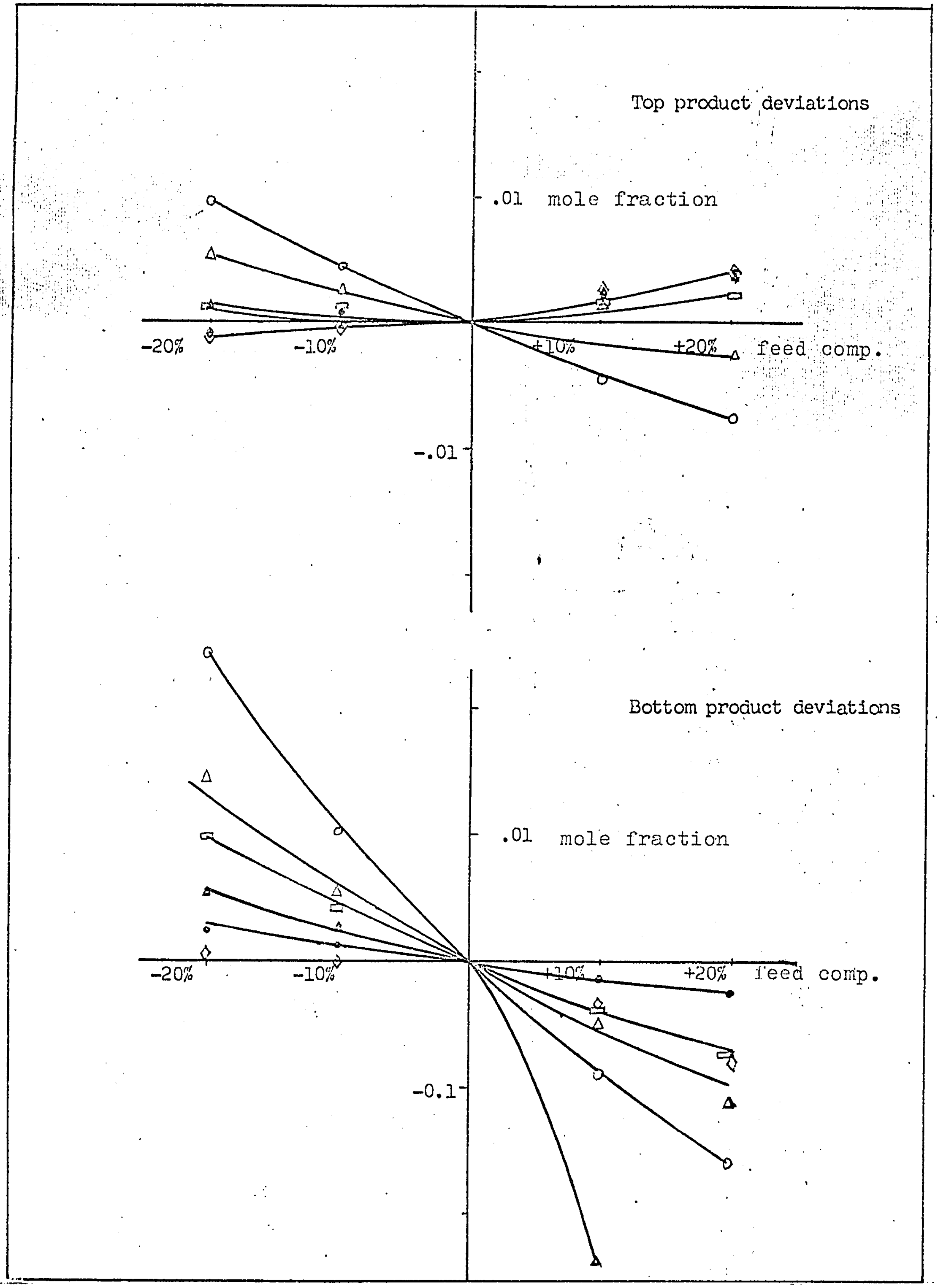


Figure 5. Deviations in x_D and x_B to change in x_F for α=2 system with

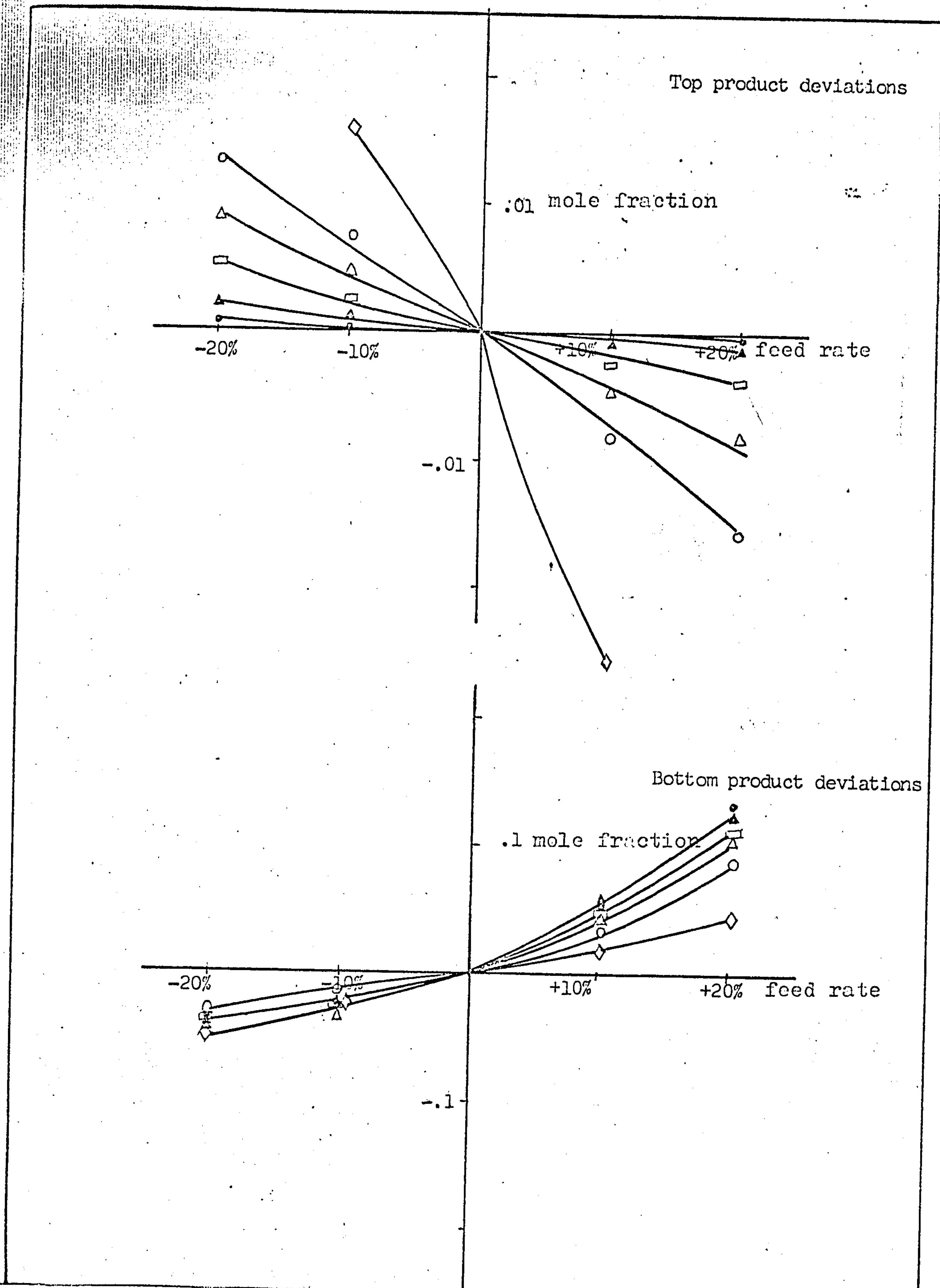


Figure 6. Deviations in x_D and x_B to changes in F for α=2 system with

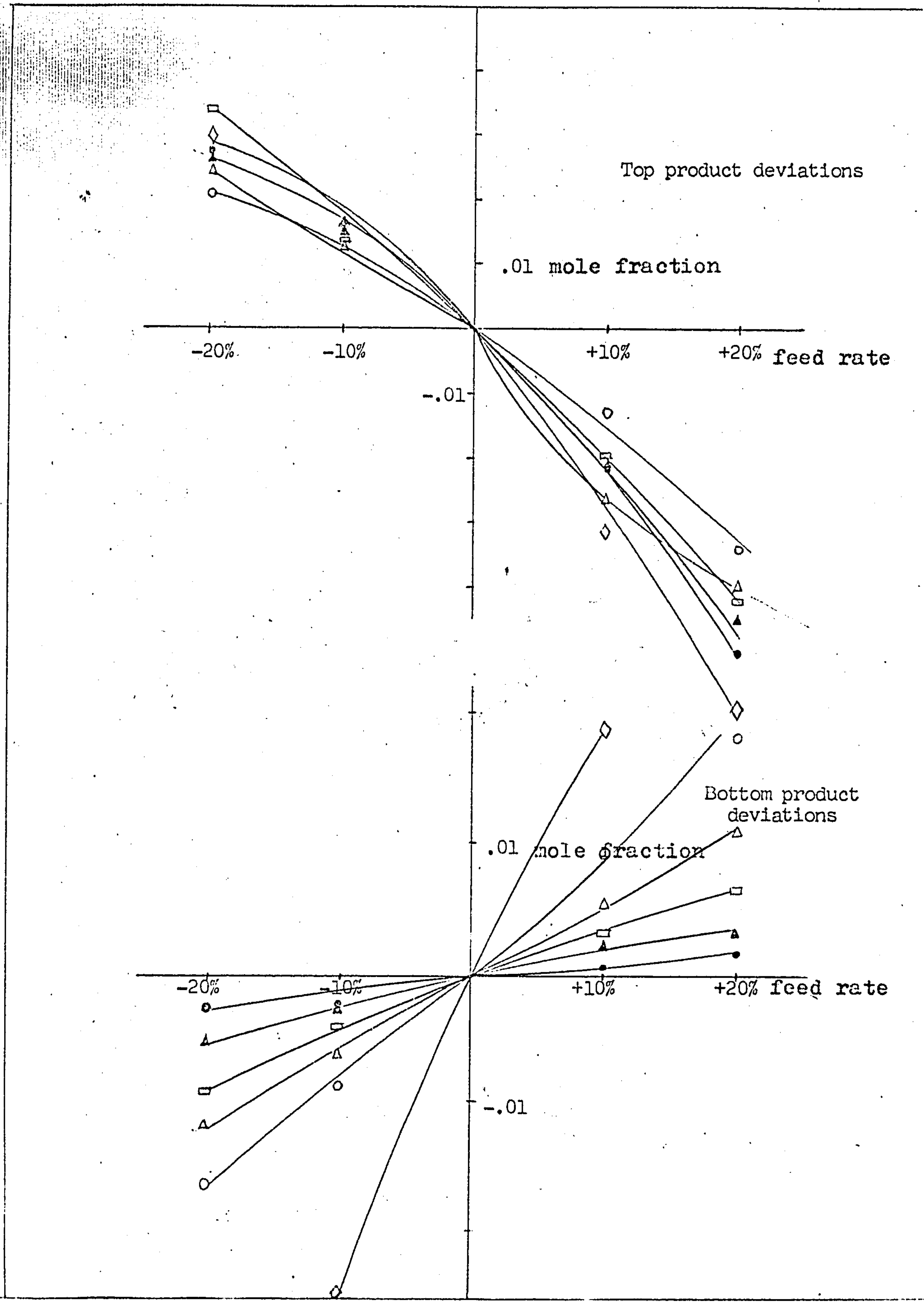


Figure 7. Deviations in x_D and x_B to changes in F for α=2 system with control tray in stripping section and feedforward control

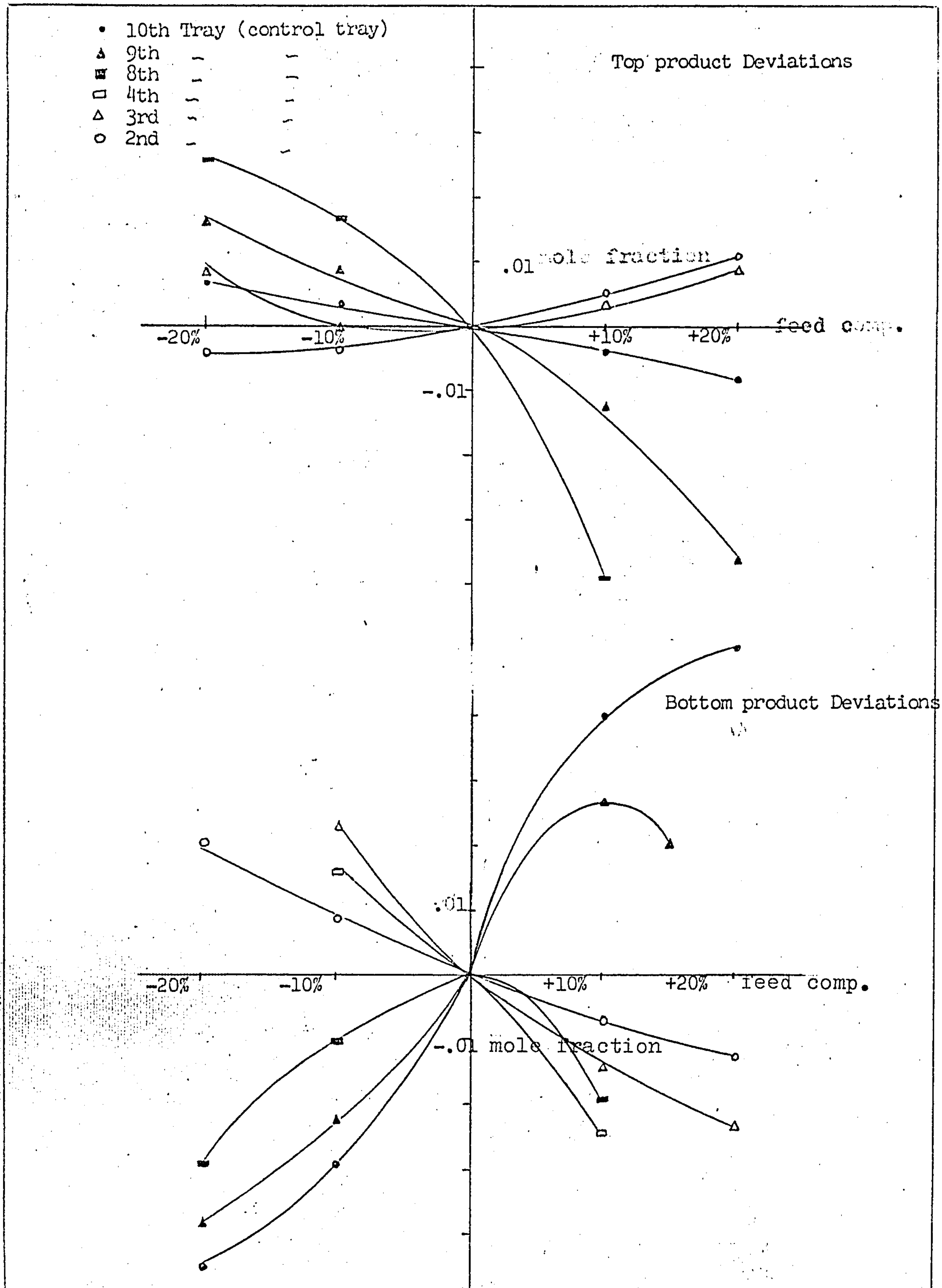


Figure 8. Deviations in x_D and x_B to changes in x_F for $\alpha=4$ system with control tray in both sections

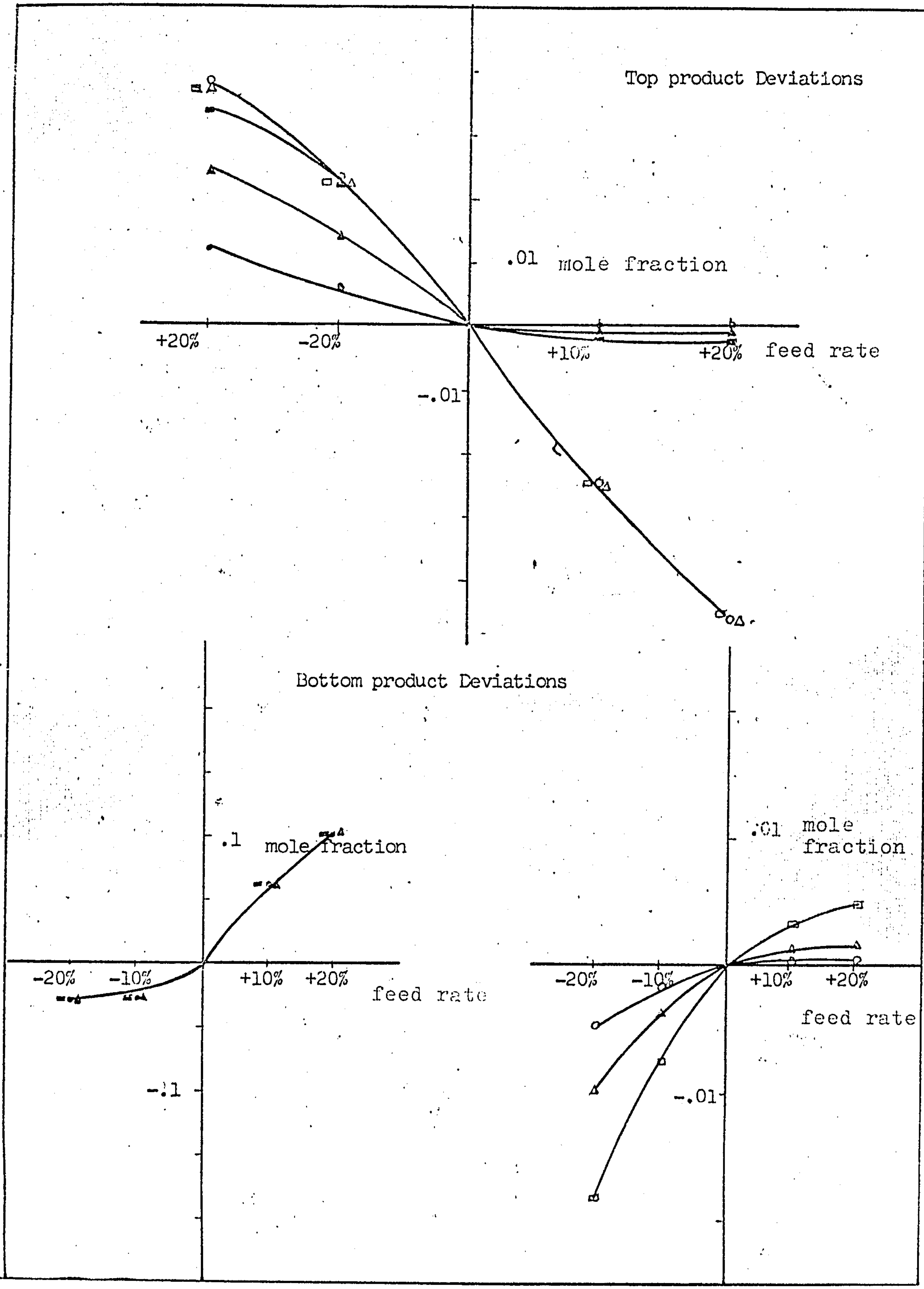


Figure 9. Deviations in x_D and x_B to changes in F for $\alpha=4$ system with control tray in both sections

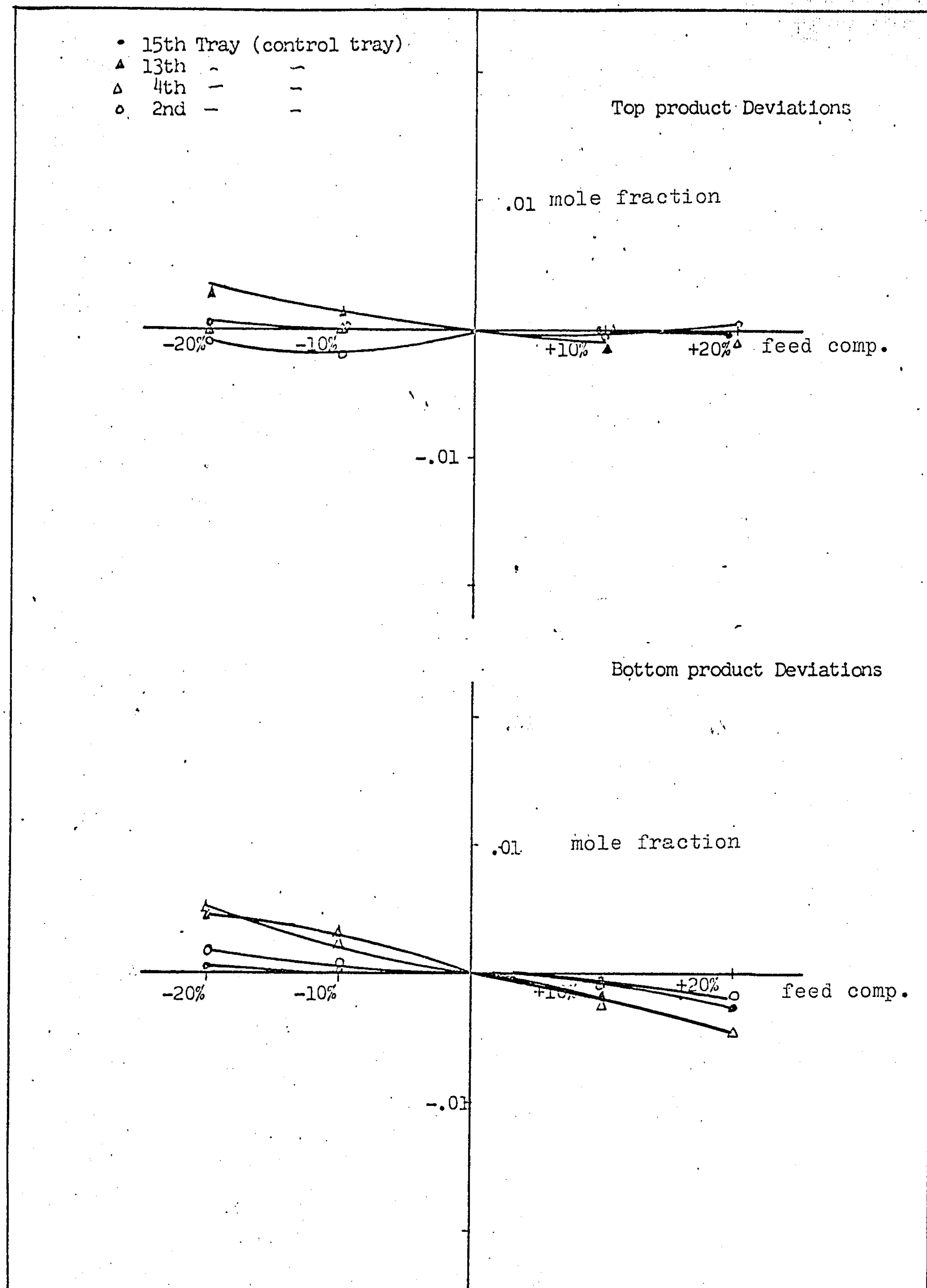


Figure 10. Deviations in x_D and x_B to changes in x_F for α=1.6 system with

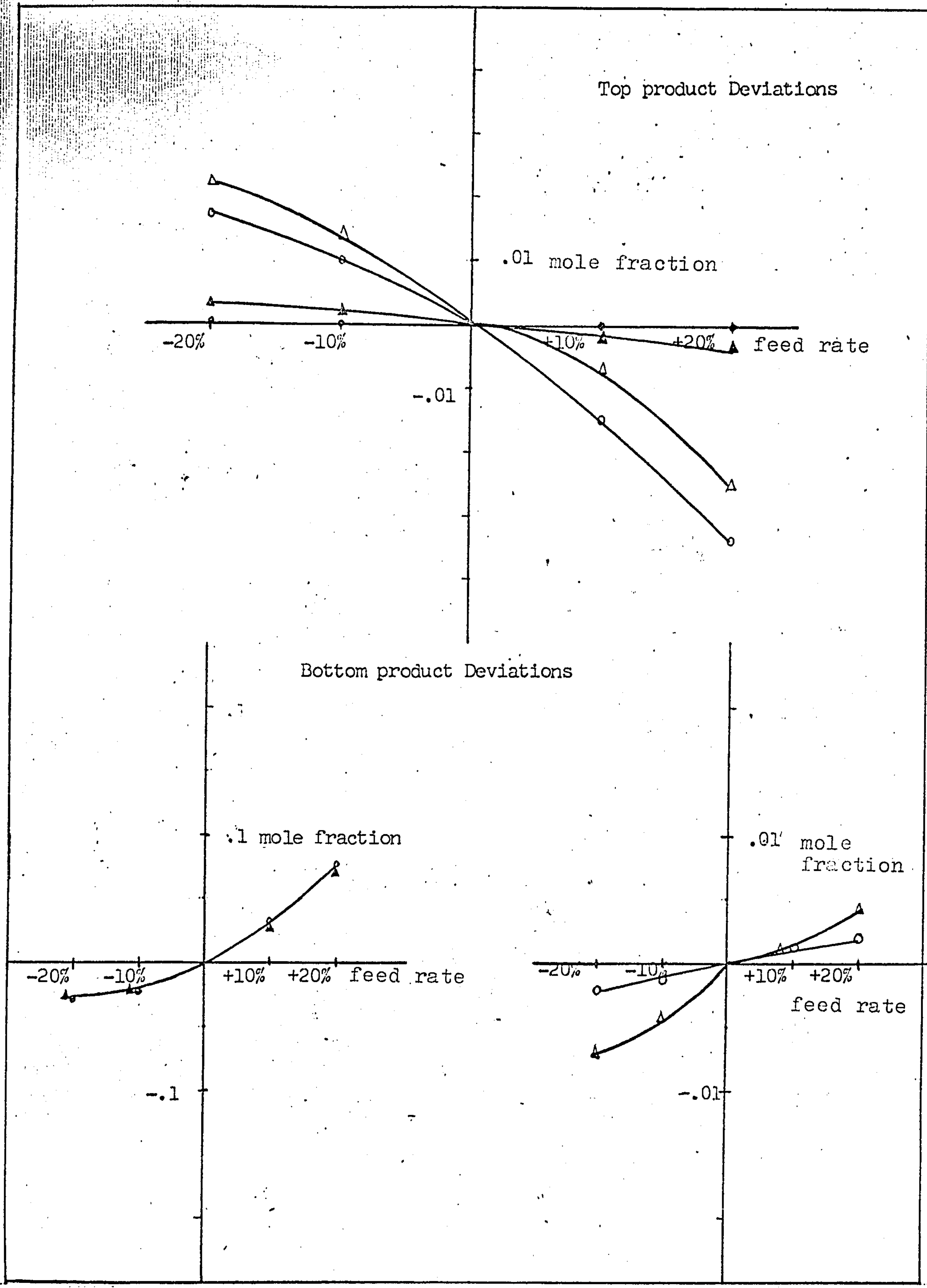


Figure 11. Deviations in x_D and x_B to changes in F for α=1.6 system with control ... in both ...

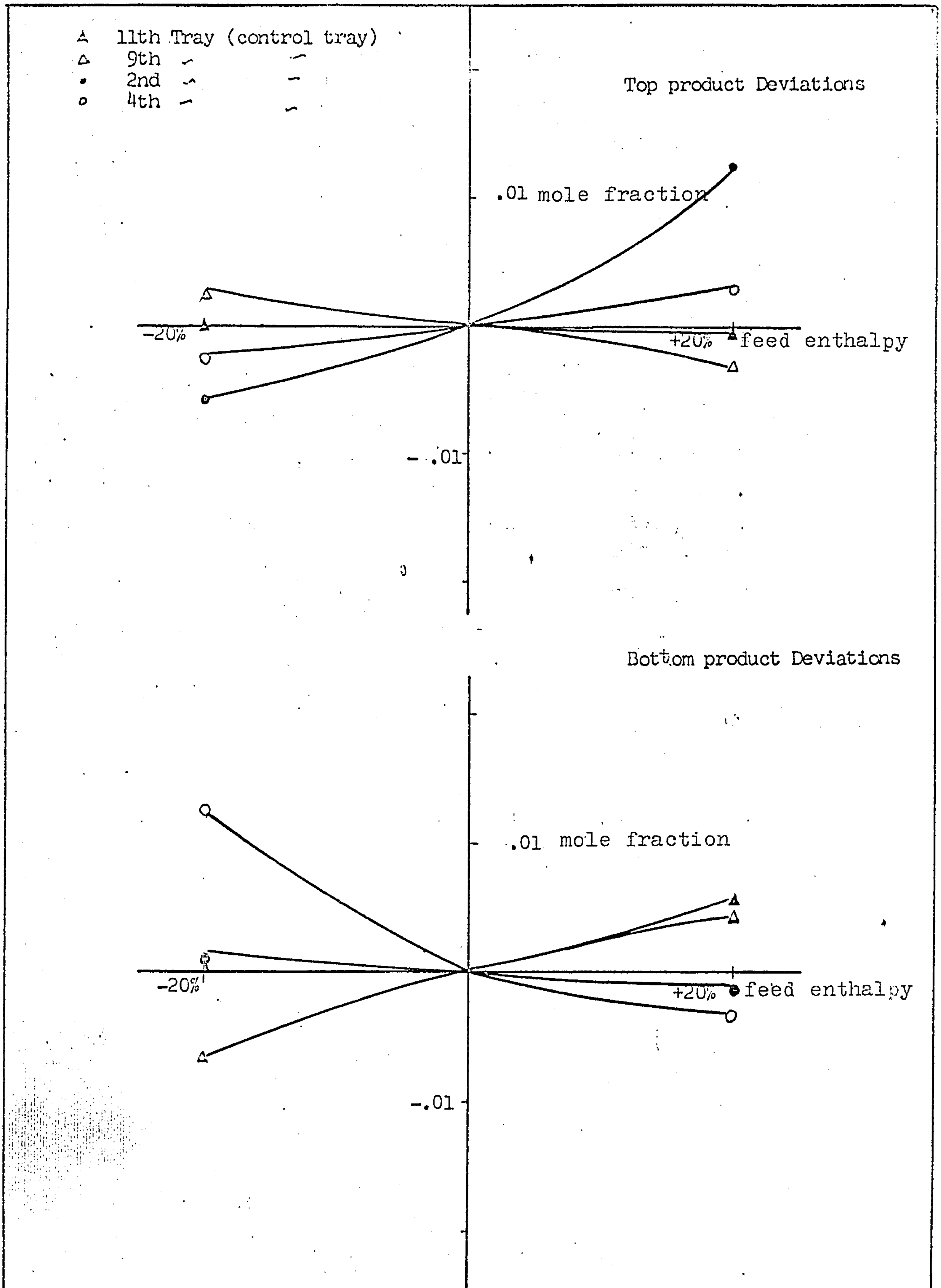


Figure 12. Deviations in x_D and x_B to changes in q for α=2 system with

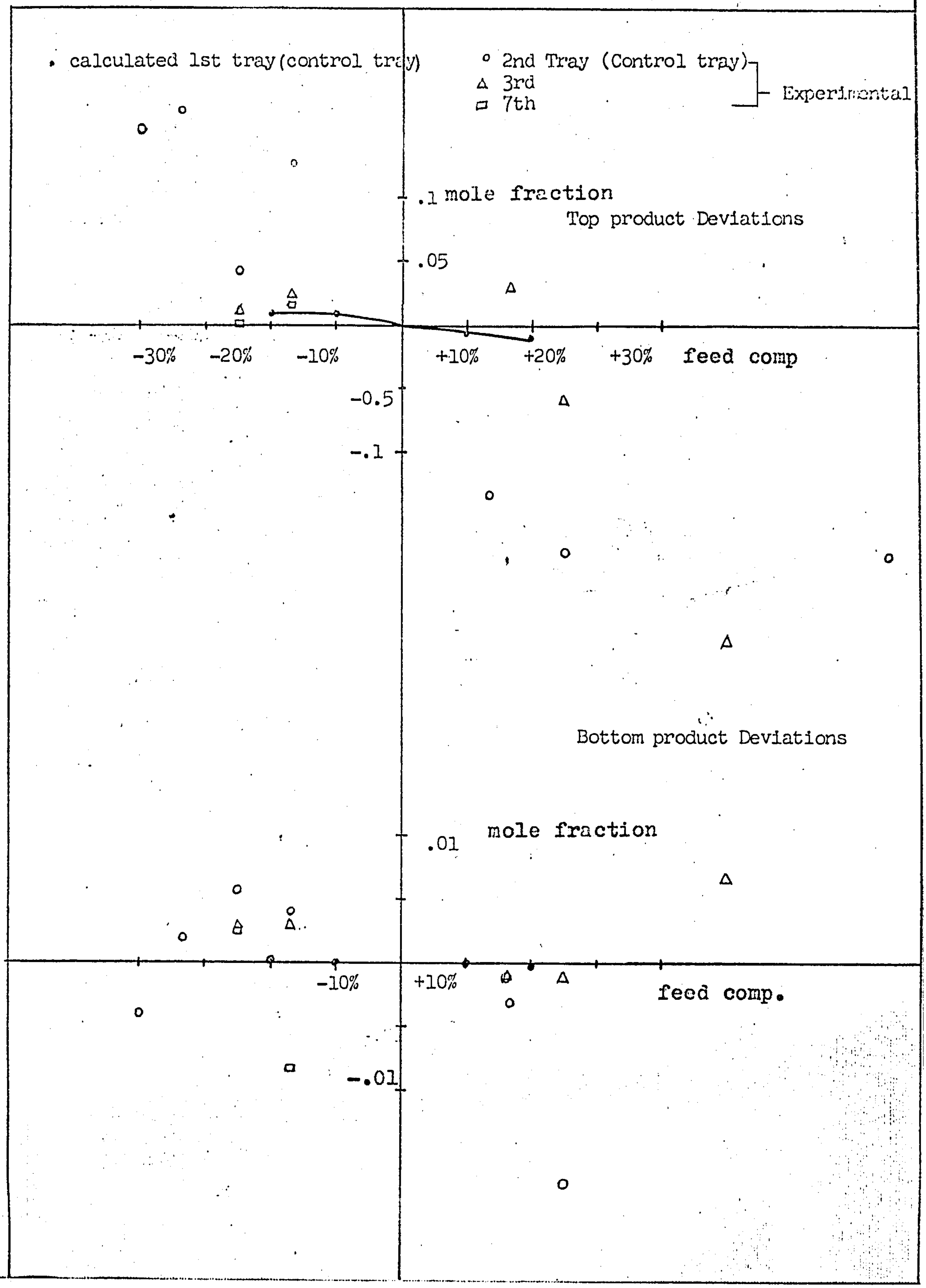


Figure 13. Deviations in x_D and x_B to changes in x_F from calculations

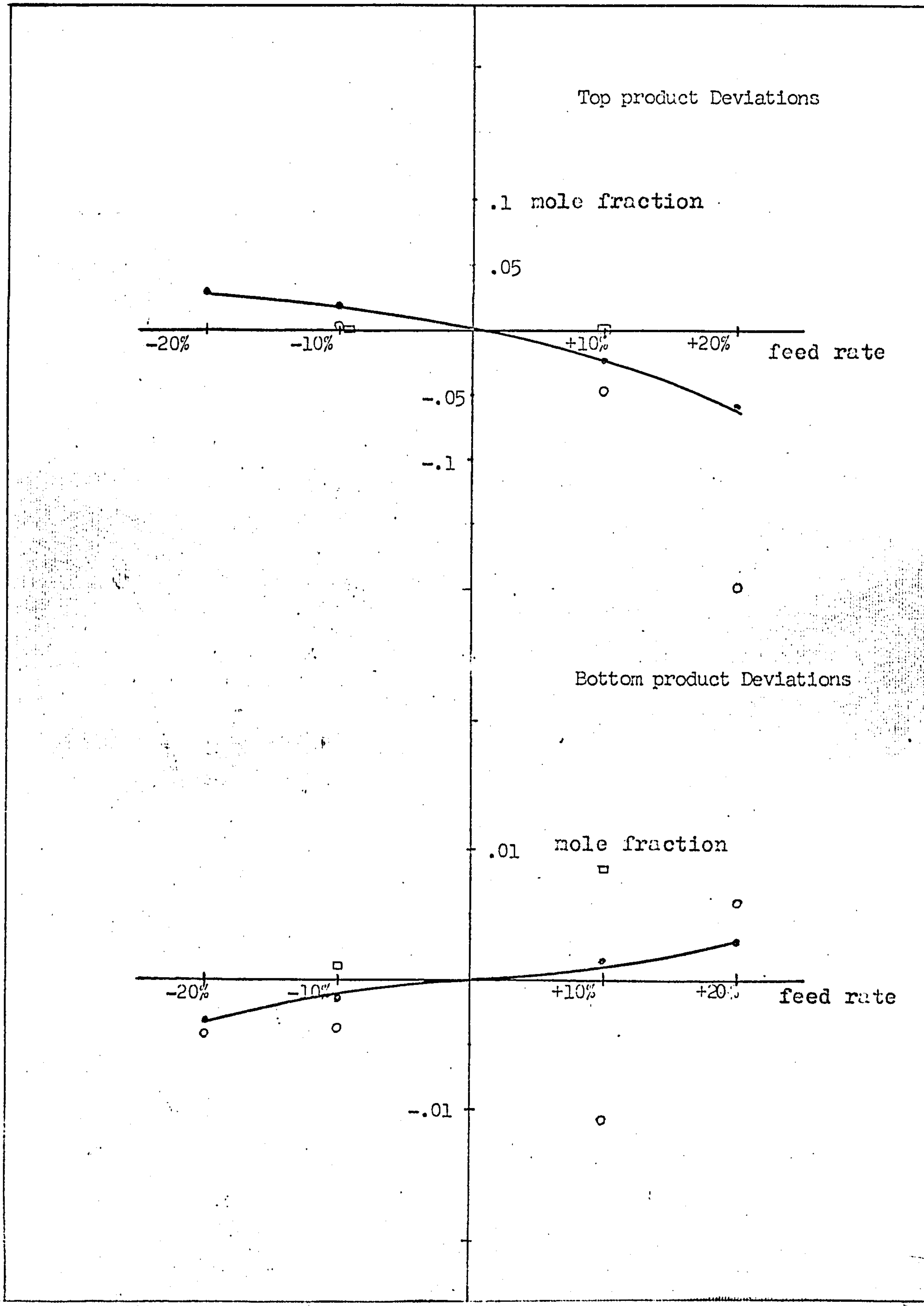


Figure 14. Deviations in x_D and x_B to change in F from calculation and experiment

of the control tray would be the middle of the stripping section. This feed-back control scheme is more suitable for systems with high relative volatility. If the top product composition is the control criterion regardless of the types of disturbance.

Generally speaking, this control scheme is better for feed composition disturbance and is less successful for feed rate disturbance.

The deviations caused by feed rate disturbance are less difficult to predict. They act more or less according to a regular trend. This can be accounted for the linear property of the rate disturbance. The deviations caused by feed composition disturbance are more irregular and more difficult to predict.

The end product deviations become more unpredictable when the control tray was located nearer the feed tray. This phenomena is more pre-dominant for feed composition disturbance. The choice of these trays as control tray should be avoided.

This control scheme can be improved by some modifications. When the reboiler heat input is the manipulated quantity the reflux rate should be changed with the change in feed rate according to a certain ratio. This is essentially a combined feedforward and feedback control and better results should be expected. Similarly when the reflux rate is the manipulated variable, the reboiler heat input should be changed according to the change in feed rate.

2) Feedforward Control Scheme

The performance of this scheme is quite satisfactory for feed composition disturbance for either one end or two end composition control. As for rate disturbance the end product deviations are large enough to regard this scheme as unsatisfactory. However this situation can be improved by making the control computer more complex as that not only the distillate withdrawal rate but also the reboiler heat input is changed according to the feed rate.

As mentioned in the very beginning of this writing, the goal of this work is to obtain a qualitative results. No attempt was made to evaluate the results quantitatively. The agreement between the experimental and calculated results show the calculated method proposed in this work is quite acceptable. The steady-state end product purity was the criterion used for comparison of the effectiveness of the control schemes. However the dynamics cannot be ignored during the evaluation of control schemes and it was assumed in this work that all the control schemes under study were satisfactory dynamically.

X Recommendation for Future Work

1) This study was concerned with binary system only. It has been reported in the literature (23) that the Smoker equation could be extended to multicomponent systems. Therefore it is suggested that the control problems of multicomponent systems also be studied using a modification of this method.

2) Only qualitative results were obtained, in this study. For some design purposes, this may not be enough. Therefore a more rigorous study using heat and material balances may be beneficial.

3) The effect of combined disturbances should also be studied. These disturbances sometimes might compensate each other and leave the end product compositions practically unchanged.

4) More investigations should be conducted on various feedforward schemes. The simple feedforward scheme proposed in this work gave an acceptable result on composition disturbance. A more sophisticated one would surely be more effective.

5) The "perfect" control scheme-combined feedforward and feedback control scheme - needs a more thorough investigation. The superior effectiveness of this scheme was beyond doubt. Yet its practical value should be determined by an economic study.

APPENDIX 1

Calculated and Experimental Results

Table 2

Steady-State Compositions and Deviations of x_D and x_B for system with

$\alpha = 2$, $q = 1$ Under Disturbance

Control tray 10th Tray (Rectifying section) | Control tray 9th Tray (Rectifying section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0 %	.950004	0	.053446	0	0	.949999	0	.056065	0
− 10%	.950236	+ .000232	.05336	− .0011	− 10	.950702	+ .000703	.053628	− .002437
− 20%	.950456	+ .000457	.051596	− .001850	− 20	.951409	+ .001410	.051976	− .004089
+ 10%	.949766	− .000248	.053362	− .000084	+ 10	.949228	− .000771	.052530	− .004535
+ 20%	.949521	− .000483	.055573	+ .001127	+ 20	.948454	− .001545	.052964	− .004101

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0%	.950004	0	.055416	0	0	.949999	0	.056065	0
− 10%	.950349	+ .000345	.029730	− .025686	− 10	.951084	+ .001085	.030529	− .025536
− 20%	.950683	+ .000679	.018159	− .036267	− 20	.952109	+ .002110	.018675	− .038390
+ 10%	.949711	− .000293	.105904	+ .050488	+ 10	.949070	− .000929	.103957	+ .057892
+ 20%	.949565	− .000439	.180090	+ .134574	+ 20	.948630	− .001369	.180790	+ .124725

Table 3

Steady-State Compositions and Deviations of x_D and x_B for System with

$a = 2$, $q = 1$ Under Disturbance

Control tray 8th Tray (Rectifying section) Control tray 7th Tray (Rectifying section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.950022	0	.056088	0	0	.950038	0	.056104	0
− 10	.951870	+ .001858	.056605	+ .000517	− 10	.953348	+ .00331	.057508	+ .001404
− 20	.953559	+ .003537	.053814	− .002274	− 20	.956465	+ .006427	.060777	+ .005634
+ 10	.948017	− .002005	.052319	− .003769	+ 10	.946309	− .003729	.050470	− .005634
+ 20	.945977	− .004045	.051203	− .004885	+ 20	.942336	− .007702	.045216	− .010888

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.950022	0	.056088	0	0	.950038	0	.056104	0
− 10	.952721	+ .002699	.030594	− .025494	− 10	.954807	+ .004769	.032826	− .033278
− 20	.955235	+ .005213	.019987	− .037101	− 20	.959020	+ .008982	.021970	− .034134
+ 10	.947602	− .002420	.104756	+ .048668	+ 10	.945259	− .004779	.097979	+ .041875
+ 20	.946113	− .003909	.172651	+ .116563	+ 20	.941963	− .008075	.162225	+ .106121

Table 4

Steady-State Compositions and Deviations of x_D and x_B for system with

$\alpha = 2$, $q = 1$ Under Disturbance

Control tray 6th Tray (Rectifying Section) Control tray 1st Tray (Stripping Section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949888	0	.053332	0	0	.95	0	.056	0
- 10%	.955496	+ .005608	.064313	+ .010981	- 10	.950874	+ .000874	.057209	+ .001209
- 20	.960195	+ .01207	.069752	+ .01642	- 20	.948843	- .001157	.058440	+ .002440
+ 10	.943889	- .005999	.047431	- .005901	+ 10	.952010	+ .002010	.054840	- .001160
+ 20	.937183	- .012705	.041774	- .011558	+ 20	.954239	+ .004239	.053727	- .002273

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949888	0	.053332	0	0	.95	0	.056	0
- 10	.957488	+ .007610	.037281	- .016051	- 10	.965599	+ .015599	.054808	- .002192
- 20	.963373	+ .013485	.023415	- .029917	- 20	.977337	+ .027337	.053539	- .002461
+ 10	.941610	- .008278	.089636	+ .036304	+ 10	.928137	- .021863	.056997	+ .000997
+ 20	.934133	- .015755	.142864	+ .089532	+ 20	.899212	- .050788	.057745	+ .001745

Table 5

Steady-State Compositions and Deviations of x_D and x_B for System With

$\alpha = 2$, $q = 1$ Under Disturbance

Control Tray 2nd Tray (Stripping section) Control tray 3rd Tray (Stripping Section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.95	0	.056	0	0	.95	0	.056	0
− 10	.949844	− .000156	.058959	+ .002959	− 10	.951130	+ .00113	.060220	+ .00422
− 20	.95154	+ .00154	.061246	+ .005246	− 20	.951130	+ .00113	.06496	+ .00896
+ 10	.952441	+ .002441	.053623	− .02377	+ 10	.95185	− .00185	.05215	− .00385
+ 20	.953631	+ .003631	.051378	− .04622	+ 20	.95212	+ .00212	.04861	− .00739

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.95	0	.056	0	0	.95	0	.056	0
− 10	.965373	+ .015373	.053567	− .002433	− 10	.964081	+ .014081	.052009	− .003991
− 20	.976554	+ .026554	.050991	− .005009	− 20	.974398	+ .024398	.048027	− .008973
+ 10	.92860	− .02140	.058280	+ .002280	+ 10	.930833	− .019167	.059614	+ .003614
+ 20	.904919	− .045081	.059901	+ .003901	+ 20	.907643	− .042357	.062738	+ .006738

Table 6

Steady-State Compositions and Deviations of x_D and x_B for System with

$\alpha = 2, \; q = 1$ Under Disturbance

Control tray 4th Tray (Stripping section) Control tray 5th Tray (Stripping Section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.95	0	.056	0	0	.95	0	.056	0
− 10	.952741	+ .002741	.062806	+ .005806	− 10	.954205	+ .004205	.066720	+ .010720
− 20	.955227	+ .005227	.070661	+ .014661	− 20	.959729	+ .009729	.080253	+ .024253
+ 10	.951287	− .001287	.050189	− .004810	+ 10	.945770	− .00423	.047179	− .00882
+ 20	.947221	− .002779	.044844	− .011166	+ 20	.942574	− .007436	.040139	− .01586

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.95	0	.056	0	0	.95		.056	0
− 10	.963130	+ .013130	.049965	− .006035	− 10	.963675		.047361	− .008639
− 20	.974136	+ .024136	.044315	− .011685	− 20	.971107		.039666	− .016334
+ 10	.932928	− .027072	.061880	+ .005880	+ 10	.936023		.65566	+ .009566
+ 20	.909491	− .040508	.067039	+ .011039	+ 20	.915170		.074575	+ .018575

Table 7

Steady-State Compositions and Deviations of x_D and x_B for System With

a = 4 , q = 1 Under Disturbance

Control tray 10th Tray (Rectifying section) Control tray 9th Tray (Rectifying section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.950048	0	.059213	0	0	.949891	0	.057739	0
− 10	.953521	+ .003473	.029295	− .29918	− 10	.958918	+ .009027	.03545	− .022289
− 20	.957027	+ .006979	.014033	− .04518	− 20	.966022	+ .016131	.018926	− .038813
+ 10	.946331	− .003717	.099802	+ .040589	+ 10	.937879	− .012012	.084206	+ .026467
+ 20	.941845	− .008203	.145073	+ .08586	+ 20	.913837	− .036054	.077978	+ .020239

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.950048	0	.059 213	0	0	.949891	0	.057739	0
− 10	.956092	+ .006044	.0 03084	− .056129	− 10	.963697	+ .013806	.003627	− .054112
− 20	.96194	+ .011892	.000747	− .058466	− 20	.974379	+ .024488	.00093	− .056809
+ 10	.949633	− .000415	.18436	+ .125147	+ 10	.949183	− .000708	.184743	+ .127004
+ 20	.949633	− .000415	.261514	+ .202301	+ 20	.949183	− .000708	.261799	+ .20406

Table 8

Steady-State Compositions and Deviations of x_D and x_B for System With

$a = 4$, $q = 1$ Under Disturbance

Control tray 8th Tray (Rectifying section) Control tray 2nd Tray (Stripping section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949853	0	.057992	0	0	.949004	0	.056029	0
− 10	.966998	+ .017145	.047961	− .010031	− 10	.945339	− .003705	.04867	+ .008838
− 20	.976061	+ .026208	.028441	− .029551	− 20	.945036	− .004008	.076432	+ .020403
+ 10	.910932	− .038921	.038710	− .019282	+ 10	.954526	+ .005482	.049052	− .006977
					+ 20	.960422	+ .011378	.043428	− .012601

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949853	0	.057992	0	0	.949044	0	.056029	0
− 10	.971928	+ .022075	.004561	− .053431	− 10	.972554	+ .02351	.054333	− .001696
− 20	.984655	+ .034802	.001228	− .056784	− 20	.987296	+ .038252	.051169	− .00486
+ 10	.948046	− .001807	.183792	+ .12580	+ 10	.92428	− .024764	.056694	+ .000665
+ 20	.948046	− .001807	.261215	+ .203223	+ 20	.903036	− .046008	.056932	+ .000903

<u>Table 9</u>

Steady-State Compositions and Deviations of x_D and x_B for system With

$\alpha = 4$ $q = 1$ Under Disturbance

Control tray 3rd Tray (Stripping section) | **Control tray 4th Tray (Stripping section)**

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949474	0	.056109	0	0	.949205	0	.056138	0
- 10	.949494	+ .00002	.079475	+ .023366	- 10	.965308	+ .016103	.141468	+ .085330
- 20	.963206	+ .013732	.124216	+ .068107					
+ 10	.953095	+ .003623	.042067	- .014042	+ 10	.951484	+ .002279	-.031575	- .024563
+ 20	.95849	+ .009016	.032912	- .023197					

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949474	0	.056109	0	0	.949205	0	.056138	0
- 10	.972284	+ .02281	.052394	- .003715	- 10	.971724	+ .022519	.04857	- .007568
- 20	.987051	+ .037577	.046129	- .009980	- 20	.986864	+ .037159	.037766	- .018372
+ 10	.924445	- .025029	.057582	+ .001473	+ 10	.924562	- .024643	.059556	+ .003418
+ 20	.903105	- .046369	.058101	+ .001992	+ 20	.903678	- .045527	.061021	+ .004883

Table 10

Steady State Compositions and Deviations of x_D and x_B for System With

$\alpha = 1.6 \quad q = 1$ Under Disturbance

Control Tray 15th Tray (Rectifying section) Control tray 13th Tray (Rectifying section)

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
				Type of Disturbance Feed Composition					
0	.94997	0	.056348	0	0	.949843	0	.055367	0
- 10	.950250	+ .00028	.056645	+ .000297	- 10	.951193	+ .00135	.058631	+ .003264
- 20	.950520	+ .00055	.057060	+ .000712	- 20	.952485	+ .002642	.060111	+ .004744
+ 10	.949687	- .000283	.055904	- .000444	+ 10	.948493	- .00135	.054066	- .001301
+ 20	.949402	- .000568	.053586	- .002762					
				Type of Disturbance Feed Rate					
0	.94997	0	.056348	0	0	.948843	0	.055367	0
- 10	.950391	+ .000421	.037118	- .01923	- 10	.951857	+ .002014	.038725	- .016642
- 20	.950793	+ .000823	.027391	- .028957	- 20	.953696	+ .003853	.029573	- .025794
+ 10	.949563	- .000407	.089341	- .032993	10	.947762	- .002081	.081235	+ .025868
+ 20	.949226	- .000744	.136328	+ .07998	+ 20	.946086	- .003757	.129862	+ .074495

Table 11

Steady State Compositions and Deviations of x_D and x_B for System With

$a = 1.6 \quad q = 1$ Under Disturbance

Control tray 2nd Tray (Stripping section)					Control tray 4th Tray (Stripping section)				
% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
Type of Disturbance Feed Composition									
0	.957591	0	.055503	0	0	.951166	0	.056267	0
- 10	.955992	- .001599	.056375	+ .000572	- 10	.951318	+ .000152	.058727	+ .00247
- 20	.956536	- .001055	.057277	+ .001774	- 20	.951894	+ .000728	.061323	+ .005066
+ 10	.957066	- .000525	.054634	- .000869	+ 10	.950949	- .000217	.053897	- .00236
+ 20	.958472	+ .000881	.053793	- .00171	+ 20	.950012	- .001154	.051621	- .004636
Type of Disturbance Feed Rate									
0	.957591	0	.055503	0	0	.951166	0	.056257	0
- 10	.968011	+ .01042	.054344	- .001159	- 10	.965295	+ .014129	.05192	- .004337
- 20	.975206	+ .017615	.053206	- .002297	- 20	.973369	+ .022203	.049277	- .00698
+ 10	.941641	- .01595	.056622	+ .001119	+ 10	.943718	- .007448	.057619	+ .001362
+ 20	.922736	- .034855	.057675	+ .002172	+ 20	.925701	- .025465	.06038	+ .004123

Table 12

Steady State Compositions and Deviations of x_D and x_B for System With

$a = 6 \quad q = 1$ Under Disturbance

Control tray 2nd Tray (Rectifying section) — Control tray 1st Tray (Stripping section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top. Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949968	0	.055619	0	0	.950216	0	.055973	0
− 10	.956493	.006525	.026352	− .029267	− 10	.944314	− .005902	.065117	− .005902
− 20	.962364	.012396	.013282	− .042337	− 20	.939917	− .010299	.077471	− .010299
					+ 10	.956927	+ .006711	.048948	+ .006711
					+ 20	.962614	+ .012398	.043296	+ .012398

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top. Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949968	0	.055619	0	0	.950216	0	.055973	0
− 10	.960949	+ .010981	.003017	− .052602	− 10	.964013	+ .013797	.044779	− .011194
− 20	.969266	+ .019298	.000813	− .054806	− 20	.977247	+ .027031	.043062	− .012911
+ 10	.949388	− .00058	.183420	− .127801	+ 10	.934316	− .01590	.046152	− .009821
+ 20	.949388	− .00058	.26089	+ .205271	+ 20	.922638	− .027578	.046427	− .009546

Table 13

Steady State Compositions and Deviations of x_D and x_B for Systems With

$\alpha = 6$ $q = 1$ Under Disturbance

$\alpha = 4.88$

Control tray 2nd Tray (Stripping section $\alpha = 6$) Control tray 1st Tray (Stripping section $\alpha = 4.88$)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
					0	.939376	0	.017766	0
0	.960073	0	.055872	0	− 10	.950513	+ .011137	.018024	+ .000258
− 10	.949013	− .00106	.092770	+ .036898	− 20	.949330	+ .009954	.018202	+ .000436
+ 10	.955063	+ .00499	.038563	− .017309	+ 10	.933506	− .00587	.01752	− .000246
+ 20	.960838	+ .010765	.028626	− .027246	+ 20	.928963	− .010413	.017282	− .000848

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.950073	0	.055872	0	0	.939376	0	.177766	0
− 10	.963736	+ .013663	.043497	− .012375	− 10	.957843	+ .018467	.016272	− .001494
− 20	.97684	+ .026767	.039184	− .016678	− 20	.967423	+ .028047	.014737	− .003029
+ 10	.935116	− .014857	.047526	− .008346	+ 10	.914574	− .024802	.019185	+ .001419
+ 20	.923132	− .026341	.048258	− .007614	+ 20	.879350	− .060026	.020484	+ .002718

Table 14

Steady-State Compositions and Deviations of x_D and x_B for System With

$\alpha = 1.4$ $q = 1$ Under Disturbance

Control tray 26th Tray (Rectifying section)					Control tray 21st Tray (Rectifying section)				
Type of Disturbance: Feed Composition									
% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.948664	0	.115971	0	0	.94951	0	.107546	0
- 10	.950524	+ .00186	.105921	- .00905	- 10	.951643	+ .002133	.108705	+ .001159
- 20	.950524	+ .0036004	.096748	- .019761	- 20	.957065	+ .007555	.107566	+ .00002
+ 10	.946624	- .00204	.121094	+ .005123	+ 10	.938094	- .011416	.100381	- .007165
+ 20	.944404	- .00426	.121238	+ .005267	+ 20	.929761	- .019749	.086987	- .020559
Type of Disturbance: Feed Rate									
0	.948664	0	.115971	0	0	.94951	0	.107546	0
- 10	.950521	+ .001857	.045495	- .070476	- 10	.95155	+ .00204	.046477	- .061069
- 20	.953084	+ .00442	.019494	- .096477	- 20	.958531	+ .009021	.021758	- .085788
+ 10	.948197	- .000467	.210689	+ .094718	+ 10	.943585	- .005925	.202238	+ .094692
+ 20	.948268	- .000396	.280999	+ .165008	+ 20	.943585	- .005925	.274026	+ .16648

Table 15

Steady-State Compositions and Deviations of x_D and x_B for System With

$\alpha = 1.4$ $q = 1$ Under Disturbance

Control tray 3rd Tray (Stripping section)					Control tray 7th Tray (Stripping section)				
Type of Disturbance: Feed Composition									
% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.90372	0	.057007	0	0	.907268	0	.059688	0
- 10	.900862	- .002858	.058515	+ .001508	- 10	.910744	+ .003476	.065526	+ .005838
- 20	.901829	- .001891	.060117	+ .00311	- 20	.918968	+ .011700	.072029	+ .012341
+ 10	.910418	+ .006698	.0556	- .001407	+ 10	.907388	- .00012	.054492	- .005196
+ 20	.916057	+ .012337	.054233	- .002774	+ 20	.873277	- .033991	.047681	- .012007
Type of Disturbance: Feed Rate									
0	.90372	0	.057007	0	0	.907268	0	.059688	0
- 10	.953216	+ .048496	.055872	- .00118	- 10	.952969	+ .045701	.055559	- .004129
- 20	.977248	+ .073528	.054269	- .002738					
+ 10	.836448	- .067271	.057433	+ .000426	+ 10	.84342	- .063848	.061565	+ .001877
+ 20	.778903	- .124817	.057480	+ .000473	+ 20	.785966	- .121302	.061940	+ .002252

Table 16

Steady-State Compositions and Deviations of x_D and x_B for Systems With

$\alpha = 2$ Under Feed Enthalpy Disturbance

Control tray 11th Tray (Rectifying section)					Control tray 9th Tray (Rectifying section)				
% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.949999	0	.056065	0	0	.950038	0	.056104	0
− 20	.950010	+ .000011	.056567	+ .000502	− 20	.952345	+ .002307	.049235	− .006868
+ 20	.949362	− .000637	.061922	+ .005857	+ 20	.946946	− .003092	.060371	+ .004267
Control tray 2nd Tray (Stripping section)					Control tray 4th Tray (Stripping section)				
0	.95	0	.056	0	0	.95	0	.056	0
− 20	.944122	− .005878	.056967	+ .000967	− 20	.947748	− .002252	.068873	+ .012873
+ 20	.956275	+ .012153	.054598	− .001402	+ 20	.952793	+ .002793	.052286	− .003714

Table 17

Steady-State Compositions and Deviations in x_D and x_B for System With

$\alpha = 2 \quad q = 1$ Under Disturbance - Feedforward Control

Type of Disturbance	Feed Composition				Type of Disturbance	Feed Rate			
% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.	% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.95	0	.056	0	0	.95	0	.056	0
- 10	.95	0	.056	0	- 10	.966	+ .016	.02648	- .02952
- 20	.948	- .001	.05662	+ .00062	- 20	.98	+ .03	.01125	- .04475
+ 10	.953	+ .003	.05229	-.00371	+ 10	.919	-.031	.07424	+ .01824
+ 20	.954	+ .004	.04978	-.00622	+ 20	.819	-.059	.1026	+ .0466

Table 18

Experimental Results 2nd Tray (Stripping section)

Control Tray

Type of disturbance : feed composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.8196	0	.01707	0
- 16.7	.9478	+ .1282	.02287	+ .00411
0	.9478	0	.01707	0
- 25	.992	+ .0442	.02287	+ .0058
0	.8196	0	.0196	0
- 33.4	.991	+ .1714	.0217	+ .0021
0	.5106	0	.02287	0
- 40	.6657	+ .1551	.01707	- .00411
0	.692	0	.02287	0
+ 13.4	.5566	- .1354	.0196	- .00327
0	.9736	0	.0287	0
+ 25	.7943	- .1793	.0113	- .0174
0	.7899	0	.0587	0
+ 75	.6101	- .1798	.02287	- .03183

Type of disturbance feed rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.9478	0	.0196	0
- 10	.9563	+ .0085	.0113	- .0038
0	.4889	0	.02287	0
- 20	.8196	+ .3307	.0196	- .00327
0	.9478	0	.0113	0
+ 10	.8976	- .0499	.0	- .0113
0	.992	0	.0113	0
+ 20	.5308	- .4592	.01707	+ .00577

Table 19

Control tray 3rd Tray (Stripping section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.9649	0	.0196	0
- 16.7	.992	+ .0271	.02287	+ .00327
0	.98	0	.0196	0
- 25	.992	+ .012	.02287	+ .00327
0	.844	0	.02287	0
+ 16.7	.8738	+ .0298	.0217	- .00117
0	.98	0	.0217	0
+ 25	.9226	- .0574	.02054	- .00116
0	.9444	0	.0113	0
+ 50	.6761	- .2683	.0182	+ .0067

Control tray 7th Tray (Rectifying section)

Type of Disturbance Feed Composition

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.9309	0	.0196	0
16.7	.9478	+ .0169	.0113	- .0083
0	.9303	0	.01707	0
- 25	.9343	+ .004	.0196	+ .00253

Type of Disturbance Feed Rate

% of Dist.	Top Comp.	Dev. in Top Comp.	Bottom Comp.	Dev. in Bottom Comp.
0	.9478	0	.0159	0
- 10	.9478	0	.01707	+ .00117
0	.9410	0	.0084	0
+ 10	.9393	- .0017	.01707	+ .00867

APPENDIX 2

Nomenclature for Computer Program

NOMENCLATURE FOR COMPUTER PROGRAM

ALPHA	Relative volatility
BRECN	b in rectifying section for normal feed condition
BRECT	b in rectifying section under disturbance
BRMAX	b in rectifying section for minimum reflux
BSTRN	b in stripping section for normal feed condition
BSTRP	b in stripping section under disturbance
BSMAX	b in stripping section for minimum reflux
CRECN	c in rectifying section for normal feed condition
CRECT	c in rectifying section under disturbance
CRECM	c in rectifying section for minimum reflux
CSTRM	c in stripping section for minimum reflux
CSTRN	c in stripping section for normal feed condition
CSTRP	c in stripping section under disturbance
DEL DELL	Increments
DELTA	Numerical constant between V' L and F
DELTR	Numerical constant between V' and F
DELTS	Numerical constant between L and F
G	Feed rate disturbance in feedforward control scheme
FN	Normal feed rate
FR	Feed rate
FD	Numerical constant between D and F
PLATA	Tray numbers between end tray and control tray
QFEEN	q-value for normal feed condition

QFEED	q-value for feed
RECKN	k in rectifying section with normal feed condition
RECTK	k in rectifying section under disturbance
RECMK	k in rectifying section with minimum reflux
RLMIN	Slope of operating line in rectifying section with total reflux
RN	Reflux ratio with normal feed condition
RR	Reflux ratio under disturbance
SRECN	Slope of operating line in rectifying section with normal feed condition
SRECT	Slope of operating line in rectifying section under disturbance
SSTRN	Slope of operating line in stripping section with normal feed condition
SSTRP	Slope of operating line in stripping section under disturbance
SLMIN	Slope of operating line in stripping section with total reflux
SRMAX	Slope of operating line in rectifying section with minimum reflux
SSMAX	Slope of operating line in stripping section with minimum reflux
STRKN	k in stripping section with normal feed condition
STRMK	k in stripping section with minimum reflux
STRPK	k in stripping section under disturbance
TRAND	Tray numbers in rectifying section with normal feed condition
TRANS	Tray numbers in stripping section with normal feed condition
TRAYD	Tray numbers in rectifying section under disturbance
TRAYS	Tray numbers in stripping section under disturbance
TRAYT	Tray numbers between end tray and control tray under normal condition

TRAYP, TRAYZ	Tray numbers between end tray and control tray under disturbance
TTN	Calculated total number of trays
TTNN	Total number of trays under normal feed condition
XBOTM	x_B with minimum reflux
XBOTL	x_B with total reflux
XDITM	x_D with total reflux
XDITL	x_D with minimum reflux
XBOTN	x_B with normal feed condition
XBOTT	x_B under disturbance
XDISN	x_D with normal feed condition
XDIST	x_D under disturbance
XFEED	x_F under disturbance
XFEEN	x_F with normal feed condition
XCNR	$x_L - k$ with normal feed condition
XTNNR	Liquid composition of control tray when in rectifying section
XXNR	XTNNR - k in rectifying section
XTNR, XCR	$x_D - k$
XNR	XTNNR - k or XNNS - k in rectifying section
XNNS	Liquid composition of feed tray under disturbance when control tray is in rectifying section
XNS	XNNS - k in stripping section when control tray is in rectifying section
XCNS	$x_F - k$ with normal feed condition
XTNNS	Liquid composition of control tray in stripping section
XBNS	$x_B - k$ with normal feed condition when control tray is in stripping section

XXNS	XMNS - k with normal feed condition
XNNS	XMNS - k in stripping section
XNS	x_S - k in stripping section when control tray is in stripping section.
XNNR	Liquid composition of feed tray when control tray is in stripping section
XXS	XNNR - k in stripping section
XI, YI	x_1, y_1
XIR, YIR	Intersecting point of q-line and equilibrium line
XIL, YIL	Intersecting point of q-line and 45° line.

APPENDIX 3

Computer Program for Feedback Control

```
C         EVALUATION OF TEMPERATURE CONTROL SCHEME OF A BINARY
C         DISTILLATION TOWER BY SMOKER EQUATION    A.L.CHAN
C         FEED RATE AND COMPOSITION DISTURBANCE
C         FIRST RATE DISTURBANCE THEN COMPOSITION DISTURBANCE
C         QFEED=1  ALPHA=2
          DOUBLE PRECISION YIID,YIH1,YIH2,YIH,YID,EF
          DOUBLE PRECISION TRANA,TRANB,TRANC,TRANS,TTNN,DELL,XI,XIH,YIIN
          DOUBLE PRECISION XFEEN,XDISN,XBOTN,QFEEN,ALPHA,PLATA,L
          DOUBLE PRECISION SSTRN,BS,BSTRN,BBSN,AASN,STRKN,CSTRN,XONS,XNNS
          DOUBLE PRECISION XONR,XTNR,TRANE,TRANG,ARG,TRANO,TRAND,SS,TT
          DOUBLE PRECISION YI,SRECN,RN,BRECN,BBRN,AARN,RECKN,CRECN
          COMMON QFEEN,ALPHA,PLATA
C         CALCULATION OF THE INTERSECTING POINT OF Q AND EQUILIBRIUM LINES
C         AT NORMAL CONDITION
          READ(1,15)XFEEN,XDISN,XBOTN,QFEEN,DELL,RN
     15   FORMAT(6F10.6)
          READ(1,13)ALPHA,PLATA,L
     13   FORMAT(2F10.6,1I2)
    301   XI=XFEEN
    425   IF(QFEEN-1.)413,432,414
    413   XIH=XI-DELL
          GO TO 415
    414   XIH=XI+DELL
    415   YIIN=ALPHA*XIH
          YIID=1.+(ALPHA-1.)*XIH
          YII=YIIN/YIID
          WRITE(3,412)
    412   FORMAT(/// 3X,3HXIH,7X,3HYII,7X,3HYIH)
          IF(QFEEN-1.)424,309,424
    424   YIH1=((QFEEN)/(QFEEN-1.))*XIH
          YIH2=XFEEN/(QFEEN-1.)
          YIH=YIH1-YIH2
    422   WRITE(3,423)XIH,YII,YIH
    423   FORMAT(1H0,3F10.6)
          YID=YII-YIH
          IF(EF-DABS(YID))431,309,309
    431   IF(YID)430,309,416
    416   XI=XIH
          GO TO 425
    430   XIH=(XI+XIH)/2.
          GO TO 415
    432   YIH=(ALPHA*XFEEN)/(1.+(ALPHA-1.)*XFEEN)
    309   YIL=XFEEN
          YI=(YIH+YIL)/2.
          WRITE(3,31)
C         CALCULATION OF THE NUMBER OF TRAYS IN THE RECTIFYING SECTION
     31   FORMAT(/// 3X,5HTRAND)
     27   SRECN=RN/(1.+RN)
          BRECN=XDISN/(RN+1.)
          BBRN=SRECN+BRECN*(ALPHA-1.)-ALPHA
          AARN=SRECN*(ALPHA-1.)
          RECKN=(-BBRN-SQRT(BBRN**2-4.*AARN*BRECN))/(2.*AARN)
          CRECN=1.+(ALPHA-1.)*RECKN
```

```
      XONR=XDISN-RECKN
      XTNR=XFEEN-RECKN
      TRANE=SRECN*CRECN**2
      TRANG=(SRECN*CRECN*(ALPHA-1.))/(ALPHA-TRANE)
      ARG=(XONR*(1.-XTNR*TRANG))/(XTNR*(1.-XONR*TRANG))
      TRANO=ALOG(ARG)
      TRAND=TRANO/ALOG(ALPHA/TRANE)
C     CALCULATION OF THE TRAY NUMBERS IN THE STRIPPING SECTION AND
C     TOTAL TRAY NUMBERS
      WRITE(3,30)TRAND
   30 FORMAT(1H0,1F10.6)
   17 WRITE(3,18)
   18 FORMAT(/// 3X,5HTRANS,5X,4HTTNN)
   25 SS=RN*XFEEN+QFEEN*XDISN-(RN+QFEEN)*XBOTN
      TT=(RN+1.)*XFEEN+(QFEEN-1.)*XDISN-(RN+QFEEN)*XBOTN
      SSTRN=SS/TT
      BS=(XFEEN-XDISN)*XBOTN
      BSTRN=BS/TT
      BBSN=SSTRN+BSTRN*(ALPHA-1.)-ALPHA
      AASN=SSTRN*(ALPHA-1.)
      STRKN=(-BBSN+SQRT(BBSN**2-4.*AASN*BSTRN))/(2.*AASN)
      CSTRN=1.+(ALPHA-1.)*STRKN
      XONS=XFEEN-STRKN
      XNNS=XBOTN-STRKN
      TRANA=SSTRN*CSTRN**2
      TRANB=(SSTRN*CSTRN*(ALPHA-1.))/(ALPHA-TRANA)
      TRANC=ALOG((XONS*(1.-XNNS*TRANB))/(XNNS*(1.-XONS*TRANB)))
      TRANS=TRANC/ALOG(ALPHA/TRANA)
      TTNN=TRAND+TRANS
      WRITE(3,36)TRANS,TTNN
   36 FORMAT(1H0,2F10.6)
C     CALCULATION OF THE VALUES OF DELTA
      WRITE(3,305)
  305 FORMAT(/// 3X,5HDELTR,5X,5HDELTS)
      DELT=((XFEEN-XBOTN)/(XDISN-XBOTN))
      DELTR=DELT*(RN+1.)
      DELTS=DELTR-DELT
      WRITE(3,306)DELTR,DELTS
  306 FORMAT(1H0,2F10.6)
      IF(L)26,12,26
   26 CALL STRIP(TTNN,TRAND,XFEEN,XDISN,XBOTN,DELTS,RN)
      GO TO 304
   12 CALL RECTY(TTNN,TRANS,XFEEN,XDISN,XBOTN,DELTR,RN)
  304 RETURN
      END

        SIZE OF COMMON  000024   PROGRAM  002824

                        END OF COMPILATION  MAIN
```

```
      SUBROUTINE RECTY(TTNN,TRAYS,XFEEN,XDISN,XBOTN,DELTA,RN)
C     CONTROL OF PLATE IN THE RECTIFYING SECTION
      DOUBLE PRECISION XBOTN,XDISN,QFEEN,RN,PLATA,L,SRECN,BN,XONR,XXNR
      DOUBLE PRECISION BRECN,BBRN,AARN,RECKN,CRECN,SCREN,AN,BND,BNF
      DOUBLE PRECISION CRECT,TRAYN,TRAYM,TRAYO,TRAYP,TTN,NBOTT
      DOUBLE PRECISION DBOTT,XBOTT,SSTRP,BSTRP,BBS,AAS,STRPK,CSTRP,XBS
      DOUBLE PRECISION SCTRP,C,D,XNS,XNNS,TRAYE,TRAYG,TRAYH,TRAYS
      DOUBLE PRECISION BH,BL,YI,SRECT,RR,BRECT,BBR,AAR,RECTK
      DOUBLE PRECISION CRETM,XTNR,XNR,TRAYW,TRAYX,TRAYY,AL,AH
      DOUBLE PRECISION AA,AB,TRAYZ,XDITL,SRMAX,BRMAX,BRM,ARM,RECMK
      DOUBLE PRECISION YII,XII,QFEED,E,TRAYD,G,XFEEN,XDITH
      DOUBLE PRECISION TRAYT,XTNNR,ALPHA,RLMIN,TTNN,DEL
      DOUBLE PRECISION TRANE,TRANO,TRANG,EF,YID
      COMMON QFEED,ALPHA,PLATA
  201 READ(1,10)RLMIN,DEL,DELL,FN,FR,XFEED
   10 FORMAT(6F10.6)
      READ(1,301)E,G,F,EF,M
C     CALCULATION OF THE LIQUID COMPOSITION IN THE CONTROL TRAY AND THE
C     LOCATION OF THE CONTROL TRAY THAT COUNTED FROM THE TOP
  301 FORMAT(4F10.6,1I2)
      WRITE(3,99)
   99 FORMAT(///3X,4HXXNR,5X,5HXTNNR)
  110 SRECN=RN/(RN+1.)
      BRECN=XDISN/(RN+1.)
      BBRN=SRECN&BRECN*(ALPHA-1.)-ALPHA
      AARN=SRECN*(ALPHA-1.)
      RECKN=(-BBRN-SQRT(BBRN**2-4.*AARN*BRECN))/(2.*AARN)
      CRECN=1.&(ALPHA-1.)*RECKN
  111 SCREN=(SRECN**PLATA)*(CRECN**(2.*PLATA))
      AN=SRECN*CRECN*(ALPHA-1.)
      BND=ALPHA-SRECN*CRECN**2
      BNF=ALPHA**PLATA-SCREN
      BN=BNF/BND
      XONR=XDISN-RECKN
      XXNR=(SCREN*XONR)/(ALPHA**PLATA-AN*BN*XONR)
      XTNNR=XXNR+RECKN
      WRITE(3,116)XXNR,XTNNR
  116 FORMAT(1H0,2F10.6)
  108 WRITE(3,104)
  104 FORMAT(/// 3X,5HSRECN,5X,5HRECKN,5X,5HTRAYT)
      TRANE=SRECN*CRECN**2
      TRANG=(SRECN*CRECN*(ALPHA-1.))/(ALPHA-TRANE)
      TRANO=ALOG((XONR*(1.-XXNR*TRANG))/(XXNR*(1.-XONR*TRANG)))
      TRAYT=TRANO/ALOG(ALPHA/TRANE)
  105 WRITE(3,103)SRECN,RECKN,TRAYT
  103 FORMAT(1H0,3F10.6)
C     CALCULATION OF THE INTERSECTION OF THE Q AND EQUILIBRIUM LINES
C     DURING DISTURBANCE
      XI=XFEED
  425 IF(QFEED-1.)413,432,414
  413 XIH=XI-DELL
      GO TO 415
  414 XIH=XI+DELL
```

```
415 YIIN=ALPHA*XIH
    YIID=1.+(ALPHA-1.)*XIH
    YII=YIIN/YIID
    WRITE(3,412)
412 FORMAT(/// 3X,3HXIH,7X,3HYII,7X,3HYIH)
    IF(QFEED-1.)424,309,424
424 YIH1=((QFEED)/(QFEED-1.))*XIH
    YIH2=XFEED/(QFEED-1.)
    YIH=YIH1-YIH2
422 WRITE(3,423)XIH,YII,YIH
423 FORMAT(1H0,3F10.6)
    YID=YII-YIH
    IF(EF-DABS(YID))431,309,309
431 IF(YID)430,309,416
416 XI=XIH
    GO TO 425
430 XIH=(XI+XIH)/2.
    GO TO 415
432 WRITE(3,307)
307 FORMAT(/// 3X,3HYIH)
    YIH=(ALPHA*XFEED)/(1.+(ALPHA-1.)*XFEED)
    WRITE(3,308)YIH
C   CALCULATION OF THE TOP PRODUCT COMPOSITION AT TOTAL REFLUX
308 FORMAT(1H0,1F10.6)
309 WRITE(3,20)
 20 FORMAT(/// 3X,5HXDITH)
    AA=RLMIN**TRAYT
    AB=(ALPHA**TRAYT-RLMIN**TRAYT)/(ALPHA-RLMIN)
    AC=ALPHA**TRAYT
    AD=RLMIN*(ALPHA-1.)
    XDITH=AC*XTNNR/(AA+AB*AD*XTNNR)
C   CALCULATION OF THE TOP PRODUCT COMPOSITION AT MINIMUM REFLUX
    WRITE(3,21)XDITH
 21 FORMAT(1H0,1F10.6)
    WRITE(3,32)
 32 FORMAT(/// 3X,5HTRAYZ,5X,5HXDITL)
    XII=XFEED
    IF(XTNNR-XII)40,40,22
 22 XDITL=XDITH-DEL
 87 SRMAX=(XDITL-YIH)/(XDITL-XII)
    BRMAX=XDITL*(1.-SRMAX)
    BRM=SRMAX+BRMAX*(ALPHA-1.)-ALPHA
    ARM=SRMAX*(ALPHA-1.)
    RECMK=(-BRM-SQRT(BRM**2-4.*ARM*BRMAX))/(2.*ARM)
    CRETM=1.+(ALPHA-1.)*RECMK
    XTNR=XDITL-RECMK
    XNR=XTNNR-RECMK
    TRAYW=SRMAX*CRETM**2
    TRAYX=(SRMAX*CRETM*(ALPHA-1.))/(ALPHA-TRAYW)
    AARGN=XTNR*(1.-XNR*TRAYX)
    AARGD=XNR*(1.-XTNR*TRAYX)
    AARG=AARGN/AARGD
    IF(AARG)13,13,14
 13 DEL=DEL/2.
    XDITL=XDITH-DEL
```

```
      GO TO 87
   14 TRAYY=ALOG(AARG)
      TRAYZ=TRAYY/ALOG(ALPHA/TRAYW)
      WRITE(3,34)TRAYZ,XDITL
   34 FORMAT(1H0,2F10.6)
      IF(E-DABS(TRAYZ-TRAYT))94,40,40
   94 EE=2.5
      IF(EE-DABS(TRAYZ-TRAYT))101,101,102
  102 DEL=DEL/2.
      GO TO 95
  101 DEL=DEL&DEL/2.
   95 IF(TRAYZ-TRAYT)86,40,85
   86 XDITL=XDITL+DEL
      GO TO 87
   85 XDITL=XDITL-DEL
C     ASSUME A TOP PRODUCT COMPOSITION BY BISECTIONAL METHOD
C     ASSUME THE INTERSECTING POINT OF THE Q AND EQUILIBRIUM LINES
C     BY BISECTIONAL METHOD
C     CALCULATE THE TRAY NUMBERS BETWEEN THE CONTROL TRAY AND THE FIRST
   40 AL=XDITL
      AH=XDITH
   48 XDIST=(AL+AH)/2.
   53 WRITE(3,42)
   42 FORMAT(///,3X,5HTRAYP,5X,5HTRAYO,5X,2HYI,7X,5HXDIST,5X,5HRECTK)
      BH=YIH
      BL=XFEED
   43 YI=(BH+BL)/2.
      IF(QFEED-1.)204,205,204
  204 QYF=XFEED/(QFEED-1.)
      QYY=QFEED/(QFEED-1.)
      XI=(YI+QYF)/QYY
      GO TO 44
  205 XI=XFEED
   44 SRECT=(XDIST-YI)/(XDIST-XI)
      RR=SRECT/(1.-SRECT)
      BRECT=XDIST/(RR+1.)
      BBR=SRECT+BRECT*(ALPHA-1.)-ALPHA
      AAR=SRECT*(ALPHA-1.)
      RECTK=(-BBR-SQRT(BBR**2-4.*AAR*BRECT))/(2.*AAR)
      CRECT=1.+(ALPHA-1.)*RECTK
      XTNR=XDIST-RECTK
      XNR=XTNNR-RECTK
      TRAYN=SRECT*CRECT**2
      TRAYM=(SRECT*CRECT*(ALPHA-1.))/(ALPHA-TRAYN)
      ARRGN=XTNR*(1.-XNR*TRAYM)
      ARRGD=XNR*(1.-XTNR*TRAYM)
      ARRG=ARRGN/ARRGD
      IF(ARRG)131,131,130
  131 AH=XDIST
      GO TO 48
  130 TRAYO=ALOG(ARRG)
      TRAYP=TRAYO/ALOG(ALPHA/TRAYN)
      WRITE(3,55)TRAYP,TRAYO,YI,XDIST,RECTK
   55 FORMAT(1H0,5F10.6)
      IF(G-DABS(TRAYP-TRAYT))60,61,61
```

```
   60 IF(TRAYP-TRAYT)64,61,66
   64 DDY=YIH-YI
      IF(.000001-DDY)112,113,113
  113 AL =XDIST
      GO TO 48
  112 BL =YI
      GO TO 43
   66 BH=YI
      GO TO 43
   61 WRITE(3,62    )
   62 FORMAT(/// 3X,5HXBOTT,5X,2HRR,7X,5HSSTRP,5X,4HXNNS,5X,5HSTRPK)
C     CALCULATION OF THE BOTTOM PRODUCT COMPOSITION
      XV=(DELTA*(YI-XI))/FR
      NBOTT=XI*(XDIST-XFEED)*FR-XDIST*XV
      DBOTT=(XDIST-XFEED)*FR-XV
      XBOTT=NBOTT/DBOTT
      IF(XBOTT)69,69,68
   69 GO TO 76
   68 IF(XBOTT-1.)81,82,82
   82 GO TO 75
   81 SSTRP=(YI-XBOTT)/(XI-XBOTT)
      BSTRP=XBOTT*(1.-SSTRP)
      BBS=SSTRP&BSTRP*(ALPHA-1.)-ALPHA
      AAS=SSTRP*(ALPHA-1.)
      STRPK=(-BBS&SQRT(BBS**2-4.*AAS*BSTRP))/(2.*AAS)
      CSTRP=1.&(ALPHA-1.)*STRPK
      XBS=XBOTT-STRPK
      SCTRP=(SSTRP**TRAYS)*(CSTRP**(2.*TRAYS))
      C=SSTRP*CSTRP*(ALPHA-1.)
      D=(ALPHA**TRAYS-SCTRP)/(ALPHA-SSTRP*CSTRP**2)
      CA=ALPHA**TRAYS
      XNS=CA*XBS/(C*D*XBS+SCTRP)
      XNNS=XNS+STRPK
      WRITE(3,63)XBOTT,RR,SSTRP,XNNS,STRPK
   63 FORMAT(1H0,5F10.6)
      IF(XNNS-STRPK)65,64,64
C     CALCULATION OF THE TRAY NUMBERS IN THE RECTIFYING SECTION AND
C     AND THE TOTAL TRAY NUMBERS
   65 WRITE(3,70    )
   70 FORMAT(/// 3X,5HTRAYP,5X,5HXDIST,5X,2HRR,7X,5HSSTRP,5X,5HXBOTT,
     15X,5HSRECT,5X,5HTRAYD,5X,3HTTN,7X,4HXNNS,5X,2HYI)
      XOR=XDIST-RECTK
      IF(XOR)76,76,91
   91 XNR=XNNS-RECTK
      TRAYE=SRECT*CRECT**2
      TRAYG=(SRECT*CRECT*(ALPHA-1.))/(ALPHA-TRAYE)
      ARG=(XOR*(1.-XNR*TRAYG))/(XNR*(1.-XOR*TRAYG))
      IF(ARG)76,76,97
   97 TRAYH=ALOG(ARG)
      TRAYD=TRAYH/ALOG(ALPHA/TRAYE)
C     CHECK THE CALCULATED TOTAL TRAY NUMBERS WITH THE DESIRED ONE
      TTN=TRAYS&TRAYD
   71 WRITE(3,72)TRAYP,XDIST,RR,SSTRP,XBOTT,SRECT,TRAYD,TTN,XNNS,YI
   72 FORMAT(1H0,10F10.6)
   73 IF(F-DABS(TTN-TTNN))74,90,90
```

```
74 IF(TTN-TTNN)75,90,76
75 AH=XDIST
   GO TO 48
76 AL=XDIST
   GO TO 48
90 IF(M)80,201,80
80 RETURN
   END
```

SIZE OF COMMON 000024 PROGRAM 006360

END OF COMPILATION RECTY

```
      SUBROUTINE STRIP(TTNN,TRAYD,XFEEN,XDISN,XBOTN,DELTA,RN)
C     CONTROL OF TRAY IN STRIPPING SECTION
      DOUBLE PRECISION SS,TT,SSTRN,BBSN,AASN,STRKN,CSTRN,SCREN,AN
      DOUBLE PRECISION ALPHA,SLMIN,TRAYT,TTNN,YII,XII,QFEED,E
      DOUBLE PRECISION XBOTN,XDISN,QFFEN,RN,PLATA,XFEEN,BND,BNF,BN,XONS
      DOUBLE PRECISION XXNS,XTNNS,TRANE,TRANG,TRANO,BS,BSTRN
      DOUBLE PRECISION TRAYD,G,L,XFEED,AA,AB,XBOTL,XBOTH,SSMAX
      DOUBLE PRECISION BSMAX,BSM,ASM,STRMK,CSTRM,XTNS,XNS,TRAYW,TRAYX
      DOUBLE PRECISION TRAYY,TRATZ,AL,AH,XBOTT,BH,BL,YIH,YIL,F,FR,XI
      DOUBLE PRECISION YI,SSTRP,BSTRP,BBS,AAS,STRPK,CSTRP,TRAYN,TRAYM
      DOUBLE PRECISION TRAYO,TRAYP,XDIST,SRECT,RR,BRECT,BBR,AAR,RECTK
      DOUBLE PRECISION CRECT,XOR,SCRET,C,D,XNR,XNNR,XOS,TRAYE,TRAYG
      DOUBLE PRECISION TRAYH,TRAYS,TTN,DEL,EF,YID
      COMMON QFEED,ALPHA,PLATA
  201 READ(1,10)SLMIN,DEL,DELL,FN,FR,XFEED
   10 FORMAT(6F10.6)
   30 READ(1,301)E,G,F,EF,M
C     CALCULATION OF THE LIQUID COMPOSITION IN THE CONTROL TRAY AND THE
C     LOCATION OF THE CONTROL TRAY THAT COUNTED FROM THE BOTTOM
  301 FORMAT(4F10.6,1I2)
      WRITE(3,199)
  199 FORMAT(/// 3X,4HXXNS,5X,5HXTNNS)
  110 SS=RN*XFEEN+QFEED*XDISN-(RN+QFEED)*XBOTN
      TT=(RN+1.)*XFEEN+(QFEED-1.)*XDISN-(RN+QFEED)*XBOTN
      SSTRN=SS/TT
      BS=(XFEEN-XDISN)*XBOTN
      BSTRN=BS/TT
      BBSN=SSTRN&BSTRN*(ALPHA-1.)-ALPHA
      AASN=SSTRN*(ALPHA-1.)
      STRKN=(-BBSN&SQRT(BBSN**2-4.*AASN*BSTRN))/(2.*AASN)
      CSTRN=1.&(ALPHA-1.)*STRKN
  111 SCREN=(SSTRN**PLATA)*(CSTRN**(2.*PLATA))
      AN=SSTRN*CSTRN*(ALPHA-1.)
      BND=ALPHA-SSTRN*CSTRN**2
      BNF=ALPHA**PLATA-SCREN
      BN=BNF/BND
      XONS=XFEEN-STRKN
      XXNS=(SCREN*XONS)/(ALPHA**PLATA-AN*BN*XONS)
      XTNNS=XXNS+STRKN
      WRITE(3,116)XXNS,XTNNS
  116 FORMAT(1H0,2F10.6)
  108 WRITE(3,101)
  101 FORMAT(/// 3X,5HSSTRN,5X,5HSTRKN,5X,5HTRAYT)
      XNNS=XBOTN-STRKN
      TRANE=SSTRN*CSTRN**2
      TRANG=(SSTRN*CSTRN*(ALPHA-1.))/(ALPHA-TRANE)
      TRANO=ALOG((XXNS*(1.-XNNS*TRANG))/(XNNS*(1.-XXNS*TRANG)))
      TRAYT=TRANO/ALOG(ALPHA/TRANE)
  102 WRITE(3,103   )SSTRN,STRKN,TRAYT
  103 FORMAT(1H0,3F10.6)
C     CALCULATION OF THE INTERSECTION OF THE Q AND EQUILIBRIUM LINES
C     DURING DISTURBANCE
      XI=XFEED
```

8

```
  425 IF(QFEED-1.)413,432,414
  413 XIH=XI-DELL
      GO TO 415
  414 XIH=XI+DELL
  415 YIIN=ALPHA*XIH
      YIID=1.+(ALPHA-1.)*XIH
      YII=YIIN/YIID
      WRITE(3,412)
  412 FORMAT(/// 3X,3HXIH,7X,3HYII,7X,3HYIH)
      IF(QFEED-1.)424,309,424
  424 YIH1=((QFEED)/(QFEED-1.))*XIH
      YIH2=XFEED/(QFEED-1.)
      YIH=YIH1-YIH2
  422 WRITE(3,423)XIH,YII,YIH
  423 FORMAT(1H0,3F10.6)
      YID=YII-YIH
      IF(EF-DABS(YID))431,309,309
  431 IF(YID)430,309,416
  416 XI=XIH
      GO TO 425
  430 XIH=(XI+XIH)/2.
      GO TO 415
  432 WRITE(3,307)
  307 FORMAT(/// 3X,3HYIH)
      YIH=(ALPHA*XFEED)/(1.+(ALPHA-1.)*XFEED)
      WRITE(3,308)YIH
  308 FORMAT(1H0,1F10.6)
  309 WRITE(3,20)
   20 FORMAT(/// 3X,5HXBOTL)
C     CALCULATION OF THE BOTTOM PRODUCT COMPOSITION AT TOTAL REFLUX
      AA=(SLMIN**TRAYT)*XTNNS
      AB=(ALPHA**TRAYT-SLMIN**TRAYT)/(ALPHA-SLMIN)
      XBOTL=AA/(ALPHA**TRAYT-AB*XTNNS)
C     CALCULATION OF THE BOTTOM PRODUCT COMPOSITION AT MINIMUM REFLUX
      WRITE(3,21    )XBOTL
   21 FORMAT(1H0,1F10.6)
      WRITE(3,32    )
   32 FORMAT(/// 3X,5HTRAYZ,5X,5HXBOTH)
      XII=XFEED
      XBOTH=XBOTL&DEL
   87 SSMAX=(YIH-XBOTH)/(XII-XBOTH)
      BSMAX=XBOTH*(1.-SSMAX)
      BSM=SSMAX&BSMAX*(ALPHA-1.)-ALPHA
      ASM=SSMAX*(ALPHA-1.)
      STRMK=(-BSM&SQRT(BSM**2-4.*ASM*BSMAX))/(2.*ASM)
      CSTRM=1.&(ALPHA-1.)*STRMK
      XTNS=XTNNS-STRMK
  202 XNS=XBOTH-STRMK
      TRAYW=SSMAX*CSTRM**2
      TRAYX=(SSMAX*CSTRM*(ALPHA-1.))/(ALPHA-TRAYW)
      TRAY=((XTNS*(1.-XNS*TRAYX))/(XNS*(1.-XTNS*TRAYX)))
      IF(TRAY)40,40,203
  203 TRAYY=ALOG(TRAY)
      TRAYZ=TRAYY/ALOG(ALPHA/TRAYW)
      WRITE(3,34    )TRAYZ,XBOTH
```

```
   34 FORMAT(1H0,2F10.6)
      IF(2-DABS(TRAYZ-TRAYT))98,98,99
   98 DEL=2.*DEL
      GO TO 95
   99 IF(E-DABS(TRAYZ-TRAYT))94,40,40
   94 DEL=DEL/2.
   95 IF(TRAYZ-TRAYT)86,40,85
   86 XBOTH=XBOTH-DEL
      GO TO 87
   85 XBOTH=XBOTH&DEL
      GO TO 87
   40 AL=XBOTL
      AH=XBOTH
      J=0
C     ASSUME A BOTTOM PRODUCT COMPOSITION BY BISECTIONAL METHOD
C     ASSUME THE INTERSECTING POINT OF THE Q AND EQUILIBRIUM LINES
C     BY BISECTIONAL METHOD
C     CALCULATE THE TRAY NUMBERS BETWEEN THE CONTROL AND THE
C     FIRST BOTTOM TRAY
   48 XBOTT=(AL&AH)/2.
   53 WRITE(3,42    )
   42 FORMAT(/// 3X,5HTRAYP,5X,5HTRAYO,5X,2HYI ,7X,5HXBOTT,5X,5HSTRPK)
      BH=YIH
      BL=XFEED
   43 YI=(BH&BL)/2.
      IF(QFEED-1.)204,205,204
  204 QYF=XFEED/(QFEED-1.)
      QYY=QFEED/(QFEED-1.)
      XI=(YI+QYF)/QYY
      GO TO 44
  205 XI=XFEED
   44 SSTRP=(YI-XBOTT)/(XI-XBOTT)
      BSTRP=XBOTT*(1.-SSTRP)
      BBS=SSTRP&BSTRP*(ALPHA-1.)-ALPHA
      AAS=SSTRP*(ALPHA-1.)
      STRPK=(-BBS&SQRT(BBS**2-4.*AAS*BSTRP))/(2.*AAS)
      CSTRP=1.&(ALPHA-1.)*STRPK
      XTNS=XTNNS-STRPK
      XNS=XBOTT-STRPK
      TRAYN=SSTRP*CSTRP**2
      TRAYM=(SSTRP*CSTRP*(ALPHA-1.))/(ALPHA-TRAYN)
      TRAYO=ALOG((XTNS*(1.-XNS*TRAYM))/(XNS*(1.-XTNS*TRAYM)))
      TRAYP=TRAYO/ALOG(ALPHA/TRAYN)
      WRITE(3,55)TRAYP,TRAYO,YI,XBOTT,STRPK
   55 FORMAT(1H0,5F10.6)
      IF(G-DABS(TRAYP-TRAYT))60,61,61
   60 IF(TRAYP-TRAYT)64,61,66
   64 DDY=YIH-YI
      IF(.000001-DDY)112,113,113
  113 AH=XBOTT
      GO TO 48
  112 BL=YI
      GO TO 43
   66 BH=YI
      GO TO 43
```

```
C         CALCULATE THE TOP PRODUCT COMPOSITION
     61   WRITE(3,62    )
     62   FORMAT(/// 3X,5HXDIST,5X,2HRR,7X,5HSRECT,5X,4HXNNR,5X,5HRECTK)
          XL =(DELTA*(YI-XI))/FR
          XDIST=(XBOTT*XL-YI*(XFEED-XBOTT)*FR)/(XL-(XFEED-XBOTT)*FR)
          IF(1.-XDIST)69,69,68
     69   GO TO 76
     68   SRECT=(XDIST-YI)/(XDIST-XI)
          RR=SRECT/(1.-SRECT)
          BRECT=XDIST/(RR&1.)
          BBR=SRECT&BRECT*(ALPHA-1.)-ALPHA
          AAR=SRECT*(ALPHA-1.)
          RECTK=(-BBR-SQRT(BBR**2-4.*AAR*BRECT))/(2.*AAR)
          CRECT=1.&(ALPHA-1.)*RECTK
          XOR=XDIST-RECTK
          SCRET=(SRECT**TRAYD)*(CRECT**(2.*TRAYD))
          C=SRECT*CRECT*(ALPHA-1.)
          D=(ALPHA**TRAYD-SCRET)/(ALPHA-SRECT*CRECT**2)
          XNR=(SCRET*XOR)/(ALPHA**TRAYD-C*D*XOR)
          XNNR=XNR&RECTK
          WRITE(3,63    )XDIST,RR,SRECT,XNNR,RECTK
     63   FORMAT(1H0,5F10.6)
          IF(XNNR-RECTK)64,64,65
C         CALCULATION OF THE TRAY NUMBERS IN THE STRIPPING SECTION
     65   WRITE(3,70    )
     70   FORMAT(/// 3X,5HTRAYP,5X,5HXDIST,5X,2HRR,7X,5HSSTRP,5X,5HXBOTT,
        15X,5HSRECT,5X,5HTRAYS,5X,3HTTN,7X,4HXNNR,5X,2HYI)
          XOS=XNNR-STRPK
          IF(XOS)91,76,76
     91   XNS=XBOTT-STRPK
          TRAYE=SSTRP*CSTRP**2
          TRAYG=(SSTRP*CSTRP*(ALPHA-1.))/(ALPHA-TRAYE)
          TRAYH=ALOG((XOS*(1.-XNS*TRAYG))/(XNS*(1.-XOS*TRAYG)))
          TRAYS=TRAYH/ALOG(ALPHA/TRAYE)
C         CHECK THE CALCULATED TOTAL TRAY NUMBERS WITH THE DESIRED ONE
C         TOTAL TRAY NUMBERS
          TTN=TRAYS+TRAYD
     71   WRITE(3,72    )TRAYP,XDIST,RR,SSTRP,XBOTT,SRECT,TRAYS,TTN,XNNR,YI
     72   FORMAT(1H0,10F10.6)
     73   IF(F-DABS(TTN-TTNN))77,90,90
     77   IF(.25-DABS(TTN-TTNN))74,74,78
     74   IF(TTN-TTNN)75,90,76
     78   IF(TTN-TTNN)83,90,84
     83   IF(J-30)81,90,90
     81   XBOTT=XBOTT&.00001
          J=J&1
          GO TO 53
     84   IF(J-30)82,90,90
     82   XBOTT=XBOTT-.00001
          J=J&1
          GO TO 53
     75   AL=XBOTT
          GO TO 48
     76   AH=XBOTT
          GO TO 48
```

```
90 IF(M)80,201,80
80 RETURN
   END
```

SIZE OF COMMON 000024 PROGRAM 006454

END OF COMPILATION STRIP

12

APPENDIX 4

Computer Program for Feedforward Control

```
C CALCULATION OF TRAY NUMBER FEEDFORWARD CONTROL A.L.CHAN
C      FEED RATE AND FEED COMPOSITION DISTURBANCE
   11 READ(1,10    )TRAYD,XFEED,ALPHA,QFEED,E,G,L
   10 FORMAT(6F10.6,1I2)
      WRITE(3,41    )
   41 FORMAT(///,3X,5HXDIST,6X,2HRR,7X,5HSSTRP,5X,5HSRECT,5X,5HXBOTT,
     15X,4HXNNR,6X,5HTRAYS,5X,3HTTN,7X,5HRECTK)
      READ(1,12) XFEEN,XDISN,QFEEN,TRANS,XBOTN
   12 FORMAT(5F10.6)
      READ(1,24)RN,FN,DELTA
   24 FORMAT(3F10.6)
C     CALCULATION OF THE NEW REFLUX RATIO
C     ASSUME A VALUE FOR THE TOP PRODUCT COMPOSITION
C     CALCULATE THE LIQUID COMPOSITION OF THE FEED TRAY
      XDIST=0.799
   70 XDIST=XDIST&.001
      IF(G-1.)80,60,80
   80 XBOTT=(XFEEN-FN*XDIST)/(1.-FN)
      RR=RN/G
      GO TO 92
   60 FD=(XFEED-XBOTN)/(XDISN-XBOTN)
      XBOTT=(XFEED-FD*XDIST)/(1.-FD)
      RR=DELTA/FD
   92 SRECT=RR/(RR+1.)
      BRECT=XDIST/(RR&1.)
      BBR=SRECT&BRECT*(ALPHA-1.)-ALPHA
      AAR=SRECT*(ALPHA-1.)
      RECTK=(-BBR-SQRT(BBR**2-4.*AAR*BRECT))/(2.*AAR)
      CRECT=1.&(ALPHA-1.)*RECTK
      XOR=XDIST-RECTK
      SCRET=(SRECT**TRAYD)*(CRECT**(2.*TRAYD))
      A=SRECT*CRECT*(ALPHA-1.)
      B=(ALPHA**TRAYD-SCRET)/(ALPHA-SRECT*CRECT**2)
      XNR=(SCRET*XOR)/(ALPHA**TRAYD-A*B*XOR)
      XNNR=XNR&RECTK
      IF(XNNR-RECTK)11,11,20
C     CALCULATE THE TRAY NUMBERS IN THE STRIPPING SECTION
C     CALCULATE THE TOTAL TRAY NUMBERS
   20 SSTRP=(RR*XFEED&QFEED*XDIST-(RR&QFEED)*XBOTT)/((RR&1.)*XFEED&
     1(QFEED-1.)*XDIST-(RR&QFEED)*XBOTT)
      BSTRP=((XFEED-XDIST)*XBOTT)/((RR&1.)*XFEED&(QFEED-1.)*XDIST-
     1(RR&QFEED)*XBOTT)
      BBS=SSTRP&BSTRP*(ALPHA-1.)-ALPHA
      AAS=SSTRP*(ALPHA-1.)
      STRPK=(-BBS&SQRT(BBS**2-4.*AAS*BSTRP))/(2.*AAS)
      CSTRP=1.&(ALPHA-1.)*STRPK
      XOS=XNNR-STRPK
      XNS=XBOTT-STRPK
      TRAYE=SSTRP*CSTRP**2
      TRAYG=(SSTRP*CSTRP*(ALPHA-1.))/(ALPHA-TRAYE)
      TRAYH=ALOG((XOS*(1.-XNS*TRAYG))/(XNS*(1.-XOS*TRAYG)))
      TRAYS=TRAYH/ALOG(ALPHA/TRAYE)
      TTN=TRAYS+TRAYD

      WRITE(3,30    )XDIST,RR,SSTRP,SRECT,XBOTT,XNNR,TRAYS,TTN,RECTK
C     CHECK THE CALCULATED TOTAL TRAY NUMBERS WITH THE DESIRED ONE
      IF(E-ABS(TRAYS-TRANS))40,50,50
   40 IF(1.-XDIST)50,70,70
   30 FORMAT(1H0,9F10.6)
   50 IF(L)16,11,16
   16 CALL EXIT
      END
```

NOMENCLATURE

B	-	Bottom product withdrawal rate, lb-moles/min.
D	-	Distillate withdrawal rate, lb-moles/min.
F, F'	-	Feed rate, lb-moles/min.
L	-	Reflux rate lb-moles/min.
L'	-	Liquid rate in stripping section lb-moles/min.
R	-	Reflux ratio.
V	-	Vapor rate in rectifying section lb-moles/min.
V'	-	Vapor rate in stripping section lb-moles/min.
x	-	Mole fraction of lighter component in liquid phase.
x_B	-	Bottom composition.
x_D	-	Distillate composition.
x_F	-	Feed composition
x_o	-	Equal to x_D in rectifying section and x_F in stripping section.
x_n	-	Equal to x_F in rectifying section and x_B in stripping section.
a, b, d	-	Numerical constants.
m	-	Slope of operating line.
n, n'	-	Number of tray
q	-	Analytical function used to describe thermal condition of feed stream.
y	-	Mole fraction of lighter component in vapor phase.
$x_i y_i$	-	x and y coordinates of intersection of q-line and equilibrium line.
x'	-	Value of x after transformation of coordinates.

REFERENCES

1. Archer, D.H., Rothfus, R.R., Chem. Eng. Prog. Symp. Series, 57, No. 36, 2-19, (1961).

2. Bertrand, L. and Jones, J.B., Chem. Eng., 139-144, February 20, (1961).

3. Buckley, p.s. "Techniques of process control", John Wiley and Son, Inc., (1964).

4. Bauer, R.L. and Orr, C.P., Chem. Eng. Prog. 50, No 6, 312-318, (1954).

5. Bennett, C.O. and Myers, J.E., "Momentum, Heat and Mass Transfer", McGraw-Hill, (1962).

6. Cadman, T.W., Rothfus, R.R. and Kermode, R.I., ISA Transactions, 6, No. 1, 74-81 (1967).

7. Calvert, S., Coulman, G., Chem. Eng. Prog., 57, 45, (1961).

8. Coughanowr, D.R., Koppel, L.B., "Process Systems Analysis and Control", McGraw-Hill, (1965).

9. Crico, A., Genie Chimique, 90, No. 4, 102-110, (1963).

10. Distefano, G.P., Ph.D. Thesis, The University of Florida, (1964).

11. Forman, E.R., Chem. Eng., Reprint - Fundamental of Process Control - 33-38, (1966).

12. Forman, E.R., Chem. Eng., Reprint - Fundamental of Process Control - 38-44, (1966).

13. Fu, B.S.J., M.Sc. Thesis, University of Ottawa, (1965).

14. Hariott, P., "Process Control" McGraw-Hill, (1964).

15. Gould, L.A., Chem. Eng. Prog. Symp. Series, 59, No. 46, 155-159, (1963).

16. Hoffman, H.L., HP/PR, 42, No. 2, 108-122, (1963).

17. Huckaba, C.E., May, F.P. and Frank, F.R., Chem. Eng. Prog. Symp. Series, 59, No. 46, 38-47, (1963).

18. Lamb, D.E., Pigford, R.L., Rippin, D.W.T., Chem. Eng. Prog. Symp. Series, 57, No. 36, 132-147, (1961).

19. Lupfer, D.E., Parsons, J.R., Chem. Eng. Prog., 58, No. 9, 37-42, (1962).

20. Luyben, W.L., Gerster, J.A., I & EC, Process Design and Development, 3, 374-381 (1964).

21. Keating, J.M., Townend, D.S., Instrumentation and Computation in Process Development and Plant Design, 29-37, London, (1959).

22. MacMullan, E.C., Shinskey, F.G., Control Engineering, 69-74, March, (1964).

23. Marshall, W.R., Pigford, R.L., "The Application of Differential Equations to Chemical Engineering Problems", University of Delaware, (1947).

24. Moczek, J.S., Otto, R.E., Williams, T.J., I & EC, Process Design and Development, 2, No. 4, 288-296, (1963).

25. Rademaker, O., Rijnsdorp, J.E., Proceeding of the Fifth World Petroleum Congress, Section VII - Paper 5, 59-79 (1959).

26. Rosenbrock H.H., Trans. Inst. Chem. Eng., 40, 35-53 (1962).

27. Shinskey, F.G., The Oil and Gas Journal, 95-103, April, (1965).

28. Shinskey, F.G., ISA Journal, 61-65, November, (1963).

29. Smith, B.D., "Design of Equilibrium Stage Processes", McGraw-Hill, (1963).

30. Smoker, E.H., Trans. A. I. Ch. E., 34, 165, (1938).

31. Lubyen, W.L., Chem. Eng. Prog., 61, No. 8, 74-78, (1965).

32. Lupfer, D.E., Oglesby, M.W., Ind. Eng. Chem., 53, No. 12, 963-969, (1961).

33. Talbot, F.D.F., Can. Cont. and Inst. May-June, 37-39, (1964).