EVALUATION OF CONTROL SCHEMES OF A BINARY DISTILLATION COLUMN

by

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ABSTRACT

Utilizing experimental transient data, of a ten-plate pilot plant binary distillation column, reported by Gerster and co-workers (26), various control schemes were investigated by means of analog computer simulations for a 10% step disturbance in feed composition. Employing reflux manipulation to maintain constant overhead product composition, the optimal location of the sensing element was determined. The utility of Rosenbrock's optimum control point policy was also considered. The so-called two-point control exhibits strong interaction between the two control loops. For close control of the overhead product composition, feedforward plus feedback compensation control was found very effective in decreasing the maximum deviation of the controlled variable from the desired value and reducing the time required for the controlled variable to return to its original steady state value.

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Finally, he wishes to state his deep appreciation, with passion, to his wife for her understanding and constant support throughout the duration of this program.

CHAPTER I INTRODUCTION

A major advancement of the chemical industries during this century has been the development of continuous process plants operating on a large scale. The advent of continuous processing in the chemical industries has brought with it a requirement for a much more stringent control of the process variables. In general, continuous processes are more sensitive to operating conditions than batch processes (36). In addition, because of their generally large size, operating at other than optimum conditions is more costly in plant returns, which further increases the need for close control.

Continuous distillation is probably the most common chemical engineering mass transfer operation. This process is designed to separate a known composition feed stream into two or more product streams within a certain specification. To maintain steady state conditions in order to achieve a specified separation, a valve or simple on-off controller has in some cases been satisfactory, but when highly specified qualities of the products are required in spite of external upsets, stringent control has become necessary (36).

The general theory of feedback control is well established and adequate for the analysis of any control system that may be selected for study, provided the transient behavior of the process is known. For the case considered in this work, the process investigated is the distillation column and its auxiliary equipment. Several mathematical models for describing the transient response of a plate-type column separating a binary mixture have been proposed (3, 20, 23, 34, 41, 52, 53, 56). Several comparisons have also been made between theoretical predictions and experiments (6, 7, 8, 26, 27, 46, 50, 51). Unfortunately, it has been found laborious and time-consuming to predict the dynamic behavior of a column. Control has been found to depend largely on effects which are secondary to the main mass transfer effect of the column and about which little information is available. For this reason it appears to be as difficult to reach general conclusions from the results of unsteady state calculations as it is to state them from practical experience (41).

For designing control systems of a distillation column, early empirical work has offered some rules, primarily of the rule of thumb type, which often give -

satisfactory but not optimal results. In the last two decades, with the appearence of digital computers, theoretical analysis (10, 23, 28, 38, 41) has become possible as far as the computing time is concerned. As a result of these studies many rules have been confirmed or revised and a few more new conclusions have also been reached, but due to the complexity of the process a completely general characterization to the problem is still not valid, even for a binary system.

A novel approach for designing control systems for a binary distillation column has been developed by Rosenbrock (41), who used the concept of "disturbance function D", which is simply the sum of the magnitudes of the rates of change of composition on all the plates and is defined as

$$\frac{1}{2} D = \sum_{j=0}^{n+1} \left| \frac{d}{dt} H_j x_j \right| \tag{1}$$

where H_j is the liquid holdup on the jth plate and x_j is the composition on the jth plate. The sum is taken over all plates, the reboiler, and the condenser system. Later, Rosenbrock (42) established that the disturbance function D has the properties of a Lyapunov function and stipulates that control should reduce the "amount of disturbance" as quickly as possible. Both Gould (18) and Rijnsdorp et. al., (35) have pointed out that Rosenbrock's criterion does not take sufficient account of the top and bottom products. Since maintaining constant product composition is the major objective of the control of distillation columns, the inadequacy of Rosenbrock's criterion is rather significant.

Most research workers in this field (10, 12, 19, 20, 36, 44, 53, 54) agree that the optimal location of the sensing element is on the top plate of the column if the objective is to achieve a constant overhead product composition by reflux rate manipulation. A direct measurement of product stream composition would be even better if the sensing element were highly sensitive. When the detector has a large "dead space" (54) compromise is required and usually intermediate plate composition is controlled (53, 36, 44). Nevertheless, according to Rosenbrock's control strategy (41), one or two of the intermediate plate compositions should be controlled. On the basis of the preceding conclusions, it appears that they are contradictory to each other and as a result of this ambiguity, some misunderstandings have occurred (21) by applying Rosenbrock's optimal control policy to a different objective i.e., constant overhead product composition.

The purpose of this study was

(1) to utilize Rosenbrock's criterion to examine the variations in the product stream composition,

(2) to investigate the performance of the so-called two-point control, and

(3) to show the effect of a feedback compensator on the feedforward control system,

by simulating a ten-plate binary pilot plant distillation column (26) plus the control systems on an electronic analog computer.

CHAPTER II LITERATURE SURVEY

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 (4). Extensive bibliography of both dynamics and control of distillation columns can also be found in a historical and critical review article by Rosenbrock (40) in 1962. One year later, a survey on the application of automatic control in the chemical industry by Rijnsdorp et. al., (44) listed several other important references on the dynamics and control of distillation columns. Gould's (18) review article deserves special mention. He not only reviewed and analyzed the research activities in the characterization and control of distillation column, but also pointed out the shortcomings of various reported studies as well as many promising aspects.

All of the studies to be mentioned in the following sections are for binary systems. Very little information about multicomponent distillation has been reported, and since it is beyond the scope of this investigation, further discussion unless otherwise specified will all refer to a binary system.

II-1. DISTILLATION COLUMN DYNAMICS

BASIC EQUATIONS :

To describe the dynamics of a plate-type distillation column, a set of basic mass transfer equations (i.e., total material balance for each plate and each component, except one) plus phase equilibrium relationships, plate efficiency, and energy balance equations for the condenser and reboiler systems and feed plate as well as the terminal plates are sufficient (4). Marshall and Pigford (29) published the first analysis of the equations of a binary distillation column suitable for study of steady state or transient behavior.

METHOD OF SOLUTION :

Analytical solutions can be obtained with customary assumptions by using the Laplace transformation and difference-differential equations or perturbation techniques (3, 29, 50) for certain cases. The analog computer has been employed to study the transient behavior for binary distillation columns (23, 26) and some other mass transfer units (1).

Rijnsdorp and Maarleveld (35) have proposed an electrical analog for a distillation column which permits the simulation of pressure and hydrodynamic transients and their interaction with the composition transients. Acrivos and Amundsen (2) have shown how matrix methods can be employed to determine the transient response of stage-by-stage mass transfer equipment. Rosenbrock (37, 38) and many others (10, 20) have used the digital computer to solve the array of equations.

Another approach, namely that of considering the plate number j as a continuous rather than a discrete variable, has been studied (17, 32, 52) and further elaboration and refinement is still in progress.

THEORETICAL AND EXPERIMENTAL WORK :

Early investigators, e.g., Marshall and Pigford (29) and Davidson (17), considered the start-up problem of a fractionating column and the rate at which the column approaches its final equilibrium state, but not the effect of a disturbance on that state. Wilkinson and Armstrong (51) calculated the transient response of a five-plate enriching column to step changes in the composition of the reboiler-vapor and comparison with experimental data showed good agreement at long-time intervals.

Voetter (50) has analytically solved the basic mass transfer equations, with certain assumptions, for step changes in feed composition and experimental data obtained from a sieve tray type column confirmed the theoretical predictions. Armstrong and Wilkinson's experimental data on the transient response of a plate-type distillation column following disturbances in feed composition and reflux rate are in agreement with the solution of the governing mass transfer equations solved by Rosenbrock (37, 38) with the digital computer method and also in accord with the theoretical predictions of Voetter (50).

Campbell (14) introduced the usefulness of the signal flow diagram concept and Rijnsdorp et. al., (34, 35) utilized it to synthesize a passive electrical network analog to represent the column dynamically. Many results, reported in transfer function form, have been obtained by Rademaker and Rijnsdorp (34), but unfortunately very few of these transfer functions have been published. Izawa and Morinaga (20) derived various transfer functions by reduction of the signal flow diagram and by use of the theorem of continued fraction to avoid the boundary value difficulties.

Several other workers (12, 56) have also pointed out the advantages of studying column dynamics via block diagram technique, which again is developed from the signal flow diagram concept.

Lamb, Pigford and Rippin (23) have derived a set of linear small perturbation type equations, with curved vapor-liquid equilibrium relationship taken into account, to describe the dynamic characteristics of a plate-type distillation column and also have shown the analog circuit for simulation of these equations. Under close control, the approach of linearization about the column's true steady state is well justified as demonstrated by the good agreement with the excellent experimental transient data reported by Gerster and co-workers (7, 8, 26, 27, 46).

II-2. DISTILLATION COLUMN CONTROL

CONTROL SCHEMES :

Numerous satisfactory and practical control schemes for distillation column control have been reported in the literature, however only a few references which are illustrative and collective will be cited in this study.

Rademaker and Rijnsdorp (34) introduced the principles of the classification for the most common types of columns and control systems. The book by Harriott (19) discusses various simple control schemes and possible combinations of these schemes while Buckley (12) in his book deals with the more practical aspects of distillation control systems. Various other more sophisticated control schemes can be found in the literature (11, 12, 25, 28, 31).

FEEDBACK CONTROL :

1. GENERAL :

In 1948, Boyd (9) reviewed various types of distillation control systems and presented recommendations for controlling column pressure and other variables. He made a plea for instrumentation and control which can maintain operations as steady as possible and prevent such upsets as slugging, for example. In 1952, Coulter (16) set down the results of experience, discussing the considerations involved in instrumenting a column and applying the principles outlined. Parkins (33) emphasized the importance of regarding the distillation column as an energy system, particularly when a sharp separation is required.

The study reported by Rademaker and Rijnsdorp (34) mentioned previously form a connecting link in the literature between the quasi-static and dynamic behavior of a distillation column. They considered that analog techniques were superior for studying control systems and by substitution of a passive network for amplifier circuits wherever possible an analysis of a column containing large number of plates was made practicable. Transfer functions including pressure effect for a section of column were summarized and certain frequency response data were shown. Rose, Williams and co-workers (36, 53, 54) used relatively simple models for a five-plate distillation column and by simulation on an analog computer studied the behavior of various types of controllers. They concluded that derivative control was ineffective, and that proportional control was generally adequate for step and sinusoidal disturbances. The conclusions of Williams et. al., (36, 53, 54) were substantiated by Brown (16) utilizing a digital computer.

An interesting approach, which may aid in understanding the effect of nonlinearities on control, is the concept of the "disturbance function", as mentioned in Chapter I, which was suggested by Rosenbrock (39, 41) in 1962. He (42) has also shown that the "disturbance function D" constitutes a Lyapunov function for binary distillation and can be used to investigate stability in the nonlinear case.

2. <u>CONTROL POINT LOCATION</u> :

It has often been stated (3, 49) that in case of reflux manipulation the sensing element should be installed at the plate where the concentration gradient is a maximum. The results of many studies (19, 20, 53, 34, 54) have shown that this criterion is based only upon steady state considerations. Consideration of the dynamics shows that if the major objective is minimum variation of product composition the detector should be installed as high as possible, with a compromise being made between the sensitivity of it and the equivalent dead time.

Rosenbrock (41), using the minimum "disturbance function D" as his criterion, has specified a different optimum control point location policy.

3. INTERACTION :

There are two kinds of interactions associated with a process, namely physical interaction and interaction between two or more independent control loops. Since the former type is concerned with the process dynamics, i.e., the parameters of each element used to describe the system cannot be specified separately, it will not be reviewed here.

Interaction between pressure control and the reflux manipulation is a known fact (15, 34). Rosenbrock (41), in 1962, was the first to demonstrate the interaction between two control loops when reflux and vapor boilup rate manipulations are used. Jafri et. al., (22) in an analog computer study of two-point control of a ten-plate laboratory column has also reported the presence of an interaction effect.

FEEDFORWARD CONTROL :

The possible application of feedforward control to a chemical process was first discussed qualitatively by Calvert and Coulman (13) in 1961. One year later, Oglesby and Lupfer (31) reported the design of a special purpose analog computer for feed enthalpy control which was the first actual application of feedforward control to a distillation column. Later, Lupfer and Parsons (25) described the synthesis of feedforward controllers of a distillation column from an empirically determined transient response of the column. Lamb, Pigford and Rippin (23) synthesized a feedforward controller using a linear model for a seven-plate column. Transfer functions for the column were computed from a frequency-domain solution of the model and matrix methods were then used to solve for the feedforward controller transfer functions. Luyben and Gerster (28) modified the approach of Lamb, Pigford and Rippin by calculating the feedforward controller transfer functions directly from an approximation on the Bode diagram. The performance of the controller was then determined by analog simulation of a ten-plate and a forty-plate column, and 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.

CHAPTER III THEORY

III-1. PROCESS

VARIABLES :

The number of variables of a typical distillation column separating an N-component mixture as shown schematically in Figure 1 is 3N + 12. A list of these variables is shown as Table 1 while tabulation of the 2N + 6 independent relationships which exist among these variables is illustrated in Table 2.

DEGREES OF FREEDOM :

The degrees of freedom inherent in a system is defined (30) as the number of variables which describe the system minus the number of independent relationships or equations which exist among these variables. Therefore, the degrees of freedom for any distillation column is N + 6. For a binary system and with fixed feed location i.e., E and S fixed, the degrees of freedom is six.

CONTROL SYSTEM :

The primary objectives of distillation column control are product quality control i.e., to maintain overhead and/or bottom product compositions at a specified value (no side streams are withdrawn), and material balance control i.e., to keep the average sum of both top and bottom product withdrawal rates to be exactly equal to the average feed rate. However, due to the physical limitations inherent in the process, certain constraints must be observed (12). For example :

(1) The column should not flood;

(2) Column pressure drop should be high enough to maintain effective column operation.

Therefore, sometimes in order to keep one end at a specified composition the composition of the other end product has to be sacrificed and can only be controlled as close to the desired value as possible without violating the physical constraints. Usually no attempt (19) is made to control both product compositions, although two-point control is theoretically possible by regulating two of the manipulated variables simultaneously. A careful analysis of the process to determine the degrees of freedom inherent in the system must be performed. The addition of controllers is equivalent to establishing functional relationships among the variables measured and controlled. These relationships are additional independent relationships among the variables, and therefore, they remove degrees of freedom. The degrees of freedom removed by the addition of a control system should not exceed the number of degrees of freedom which are inherent in the system or the system will be overdetermined.

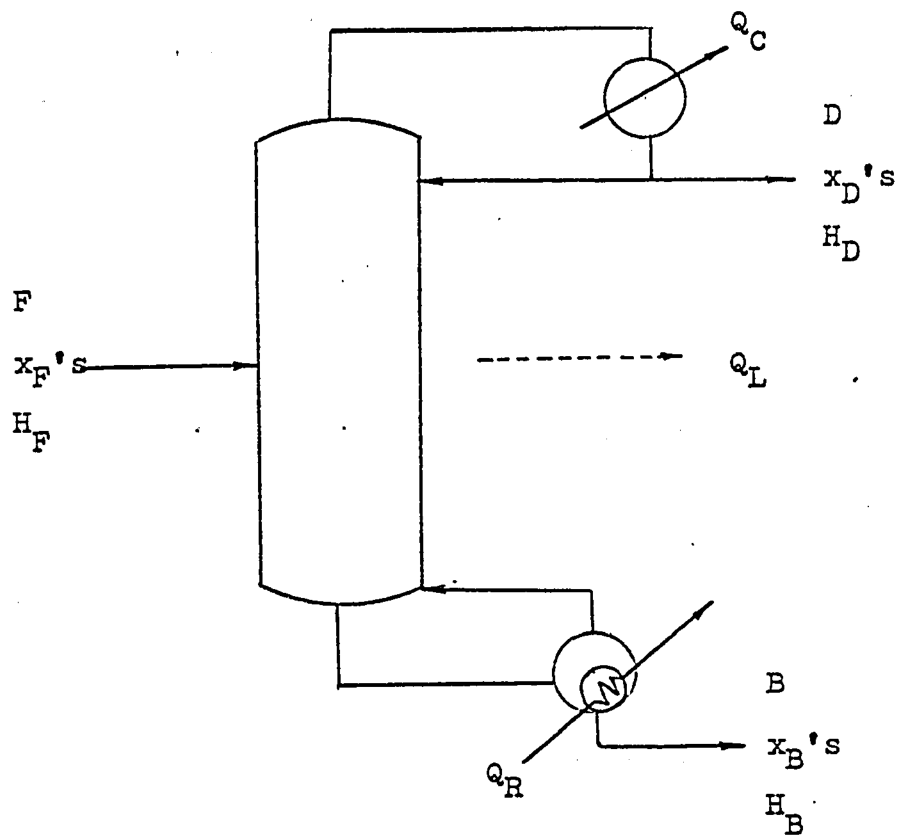


Fig. 1 Variables of a distillation column

TABLE 1

TABULATION OF DISTILLATION VARIABLES

E (number of enriching stages)	1
S (number of stripping stages)	1
Q_R	1
Q_C	1
Q_L	1
F	1
H_F	1
x_F	N
D	1
H_D	1
x_D	N
B	1
H_B	1
x_B	N
P	1
Total	3N + 12

TABLE 2

TABULATION OF THE INDEPENDENT RELATIONSHIPS AMONG THE VARIABLES

Equilibrium relationships	N
Material balance	N
$\sum x$'s = 1.0	3
Total energy balance	1
Condenser energy balance	1
Reboiler energy balance	1
Total	2N + 6

III-2. CONTROL

FEEDBACK CONTROL :

The design of a control system usually starts with certain specific information such as a mathematical description of the plant or process to be controlled, specified performance requirements under dynamic and static conditions, constraints and limitations on the system components, the inputs to the process and the variables, the statistical properties of the external disturbances, and a statement of the output variables which can be measured directly or can be estimated. In addition, extra requirements such as the introduction of sampling and quantization, the use of special equipment such as a digital computer, and the specification of certain control system configurations or models of control may be imposed. In some cases the solution of the problem may be unique; i.e., there is a system which is best; in other cases compromises may lead to several satisfactory solutions.

The methods for the design and synthesis of control systems may be classified into three categories. The first is the well known method based upon transform techniques, including the s domain and z domain, or upon the root-locus technique. This is often referred to as the trial-and-error design procedure (48). The designer is given a set of rather arbitrary specifications in the time and/or in the frequency domain and possibly a system configuration. Gain margin, phase margin, peak overshoot, decay ratio, etc., are the most commonly used specifications. The designer seeks to satisfy the specifications by altering the gain and/or other controller parameters. The second method, which has been referred to as analytical design, uses an infinite-time integral of squared system error or a mean-square error as a measure of performance. The design procedure yields a compensation network for a linear control system configuration upon application of classical variational methods to minimize the performance measure. During World War II and in the decade following, the first and second method constituted the major technique for the design of control systems. These techniques, which are employed extensively by practicing control engineers, may be referred to as the classical approach.

The block diagram of a typical feedback control system is illustrated in Figure 2. The classical approach of system design is confronted by severe limitations and difficulties when applied to the design of multivariable and time-varying systems. Applications usually are limited to idealized and relatively simple feedback control systems. The outline of the two methods is given in Tables 3 and 4 respectively.

The third method is a broad generalization of the second method and has been developed in many different ways. If, however, one restricts his interest to the control of linear processes, a rather clear pattern of design procedures appears as pointed out in the book by Tou (48). The essential idea is that a functional on the plant variables and control variables is specified or selected, constraints are imposed on the variables, and a control law or sequence is derived by finding an extremum of the functional by some variational technique. This method is rapidly becoming the major technique for the design of control systems in many fields other than chemical engineering. In 1962, Rosenbrock (43) introduced a concept similar to the third method for application to chemical processes and as Rijnsdorp et. al., (44) pointed out, one year later, it will be interesting to see Rosenbrock's method applied to a specific example but as yet no work toward that goal has been reported in the

TABLE 3

TRIAL-AND-ERROR METHOD

I. SPECIFICATIONS

1. Dynamics of the process
2. Desired values of the controlled variables
3. Load disturbances
4. Tolerable error and degree of stability (e.g., decay ratio as a criterion)

II. DESIGN PROCEDURE

1. Determine the system (the maximum controller gain) gain on the basis of error specification.
2. Design or choose the type of controller action to meet the stability requirement
3. By analysis, check to see if the specifications are satisfied
4. If not, repeat procedure, using a more elaborate controller

TABLE 4

ANALYTICAL DESIGN PROCEDURE

I. SPECIFICATIONS

1. Dynamics of the process
2. Desired values of the controlled variables
3. Load disturbances
4. Performance criterion
5. Physical constraints or limitations of the process

II. DESIGN PROCEDURE

1. Determine the classification of the problem--- fixed, free, or semifree configuration (48)
2. If the configuration is fixed, express the performance index in terms of the free parameters (parameters of the controller); then minimize or maximize the performance index by appropriate adjustment of these parameters
3. If the configuration is free or semifree, apply the method of spectral factorization to design the compensator (special type controller) which minimizes or maximizes the performance index
4. Check to see if the compensator (controller) thus

(Cont. of Table 4)

determined satisfies the performance criterion. If so, the theoretical design is completed and physical realization may begin. If not, the specification cannot be met and must be altered.

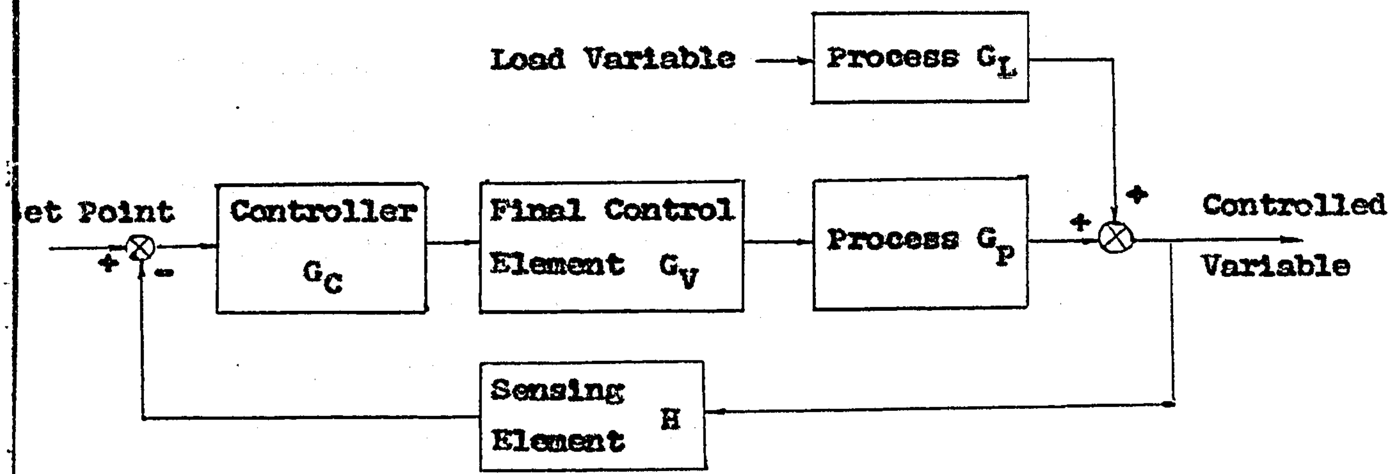


Fig. 2 Block Diagram of a typical feedback control system

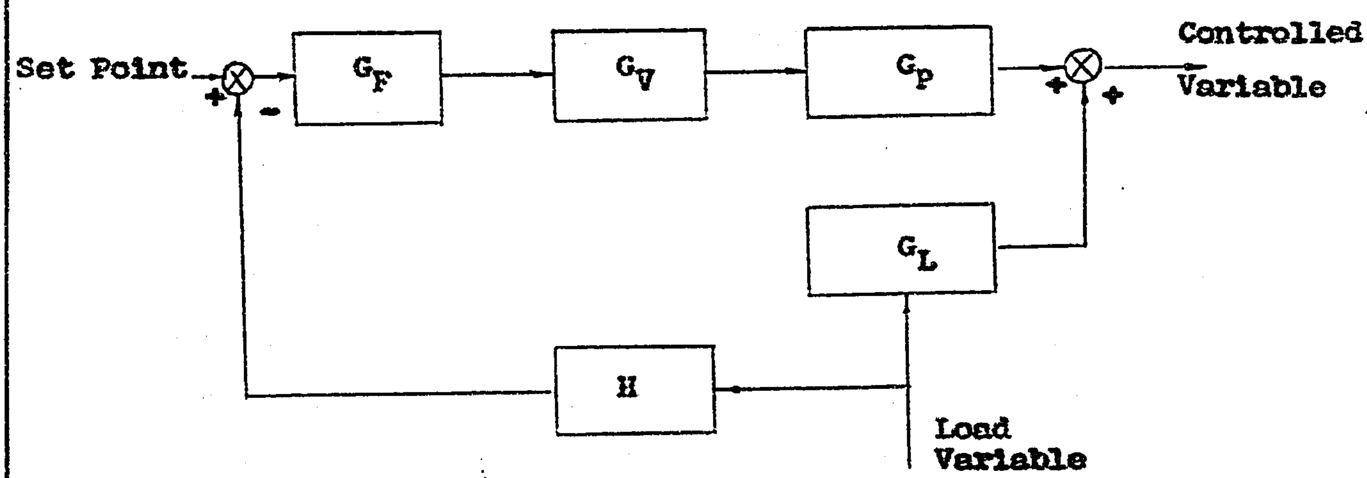


Fig. 3 Block Diagram of a simple feedforward control system

literature.

In the present study, since each individual control loop of the distillation column was examined independently, it was satisfactory to follow the classical approach.

There are some distinct disadvantages of feedback control, namely

(1) corrective action is only taken after product qualities have deviated from their desired values,

(2) a long time is required for the controlled variables to return to their desired values, in many processes such as large distillation columns,

(3) undesirable interaction of feedback control loops may occur as most chemical processes are multivariable.

<u>FEEDFORWARD CONTROL</u> :

Automatic control was first developed by the mechanical engineers and because of this, the emphasis in theoretical research has been on feedback systems. Chemical engineers utilizing the similarity between mechanical systems and chemical process systems started to develop their own theory, however, there are some fundamental differences between those two systems (13). Due to these differences of the two systems, and from the accumulated experience of chemical engineers, the concept of feedforward control appears to be attractive to process control engineers.

Feedforward control, as shown by the block diagram in Figure 3, in principle eliminates the main disadvantages of feedback control by sensing the input disturbances as they enter the process and taking the proper corrective action. This implies knowledge of all significant inputs and of an adequate causal relationship between inputs and outputs. Consequently, the desired outputs may be obtained by adjusting the inputs to the predicted settings.

Feedforward control in pure form is the opposite of feedback control in that in feedforward control the control action is initiated by input variables rather than by output variables as in feedback control. Feedforward control can be utilized in a varied range of applications from individual unit control up to the over-all control of an entire plant system. Individual unit feedforward control will be attractive where (13) :

(1) Input analyses are most economically provided by analytical devices.

(2) ~~Feedback dynamics may introduce severe instability.~~

(3) On-line continuous computer control (as distingushed from optimizing control at discrete intervals) is economically justifiable.

There are also shortcomings of feedforward control, namely

(1) the process may not correspond exactly to the mathematical or empirical models used to describe it's dynamic behavior,

(2) changes in controlled variables caused by unmeasured input disturbances go uncorrected, and

(3) physical and process limitations may constrain manipulative inputs so that they can not be adjusted to the required values by the feedforward controllers.

FEEDFORWARD PLUS FEEDBACK COMPENSATION CONTROL :

Since feedforward control and feedback control each has it's own disadvantages, a compromised control system apparently appears to be the combination of the two control philosophies, i.e., use feedforward control to eliminate measurable input disturbances and add a feedback compensator to compensate for any variations of quality in the desired products caused by some unpredictable and unmeasured disturbances. To illustrate this concept, a block diagram is shown in Figure 4.

For a multivariable system, it is not necessary to have a feedback compensator on every controlled variable. Adding feedback compensators to those variables which are critical is sufficient. Nevertheless, an analysis of the degrees of freedom inherent in the system is still necessary.

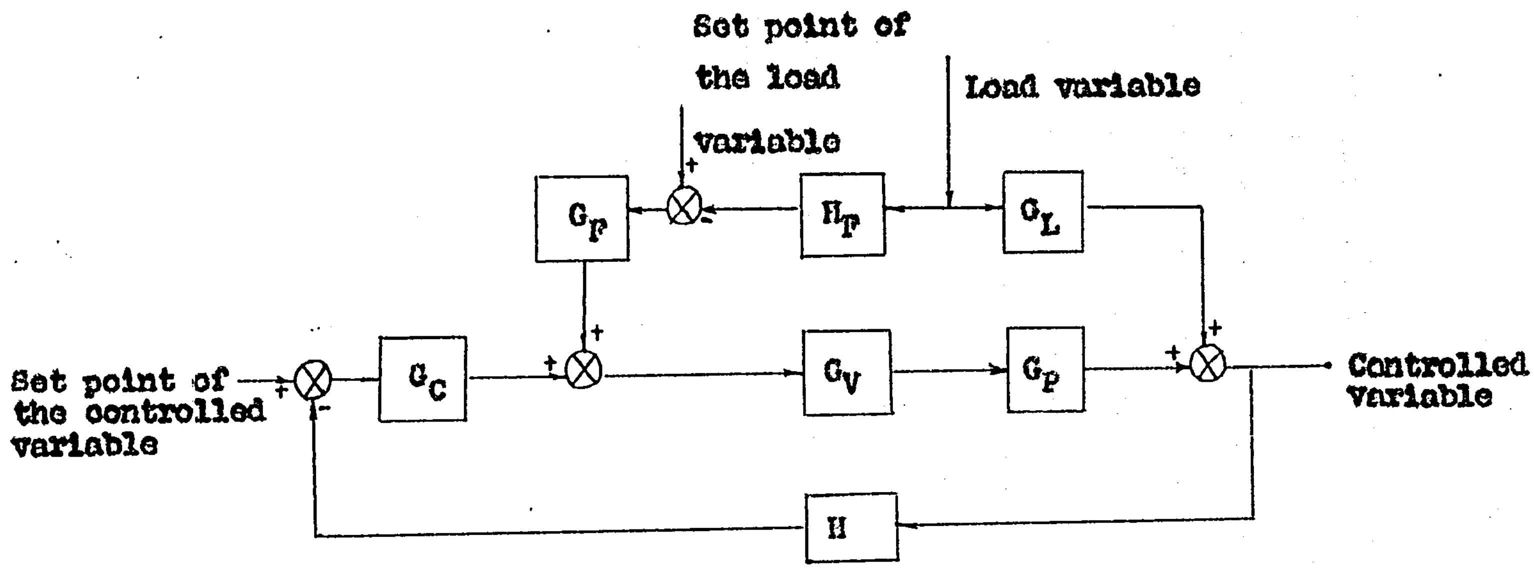


Fig. 4 Block diagram of a typical feedforward plus feedback compensation control system

CHAPTER IV PROBLEM FORMULATION

IV-1. PROCESS IDENTIFICATION

The major requirement in the design of control systems for distillation columns is the knowledge of the dynamic behavior of the system. In this study, empirical mathematical models for describing the dynamic behavior of the system were obtained by approximating the experimental transient data of Gerster et. al., (26) by second order overdamped type transfer functions. Gerster and co-workers obtained their data from a ten-plate pilot plant distillation column. A detailed description of the equipment used by Gerster and co-workers and the operating conditions selected for this study are tabulated in Table 5.

TABLE 5

PROCESS IDENTIFICATION

SYSTEM : Acetone-benzene mixture

EQUIPMENT :

Column :

Inside diameter : 2 ft.

Number of plates : 10

Feed plate location : 5th plate from bottom

Caps : 17-3 in. round bubble caps per plate on 4.5 in. spacing in 2.45 sq. ft. area

Outlet weir height : 2 in.

Downcomers : 0.345 sq. ft. cross section

Plate spacing : 18 in.

Condenser :

Downflow of vapor through 121 vertical tubes, 5/8 in. O.D., 75 in. long

Reflux drum :

100 gal.

Reboiler :

Natural circulation, thermosiphon.

Upflow of liquid through 163 vertical tubes. 3/4 in. O.D., 122 in. long, discharging to 50 gal. separator.

(Cont. of Table 5)

OPERATING CONDITIONS :

System pressure, P : 25 psig

Feed rate, F : 1.096 lb-moles/min.

Overhead product rate, D : 0.560 lb-moles/min.

Bottoms product rate, B : 0.536 lb-moles/min.

Reflux rate, R : 1.155 lb-moles/min.

Reflux ratio, R/D : 2.05

Feed composition, x_F : 44.2 mole % acetone

Vapor flow rate below feed plate, V : 2.07 lb-moles/min.

Overhead product composition, x_D : 82.2 mole % acetone

Bottoms product composition, x_B : 2.5 mole % acetone

IV-2. OVERALL CONTROL OF THE PROCESS

CONTROLLED VARIABLES :

1. Compositions of both end products or any two plates of the column
2. Feed enthalpy
3. Feed rate
4. Heat input to the reboiler
5. Pressure of the column

DISTURBANCES :

1. Feed composition
2. Ambient temperature
3. Mechanical failures

OBJECTIVE :

The overall control is designed to keep all of the above six variables at constant values in spite of the external disturbances. However, since disturbances such as ambient temperature and mechanical failure are unpredictable, the major objective of control is to ensure constant quality when there are disturbances in feed composition. The other four variables could be predetermined on the basis of a statistical evaluation of the disturbances and steady state calculations and controlled at those values. For this reason, the present investigation concentrated on the control of product quality. The control of the other four variables are usually referred to as "secondary control" (19), and control is designed primarily to stabilize the process under a specific operating condition. Buckley (12) and Harriott (19) outlined the various aspects of such control schemes in their books, and some of the specific features of these control schemes were discussed in many other sources (25, 31, 34). For the purpose of this study, it has been assumed that feed enthalpy, feed rate, heat inputs to the reboiler and pressure of the column are closely controlled. A typical overall control scheme for a column operating under such conditions is shown in Figure 5.

ANALYSIS OF THE DEGREES OF FREEDOM :

As established in Chapter III, the degrees of freedom inherent in a binary distillation column is eight (d.v. $= N + 6 = 8, \quad N = 2$).

Tabulation of the degrees of freedom removed by the control of those six variables mentioned previously is shown in Table 6. From the results shown in Table 6, it is obvious that the system under consideration is not overdetermined.

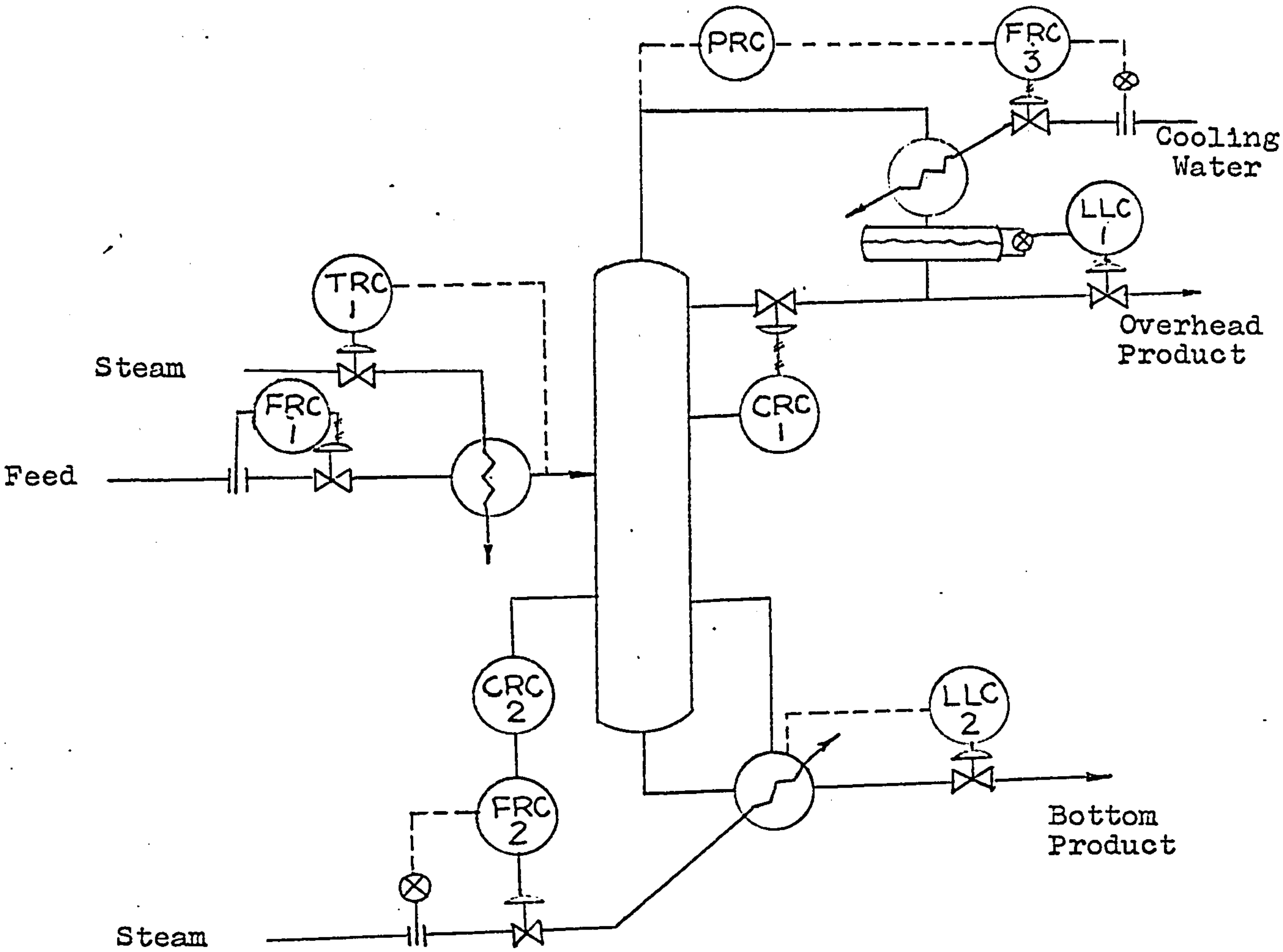


Fig. 5 A typical overall control scheme for a binary distillation column

TABLE 6

DEGREES OF FREEDOM ANALYSIS

Condition	Degrees of Freedom Removed
Operation with a fixed feed stage i.e., E and S fixed	2
FRC-1 sets feed rate	1
TRC-1* is sufficient to produce a constant enthalpy feed	1
FRC-2 is sufficient to yield a constant boilup rate	1
CRC-1 establishes a relationship between composition (under constant pressure) and reflux rate	1
CRC-2 establishes a relationship between composition (under constant pressure) and boilup rate	1

(Cont. of Table 6)

PRC and FRC-3 together establish a relationship between column pressure and the rate of heat removal from the condenser	1
Total degrees of freedom removed	8

* More sophisticated methods (11, 31) to control the feed enthalpy can be found in the literature.

IV-3. CONTROL SCHEMES TO BE STUDIED

1. ONE-POINT CONTROL :

Criterion :

To maintain constant overhead product composition, x_D.

Control Scheme :

A plate composition is controlled by manipulating reflux rate when there is any variation in the specified feed composition.

Object :

To find the optimum location of the sensing element, i.e., the plate whose composition should be controlled, in order to satisfy the above mentioned criterion.

2. TWO-POINT CONTROL :

Criterion :

To maintain constant overhead and bottom products compositions.

Control Scheme :

To maintain two plate compositions constant by manipulating reflux rate and boilup rate to compensate for any disturbances in feed composition.

Object :

To determine if the two control loops interact.

3. <u>FEEDFORWARD PLUS FEEDBACK COMPENSATION CONTROL</u> :

<u>Criterion</u> :

To maintain constant overhead product composition, x_D.

<u>Control Scheme</u> :

Overhead product composition is closely controlled by applying an instantaneous corrective action to the reflux rate, when there is a disturbance in feed composition, and a feedback compensator is employed to eliminate off-set due to imperfect models or unpredicted disturbances and to speed up the response of the control system. This control scheme is illustrated schematically by the block diagram shown as Figure 6.

<u>Object</u> :

(1) To compare the performances of feedback, feedforward and feedforward plus a feedback compensator control, and

(2) To synthesize a perfect feedforward controller (a mathematical model for G_F) which would allow no variation to appear in product composition at any time.

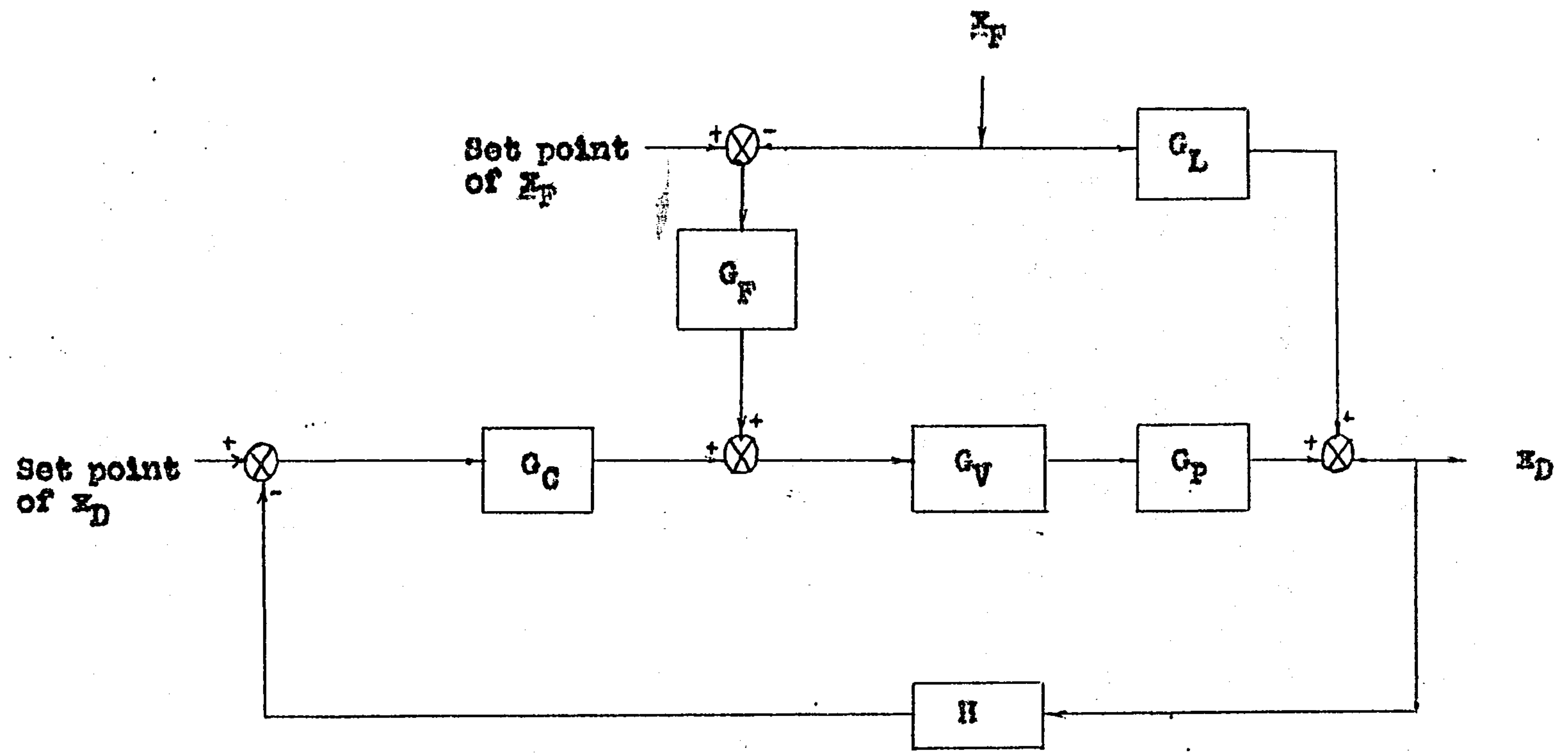


Fig. 6 Block diagram of the feedforward plus feedback compensation control scheme

CHAPTER V

SIMULATION OF CONTROL SCHEMES

The performance of the control schemes listed in the previous chapter were simulated on an electronic analog computer. The specific details of all the simulations in this study are given in Appendix 2.

PROCESS TRANSFER FUNCTIONS :

It was assumed that the column under close control behaves linearly and the relationships between variables can be represented by transfer functions. Transfer functions were developed by considering the experimental transient data of the type of second order overdamped as

$$G(s) = \frac{K}{(T_1 s + 1)\,(T_2 s + 1)} \qquad (2)$$

Such experimental transient data were obtained by Gerster and co-workers (26) on a ten-plate pilot plant distillation column.

The two time constants and gain for each of the transfer functions that were employed in this study are listed in Tables 7 and 8.

CONTROLLER :

On the basis of analog computer studies by Williams and co-workers (36, 53, 54, 55) on feedback control

TABLE 7

NOTATIONS FOR THE TRANSFER FUNCTIONS

INPUTS / OUTPUTS	x_F	R	V
x_D	G_{DL}	G_{DP}	-
x_9	G_{9L}	G_{9P}	-
x_7	G_{7L}	G_{7P}	G_{7B}
x_3	G_{3L}	G_{3P}	G_{3B}
x_1	G_{1L}	G_{1P}	-

TABLE 8

PARAMETERS OF THE TRANSFER FUNCTIONS

G	T (min.)	T (min.)	K — For G_P, $K_P = (\underline{X}_j)_f/(\underline{R})_f/(R)ss$ For G_L, $K_L = (\underline{X}_j)_f/(\underline{X}_F)_f$
G_{DL}	7.55	0.15	0.5800
G_{DP}	8.20	0.10	0.3685
G_{9L}	7.80	0.65	1.1700
G_{9P}	6.65	0.27	0.6920
G_{7L}	6.60	0.60	1.6700
G_{7P}	8.00	0.46	1.0520
G_{3L}	7.90	0.10	1.9400
G_{3P}	8.25	0.65	1.1750
G_{1L}	9.20	2.50	1.1250
G_{1P}	7.95	3.25	0.9650
G_{7B}	7.95	0.60	1.6500
G_{3B}	7.56	1.14	2.5200

systems, a proportional-integral controller was chosen for this investigation. The transfer function G_C of an ideal proportional-integral controller can be stated as

$$G_C(s) = K_C \left(1 + \frac{1}{T_R s} \right) \qquad (3)$$

where K_C = Controller gain (Δpsi/12 psi/mole % acetone)

T_R = Integral time (min.)

CONTROL VALVE :

The valve was considered to be pneumatic and to exhibit linear characteristices and therefore its behavior could be represented in transfer function notation as

$$G_V = K_V = (0.182/1.155)/(1/12) = 2 \text{ (dimensionless)}$$

SENSING ELEMENT :

A first order lag, which has a time constant equal to 1 min., was assumed for the sensing element in the feedback loop and its transfer function is

$$H = \frac{K_m}{T_m s + 1} = \frac{1}{s + 1} \qquad (4)$$

where K_m = Gain of the sensing element = 1

T_m = Time constant of the sensing element = 1 min.

ANALOG COMPUTER SIMULATION :

Simulation of the control systems was done on an electronic analog computer by employing a step-by-step approach whose steps are as follows :

(1) Insert the proper transfer functions which are expressed in the Laplace transformation operator s domain, into each of the blocks of the control system block diagram.

(2) Synthesize a non-scaled computer circuit on the basis of the block diagram together with the transfer functions.

(3) Scale the variables (including the time variable) carefully and properly to avoid overloading any of the amplifiers used in the circuit as well as to use the computer to its maximum efficiency. Scaling of the analog simulation may be done by preparing a digital computer program for it, but for a simple project the trail-and-error method is more convenient.

(4) Patch the scaled circuit on the analog computer and, if none of the amplifiers are overloaded (which will usually necessitate rescaling the problem), proceed to record the output voltages of the desired variables for the input specified.

SCALE FACTOR :

Two types of scale factors are utilized in analog computer simulation, namely a time scale factor and a magnitude scale factor. In this study, the time scale factor was chosen as one minute of real problem time equals one second of computer time;, and different magnitude scale factors, depending upon the behavior of each variable, were introduced.

V-1. ONE-POINT CONTROL

CONTROLLED VARIABLE :

Overhead product composition, x_D, or any intermediate plate composition, x_j.

MANIPULATED VARIABLE :

Reflux rate, R (lb-moles/min.).

DISTURBANCE :

A 10 % step change in feed composition, x_F.

BLOCK DIAGRAM :

The overhead product stream, 9th, 7th, 3rd and 1st plate were the five locations of the sensing element studied in this investigation. The block diagram for the case in which the overhead product composition is controlled is shown in Figure 7. The block diagram representation of the other four cases is shown in Figure 8, which also shows the overhead product composition response by assuming a linear relationship between x_D and R and x_F.

ANALOG COMPUTER SIMULATION :

The details of the procedure of the synthesis of the scaled analog computer circuit can be found in Appendix 2. Optimal values for the two parameters, i.e., controller gain K_C and integral time T_R, of the controller were determined by using the integral-squared-error criterion ($I = \int_{0}^{\infty} e^2 dt$, where e is the error function of the controlled variables). The performance criterion function was also simulated on the analog computer. Performance of the control systems with different locations of control point were than evaluated with the controller operating under optimal settings. The transient response of both overhead product composition, x_D, and intermediate plate composition, x_j, (j = 1, 3, 7, 9) to a 10 % step change in feed composition x_F were recorded.

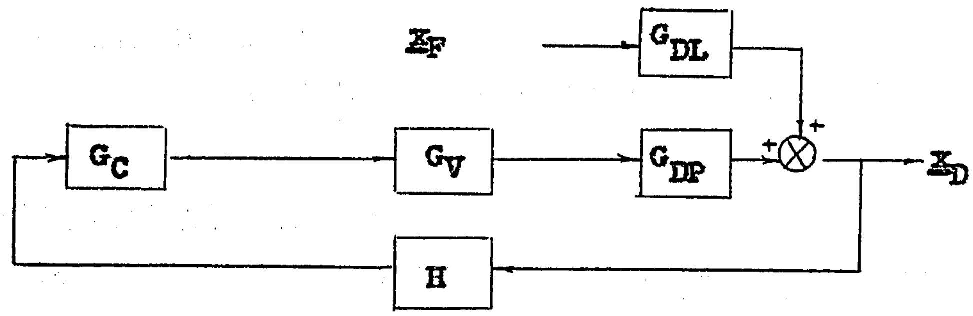


Fig. 7 Block diagram of one-point control scheme
(controlled variable : x_D)

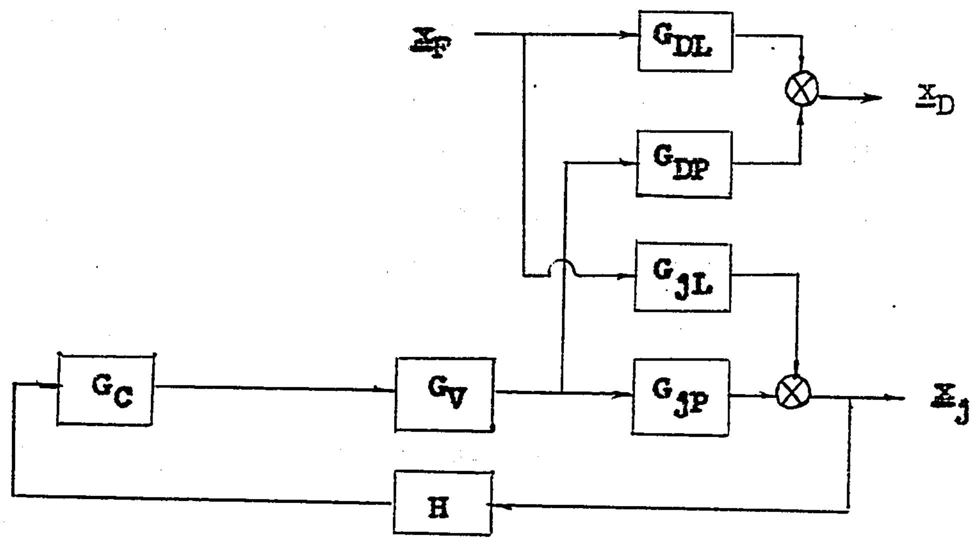


Fig. 8 Block diagram of one-point control scheme
(controlled variable : x_j, where $j = 1, 3, 7, 9$)

V-2. TWO-POINT CONTROL

CONTROLLED VARIABLES :

Compositions of the third and seventh plate, i.e., x_3 and x_7.

MANIPULATED VARIABLES :

Boilup rate V (to control x_3 at a constant value) and reflux rate R (to control x_7 at a constant value).

DISTURBANCE :

A 10 % step change in feed composition, x_F.

BLOCK DIAGRAM :

The block diagram of the two-point control scheme is shown in Figure 9. In this study, the dynamic lag due to heat transfer in the reboiler unit was assumed to be negligible, and consequently is simply represented by a gain. There is no lose in generality of the problem considered since the dynamics of this heat transfer unit could be included with no difficulty, however, it will only change the range of stability of the bottom loop.

ANALOG COMPUTER SIMULATION :

As for the one-point control scheme proportional-integral controllers were used for both loops. The optimal settings of the two controllers were determined independently, that is with only one of the two loops in action, using the integral-squared-error criterion. The performance of two-point control was then evaluated by operating the top loop (reflux manipulation) at the predetermined optimal settings and then bringing the other loop into operation by gradual application of the predetermined values of the controller settings. Transient response of both controlled variables x_7 and x_3 were recorded.

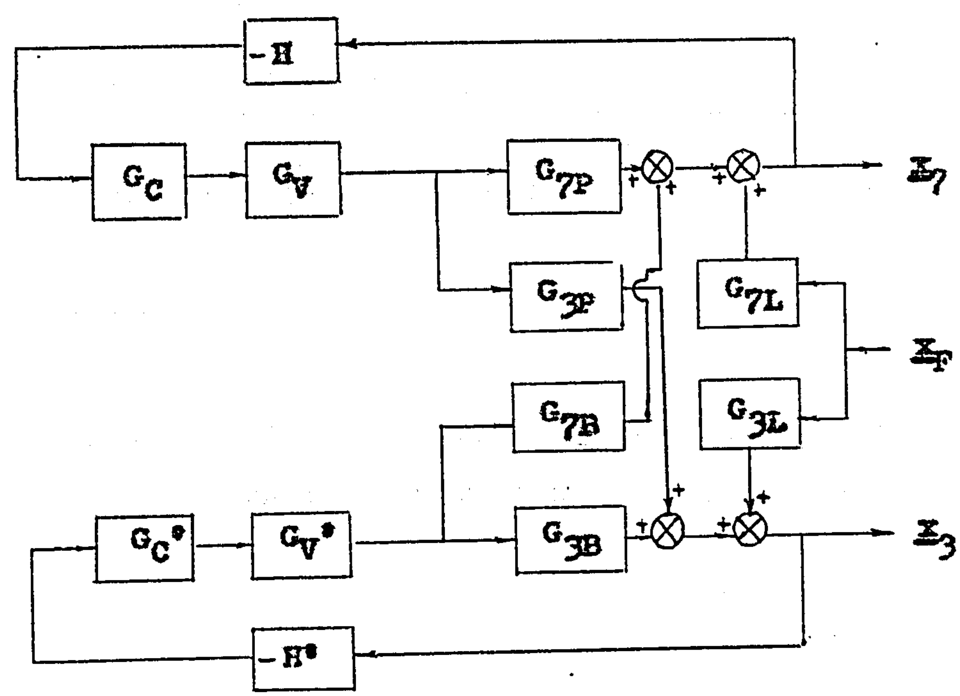


Fig. 9 Block diagram of two-point control scheme

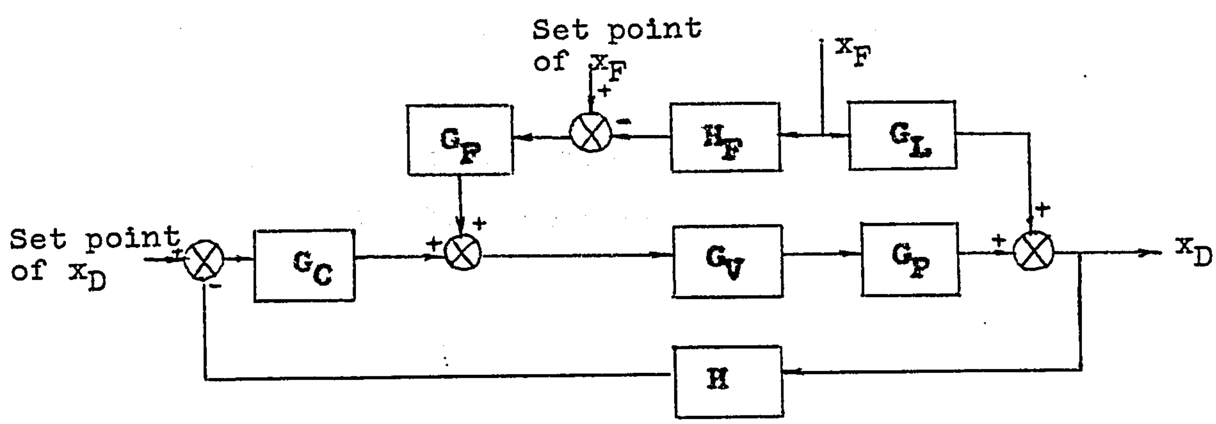


Fig. 10 Block diagram of feedforward plus feedback compensation control scheme

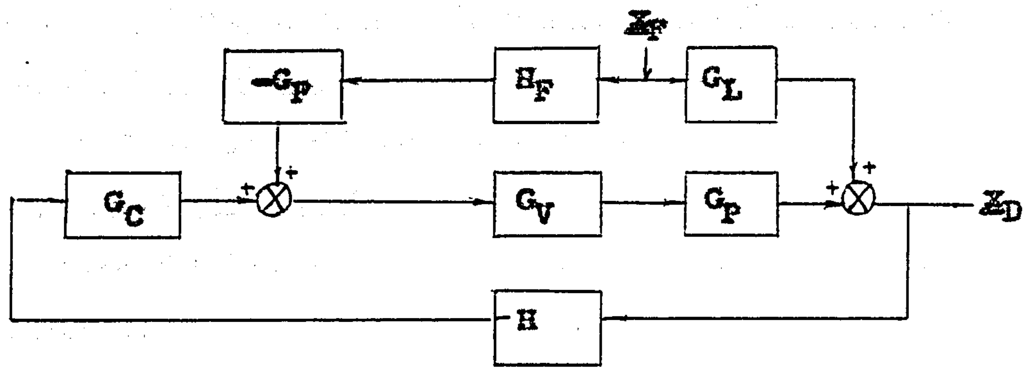


Fig. 11 Simplified block diagram of feedforward plus feedback compensation control scheme

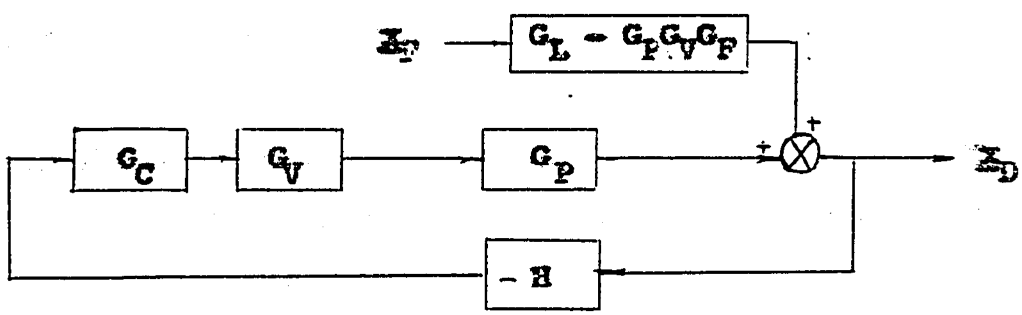


Fig. 12 Equivalent block diagram of feedforward plus feedback compensation control scheme

V-3. FEEDFORWARD PLUS FEEDBACK COMPENSATION CONTROL

CONTROLLED VARIABLE :

overhead product composition, x_D.

MANIPULATED VARIABLE :

Reflux rate, R.

DISTURBANCE :

A 10 % step change in feed composition, x_F.

BLOCK DIAGRAM :

The block diagram of feedforward plus feedback compensation control is shown in Figure 10. A proportional-integral controller was used in the feedback loop. Since the steady state value of feed composition, $(x_F)ss$, is usually a predetermined constant, the block diagram can be simplified to Figure 11. When the measurement lag H_F is negligible Figure 11 can be simplified further to Figure 12. (The synthesis of Figure 12 may be found in Appendix 3). The advantage of representing the block diagram as Figure 12 becomes evident when the perfect feedforward functional G_F must be specified. The gain K_F of G_F is determined from the steady state calculation as

$$K_F = \frac{K_L}{K_P K_V} \qquad (5).$$

A "perfect" feedforward functional G_F was found as

$$G_F = K_F \frac{(T_{DP1}\, s + 1)\ (T_{DP2}\, s + 1)}{(T_{DL1}\, s + 1)\ (T_{DL2}\, s + 1)} \qquad (6)$$

The synthesis of the perfect feedforward functional G_F is outlined in Appendix 3, where a tabulation of specific values is available.

ANALOG COMPUTER SIMULATION :

Four control schemes, namely feedback control with the controller operating at the optimal settings which were determined previously, instantaneous corrective action feedforward control, dynamically "perfect" corrective rate feedforward control, and feedforward plus feedback compensation control, were simulated on the analog computer.

CHAPTER VI

RESULTS FROM THE SIMULATION

VI-1. RESULTS

1. ONE-POINT CONTROL :

The ranges of stability of the controller settings are shown graphically in Figures 13-17 and the recommended controller settings for different sensing element locations are tabulated in Table 9. Figures 18-22 show the integral-squared-error, $\int_0^\infty e^2 dt$, plotted against $1/T_R$ with K_C as the parameter.

When intermediate plate composition is controlled, the off-set of overhead product composition x_F may be calculated from

$$\frac{x_D}{x_F} = \frac{G_{DL} + (G_{jP}G_{DL} - G_{jL}G_{DP})G_C G_V H}{1 + G_{jP}G_C G_V H} \qquad (7)$$

and utilizing the final value theorem, for a step change $x_F(s) = L/s$

$$\text{Off-set of } x_D = \lim_{t \to \infty} x_D$$

$$= \lim_{s \to 0} s\,(L/s)\,\frac{G_{DL} + ((G_{jP}G_{DL} - G_{jL}G_{DP})G_C G_V H}{1 + G_{jP}G_C G_V H}$$

$$= L\,(K_{DL} - (K_{jL}/K_{jP})K_{DP}) \qquad (8)$$

Calculated values of the resulting off-set in x_D are shown in Table10.

TABLE 9

RECOMMENDED CONTROLLER SETTINGS

Control Point	$1/T_R$	K_C	Min. $\int_0^\infty e^2 dt$ (volts)
x_1	0.21	4	0.58
x_3	0.40	2	2.68
x_7	0.60	2	1.24
x_9	0.75	4	1.70
x_D	0.60	15	0.078

TABLE 10

OFF-SET OF OVERHEAD PRODUCT COMPOSITION

Control point	K_L	K_P	Off-set of x_D (mole % acetone)
x_D	0.580	0.3685	0.000
x_9	1.170	0.692	0.427
x_8	1.449	0.909	0.074
x_7	1.670	1.052	0.050
x_6	1.784	1.067	0.361
x_5	1.772	0.728	3.169
x_4	1.932	0.986	1.420
x_3	1.940	1.175	0.284
x_2	1.620	1.204	-0.807
x_1	1.125	0.965	-1.500

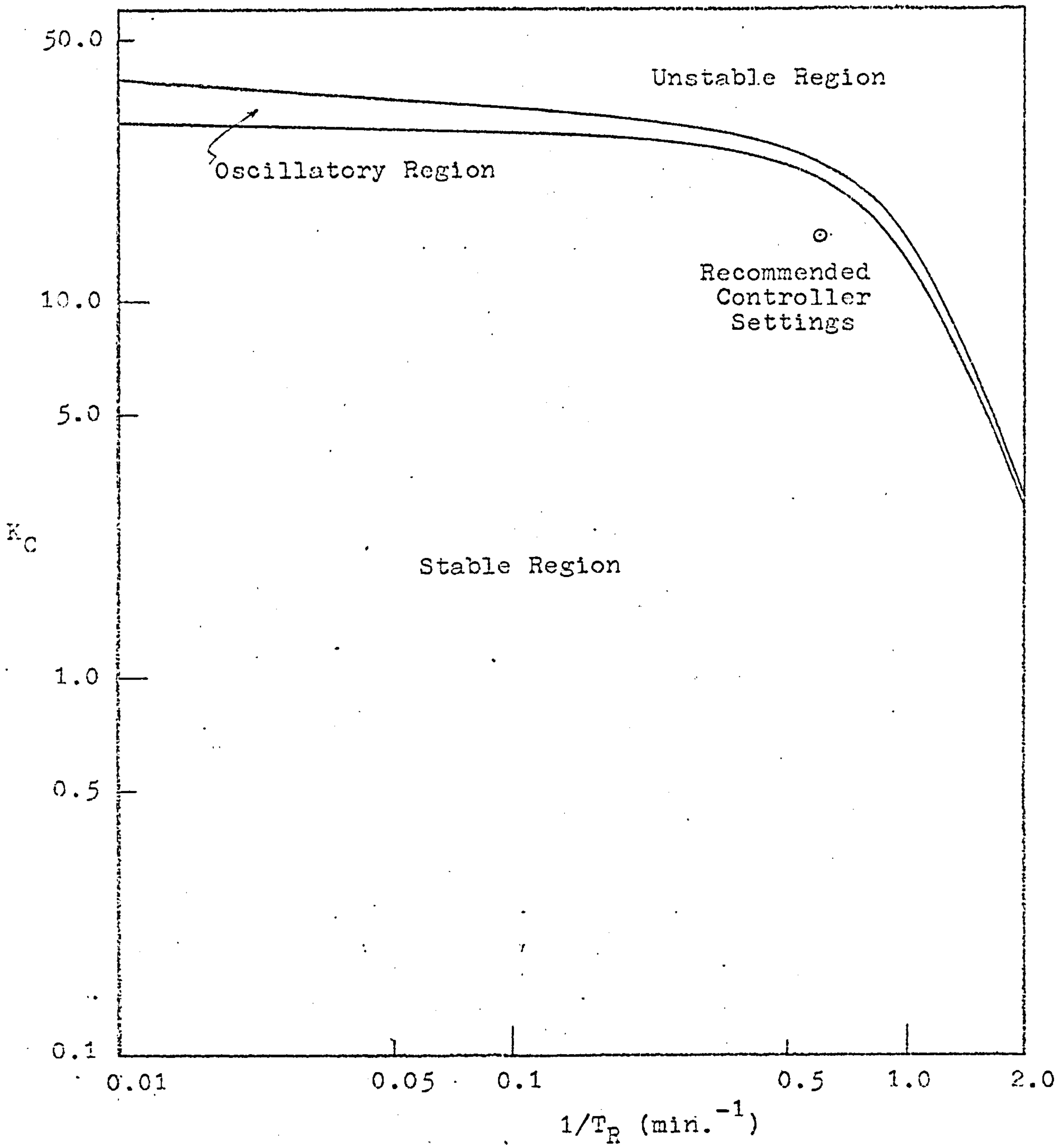


Fig. 13 The stability range of the controller when x_D is controlled

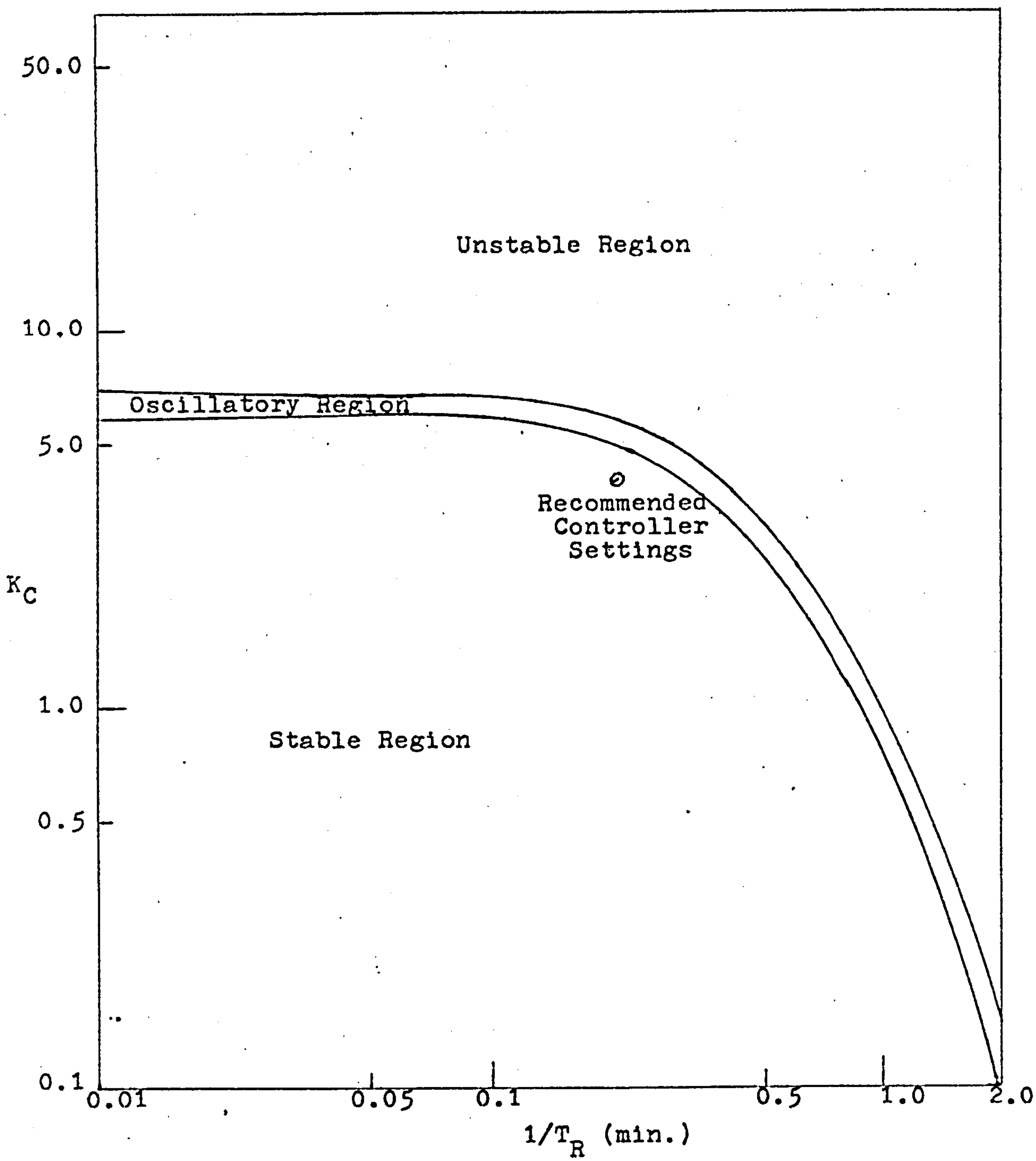


Fig. 14 The stability range of the controller when x_1 is controlled

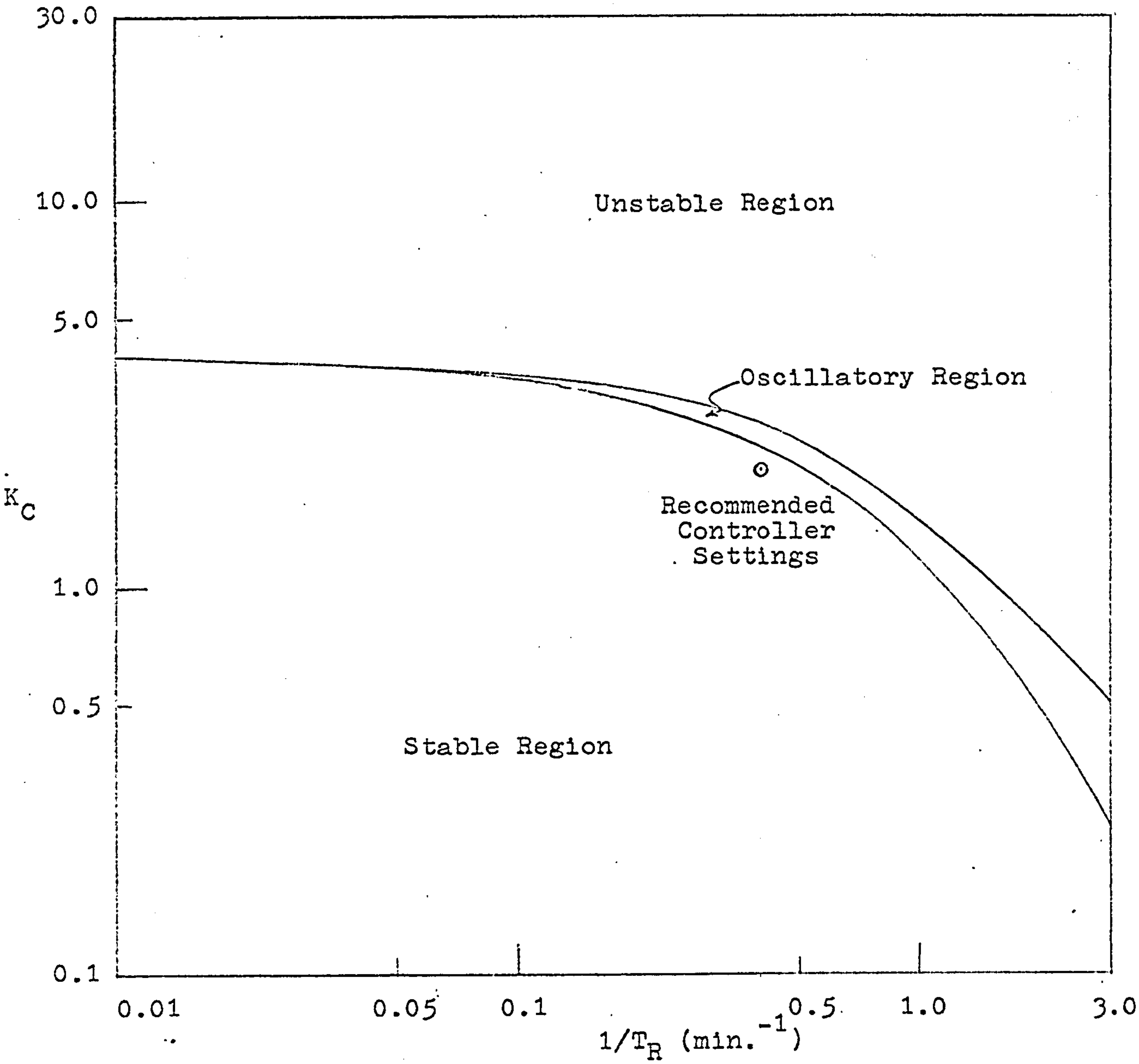


Fig. 15 The stability range of the controller when x_3 is controlled

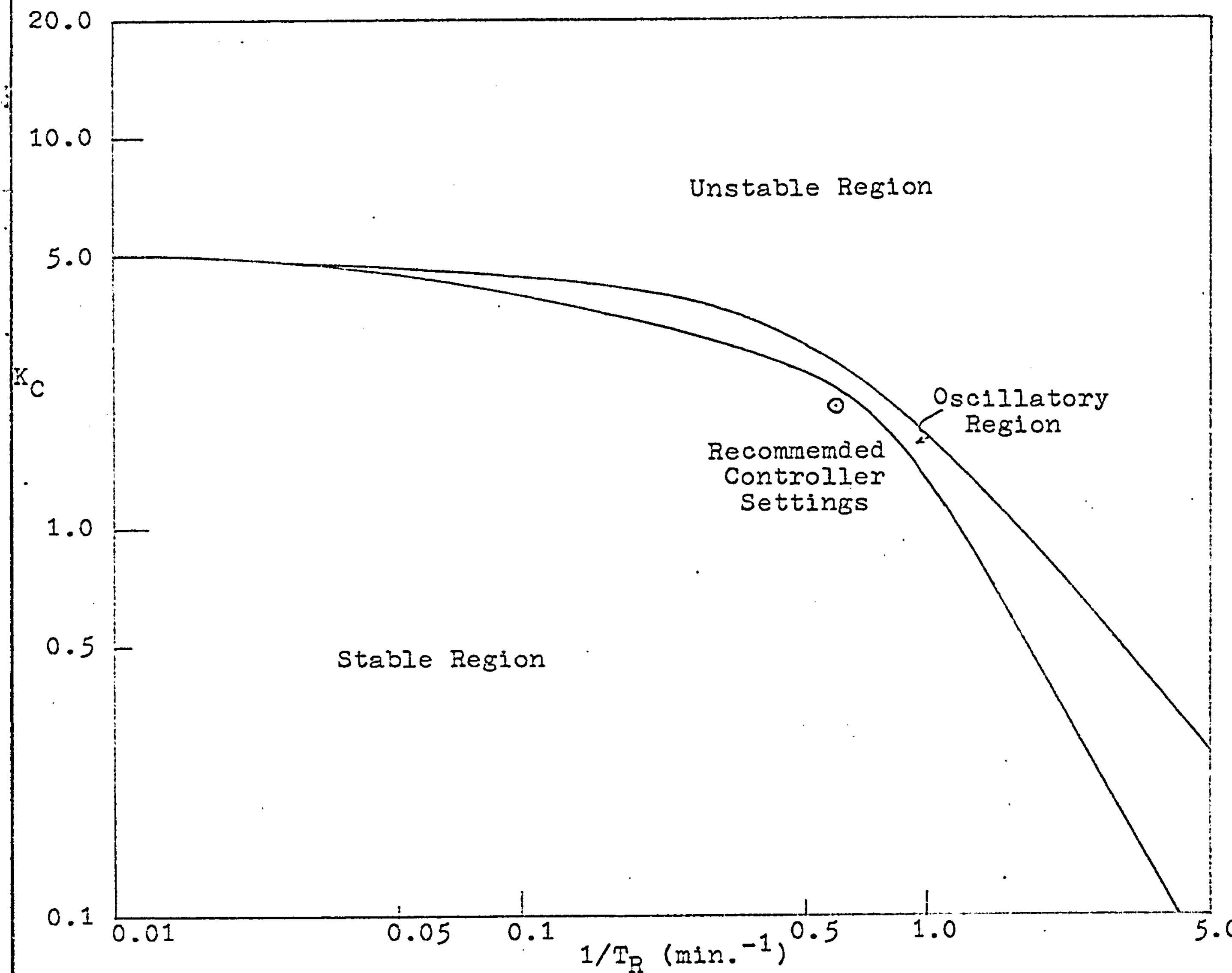


Fig. 16 The stability range of the controller when x_7 is controlled

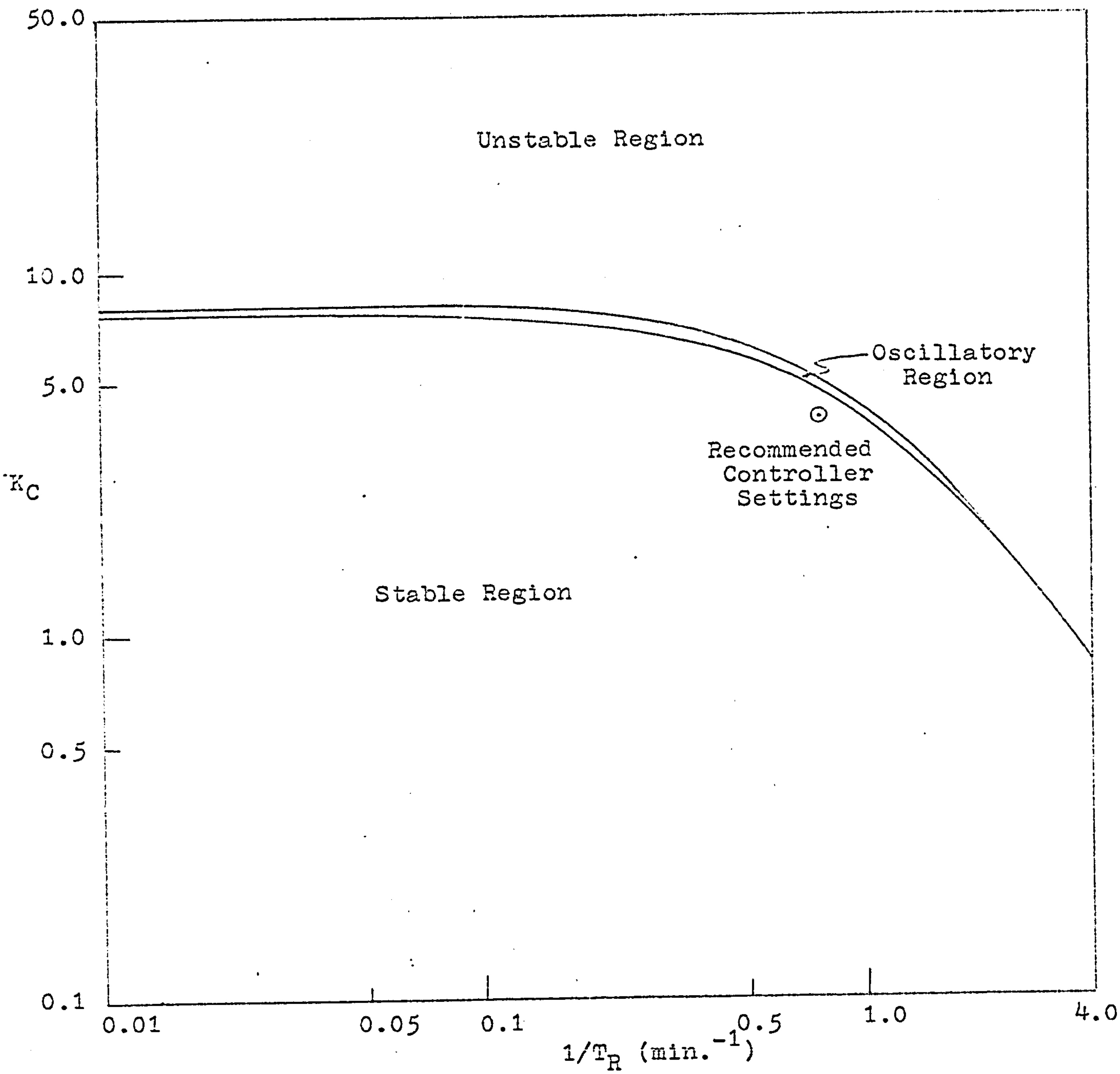


Fig. 17 The stability range of the controller when x_9 is controlled

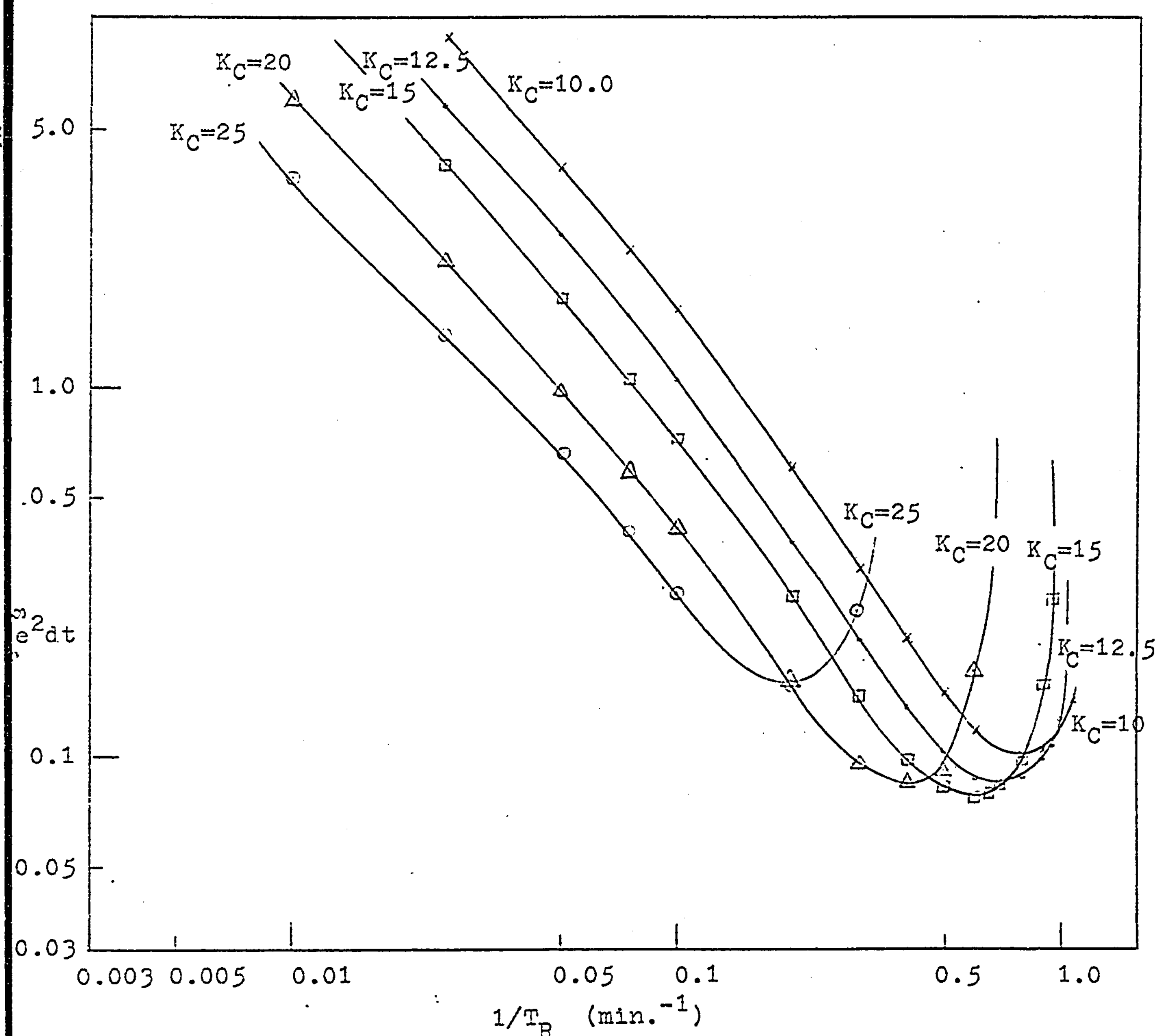


Fig. 18 $\int_0^\infty e^2dt$ vs. $1/T_R$ (when x_D is controlled)

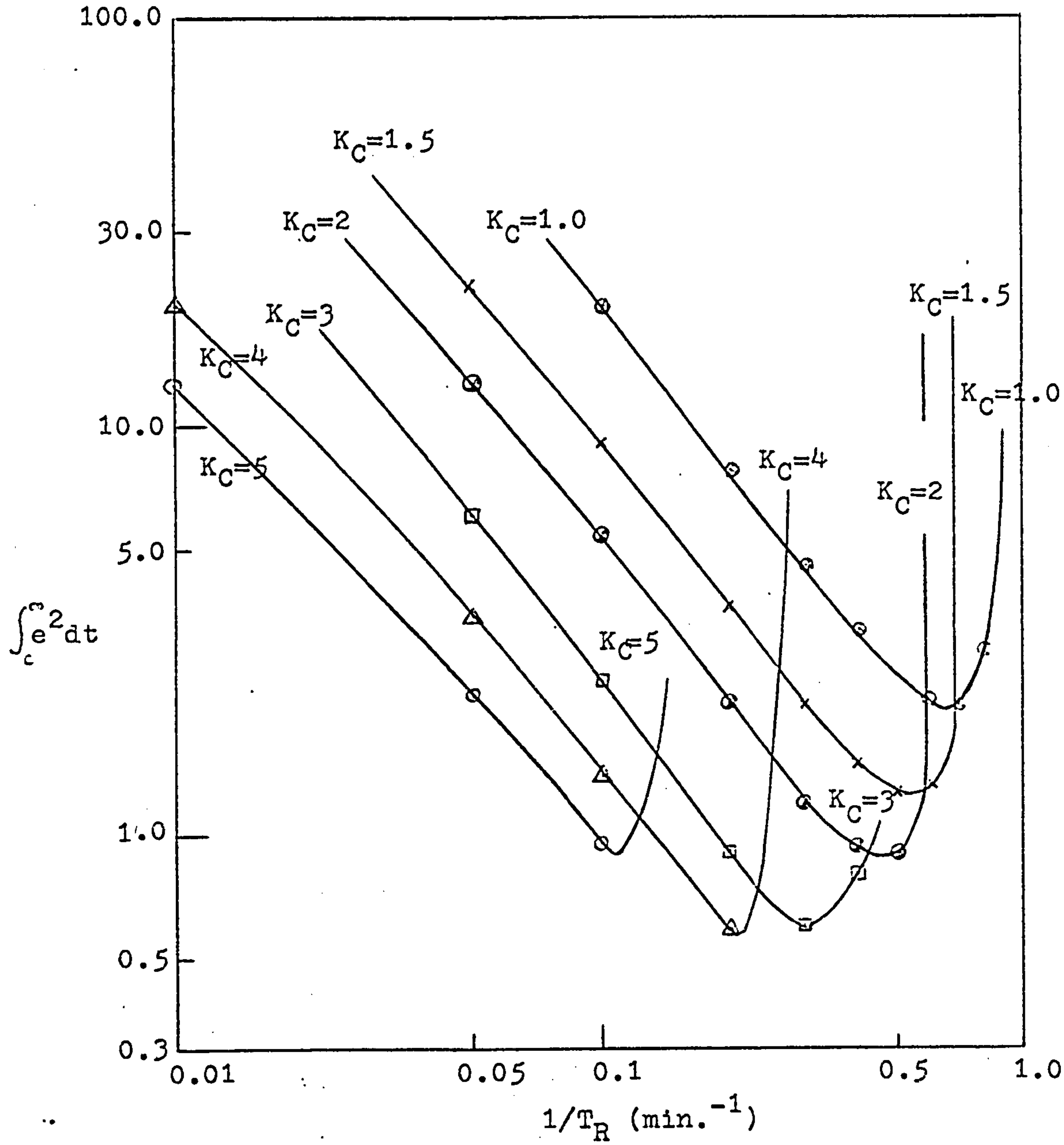


Fig. 19 $\int_0^\infty e^2 dt$ vs. $1/T_R$ (when x_1 is controlled)

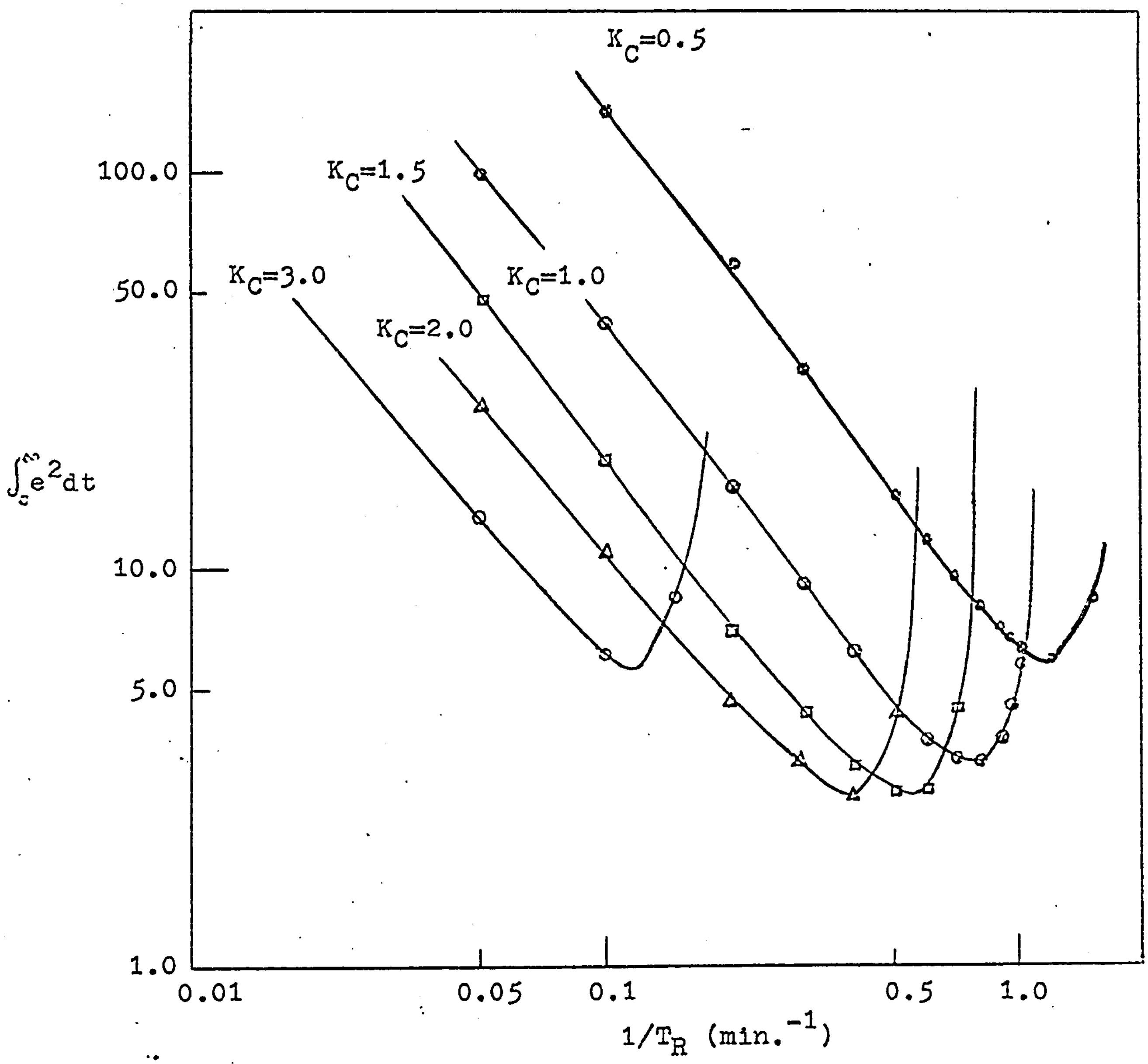


Fig. 20 $\int_0^\infty e^2 dt$ vs. $1/T_R$ (when x_3 is controlled)

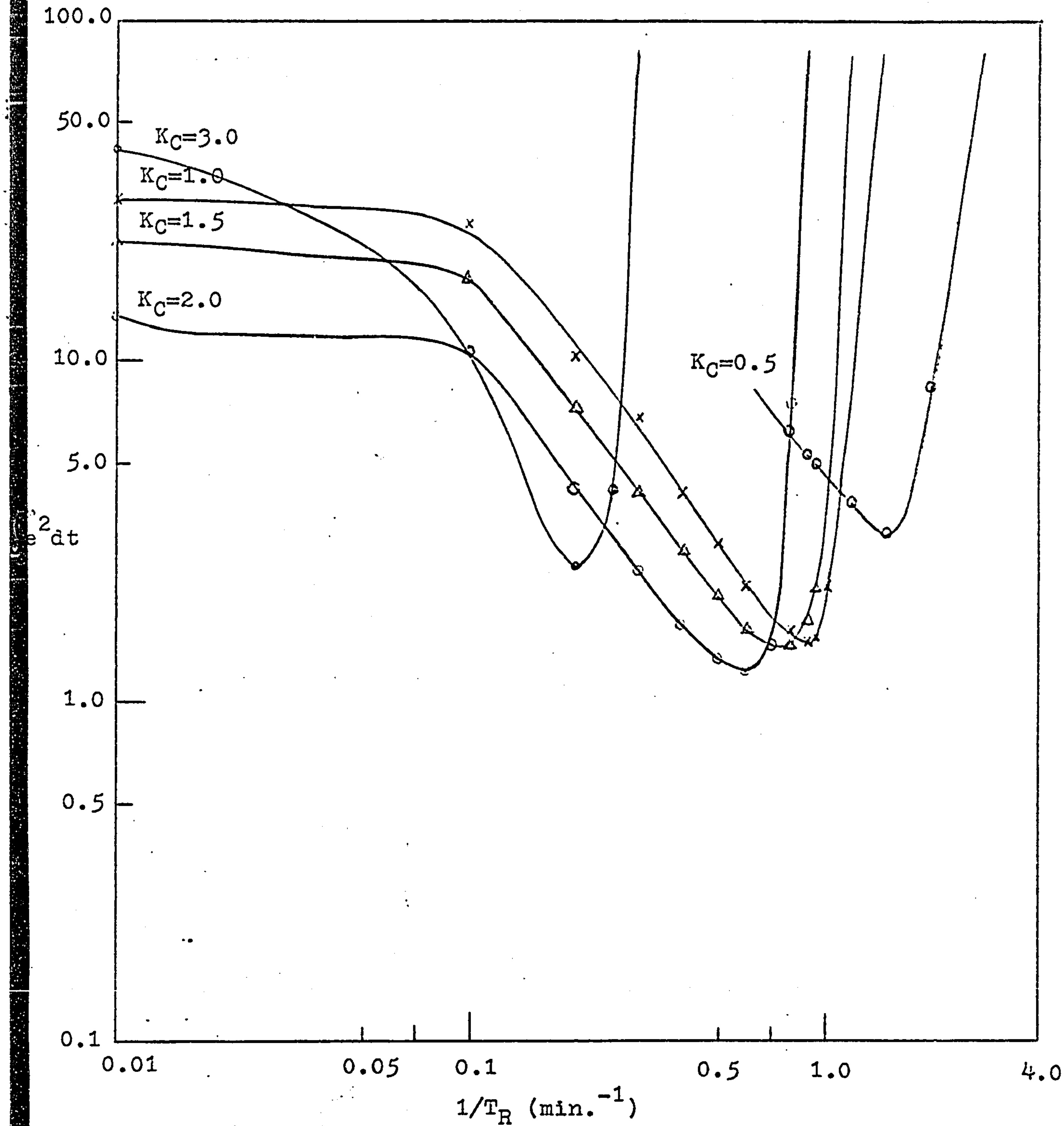


Fig. 21 e^2dt vs. $1/T_R$ (when x_7 is controlled)

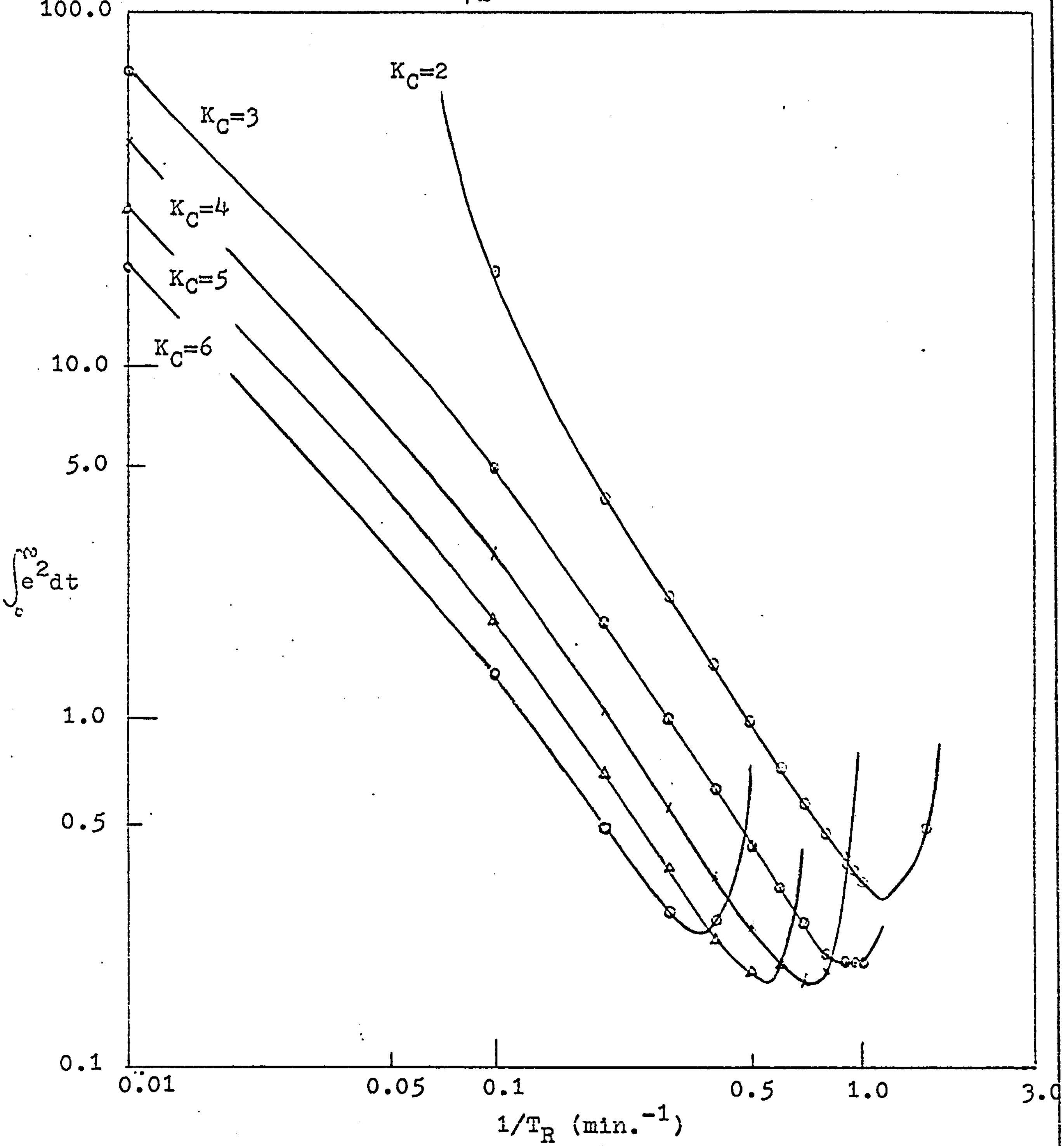


Fig. 22 e^2dt vs. $1/T_R$ (when x_9 is controlled)

2. <u>TWO-POINT CONTROL</u> :

The optimal settings for the controller used in the upper control loop (reflux manipulation) are $K_C = 2.0$, $1/T_R = 0.6$ min.$^{-1}$ and for the controller used in the lower loop (boilup manipulation) are $K_C = 2.1$, $1/T_R = 0.1$ min.$^{-1}$.

Responses of the two controlled variables x_3 and x_7 to a 10 % step change in feed composition x_F for different controller parameters are illustrated in Figures 23-35. The conditions that were investigated are tabulated in Table 11.

3. <u>FEEDFORWARD PLUS FEEDBACK COMPENSATION CONTROL</u> :

Response of the controlled variable, x_D, for feedforward control, feedback control and feedforward plus feedback compensation control is shown in Figures 36-40.

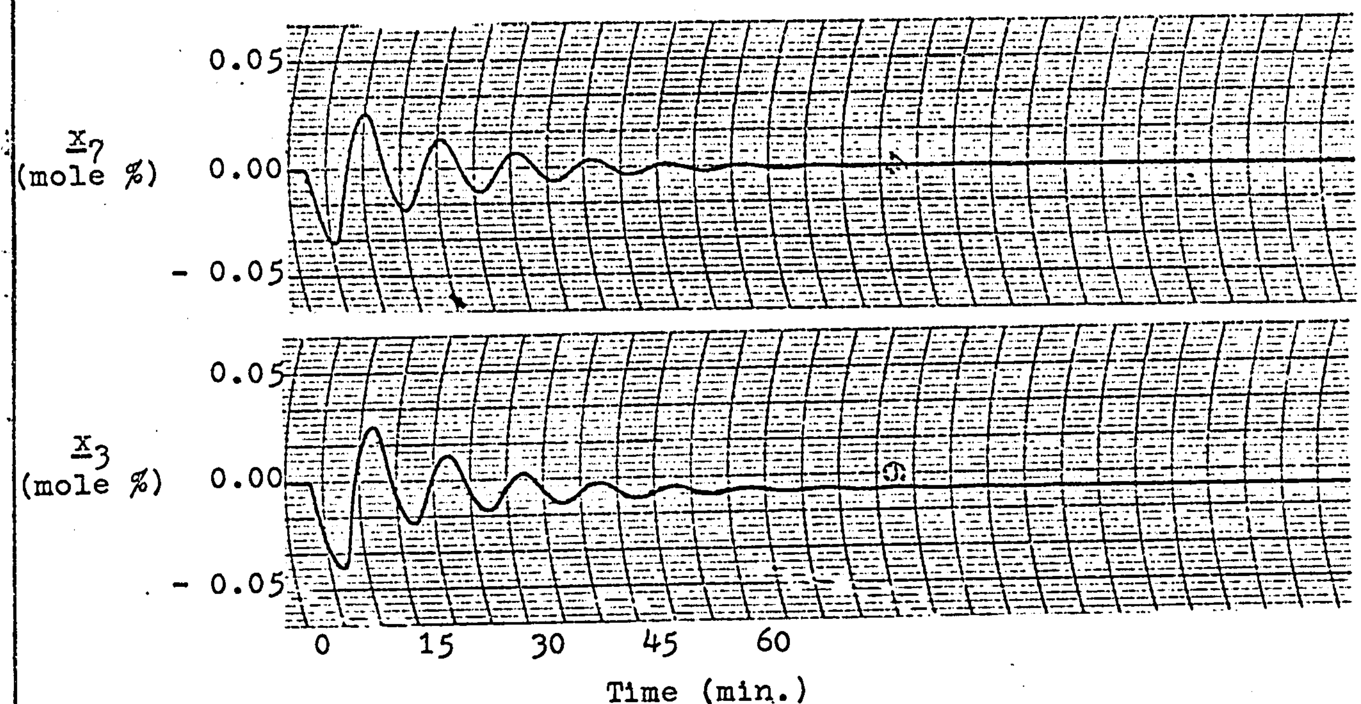
TABLE 11

CONDITIONS OF TWO-POINT CONTROL INVESTIGATED

No. of Figure	Conditions
23	Only the upper loop is in action (optimal settings : $K_C = 2.0$, $1/T_R = 0.6$)
24	Only the lower loop is in action (optimal settings : $K_C' = 2.1$, $1/T_R' = 0.1$)
25	Both loops are in action (optimal settings)
26	The upper loop is in action (optimal settings) and a proportional controller ($K_C' = 0.75$) is used in the lower loop
27	Same as above, except $K_C' = 1.80$
28	Same as above, except $K_C' = 2.10$
29	Same as above, except $K_C' = 2.40$
30	Same as above, except $K_C' = 2.70$

(Cont. of Table 11)

31	The upper loop is in action (optimal settings) and a proportional-integral controller ($K_C' = 0.75$, $1/T_R' = 0.3$) is used in the lower loop
32	Same as above, except $1/T_R' = 0.4$
33	Same as above, except $1/T_R' = 0.5$
34	Same as above, except $K_C' = 0.325$, $1/T_R' = 0.6$
35	Same as above, except $K_C' = 0.325$ $1/T_R' = 0.8$


0.05
$\underline{x}_7$
(mole %)
0.00
- 0.05
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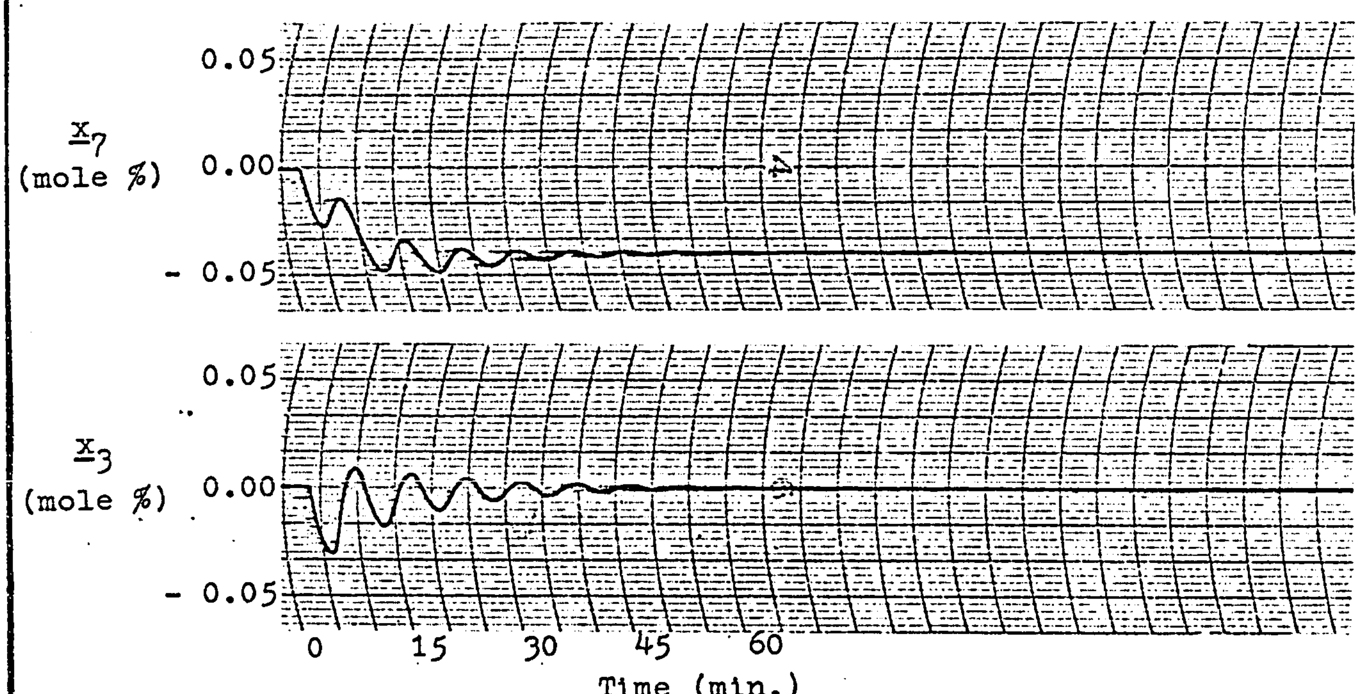
$\underline{x}_3$
(mole %)
0.00
- 0.05
0
15
30
45
60
Time (min.)

Fig. 23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$\underline{x}_7$
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(mole %)
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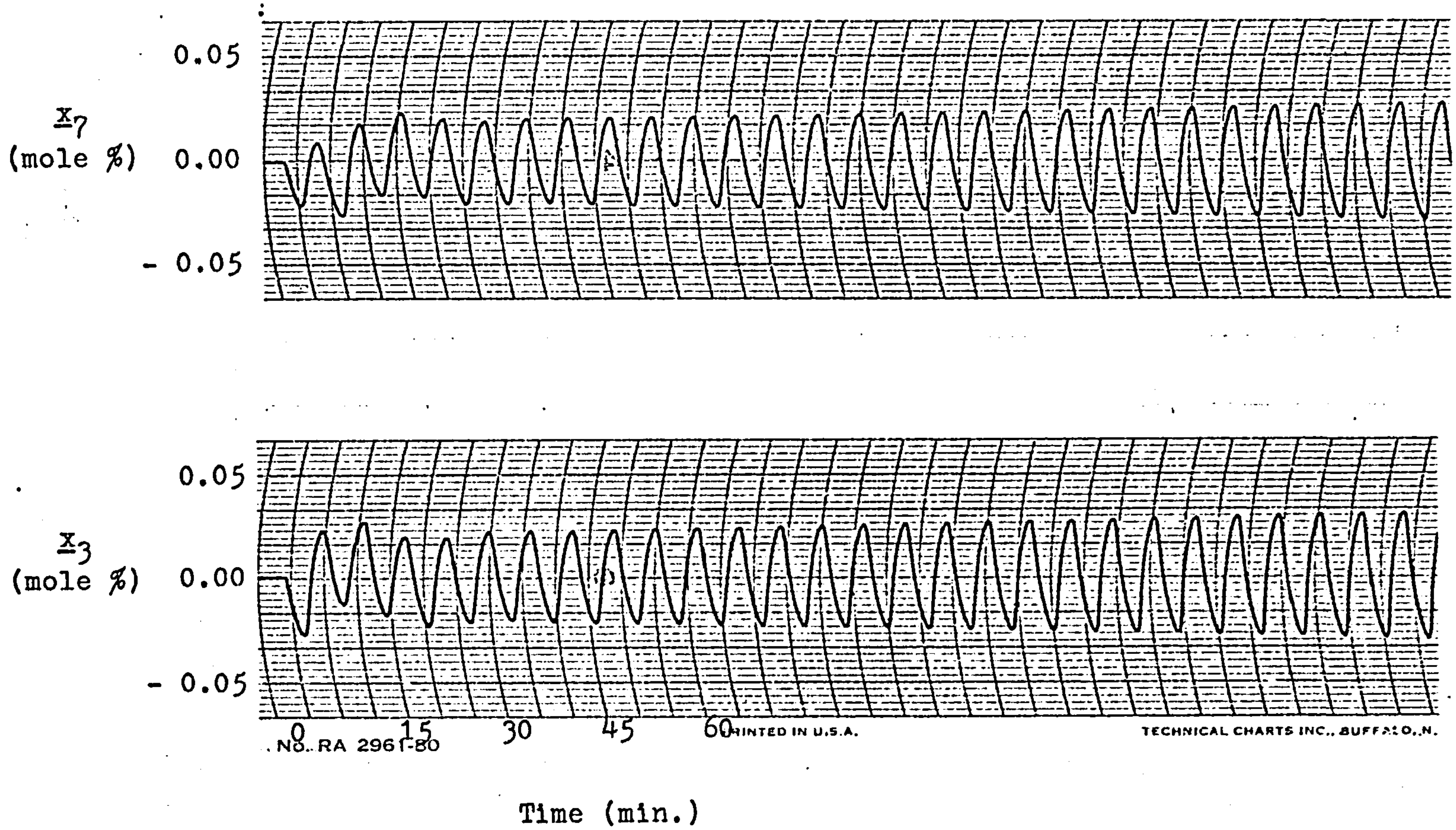
Time (min.)

Fig. 24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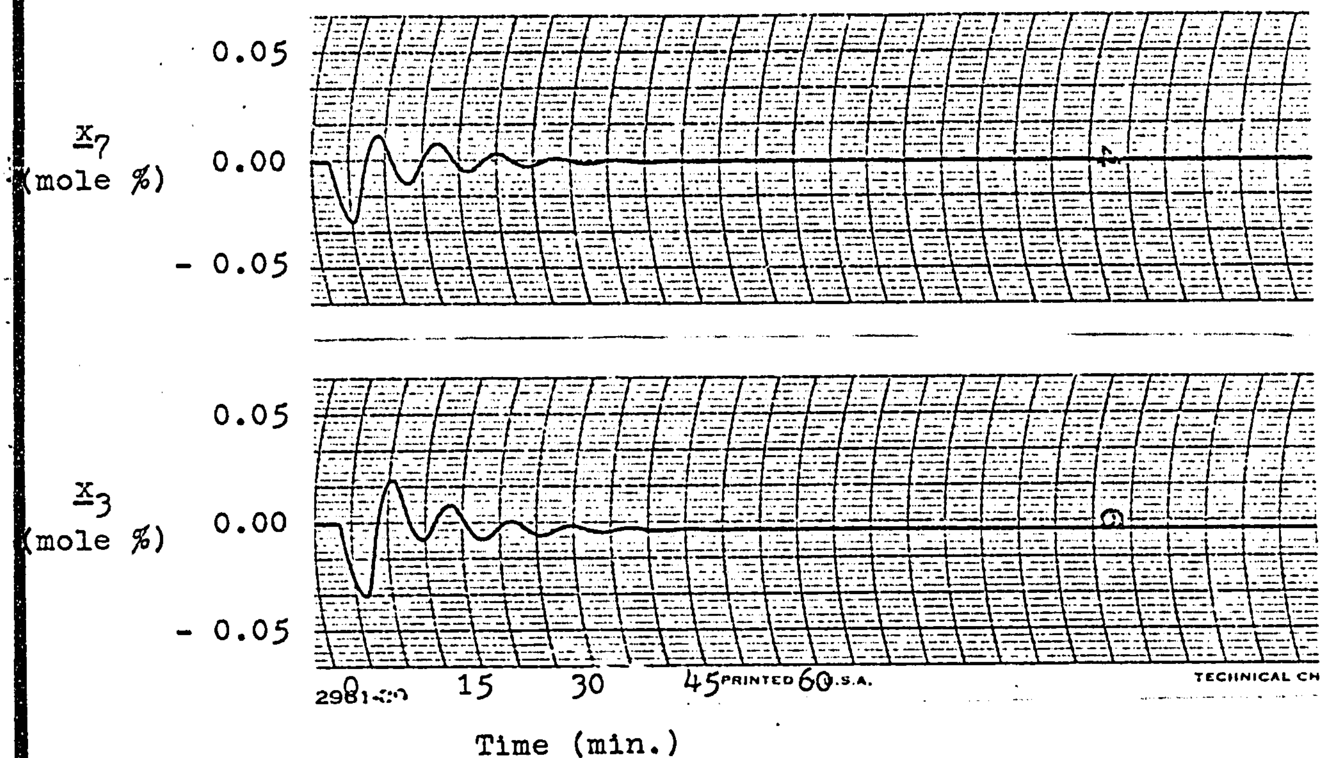
TECHNICAL CHARTS INC., BUFFALO, N.
Time (min.)

Fig. 25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Time (min.)

Fig. 26

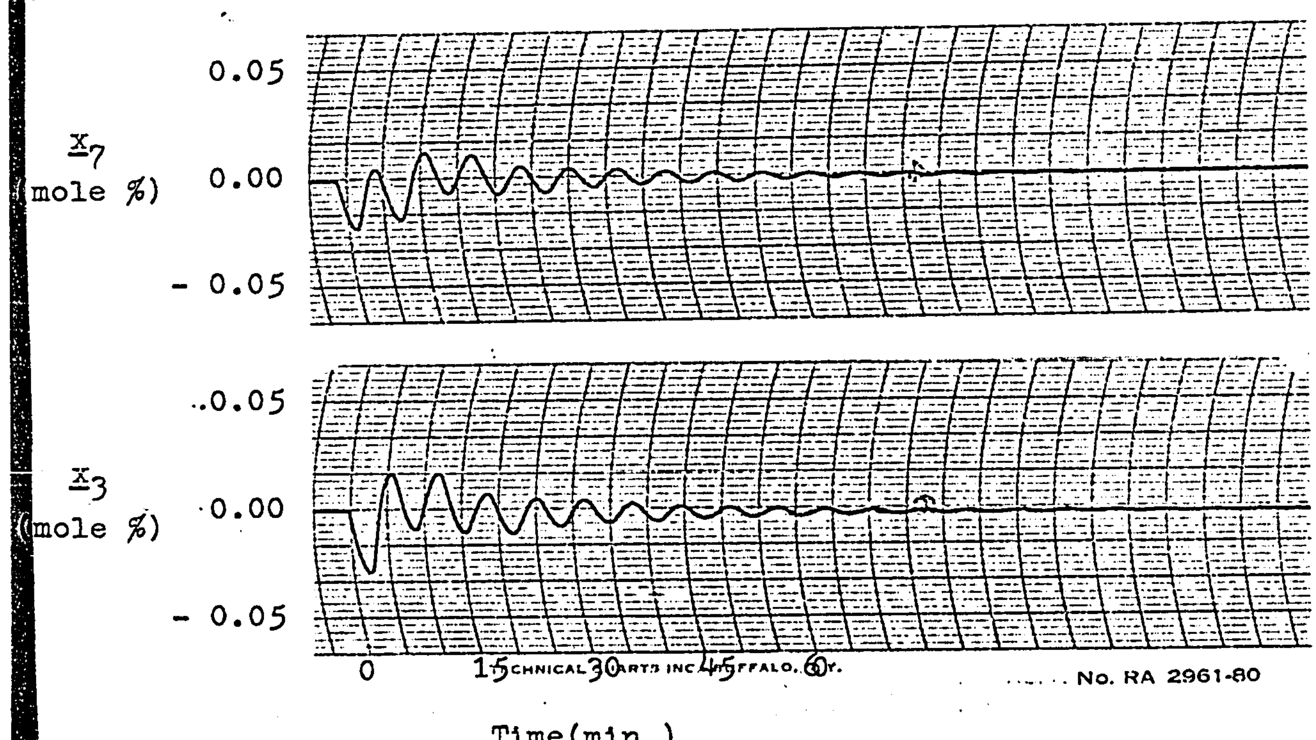
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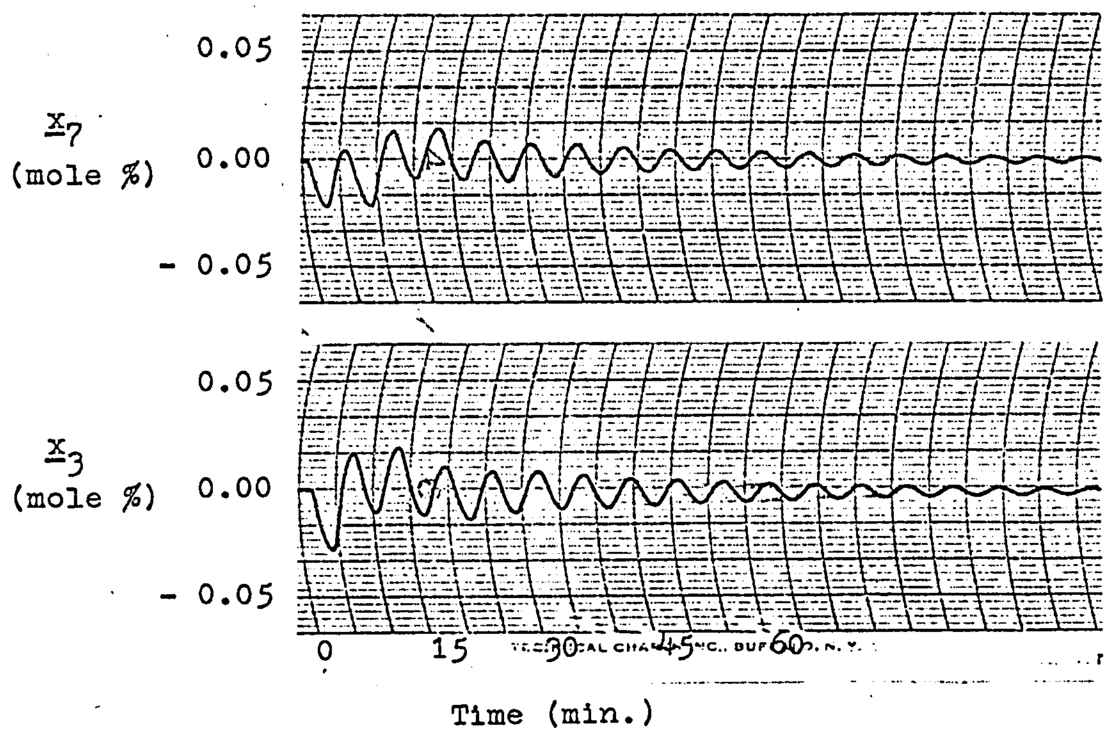
Fig. 27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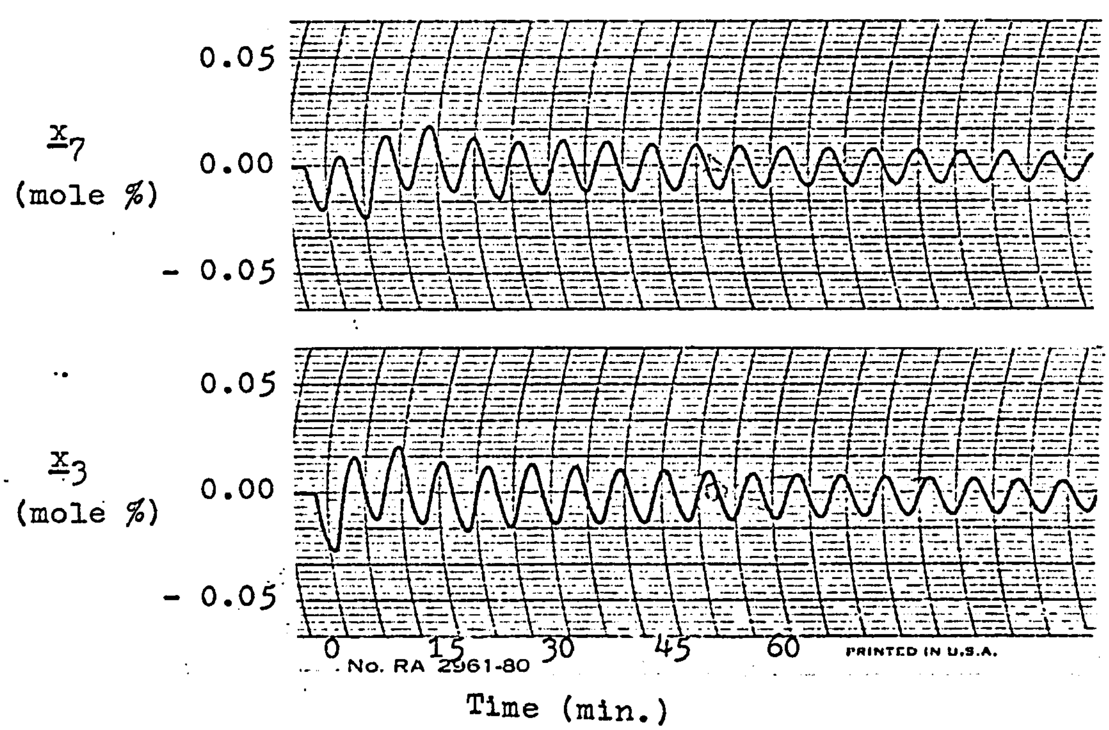
Time (min.)

Fig. 28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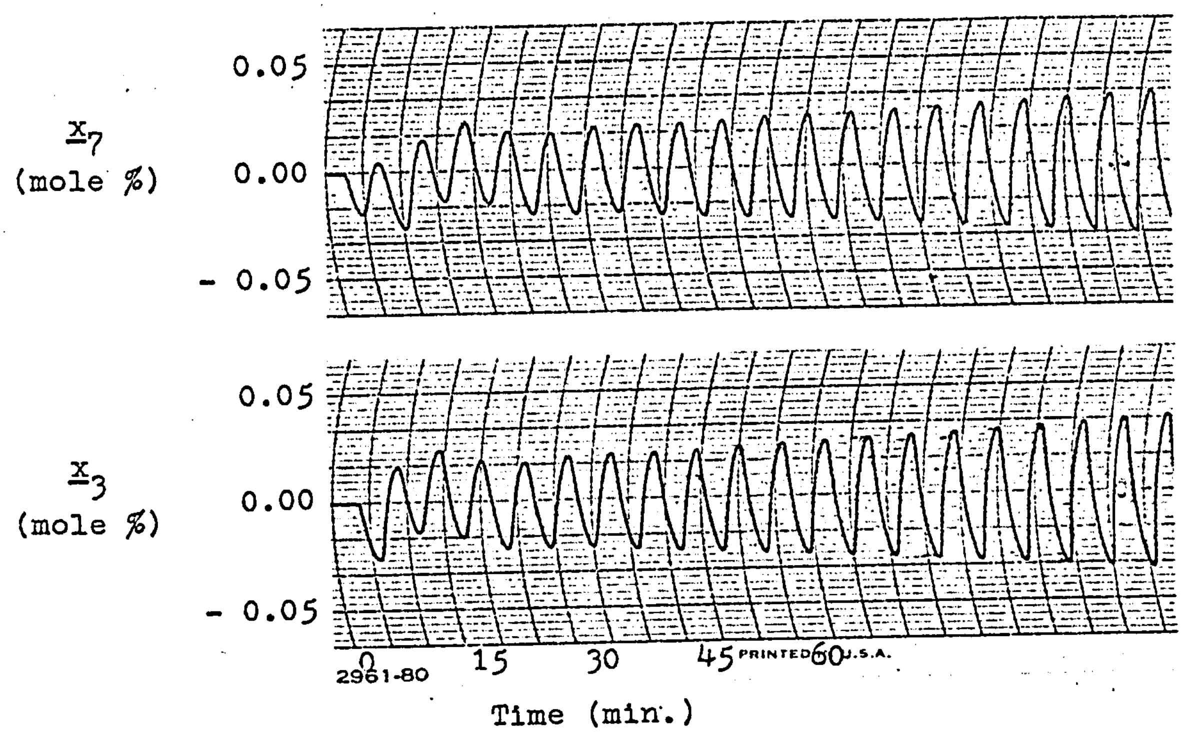
Time (min.)

Fig. 29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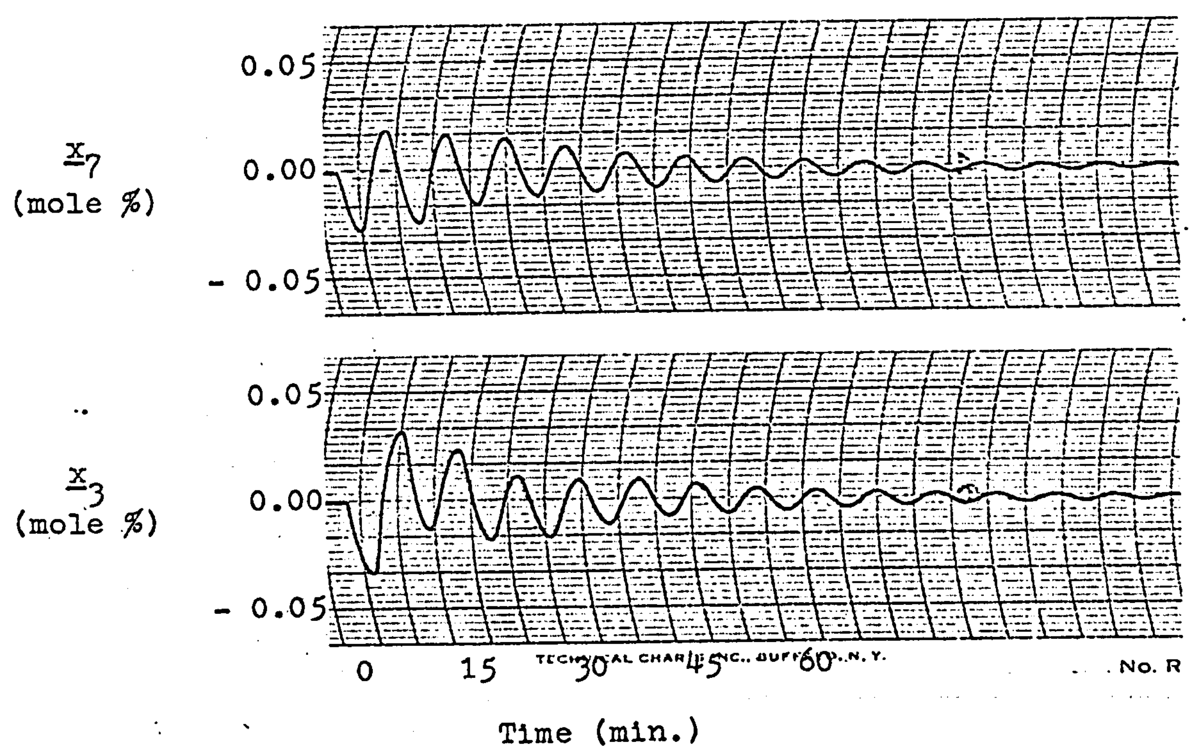
Time (min.)

Fig. 30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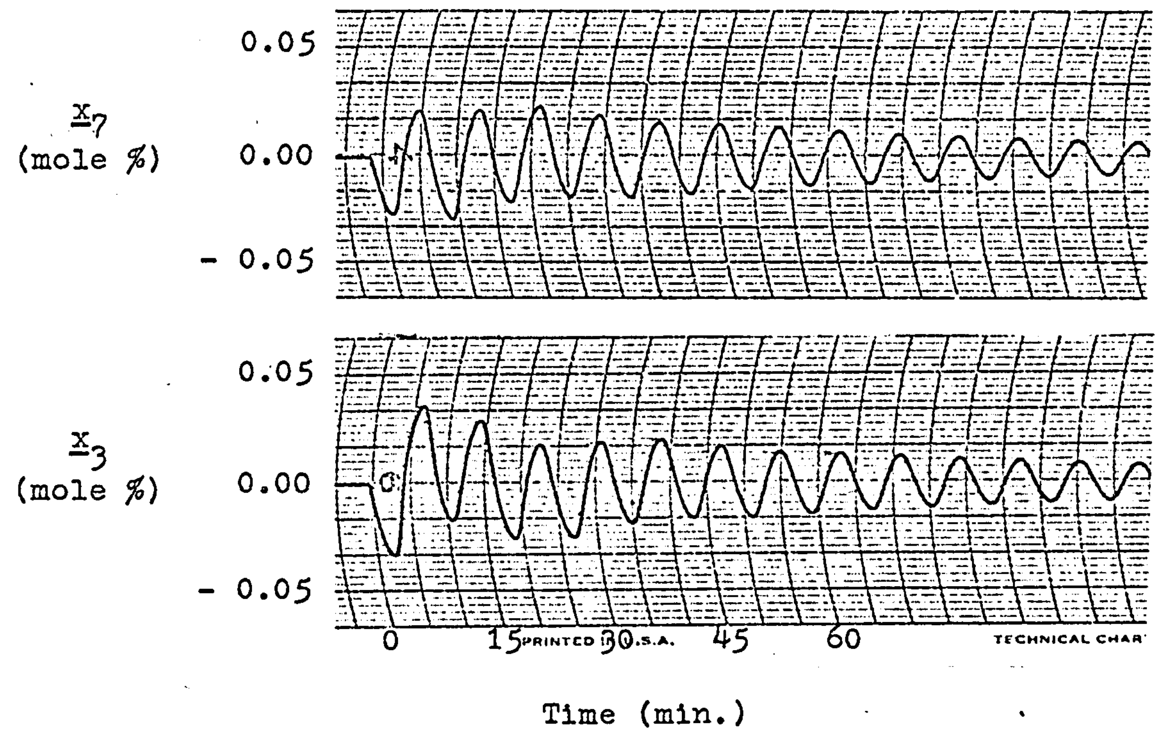
No. R
Time (min.)

Fig. 31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$\underline{x}_7$
(mole %)
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0.05
$\underline{x}_3$
(mole %)
0.00
- 0.05
0
15
30
45
60
Time (min.)

Fig. 32

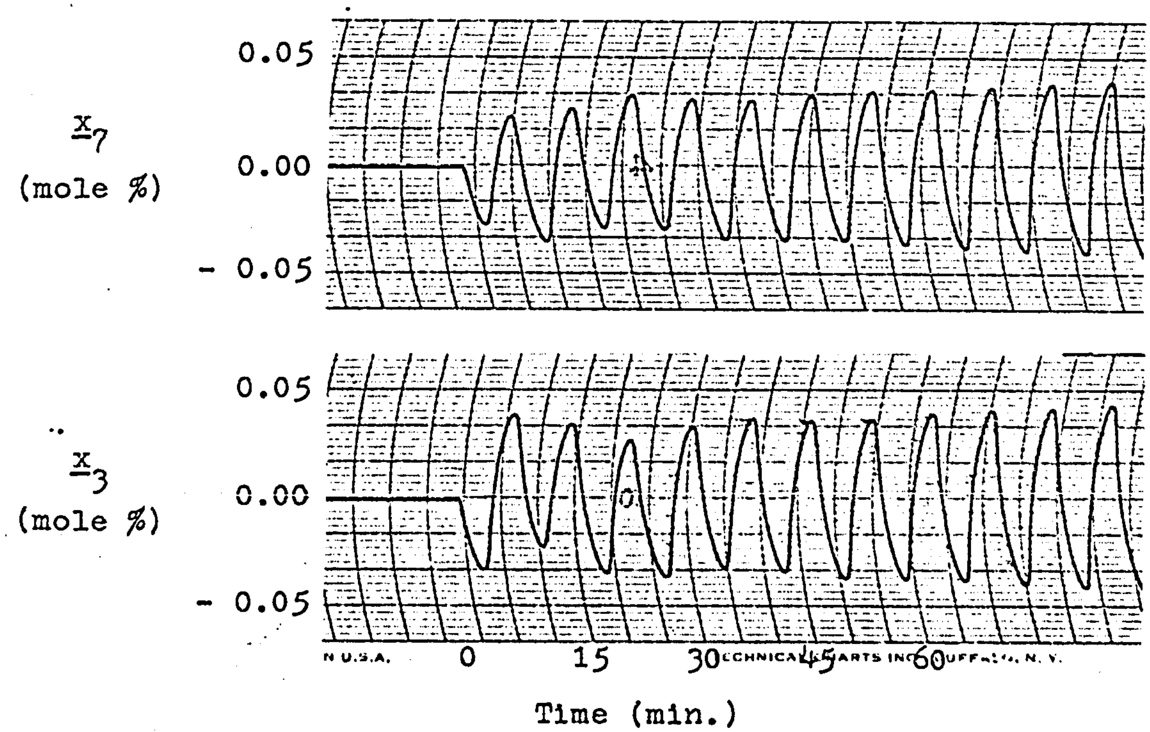

0.05
$\underline{x}_7$
(mole %)
0.00
- 0.05
0.05
$\underline{x}_3$
(mole %)
0.00
- 0.05
0
15
30
45
60
Time (min.)

Fig. 33

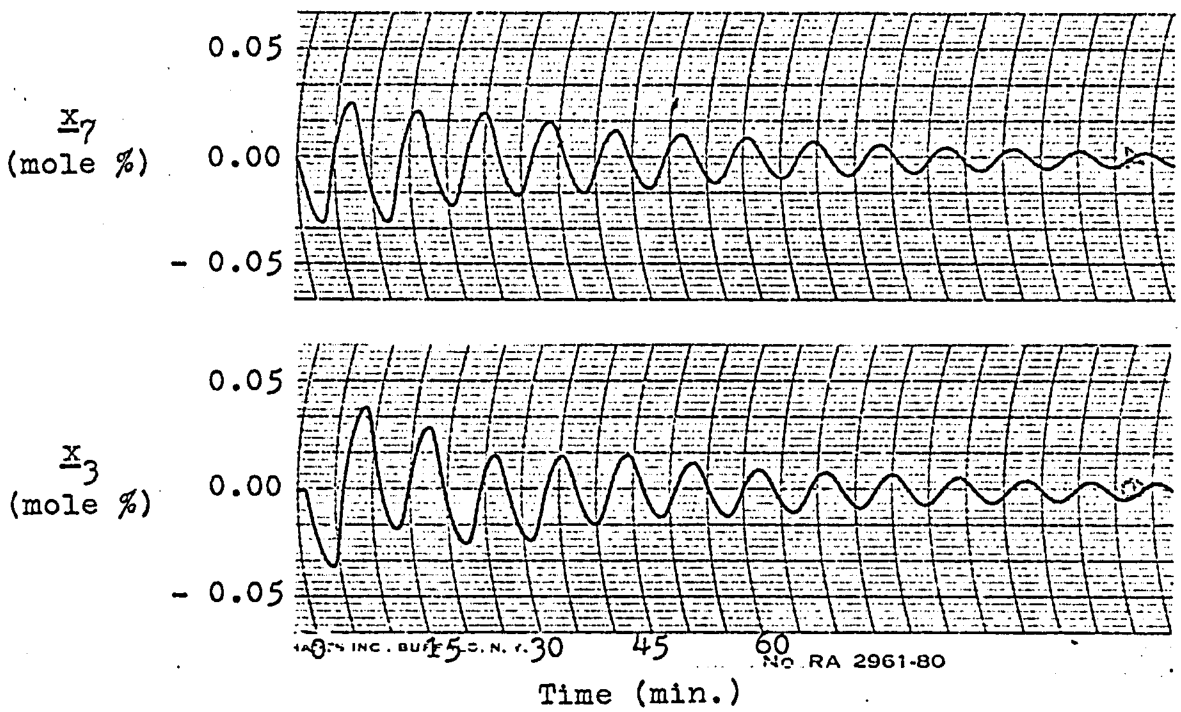

0.05
$\underline{x}_7$
(mole %)
0.00
- 0.05
0.05
$\underline{x}_3$
(mole %)
0.00
- 0.05
0
15
30
45
60
No. RA 2961-80
Time (min.)

Fig. 34

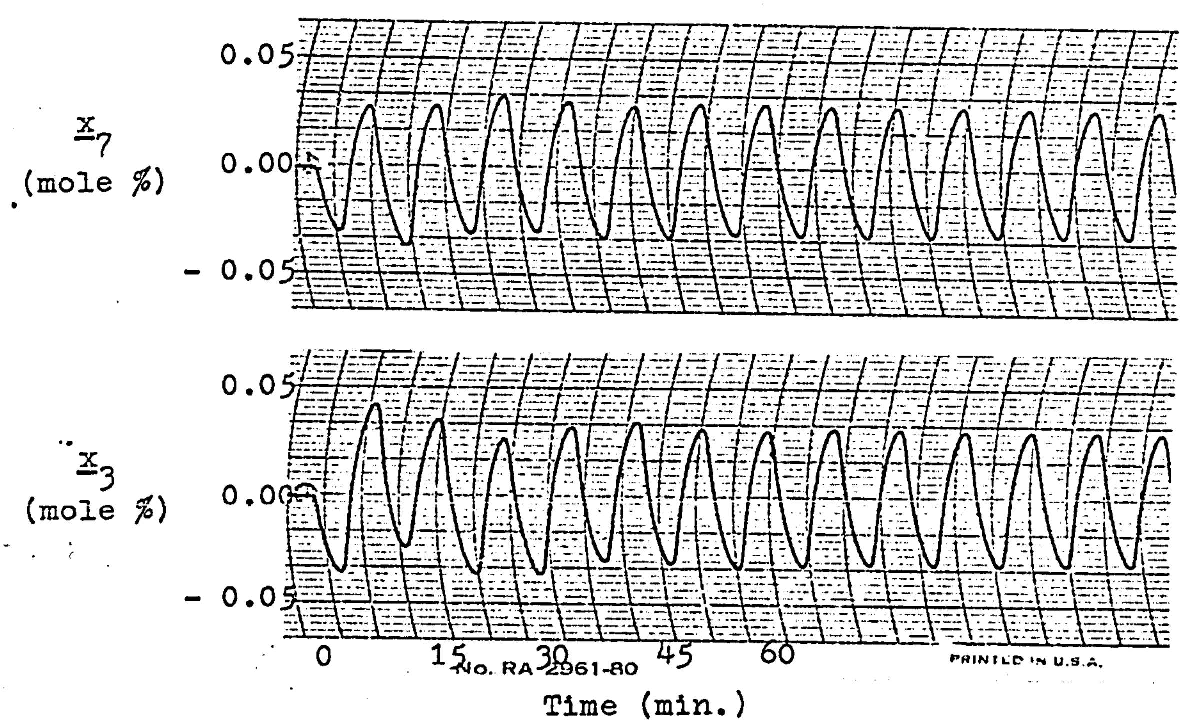

0.05
$\underline{x}_7$
(mole %)
0.00
- 0.05
0.05
$\underline{x}_3$
(mole %)
0.00
- 0.05
0
15
30
45
60
No. RA 2961-80
PRINTED IN U.S.A.
Time (min.)

Fig. 35

TABLE 12

COMPARISON OF CONTROL SCHEMES

Control Schemes	FBC	FFC	FFC +FBC
Maximum deviation (mole % acetone)	-1.09	-0.30	-0.13
Time required to return to $(x_D)ss$ (min.)	50	60	25
Period of oscillation (min.)	6.0	-	6.0
Decay ratio	0.385	-	0.5

Note : FBC = Feedback control

FFC = Feedforward control

FFC + FBC = Feedforward plus feedback compensation control

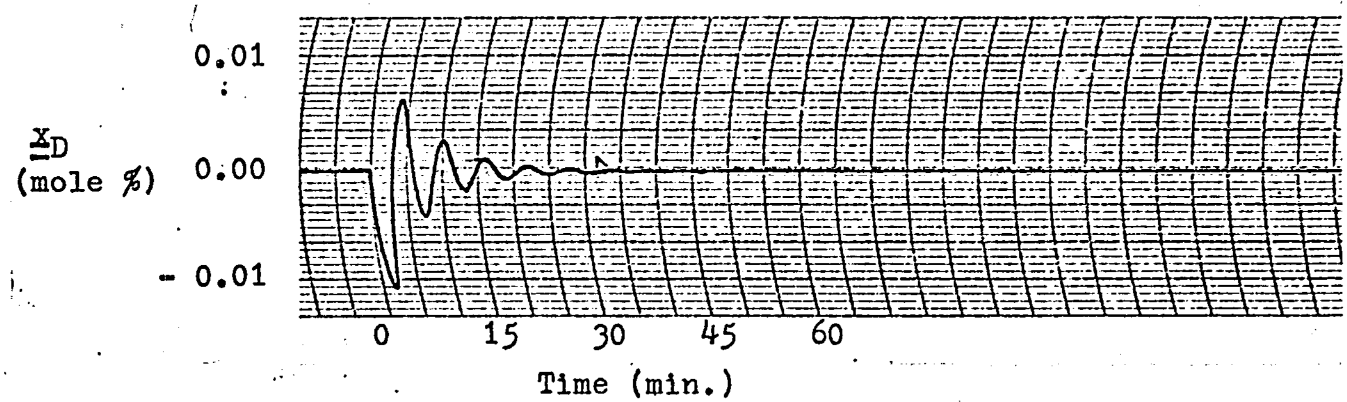


Fig. 36 Response of x_D to a step change in x_F using feedback control

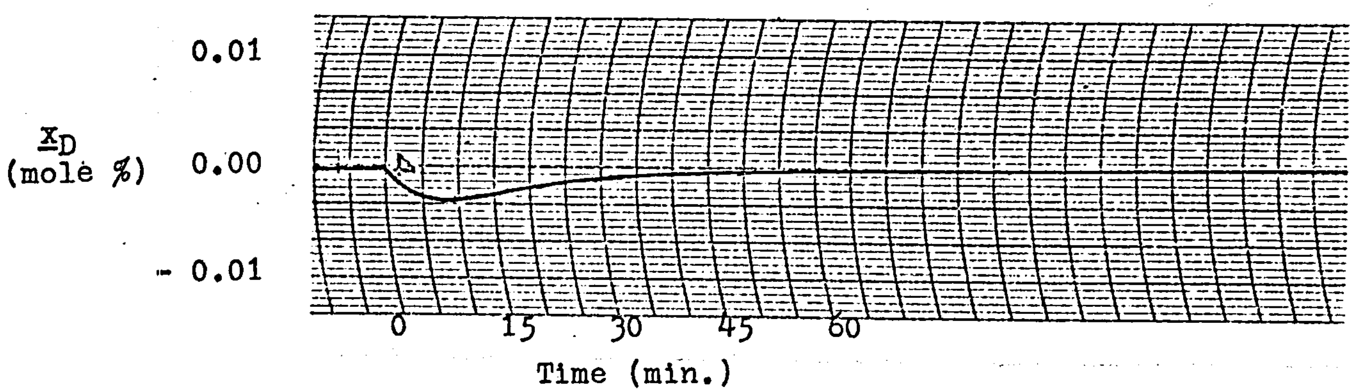
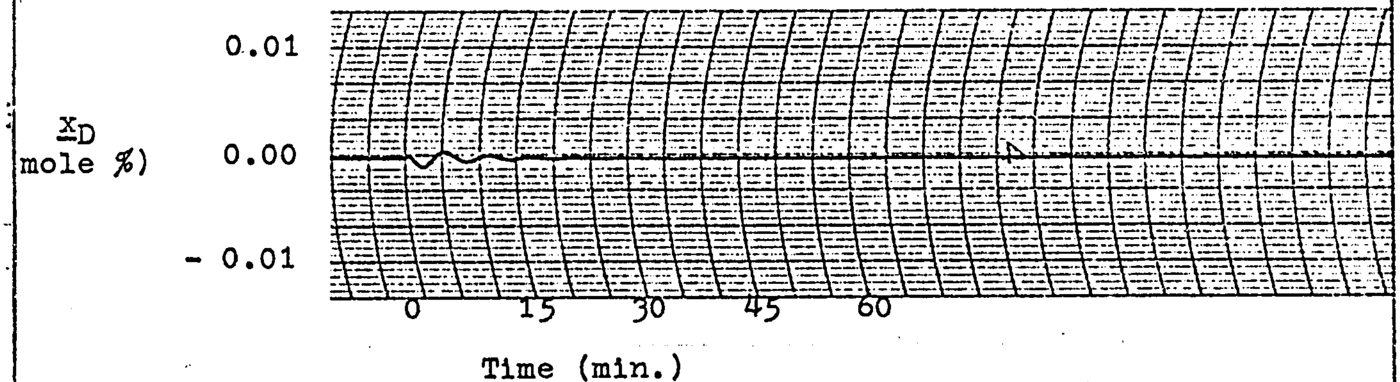


Fig. Response of x_D to a step change in x_F using instantaneous feed-forward control



ig. 38 Response of x_D to a step change in x_F using feedforward plus feedback control

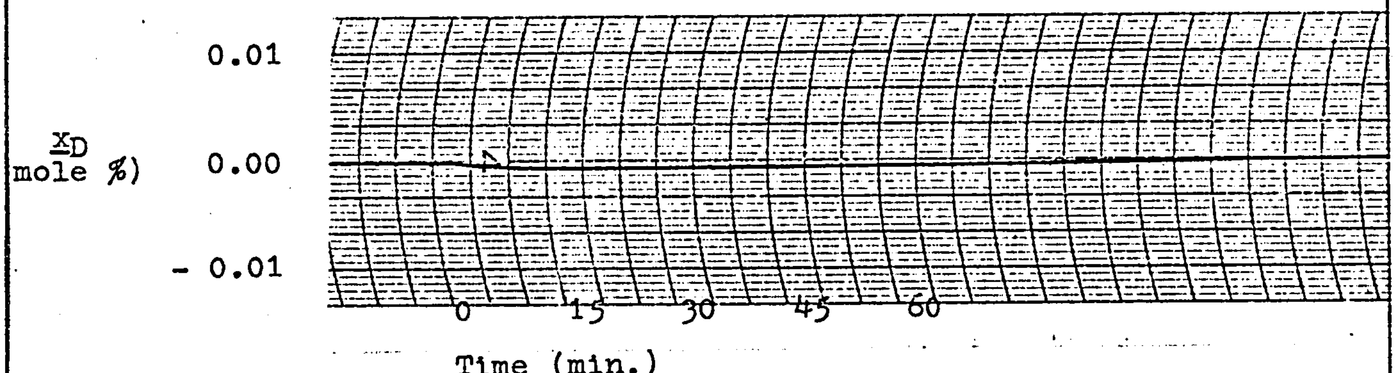
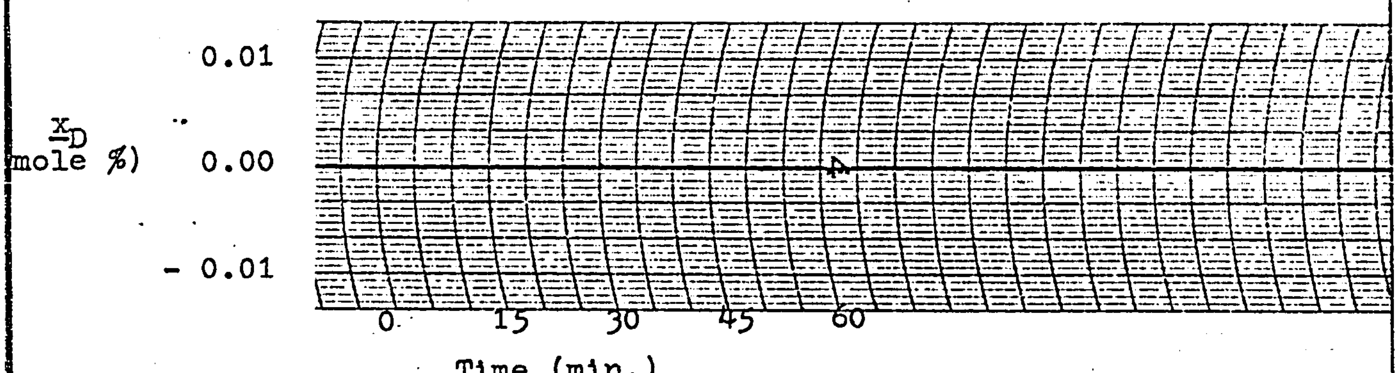


Fig. 39 Response of x_D to a step change in x_F using "perfect" feedforward control



ig. 40 Response of x to a step chang in x using "perfect" feed-forward plus feedback control

VI-2. <u>ANALYSIS OF RESULTS</u>

1. It should be noted that the boundaries of the controllability regions shown in Figures 13-17 are somewhat qualitative, because the search technique used in this study was discrete instead of continuous. This is of no consequence since for the determination of the optimal controller parameters, it is not necessary to have detailed knowledge of the location of the exact region boundaries.

2. For selection of the controller settings, the integral-squared-error criterion was employed. It can be seen from Figures 18-22 that the minimum in the integral-squared-error curve was quite flat meaning that a range of settings near the optimum is possible without significantly affecting the quality of control. The optimum controller parameters for the case considered may not be the optimal values for an actual column, since an ideal controller and linearity of the process were assumed. Since a criterion of constant overhead product composition was used to locate the optimum control point, the use of controller parameters near the true optimum

will have no effect.

3. The values of the off-set of overhead product composition x_D caused by controlling the intermediate plate compositions were calculated from equation (8). It should be noted that the values predicted by equation (8) depend on the gains of the transfer functions. The values shown are not exact off-set values for a step change in feed composition x_F because the gains of the transfer functions are basically unreliable (18). Nevertheless, these values indicate that by controlling intermediate plate composition x_j off-set in overhead product composition x_D does exist.

4. From the previous argument, it is obvious that for the objective of constant overhead product composition x_D the optimal control point should be the overhead product composition x_D or the vapor composition y_{10}, leaving the top plate. This is in contrast to the control strategy of Rosenbrock (41), who proposed a strategy based on the maximum rate of disturbance function D annulment after an isolated disturbance as the criterion. In order to apply this control strategy,

an excessive amount of information, since x_j and d/dt $H_j x_j$ have to be determined for all n+2 of the values of j, is desired. To avoid this impractical requirement and to ensure that the steady state, to which the column tends in the absence of further disturbance, is the one desired, on the basis of some convenient assumptions, he suggested that one or some more special plate compositions should be controlled. This optimal control point policy recommends, for the case of reflux manipulation, i.e., the so-called one-point control, that the composition of the plate where the desired composition is $x_F + \frac{1}{4}(x_{n+1} - x_0)$ or on that where it is $x_F - \frac{1}{4}(x_{n+1} - x_0)$ should be controlled. For the column utilized in this investigation, the 7th (or 8th) and the 3rd plate correspond to the above mentioned two plates respectively. Confusion regarding the use of various control criterion has already occured as exemplified by the work of Jafri (21) who adopted Rosenbrock's control point policy to an objective of constant overhead product composition instead of its own objective of minimum disturbance function D at any instance. This ~~error may not be serious when a small off-set of the~~ overhead product quality is tolerable, however, it must be stressed that conclusions can only be meaningful when the criterion is applied in the proper manner.

5. It is interesting to observe from Table 10 that the off-set in overhead product composition x_D which results by utilizing reflux manipulation to control x_7 (x_8) or x_3 to compensate a step change in feed composition x_F is much smaller than would result by controlling other intermediate plate compositions. This fact may or may not be purely coincidental; but this behavior is a good indication of the utility of Rosenbrock's disturbance function D criterion.

6. Figure 23 shows the response of the plate composition x_7 to a step change in feed composition when only the upper control loop was in action at the predetermined optimal settings. Control is good, and the damped oscillation response of the concentration has a period of approximately 10 minutes and a decay ratio of approximately 0.5. Figure 24 shows a similar result for the lower plate composition x_3 with only the lower control loop in action at the predetermined settings; the period of the damped oscillation is about 8 minutes and the decay ratio is about 0.55. The responses of x_7 and x_3 to a step change in feed composition x_F, when both

control loops were operative at the same settings as Figure 23 and Figure 24, are shown in Figure 25. Control becomes unstable, the responses of both controlled variables showed a period of damped oscillation of about 5-6 minutes and a decay ratio of 1.01 to 1.03. Figures 26-35 show the responses of both x_7 and x_3 when the upper controller alone was connected (at the predetermined settings) and then the lower controller brought into use by applying various degrees of proportional and integral action. The results indicate that increasing the gain of the lower controller causes an increase in the decay ratios of both controlled variables and addition of integral action results in poorer stability characteristics compared to when there is no control action in the lower loop. All these results demonstrate that the two control loops employed showed strong interaction against each other.

7. Employing an analog computer, Rosenbrock (41) used arbitrary transfer functions to demonstrate the possibility of interaction between two loops for two-point control of a distillation column by applying a step change in set point (desired values of x_3 and x_7) of either of the two controlled variables. His simulation

results showed that the addition of a second controller increased the period of damped oscillation substantially, whereas in this study the results are somewhat different, that is the effect of interaction was found to narrow the stability range of the control systems rather than enlarging the damped oscillation period. This difference is presumably because of the very different transfer functions that were used in both studies. Rosenbrock was interested in illustrating the possible difficulty of two-point control, therefore, he employed arbitrary time constants and gains of the transfer functions. The process gains were very small, close to unity, hence a high value of controller gain was allowable. But, the purpose of this study was undertaken to determine the type of interaction existing between two control loops for an actual process. The overall process gain of the column is much larger than the one arbitrarily chosen by Rosenbrock and consequently, the allowable value of the controller gain, in order to fulfill the stability requirement, is decreased. This resulted in narrowing the range of stability of the control systems. Therefore the system would become unstable long before the effect of enlarging the damped oscillation period appears substantially. This may be the reason for the difference showed between Rosenbrock's investigation and the present work. As a consequence of this, it is confirmed that the interaction between two control loops depends strongly on the dynamic characteristics of the system under consideration (41).

Jafri (21), who employed a set of empirical transfer functions, simulated two-point control for a step change in feed composition x_F (usually referred as load variable). Interaction was found to exist despite the fact that the effect of reflux rate on the lower plate composition was neglected.

No general statement concerning such possible interaction of two control loops can be made, since no analytical study of this matter has been undertaken to date.

8. Methods of improving the characteristics of an interacting control system are still not fully understood but techniques for making the two control loops noninteracting have been developed (41, 48). One approach to noninteracting control is to construct a linearized model of the process. Assuming that two variables are to be manipulated to control two outputs,

one can visualize a two by two matrix of transfer functions relating the outputs to the manipulated variables which results in the use of four controllers as opposed to the conventional control scheme with two controllers. The additional controllers are chosen so as to diagonalize the process transfer function matrix and thus effectively cancel interaction effects. In principal this idea seems promising, but there has been little investigation of the residual interaction due to nonlinearities. Obviously, a linear control philosophy of noninteraction can solve only part of the problem when the controlled process is in fact nonlinear.

9. Figure 36 shows the response of overhead product composition x_D to a step change in feed composition x_F when only the conventional feedback control loop, using a proportional-integral controller, was operative. Even employing the predetermined optimal controller settings, the maximum deviation from the desired value of x_D is - 1.09 mole % and about 50-55 minutes are required to damp out the oscillations. Figure 37 shows that when an instantaneous feedforward control is used alone, the response of x_D is somewhat like a ramp response and the maximum deviation from the desired value of x_D is - 0.3

mole % and takes about 60 minutes to return to the steady state value. This illustrates that the controllability, as far as the maximum deviation from the specified value of x_D is concerned, has been improved as compared to the feedback control, but the time required to return to the original steady state value has not been improved. Utilization of a combination of these two, feedforward plus feedback compensation control, resulted in better control as shown in Figure 38. The response of x_D has a maximum deviation of - 0.13 mole % and only 25 minutes are required for return to the steady state.

10. Control can be even improved further as shown by Figure 39 which is a record of the response of x_D when a "perfect" feedforward control was employed. The response of x_D when "perfect" feedforward plus feedback compensation control was used is shown in Figure 40. It can be seen that almost perfect control was achieved by using "perfect" feedforward, but whether it is worthwhile to use the so-called "perfect" feedforward control has to be justified economically and practically. And, the synthesis of the feedforward functional G_F requires an exact detailed knowledge of the process

dynamics which still can not be predicted precisely.

CHAPTER VII

EVALUATION OF CONTROL SCHEMES INVESTIGATED

1. Conventional feedback control of constant overhead product composition by manipulating reflux rate is usually satisfactory when a proportional-integral controller is used in the feedback loop.

The determination of the optimal control point, for feedback control, depends on the objective of control. Confusion or misunderstanding of the significance of the control objective in selecting the optimal location of the sensing element has already occured (21). On the basis of transient data, of a ten-plate pilot plant distillation column, reported by Gerster and co-workers (26), five cases for different control points of the reflux manipulation were simulated on the analog computer. The simulation results showed that for the criterion of constant overhead product composition the optimal location of the sensing element should be in the overhead product stream or in the vapor stream leaving the top plate if the sensing element is highly sensitive. Generally in the region of the intermediate plates the composition has a greater gradient as compared to near the top plate, and consequently a sensing element that has a large "dead space", on the basis of the work of Williams and co-workers (53, 54), ahould be located at intermediate plate. But, intermediate plate composition control results in a permanent off-set in overhead product composition when a disturbance enters the process from the feed composition. This off-set in the overhead product composition is critical to the objective of constant overhead product composition. Calculated values of off-set in overhead product composition due to the control of intermediate plate composition, for a particular set of operating conditions, showed that the magnitude of off-set for the cases of controlling the 7th (8th) or 3rd plate composition was much less than the off-set which resulted by controlling any other plate composition. The above mentioned two plates correspond to Rosenbrock's recommendation of the optimal control point for the case of reflux manipulation. This may or may not be purely coincidental, but it offered a good indication of the utility of Rosenbrock's "disturbance function" criterion.

2. The control of both product compositions at constant values is usually referred to as "two-point" control. Although "two-point" control has a potential advantage over one-point control (19), it has been pointed out that control is sometimes difficult. This is because of the existince of the interaction between the two control loops. The results of this study indicated a different type of interaction occurring between the control loops than Rosenbrock has noted. This investigation showed that interaction affects the stabilities of the two control loops, whereas the interaction effect noted by Rosenbrock had the effect of enlarging the damped oscillation period (41). Therefore, the so-called "two-point" control scheme should not be applied to a binary distillation column without investigating the severity of the interaction. Even though methods for the design of a noninteracting control scheme (41) have been suggested, it is prudent to make that investigation, for the sake of economical justification. Interaction depends strongly upon the dynamics of the process and with a particular column no difficulty may arise; while with another apparently similar, the transfer functions may be such as to make

the column virtually uncontrollable. This confirms the strong dependence of the interaction between control loops on the dynamics of the process. Nevertheless, a general theoretical explanation of the phenomenon of interaction between control loops is still not valid.

3. Compared to conventional feedback control, instantaneous feedforward control decreases the magnitude of the maximum deviation from the desired value of the controlled variables substantially, but does not decrease the time required to return to the desired value. This is illustrated by the results from analog computer simulations. Instantaneous feedforward control of the overhead product composition x_D reduced the maximum deviation of x_D from - 1.09 mole % to - 0.30 mole % but both schemes required about 50-60 minutes for return to the desired value. Feedforward plus feedback compensation control was found not only to decrease the magnitude of maximum deviation but also to decrease the time required for the overhead product composition to return to the desired value by almost one-half of the time required when only conventional feedback control was used. When feedforward plus feedback compensation control was employed, the

maximum deviation of the overhead product composition from its desired value changed from - 1.09 mole % to - 0.13 mole % and the time required for x_D to return to its desired value decreased from 50 minutes to about 25 minutes compared to feedback control. Also, since feedforward control alone cannot compensate for any variation of the controlled variable due to unpredicted disturbances, such as ambient temperature, steam quality, the feedback compensator can be employed as a monitor and to control the deviation of the controlled variable caused by these unpredicted disturbances.
The results have shown that for close control of a critical variable feedforward plus feedback compensation control should be employed. Although the initial investment for such a control scheme involve the cost of an extra controller the savings achieved by the reduction of off specification product would no doubt justify the additional cost in a very short time.

CHAPTER VIII

RECOMMENDATIONS FOR FUTURE WORK

1. For design of control systems of a distillation column, a set of linearized differential equations to describe the dynamic characteristics of the column is sufficient(23). In order to avoid the tedious calculations involved in solving these equations, it would be worthwhile to attempt to express the process by transfer functions in the form of nomograph with the operating conditions as the parameters. Another possible approach would be to develop quick, simple but sufficiently accurate methods for estimating the parameters of the transfer functions involved in the design of control systems for different operating conditions.

2. Further study on the theory of interacting control loops is needed. The approach to noninteracting control for a linear system has been established quite extensively. Obviously, a linear control philosophy of noninteraction can only solve part of the problem when the controlled process is in fact nonlinear. The effect of nonlinearities is still not

fully understood, although Rosenbrock's "disturbance function" seems promising. Therefore, investigation of the residual interaction due to nonlinearity should be conducted.

3. Extension of the type of result obtained by means of the "disturbance function" is evidently desirable. A statistical description of the rate of "disturbance function" entering the process and a rigrous mathematical expression of the rate of "disturbance function" vanishing due to the control action applied are required, to obtain a closer physical understanding of the application of Rosenbrock's theory.

4. Direct digital computer control is possible, although very limited information has been reported in the literature. Recently, in 1965, Thompson (47) reported the operating experience of Imperial Chemical Industries Ltd. with direct digital computer control of a small size chemical plant and concluded that if certain requirements can be fulfilled direct digital computer control is the pattern of instrumentation for the future. The Institution of Chemical Engineers of Great Briton have compiled (5) a list giving details of current process optimization

research projects in progress in universities and colleges of advanced technology throughout the United Kingdom and in goverment research stations. From the list, it can be seen that many projects related to the applications of direct digital computer control to chemical process have already been started. An actual application of direct digital computer control, utilizing the avaliable design techniques for a multivariable system, to a chemical process such as a distillation column would be invaluable for standarizing the design approach to direct digital computer control.

5. Investigation of both the dynamics and control of multicomponent distillation columns should be conducted.

6. The control scheme of manipulating boilup rate to maintain the bottom product composition at a specified value has already been adopted in the chemical industries. However, no systematic work concerning the optimal sensing element location for the boilup manipulation control scheme has been reported in the literature. This study would not be difficult to achieve on an analog computer once the criterion of control has been determined and the process is properly characterized.

APPENDIX 1

Comparison of Gerster's data and approximated models

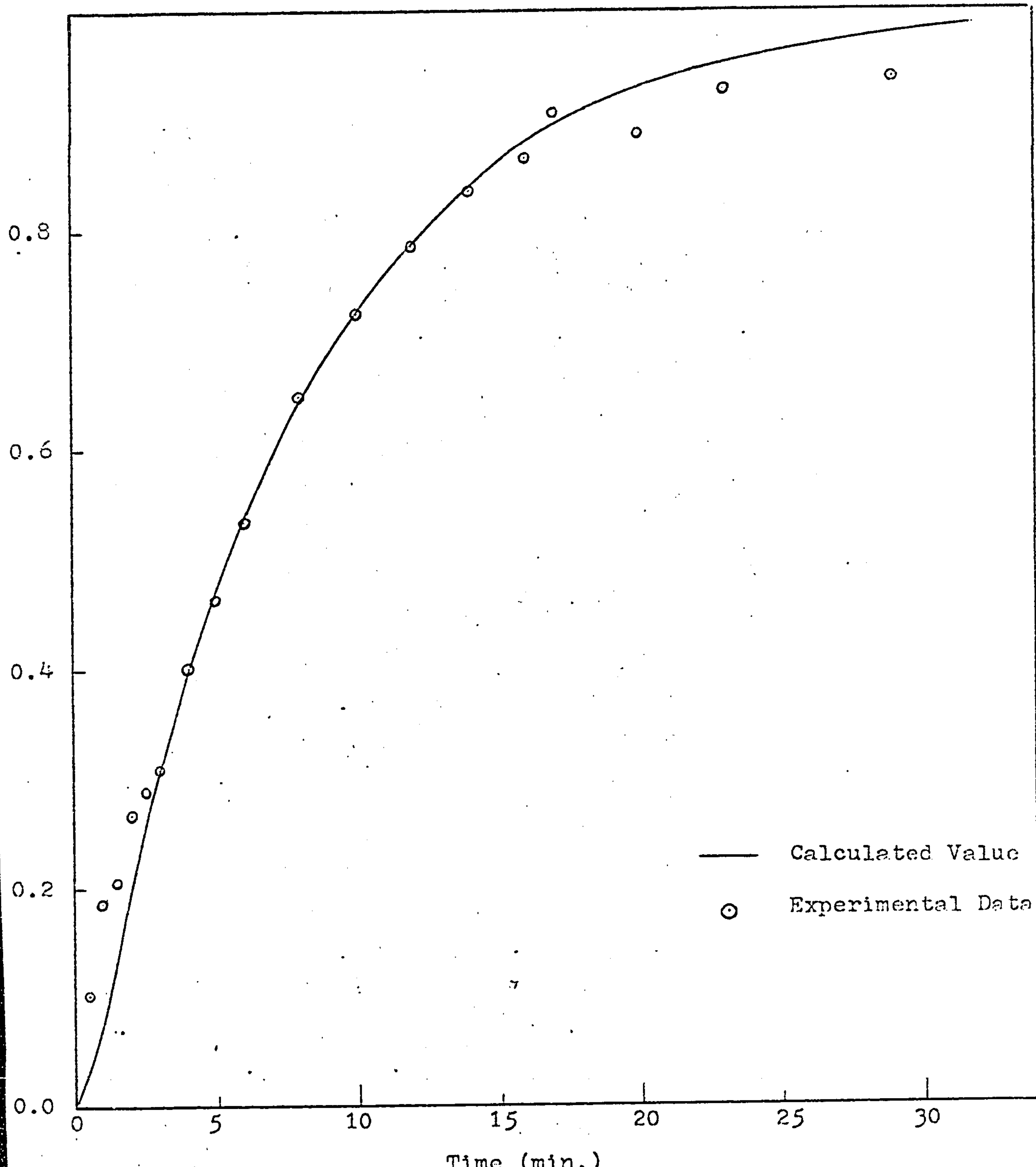


Fig. 41 Transient response of x_D to a step change in x_F

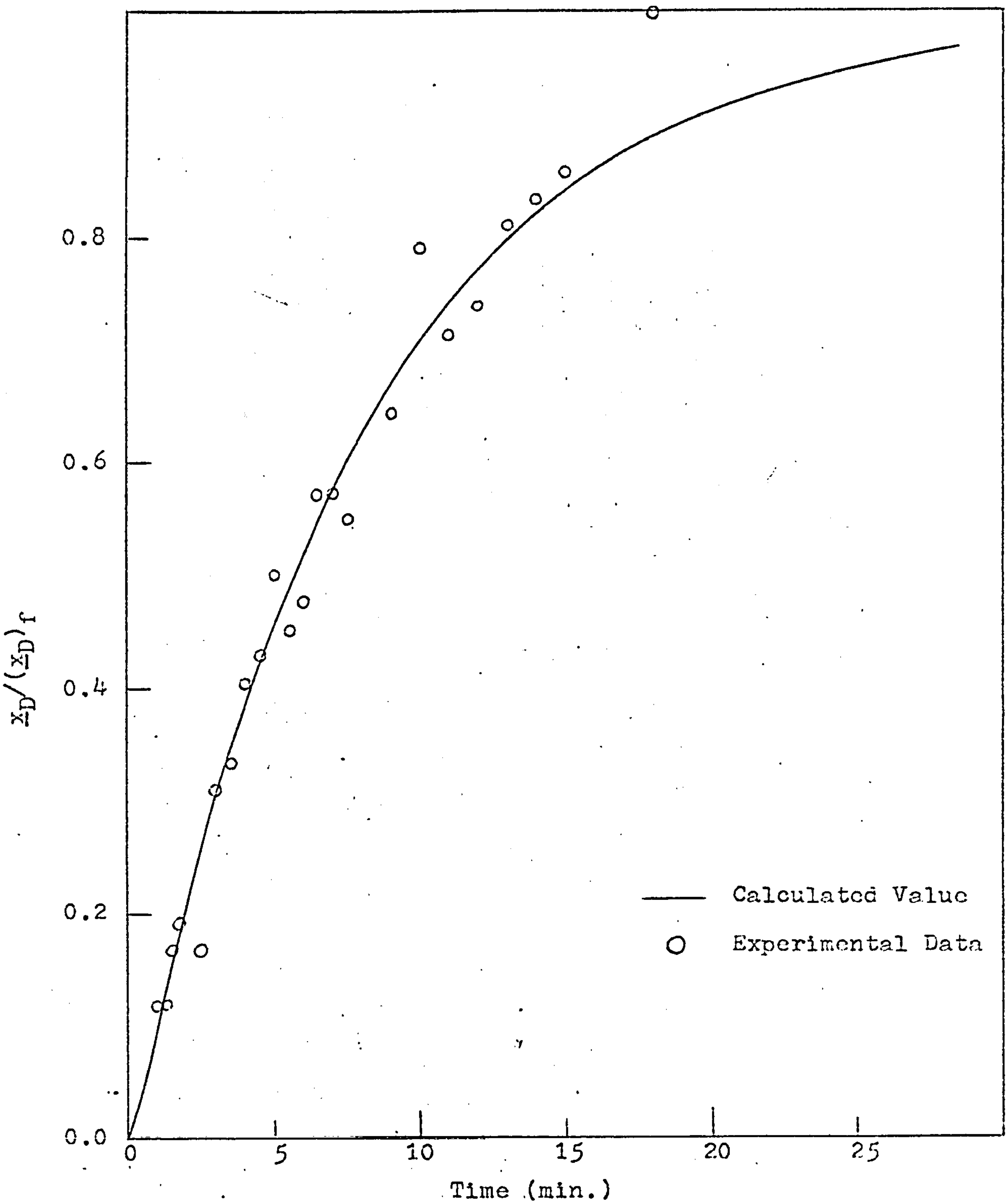


Fig. 42 Transient response of x_D to a step change in R

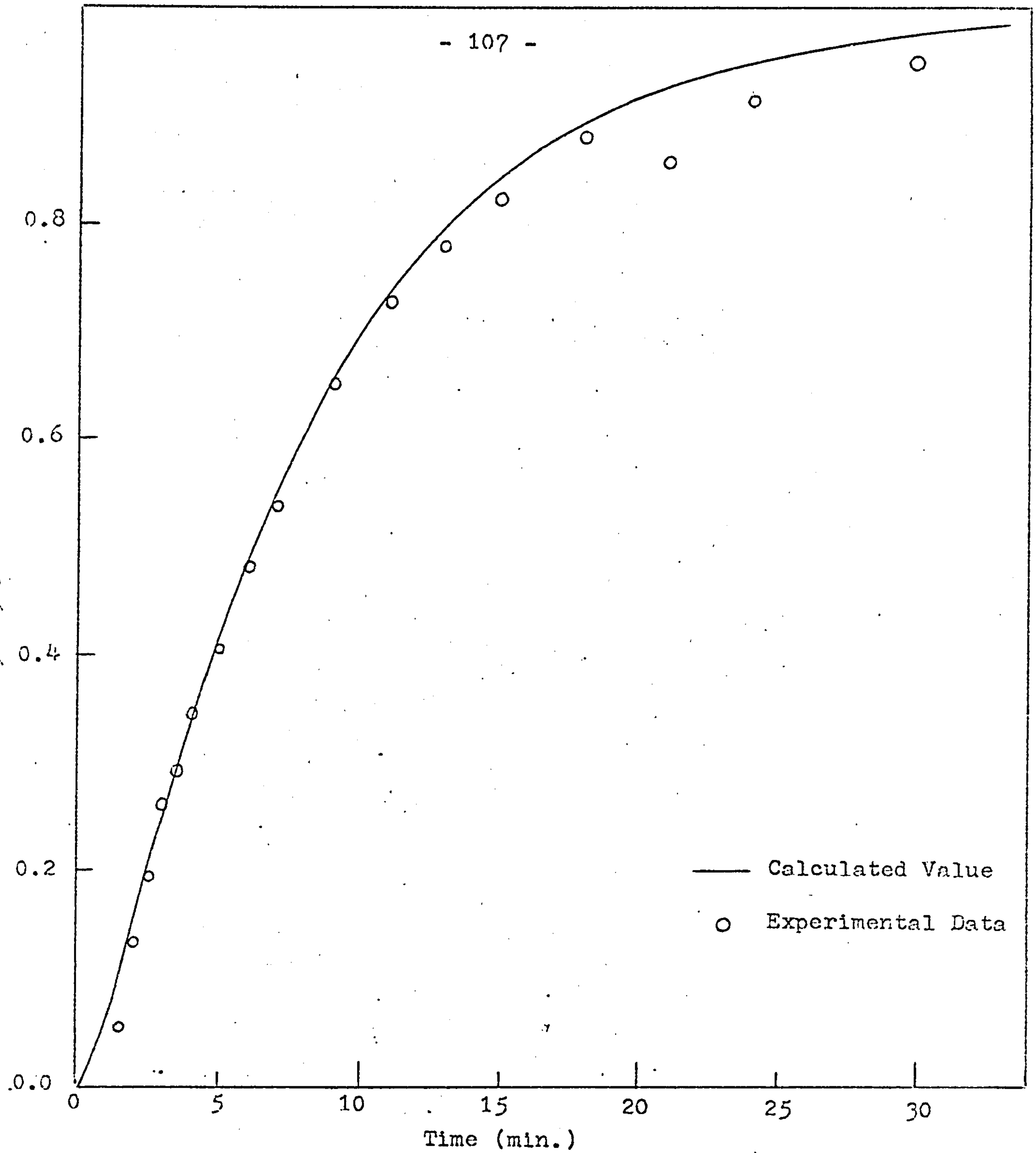


Fig. 43 Transient response of x_9 to a step change in x_F

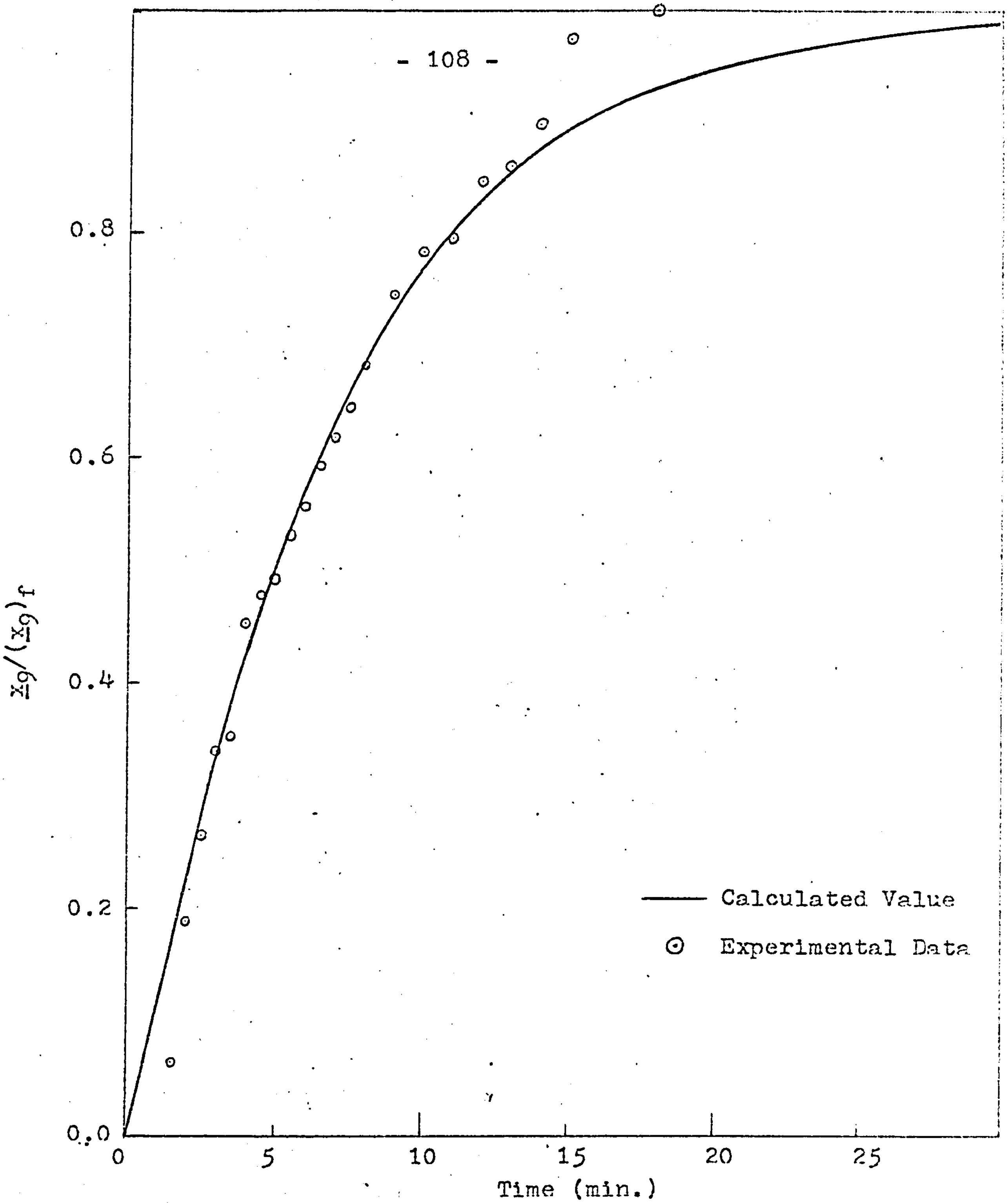


Fig. 44 Transient response of x_9 to a step change in R

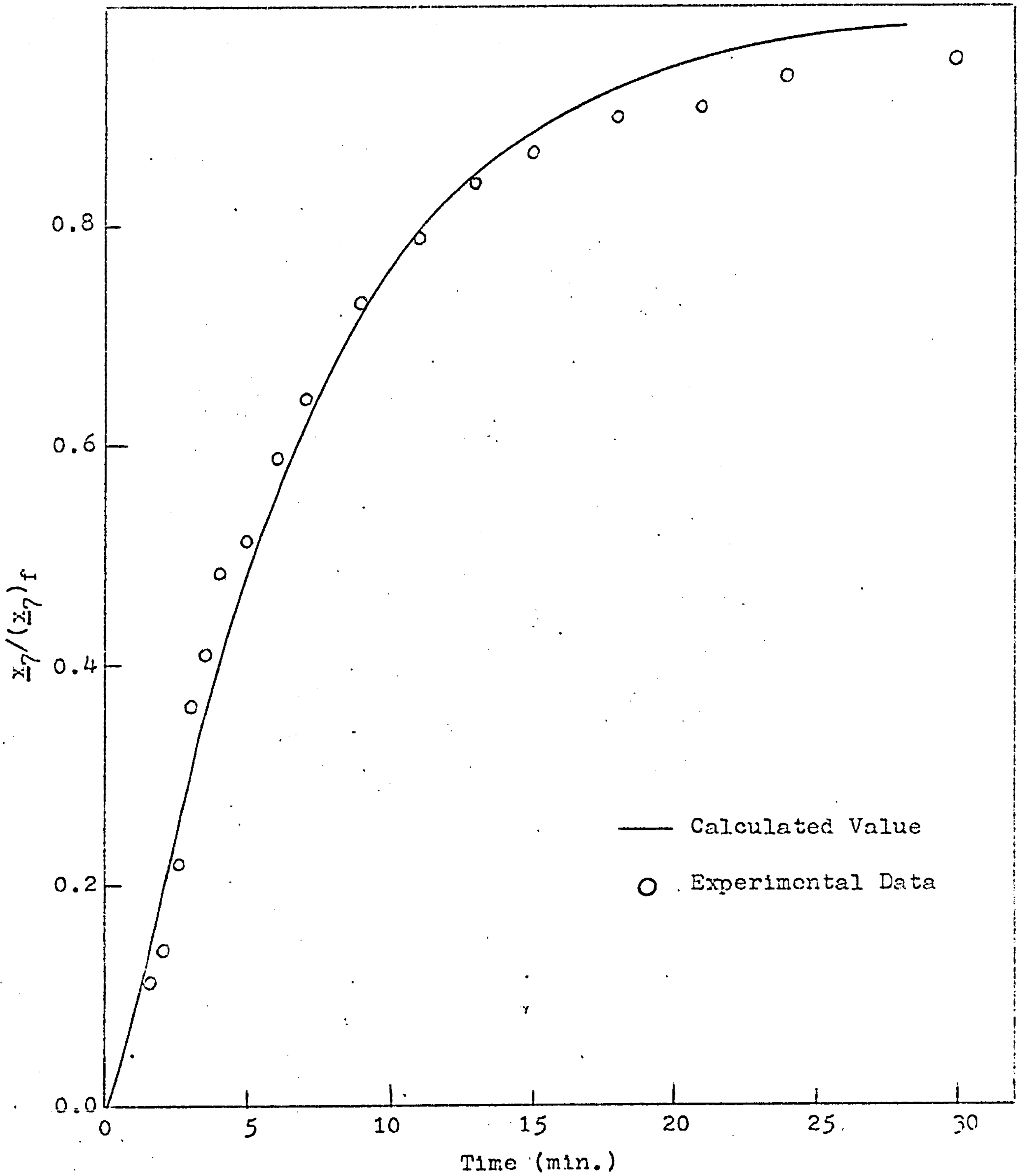


Fig. 45 Transient response of x_7 to a step change in x_F

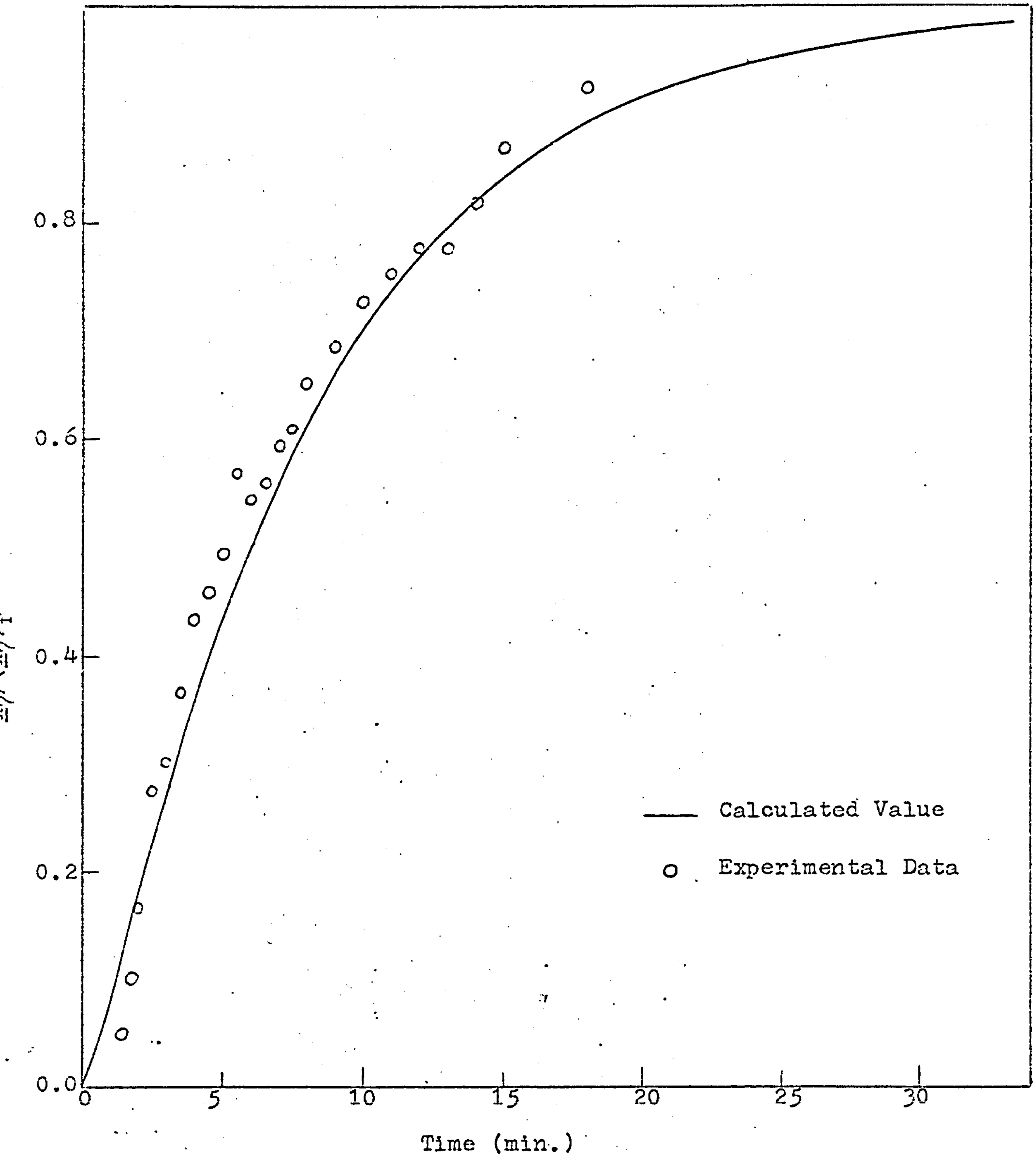


Fig. 46 Transient response of x_7 to a step change in R

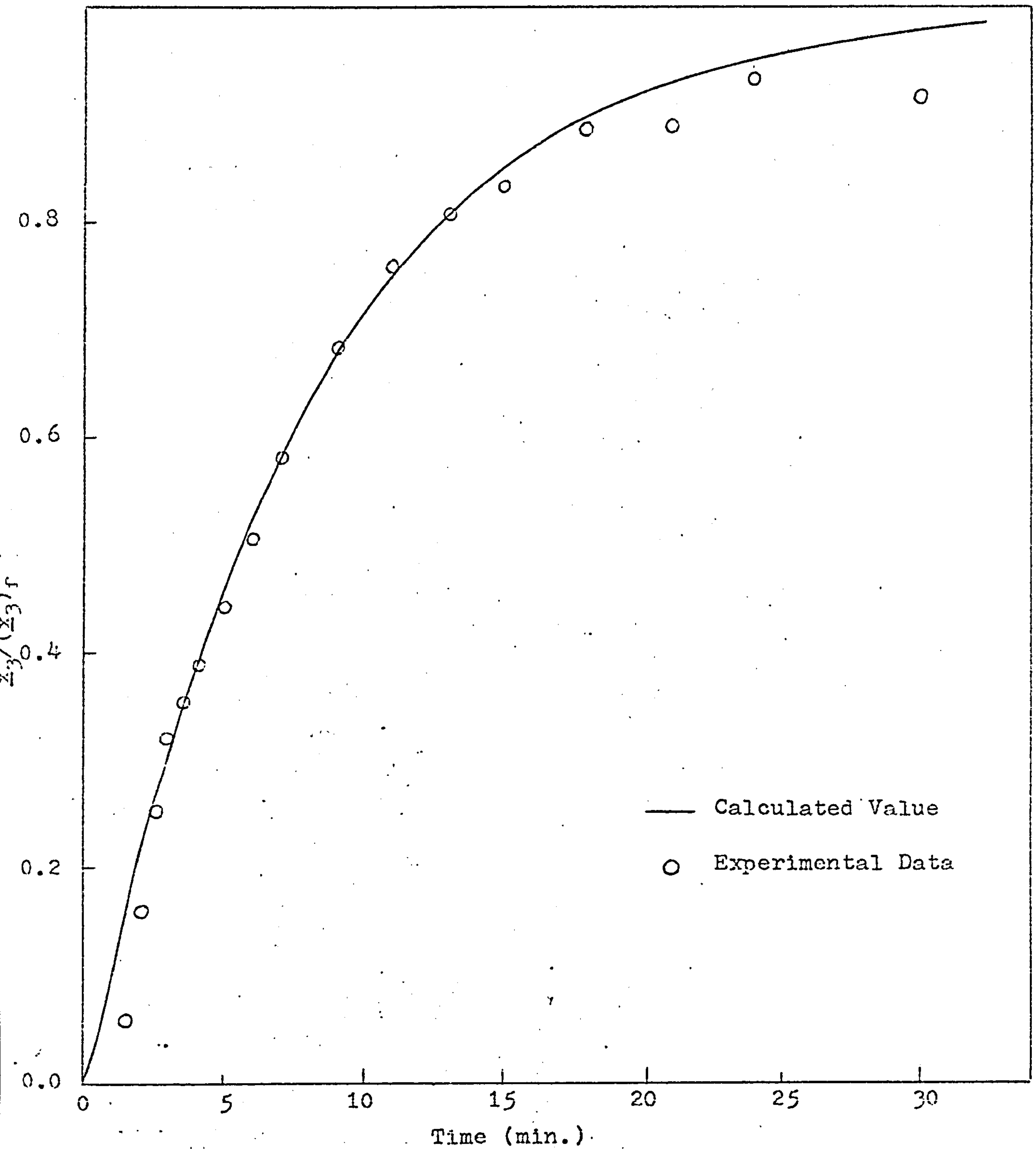


Fig. 47 Transient response of x_3 to a step change in x_F

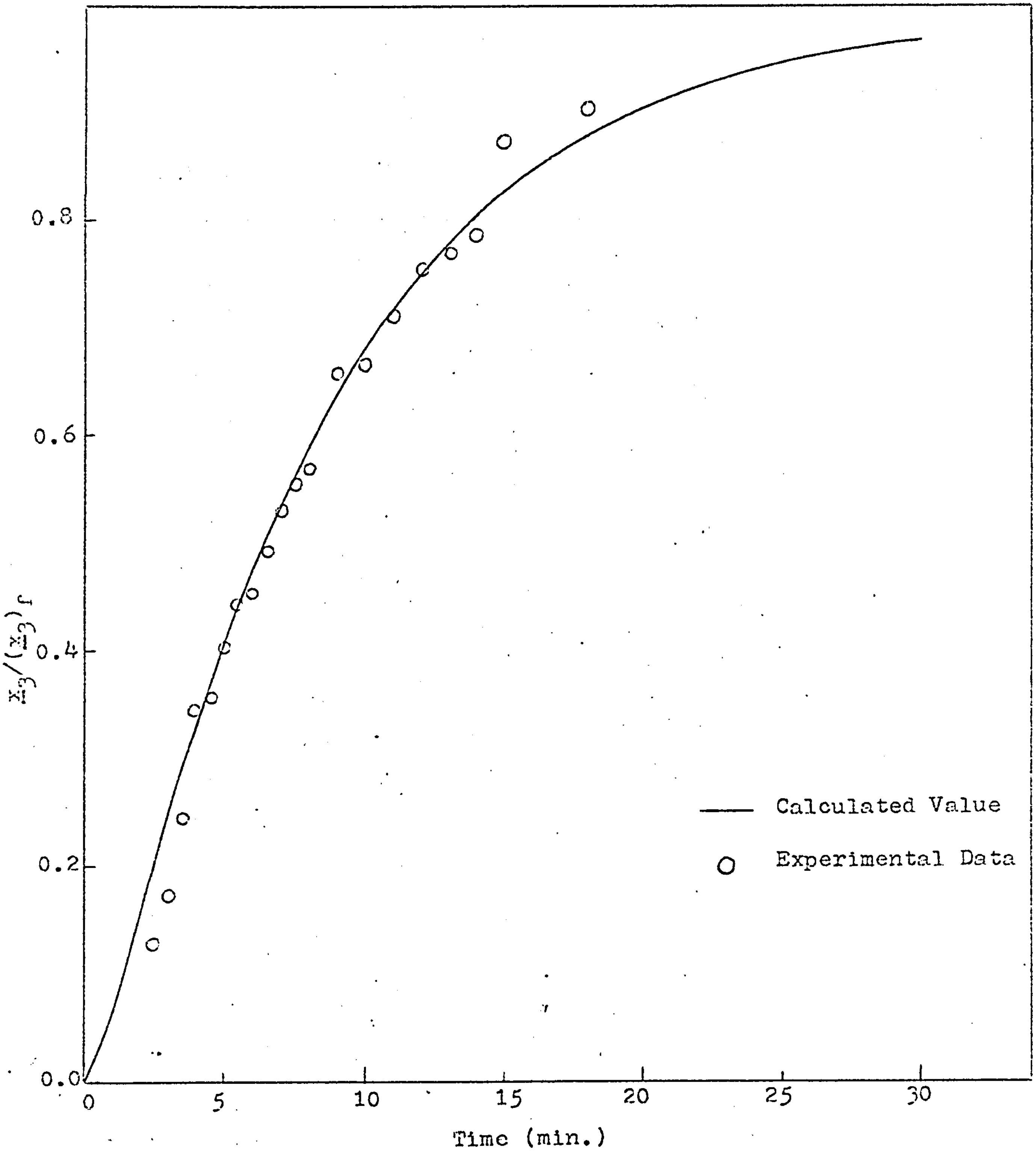


Fig. 48 Transient response of x_3 to a step change in R

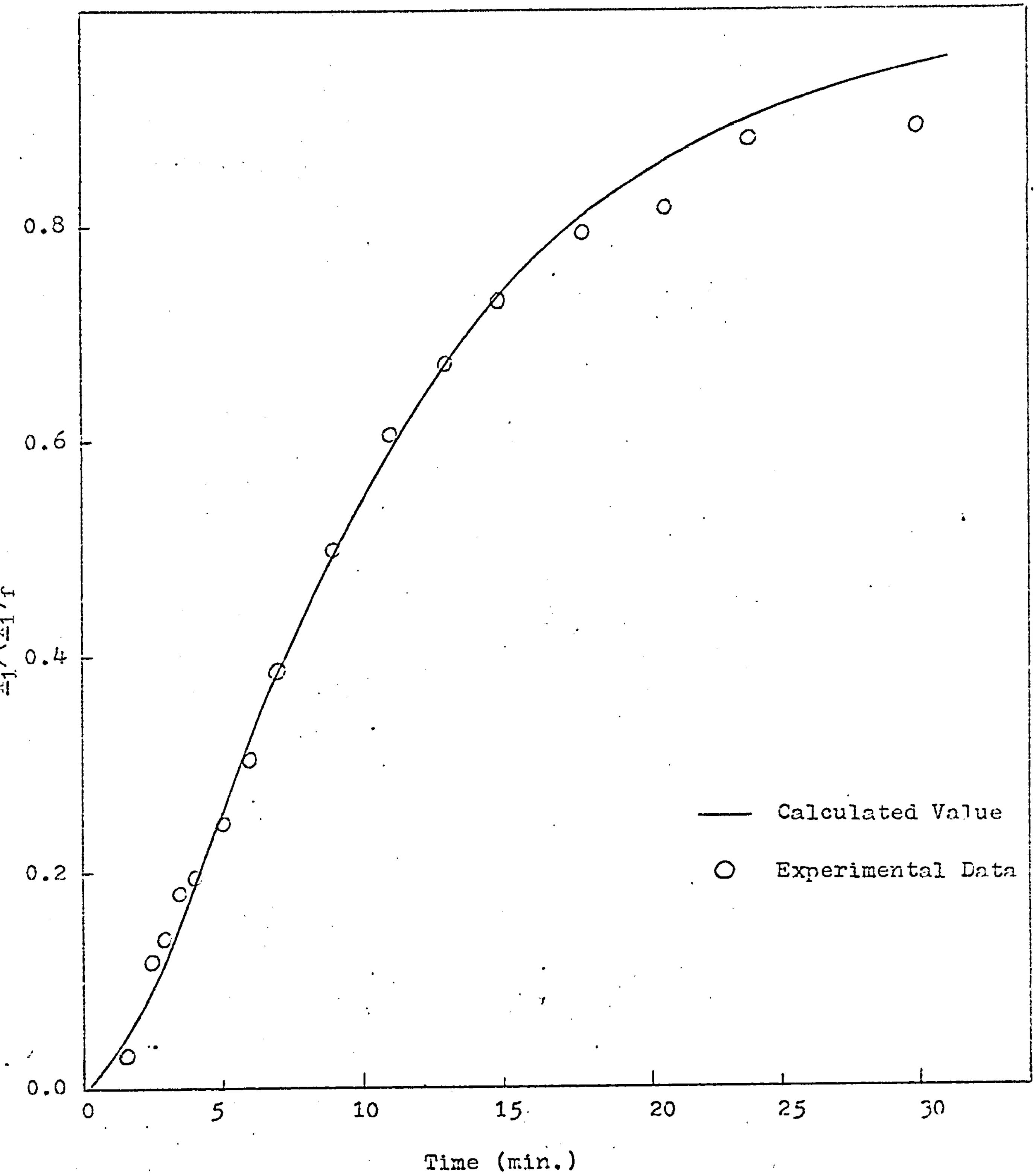


Fig. 49 Transient response of x_1 to a step change in x_F

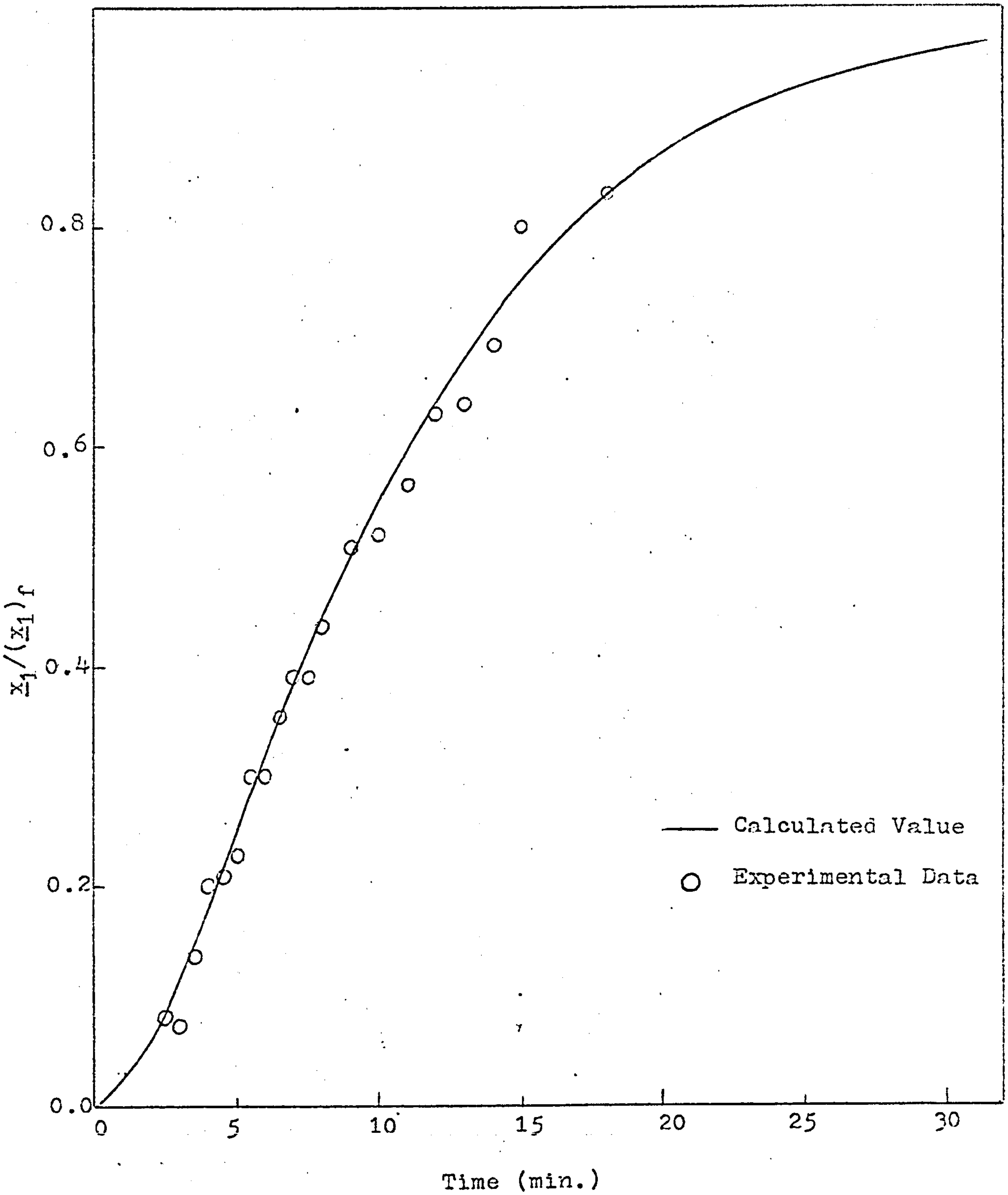


Fig. 50 Transient response of x_1 to a step change in R

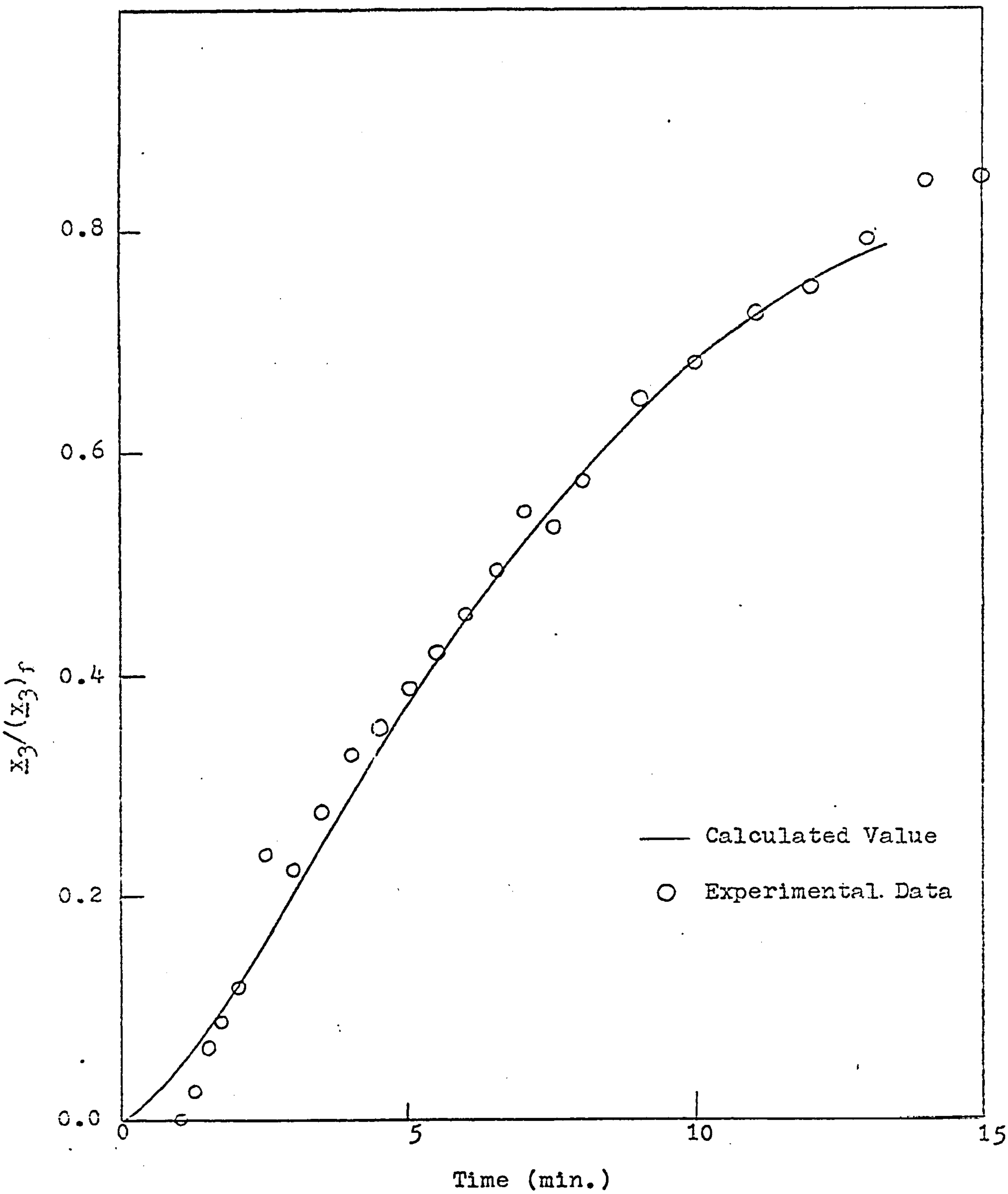


Fig. 50a Transient response of x_3 to step change in boilup rate

Digital Computer Program

```
C TRANSIENT RESPONSE OF AN OVERDAMPED SECOND ORDER
  SYSTEM
5 DEAD 25, DETAT, TTH, TFO
  PRINT 30, DETAT, TTH, TFO
  THETA = 0
10 THETA = THETA + DETAT
  ZONE = THETA/TTH
  ZTWO = THETA/TFO
  BETA = TTH/(TTH-TFO)
  Y = 1. - (BETA/(2.73**ZONE)-((BETA-1.0)/(2.73**ZTWO))
  PRINT 35, THETA, Y
  IF (Y - 0.98), 10, 15, 15
25 FORMAT (3F10.4)
30 FORMAT (//1H , 3F10.4)
35 FORMAT (1H , F10.4, F10.8)
15 IF (DETAT - 0.) 20, 20, 5
20 CALL EXIT
  END
```

APPENDIX 2

<u>Analog computer circuits and potentiometer sheets</u>

TABLE 13

TABULATION OF THE CONDITIONS SIMULATED

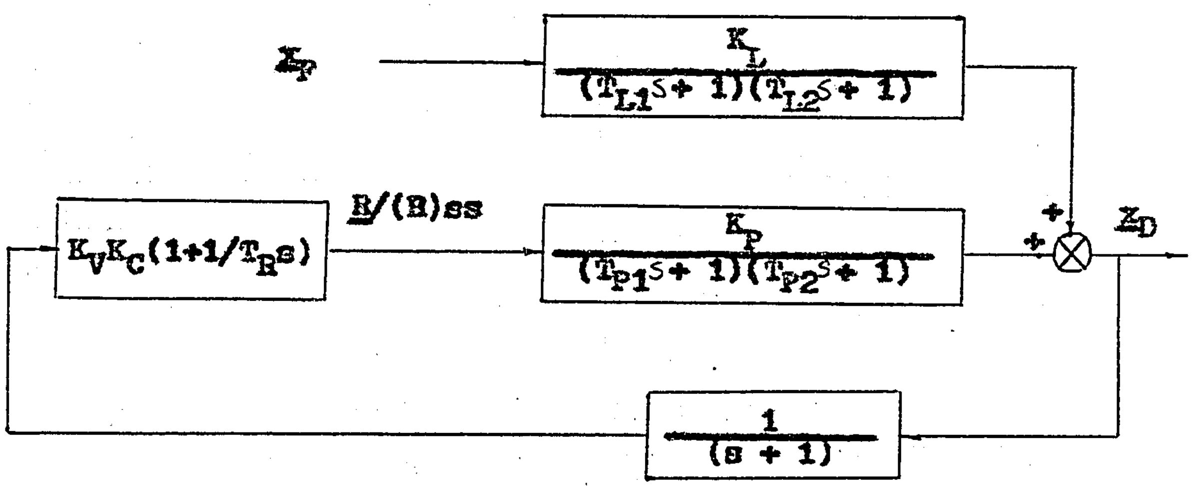
Case Number	Conditions
1A	Reflux manipulation, x_D is controlled
2B	Reflux manipulation, x_9 is controlled
3B	Reflux manipulation, x_7 is controlled
4B	Reflux manipulation, x_3 is controlled
5B	Reflux manipulation, x_1 is controlled
6C	Two-point control, x_7 and x_3 are controlled
7C	Boilup manipulation, x_3 is controlled
8D	Instantaneous feedforward plus feedback compensation control
8D	"Perfect" feedforward plus feedback compensation control

Example of the scaling of the problem :

Chose a time factor as :

1 minute problem time = 1 second machine time.

Set up an unscaled block diagram for the control system studied as :



Chose the proper magnitude scale factors and then set up a scaled block diagram as :

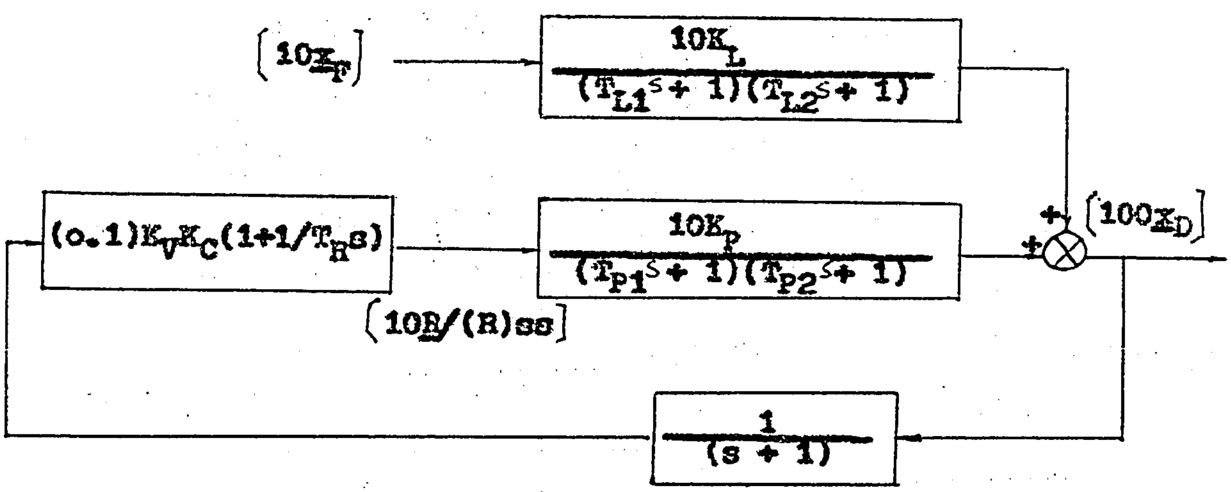


Fig. 51

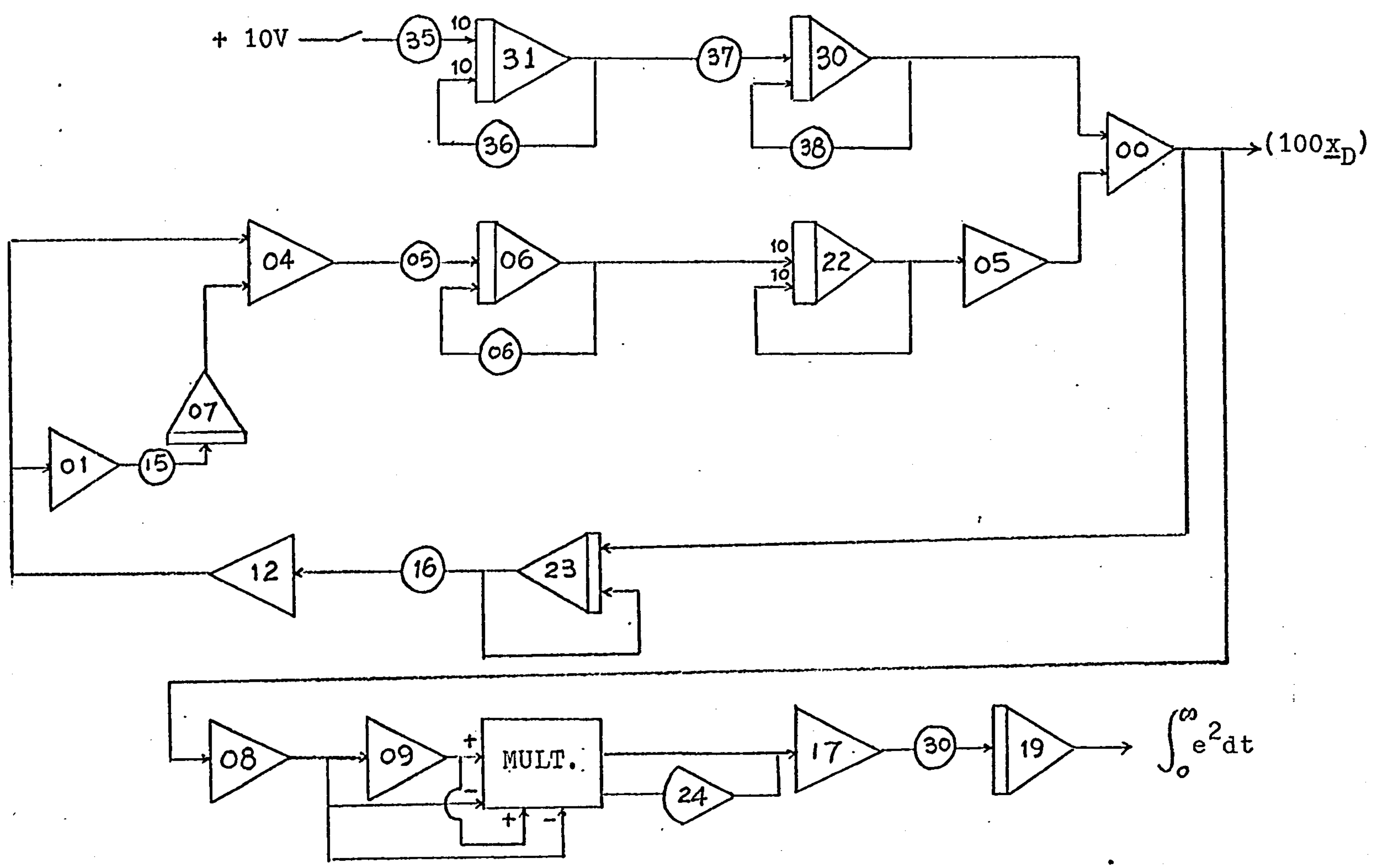


Fig. 52 Analog computer circuit for case 1A

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
1A	35	5.8/(0.15)(10)	= 3.87	31	10	0.387	+ 10V
1A	36	1/(0.15)	= 6.67	31	10	0.667	
1A	37	1/(7.55)	= 0.132	30	1	0.132	
1A	38	1/(7.55)	= 0.132	30	1	0.132	
1A	05	3.69/(8.2)	= 0.450	06	1	0.45	
1A	06	1/(8.2)	= 0.122	06	1	0.122	
1A	15	$1/T_R$	Various				
1A	16	$(0.2)K_C$	Various				

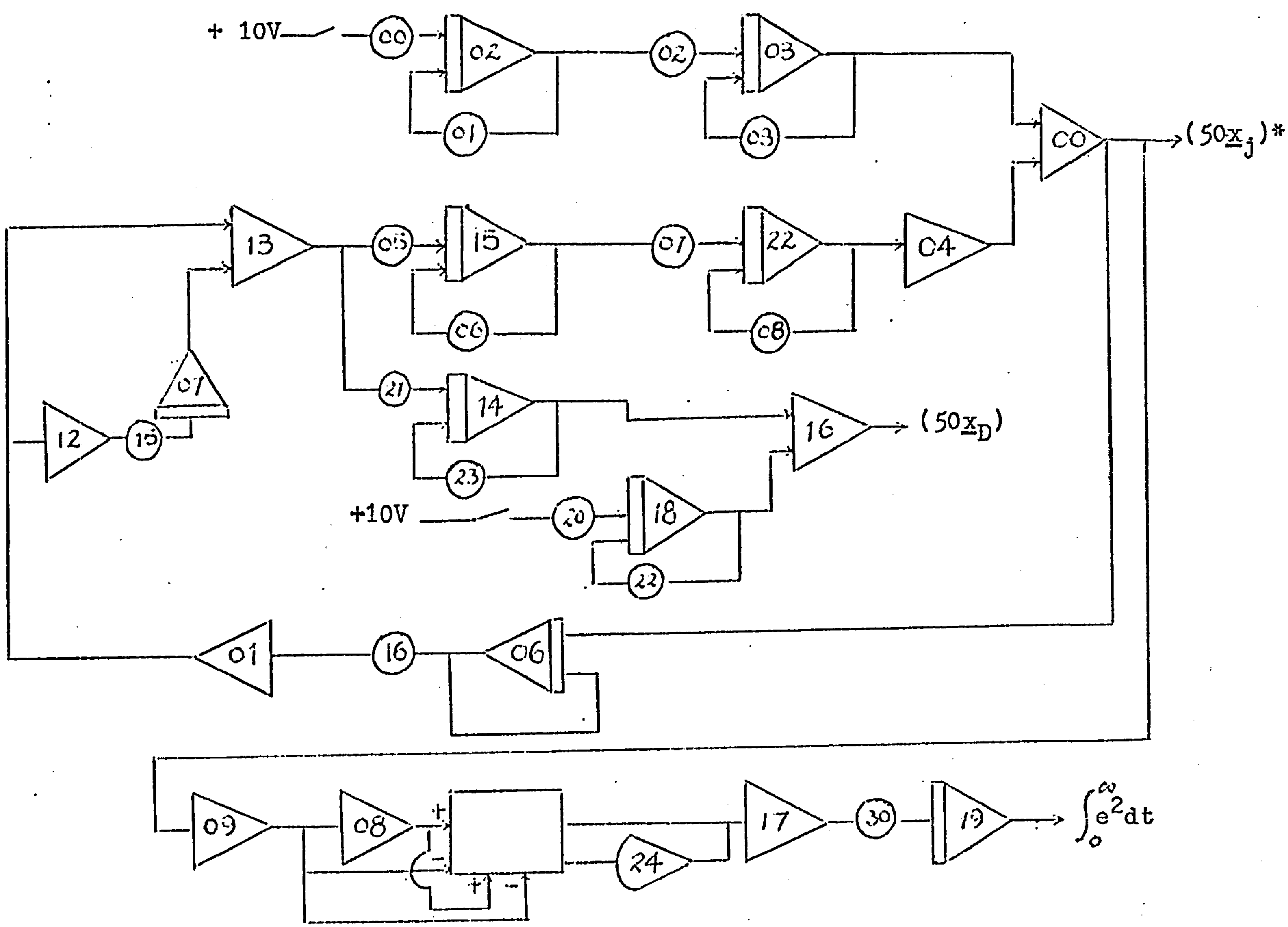


Fig. 53 Analog computer circuit for cases 2B-5B

* where j = 1, 3, 7, 9.

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
2B	00	1/(0.65)(10)	= 0.154	02	1	0.154	+10V
"	01	1/(0.65)	= 1.540	02	10	0.154	
"	02	5.85/7.80	= 0.750	03	1	0.750	
"	03	1/7.8	= 0.128	03	1	0.128	
"	05	3.46/6.65	= 0.520	15	1	0.520	
"	06	1/6.65	= 0.150	15	1	0.150	
"	07	1/0.27	= 3.700	22	10	0.370	
"	08	1/0.27	= 3.700	22	10	0.370	
"	15	$1/T_B$	Various				
"	16	$K_C(0.4)$	Various				
"	20	2.9/(7.8)(10)	= 0.037	18	1	0.037	+10V
"	21	1.84/8.2	= 0.220	14	1	0.220	
"	22	1/7.8	= 0.128	18	1	0.128	
"	23	1/8.2	= 0.122	14	1	0.122	

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
3B	00	8.35/(0.6)(10)	= 1.390	02	10	0.139	+10V
"	01	1/0.6	= 1.670	02	10	0.167	
"	02	1/6.6	= 0.152	03	1	0.152	
"	03	1/6.6	= 0.152	03	1	0.152	
"	05	5.26/8.0	= 0.658	15	1	0.658	
"	06	1/8.0	= 0.125	15	1	0.125	
"	07	1/0.46	= 2.170	22	10	0.217	
"	08	1/0.46	= 2.170	22	10	0.217	
"	15	$1/T_R$	Various				
"	16	$K_C(0.4)$	Various				
"	20	2.9/(7.8)(10)	= 0.037	18	1	0.037	+10V
"	21	1.84/8.2	= 0.220	14	1	0.220	
"	22	1/7.8	= 0.128	18	1	0.128	
"	23	1/8.2	= 0.122	14	1	0.122	

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
4B	00	9.7/7.9(10)	= 0.123	02	1	0.123	+10V
"	01	1/7.9	= 0.1265	02	1	0.127	
"	02	1/0.1	= 10.0	03	10	1.000	
"	03	1/0.1	= 10.0	03	10	1.000	
"	05	5.875/8.25	= 0.712	15	1	0.712	
"	06	1/8.25	= 0.121	15	1	0.121	
"	07	1/0.65	= 1.540	22	10	0.154	
"	08	1/0.65	= 1.540	22	10	0.154	
"	15	$1/T_R$	Various				
"	16	$K_C(0.4)$	Various				
"	20	2.9/(7.8)(10)	= 0.037	18	1	0.037	
"	21	1.84/8.2	= 0.220	14	1	0.220	
"	22	1/7.8	= 0.128	18	1	0.128	
"	23	1/8.2	= 0.122	14	1	0.122	

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
5B	00	5.625/(2.5)(10)	= 0.225	02	1	0.225	+10V
"	01	1/2.5	= 0.400	02	1	0.400	
"	02	1/9.2	= 0.109	03	1	0.109	
"	03	1/9.2	= 0.109	03	1	0.109	
"	05	4.825/7.95	= 0.607	15	1	0.607	
"	06	1/7.95	= 0.126	15	1	0.126	
"	07	1/3.25	= 0.308	22	1	0.308	
"	08	1/3.25	= 0.308	22	1	0.308	
"	15	$1/T_R$	Various				
"	16	$K_C(0.4)$	Various				
"	20	2.9/(7.8)(10)	= 0.037	18	1	0.037	
"	21	1.84/8.2	= 0.220	14	1	0.220	
"	22	1/7.8	= 0.128	18	1	0.128	
"	23	1/8.2	= 0.122	14	1	0.122	

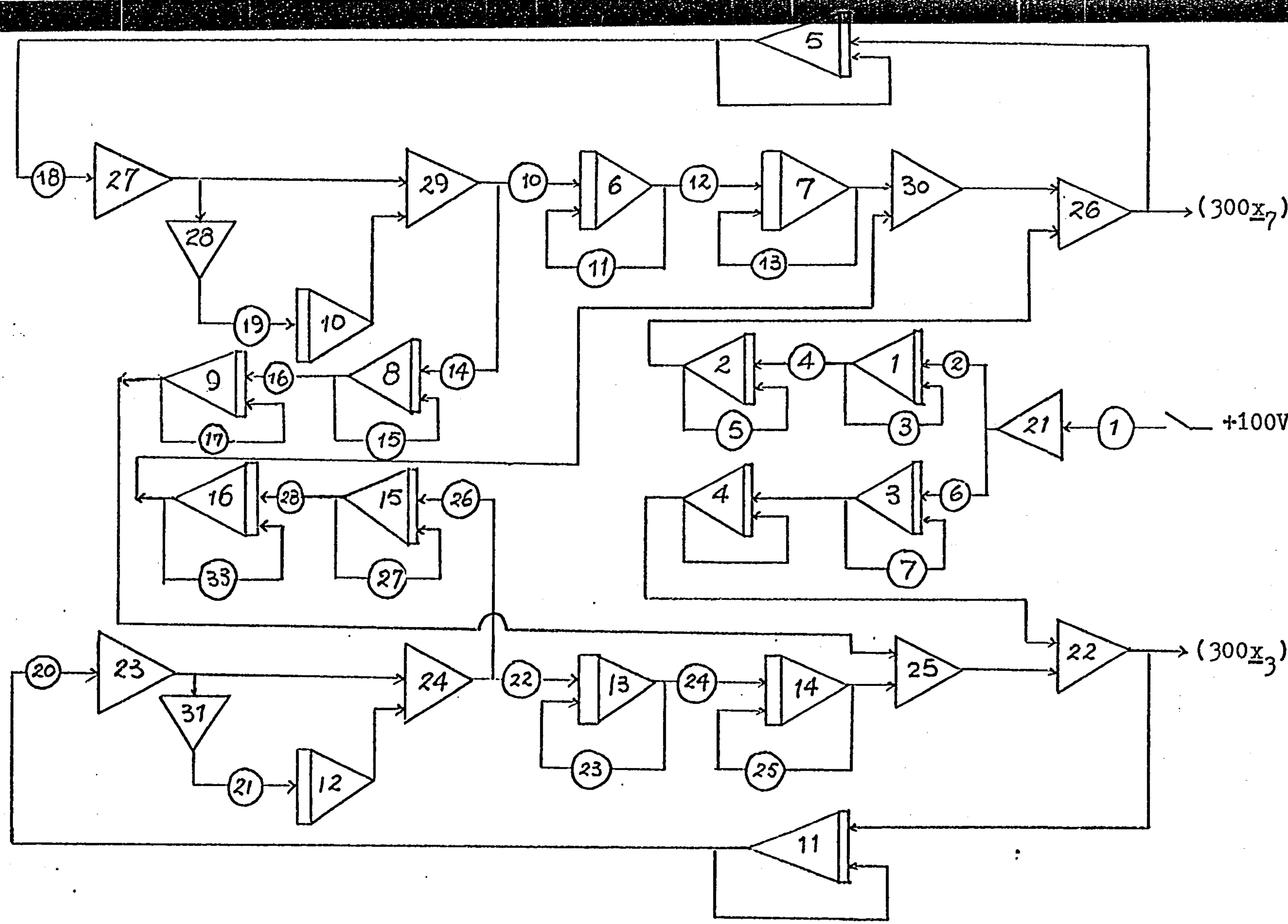


Fig. 54 Analog computer circuit for case 7C (two-point control)

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
6C	01	1/10	0.1000	21	1	0.100	+100V
"	02	5.01/6.6	0.7590	1	1	0.759	
"	03	1/6.6	0.1515	1	1	0.152	
"	04	1/0.6	1.6666	2	5	0.333	
"	05	1/0.6	1.6666	2	5	0.333	
"	06	5.82/7.9	0.7367	3	1	0.737	
"	07	1/7.9	0.1270	3	1	0.127	
"	08	1/0.1	10.0000	4	10	1.000	
"	09	1/0.1	10.0000	4	10	1.000	
"	10	3.156/8.0	0.3945	6	1	0.395	
"	11	1/8.0	0.125	6	1	0.125	
"	12	1/0.46	2.1739	7	5	0.435	
"	13	1/0.46	2.1739	7	5	0.435	
"	14	3.525/8.25	0.4273	8	1	0.427	
"	15	1/8.25	0.1212	8	1	0.121	
"	16	1/0.65	1.53846	9	5	0.308	
"	17	1/0.65	1.53846	9	5	0.308	
"	18	(2)(2)/3	1.3333	27	5	0.267	
"	19	0.6	0.6000	10	1	0.600	
"	20*	(2)(2.1)/3	1.4000	23	5	0.280	
"	21*	0.1	0.1000	12	1	0.100	
"	22	7.55/7.56	0.9987	13	5	0.199	
"	23	1/7.56	0.1323	13	1	0.132	
"	24	1/1.14	0.8777	14	1	0.877	
"	25	1/1.14	0.8777	14	1	0.877	

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
6C	26	4.95/7.95	0.6225	15	1	0.623	
"	27	1/7.95	0.1257	15	1	0.126	
"	28	1/0.6	1.6667	16	5	0.333	
"	33	1/0.6	1.6667	16	5	0.333	

* Various values of Pot No. 20 and 21 were used for different settings of the controller in the lower loop.

POTENTIOMETER SHEET

Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
7C*	35	5.82/0.1(10)	5.820	31	10	0.582	
"	36	1/0.1	10.000	31	10	1.000	
"	37	1/7.9	0.1267	30	1	0.127	
"	38	1/7.9	0.1267	30	1	0.127	
"	05	7.56/7.56	1.000	06	1	1.000	
"	06	1/7.56	0.1320	06	1	0.132	
"	07	1/1.14	0.8777	22	1	0.877	
"	08	1/1.14	0.8777	22	1	0.877	
"	15	$1/T_R$	Various				
"	16	$K_C(2/3)$	Various				

* The analog computer circuit is the same as for case 1A (Fig. 52) except that two more pots were added to A22 (Pot No. 07 and 08).

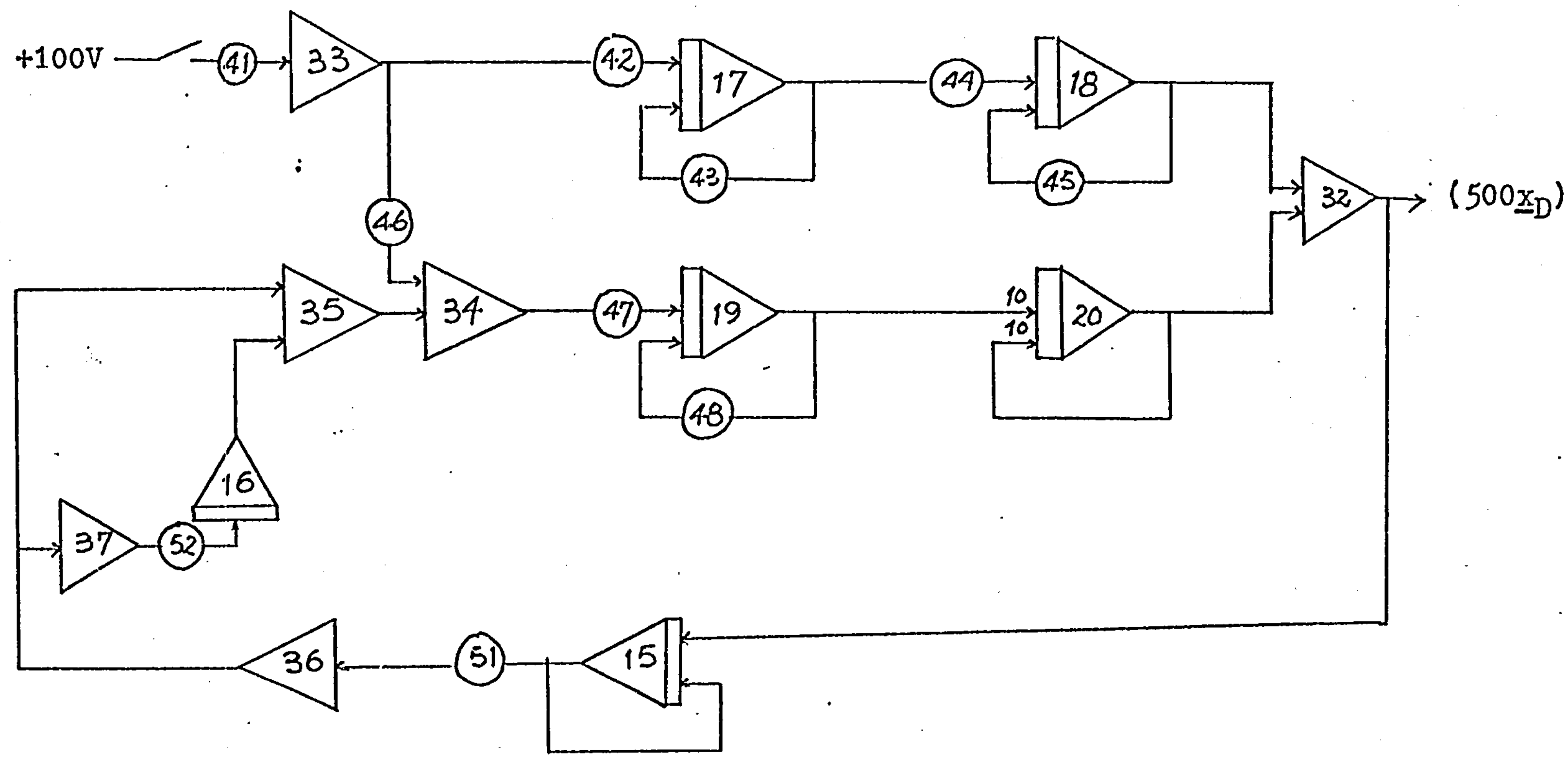


Fig. 55 Analog computer circuit for case 8D (instantaneous feedforward plus feedback compensation control)

POTENTIOMETER SHEET

ase	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
BD	41	1/10	0.1000	33	1	0.100	+100V
"	42	2.9/7.55	0.3841	17	1	0.3841	
"	43	1/7.55	0.1325	17	1	0.1325	
"	44	1/0.15	6.6666	18	10	0.666	
"	45	1/0.15	6.6666	18	10	0.666	
"	46	2.9/3.69	0.7850	34	1	0.785	
"	47	3.69/8.2	0.4500	19	1	0.450	
"	48	1/8.2	0.1222	19	1	0.122	
"	49	1/0.1	10.0000	20	10	1.000	
"	50	1/0.1	10.0000	20	10	1.000	
"	51	$K_C/5=15/5$	3.0000	36	5	0.600	
"	52	$1/T_R = 0.6$	0.6000	16	1	0.600	

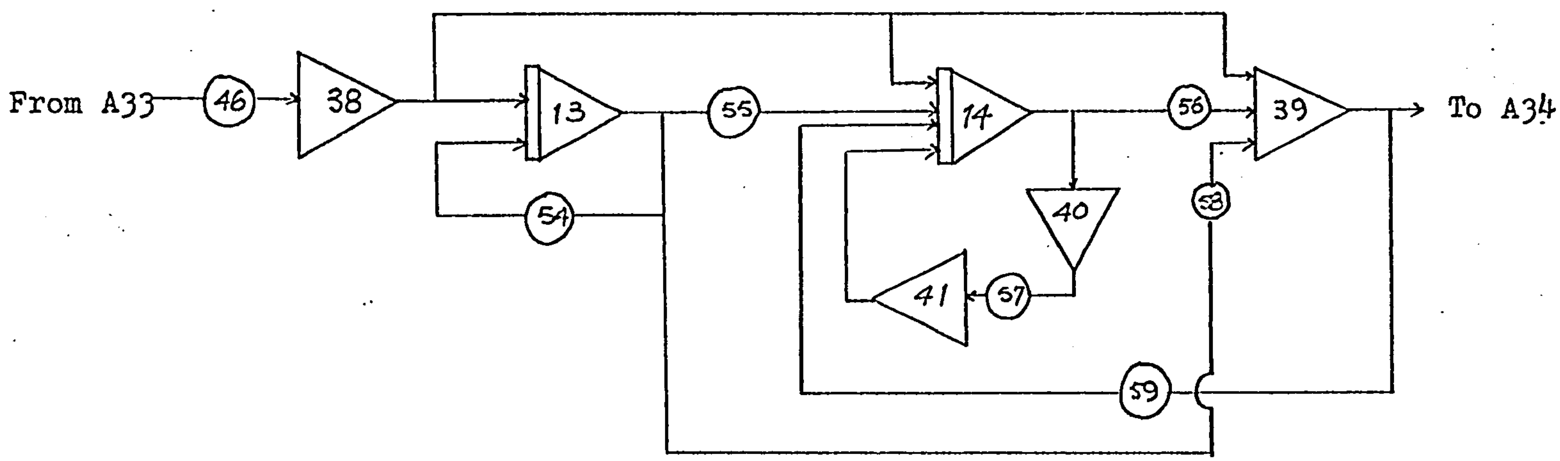


Fig. 56 Analog computer circuit for the "perfect" feedforward functional G_F

POTENTIOMETER SHEET

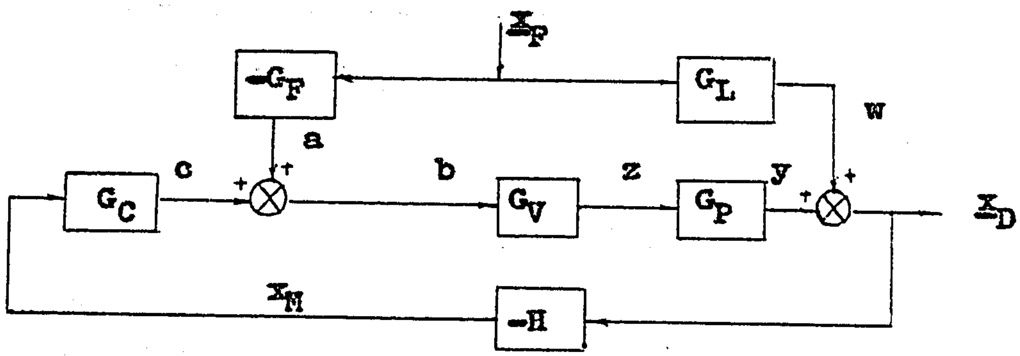
Case	Pot No.	Coeff.	Value	Amplifier No.	Amplifier Gain	Pot setting	Ref.
9D	41	1/10	0.1000	33	1	0.100	+100V
"	42	2.9/7.55	0.3841	17	1	0.384	
"	43	1/7.55	0.1325	17	1	0.1325	
"	44	1/0.15	6.6666	18	10	0.666	
"	45	1/0.15	6.666	18	10	0.666	
"	46*		0.5651	38	1	0.5651	
"	47	3.69/8.2	0.4500	19	1	0.4500	
"	48	1/8.2	0.1222	19	1	0.122	
"	49	1/0.1	10.0000	20	10	1.000	
"	50	1/0.1	10.0000	20	10	1.000	
"	51	15/5	3.0000	36	5	0.600	
"	52	0.6	0.6000	16	1	0.600	
"	54*		0.1325	13	1	0.1325	
"	55*		0.0105	14	1	0.0105	
"	56*		3.3333	39	10	0.333	
"	57*		2.6666	41	10	0.2667	
"	58*		0.0105	39	1	0.0105	
"	59*		2.0000	14	10	0.2000	

* See reference 45, page 81.

APPENDIX 3

Synthesis of the "perfect" feedforward functional G_F

Redraw Figure 11 as



From it, we may obtain

$$\underline{X}_D = w + y$$
$$= G_L\underline{X}_F + zG_P$$
$$= G_L\underline{X}_F + G_P G_V b$$
$$= G_L\underline{X}_F + G_P G_V(a + c)$$
$$= G_L\underline{X}_F + G_P G_V(-G_F\underline{X}_F + G_C X_M)$$
$$= G_L\underline{X}_F + G_P G_V(-G_F\underline{X}_F + G_C(-H)\underline{X}_D)$$
$$\underline{X}_D/\underline{X}_F = (G_L - G_P G_V G_F)/(1 + G_C G_P G_V H)$$

The same result can also be obtained from Figure 12. Therefore, Figure 12 is equivalent to Figure 11.

A "perfect" feedforward functional G_F should make $G_L - G_P G_V G_F \doteq 0$, therefore

$$G_F = (K_L/K_V K_P)\left[(T_{P1}s + 1)(T_{P2}s + 1)/(T_{L1}s + 1)(T_{L2}s + 1)\right]$$

TABLE 14

PARAMETERS OF THE "PERFECT" G_F

K_L	0.580
K_P	0.369
K_V	2.000
K_F	0.785
T_{P1} (min.)	8.200
T_{P2} (min.)	0.100
T_{L1} (min.)	7.550
T_{L2} (min.)	0.150

NOMENCLATURE

B - bottom product withdrawal rate, lb-moles/min.

D - overhead product withdrawal rate, lb-moles/min.

E - number of enriching stages

F - feed rate, lb-moles/min.

G - transfer function

G_C - transfer function of the controller

G_F - feedforward functional

G_L - transfer function between the load variable and the controlled variable

G_P - transfer function of the process

G_V - transfer function of the control valve

H - transfer function of the measuring element

H_B - enthalpy of the bottom product

H_D - enthalpy of the overhead product

H_F - enthalpy of the feed

I - performance criterion function

K - gain

K_C - controller gain

K_F - gain of G_F

K_L - gain of G_L

K_P - process gain

K_V - control valve gain

L - magnitude of the step change in x_F

P - pressure

Q_C - condenser heat duty

Q_L - total heat lost

Q_R - reboiler heat duty

R - reflux rate, lb-moles/min.

$\underline{R}$ - perturbation quantity of R

$(R)_{ss}$ - steady state value of R

S - number of stripping stages

T - time constant (min.)

T_R - integral time (min.)

d - differentiation operator

e - error function

j - plate number

n - total number of plates

s - Laplace transformation operator

t - time

x - mole fraction of acetone

x_B - bottom product composition

x_D - overhead product composition

x_F - feed composition

x_j - jth plate composition

$\underline{x}$ - perturbation quantity of x

$(\underline{x})_f$ - final deviation of x from the original steady state value

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