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DYNAMIC SIMULATION OF AN EXTRACTIVE DISTILLATION COLUMN

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by

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ABSTRACT

A mathematical model, based on heat transfer, mass transfer and fluid dynamics principles, has been developed to represent the dynamics of an extractive distillation column. The column simulation subroutine was incorporated into the recent DYNamic System Simulator executive package (DYNSYS 2.1), created at McMaster University, which uses Gear's method of integration.

Simulation runs were performed with step and pulse disturbances in the feed composition, feed flowrate and reflux ratio variables. Calculated results were compared with available experimental data obtained from a laboratory-scale (2 inch diameter), fifteen-stage extractive distillation column which used distilled water as the solvent to separate a binary mixture of acetone-methanol. Fairly good consistency was found within the normal operating conditions of the column.

A preliminary simulation study of the dynamics of the column (overhead composition transient behaviour) subject to controlled-cycling of the feed flowrate showed a substantial improvement in the separation efficiency of the column.

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TABLE OF CONTENTS

INDEX OF FIGURES

PAGE

INDEX OF TABLES

CHAPTER 1

INTRODUCTION

Because of economic pressure, any industrial company is forced towards greater efficiency and productivity. In many areas, processes can be improved resulting in a favourable return on investment. One area for potential improvement is in process control.

Advances have been slow in industry because of the lack of mathematical techniques to analyze complex control systems and also because of the costly, sometimes impossible, implementation of advanced control schemes.

With the introduction of digital computers into the chemical process industries, more efficient control strategies have been investigated by major chemical companies. However, the approach, if it is to be widely used, has to be simple and fundamental in nature for practical application.

1.1 Computer control of distillation columns

Distillation has been one of the particular processes which attracted the many industrial control designers as well as university research control groups. Distillation can be simply understood as the process of vaporizing then condensing a mixture of components. Precise control of distillation is difficult to achieve because:

(1) Factors affecting separation are not readily interpreted in terms of control system requirements.

(2) Distillation processes are known to interact, sometimes severely.

(3) Measurements used for control may be non-existent or difficult to incorporate into the control strategy.

(4) Distillation units are usually the last in a chain of unit operations, therefore subject to upstream upsets.

(5) Towers with many trays respond very slowly, thus conventional negative feedback control does not seem to be the most efficient solution.

Poor control of a distillation column reduces the throughput capacity, makes the recovery of product difficult and results in excessive utility consumption.

The following sections will present a very practical approach to distillation control resulting from the study of Bojnowski (6).

1.1.1 Relative gain concept

Distillation is a multivariable, highly interactive process. From the control standpoint, a multivariable process is known to have a number of input variables called manipulators such as reflux rate, steam rate or bottom rate, and a number of output variables called controlled variables such as distillate product composition, column pressure or bottom composition.

In such a complex process, in order to achieve low operating cost, good operability and high product quality, the question should be "what variables do I want to control?". The answer is not so evident, the reason is that the controllers interact. Therefore, when looking at a multivariable process, often the question is asked: "If I change a manipulator, what essential will change in the system?". Obviously, the problem is to determine the best pairing of controlled and manipulated variables. The expert control engineer can often choose the right pairs by experience, but sometimes intuitive solutions can lead to dangerous consequences.

The technique of relative gain is commonly used to find the best pairing of variables. The approach bases the selection on a steady-state analysis which is probably the only drawback, since results must be checked against dynamic considerations. As said by Bojnowski (6), for most systems, the results are consistent with dynamic considerations.

The relative gain is defined as:

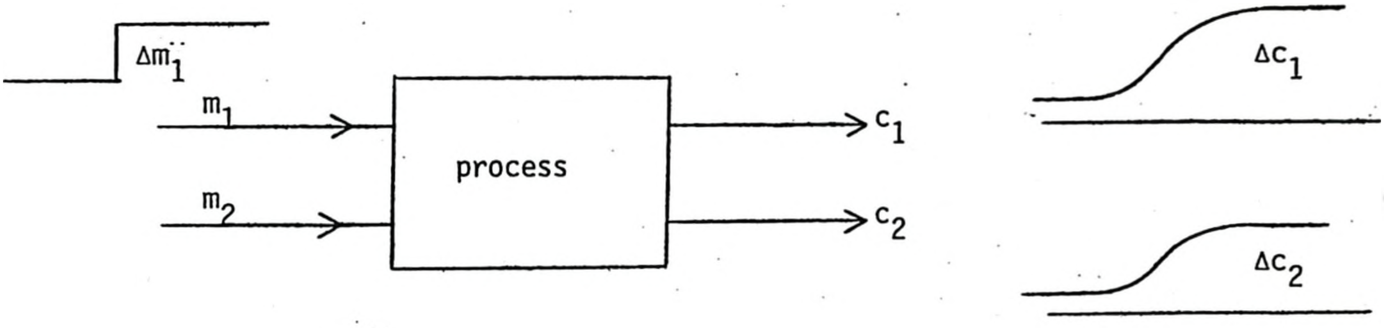
$$\lambda_{ij} = \frac{\partial c_i/\partial m_j|_m}{\partial c_i/\partial m_j|_c} = \frac{\text{open loop gain}}{\text{closed loop gain}} \tag{1.1}$$

Physical meaning for a system of 2 × 2 can be found on fig.1.1. A change is forced in m_j by keeping all other m's constant, then the change in c_i is measured. This is the open-loop gain. The same change is applied to m_j, but all the c's, except c_i, are kept constant by closed-loop control; the change in c_i is observed then the closed-loop gain results.

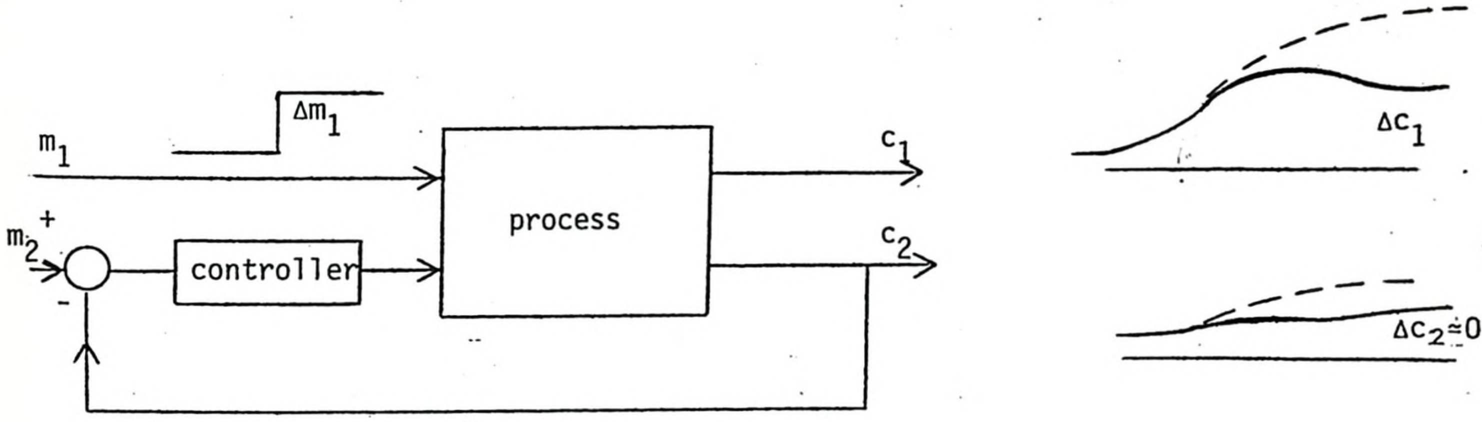
Therefore, for a multivariable process, the relative gain matrix can be obtained:

$$\begin{array}{c} \\ c_1 \\ c_2 \\ c_i \end{array} \begin{array}{c} \begin{array}{ccc} m_1 & m_2 & \quad m_j \end{array} \\ \begin{bmatrix} \lambda_{11} & \lambda_{12} & \\ \lambda_{21} & \lambda_{22} & \\ & & \lambda_{ij} \end{bmatrix} \end{array} \tag{1.2}$$

Fig. 1-1 PHYSICAL MEANING OF THE RELATIVE GAIN IN THE MULTIVARIATE PROCESS



$$\left.\frac{\partial c_1}{\partial m_1}\right|_{m_2} = \left.\frac{\Delta c_1}{\Delta m_1}\right|_{m_2}$$ = open loop gain



$$\left.\frac{\partial c_1}{\partial m_1}\right|_{c_2} = \left.\frac{\Delta c_1}{\Delta m_1}\right|_{c_2}$$ = closed loop gain

One property of the matrix is that the algebraic sum of any row or column is one. One must note that when λ_{ij} is one, m_j controls c_i and no interaction exists.

One does not have to perform both open-loop and closed-loop experiments. Obtaining the open-loop gains is sufficient since matrix manipulation will give the whole relative gain matrix. Techniques and examples are described in reference 6.

Applications can be based on the following rules:

(1) The pairs with the largest positive numbers should be selected for closed-loop control.

(2) Relative gain values approaching 0.5 indicate strong mutual coupling and will be extremely difficult to control.

(3) Negative numbers indicate a divergent control system, and if these pairs are chosen, control is impossible.

(4) One operating condition for which the c-m pairs were chosen may not be valid for another operating condition.

1.1.2 Factors affecting distillation performance

The control strategy should keep a column functioning as designed, despite various upsets. Two very important facts must be retained for control design objectives:

(1) the composition of the distillate and bottoms streams are intimately affected by the product flowrates. If 100 lbs. of a component is fed to a column, 100 lbs. must be withdrawn. If the 100 lbs. is not withdrawn in the correct amount in each product stream, the desired composition cannot

be obtained even though the column is capable of making the separation. This is called <u>the material balance problem</u>.

(2) Energy is required for separation. If there is not enough energy available to carry out a separation, even though the material flows are correct, the desired compositions will not be obtained. This is known as <u>the separation problem</u>.

Usually, control of the material balance is done with flow variables, and control of separation is achieved with reflux rate variable or feed point location. Unfortunately, in most cases, these two control loops severely interact. The problem is then to decouple the interaction by choosing the right pairs of controlled and manipulated variables.

Applying the relative gain technique and checking the performance on many industrial columns, Bojnowski mentioned many general rules for control of distillation:

(1) When only one product quality is to be controlled closely, the material balance must be manipulated.

(2) The lesser of two product streams should be manipulated to control product quality.

(3) The material balance should be manipulated to control the quality of the less pure product, and the separation should be manipulated to control the quality of the more pure product.

(4) The pressure must be tightly controlled for two reasons. If the pressure is increased, the liquid on the trays and in the reboiler is below its boiling point; the vapor is condensed to heat the liquid to its

boiling point; once at its boiling point, the vapor rate snaps back causing an increase in pressure. This causes column instability. Furthermore, if the column pressure is not held constant, measurements for control purposes are invalid. Suggested control schemes are described in reference 6.

1.1.3 Special features for extractive distillation control

In extractive distillation, the solvent is fed near the top of the column and entrains one or more heavy components on the way down the column. Light components concentrate up and are removed near the top. Heavy components and the solvent are withdrawn at the base and sent to a recovery column to recover the solvent which is recycled back to the extractive column.

When dealing with extractive distillation, one must take into account the following facts:

(1) Since the solvent is an extractant, the correct ratio between feed and solvent must be maintained. Insufficient solvent will cause poor separation while excessive solvent will result in the dilution of light components.

(2) Reflux greatly influences the separation. With higher reflux, better separation occurs, but excessive reflux dilutes the solvent thus causing poor separation. Thus a designer must try to control the solvent concentration at the solvent addition tray.

(3) Usually, extractive distillation is most sensitive to the solvent composition and temperature. Since the heat content of the solvent is a large percentage of total heat requirements, control of solvent temperature

is critical.

1.1.4 Conclusion

In summary, many methods of control other than feedback control have been used in distillation processes with certain special features. One example which can be mentioned is that of ultra high reflux and high reflux ratio separations; because of the extremely long response times, bang-bang control is the recommended control strategy since it is inexpensive, it gives extremely tight control, it gives fast responses to disturbances, it eliminates non-linearity problems such as sticking control valves and it eliminates tuning problems.

Feedforward control is usually applicable where:

(1) one column is noticeably limiting the capacity of the entire unit.

(2) a small improvement in recovery efficiency on a large refining column can mean substantial additional production.

(3) feed rate and composition changes upset a column early in a distillation train and cause problems in the rest of the operation.

(4) safety considerations demand that compositions stay within certain limits.

A fairly detailed presentation of different control schemes along with discussions can be found in reference 6.

1.2 Project objectives

In general terms, the present project is actually based on a fifteen stage column with thirteen trays and condenser and reboiler systems. Two feed streams enter the column: water, used as an extractant, enters near the top of the column, and a mixture of acetone-methanol is fed near the bottom. Simple binary distillation to separate acetone from methanol is not possible since the mixture forms an azeotrope of .8 mole fraction acetone at atmospheric pressure. As mentioned by D.O'Neill (10), at approximately 0.3 mole fraction water, the azeotrope is completely broken.

1.2.1 Long term objective on the equipment

A complete extractive distillation process must include two columns in series. The first one, usually called the extractive column, extracts the light component, in this case acetone. The second one, often called solvent recovery column, recovers the solvent which is recycled back to the extractive column.

The long term objectives will consist of building the second column, and studying some advanced control configurations on the complete process and checking the performance of each control scheme.

However, before doing any investigation of this kind on such a complex process, one must do more specific work to obtain the minimum requirements, e.g. to find out suitable models at different levels of sophistication, to achieve some simple control of input stream variables such as temperature or flowrate, and to understand the dynamic behavior of the process.

1.2.2 Summary of previous work

Jackson (11) undertook the initial step of this project. His work consisted of building the fifteen stage extractive column, evaluating its steady-state performance, interfacing basic column instrumentation to the mini-computer and developing a basic real time software package mainly for data logging and input stream temperature control. The software was made using the original version of the Real Time Operating System. Major operating problems occurred, due to tray priming at the solvent recovery section and tray weeping near to the bottom of the column.

D.O'Neill (10) rebuilt the column completely and evaluated its operating range as well as its steady-state performance. A steady-state simulation was also developed to better understand the process. A preliminary study on the dynamic behavior was made by disturbing input variables, such as feed composition and feed rate, by conventional step change and pulse change. Transient responses of overhead and bottom compositions were analyzed and satisfactorily fitted by first order transfer models with dead time.

1.2.3 Present objectives

The agreed long term objective was to investigate different control strategies to obtain the best separation performance. Digital computers become more and more an important tool for process designers. Industrial practice tends to use digital simulation to estimate roughly the performance of operations, rather than carrying out experiments on pilot plants. This, results in substantial time and research cost savings. Therefore, before investigating experimentally a control strategy, it is always economic to obtain a rough result by simulating this strategy on a computer. Obviously, the next step will be to develop a suitable mathematical model which can dynamically represent the extractive column since one needs process models to do any simulation work.

As far as the dynamic study is concerned, input disturbances must be achieved manually. With this technique, the study of oscillatory disturbances or random disturbances is difficult. Therefore, one of the main experimental steps will be to connect control valves into the two feedstreams. Consequently, input flowrates can be controlled automatically by the computer or can be disturbed, as programmed, by the computer.

A summary of the present objectives is:

(1) To develop a dynamic model representing the extractive distillation column

(2) To compare calculated step and pulse responses with available experimental data

(3) To perform a rough study with oscillatory disturbances, given the flexibility offered by the installation of the control valves.

CHAPTER 2

PROCESS EQUIPMENT AND FACILITIES

As briefly mentioned in chapter 1, section 1.2, the column was totally rebuilt by D.O'Neill with the aim of eliminating liquid priming in the solvent recovery section, due to tray weeping and also due to technical problems caused by the vibrostaltic feed pumps during the operation.

The present chapter will give a brief description of the process equipment followed by the specific modifications made to improve the flexibility of the column for the dynamic study. Control software alterations will also be described as well as the specific possibilities of the modified process arrangement.

2.1 Previous apparatus

The complete details of the apparatus could be found in reference 10 chapter 4. A general description of the experimental equipment will be presented in this section.

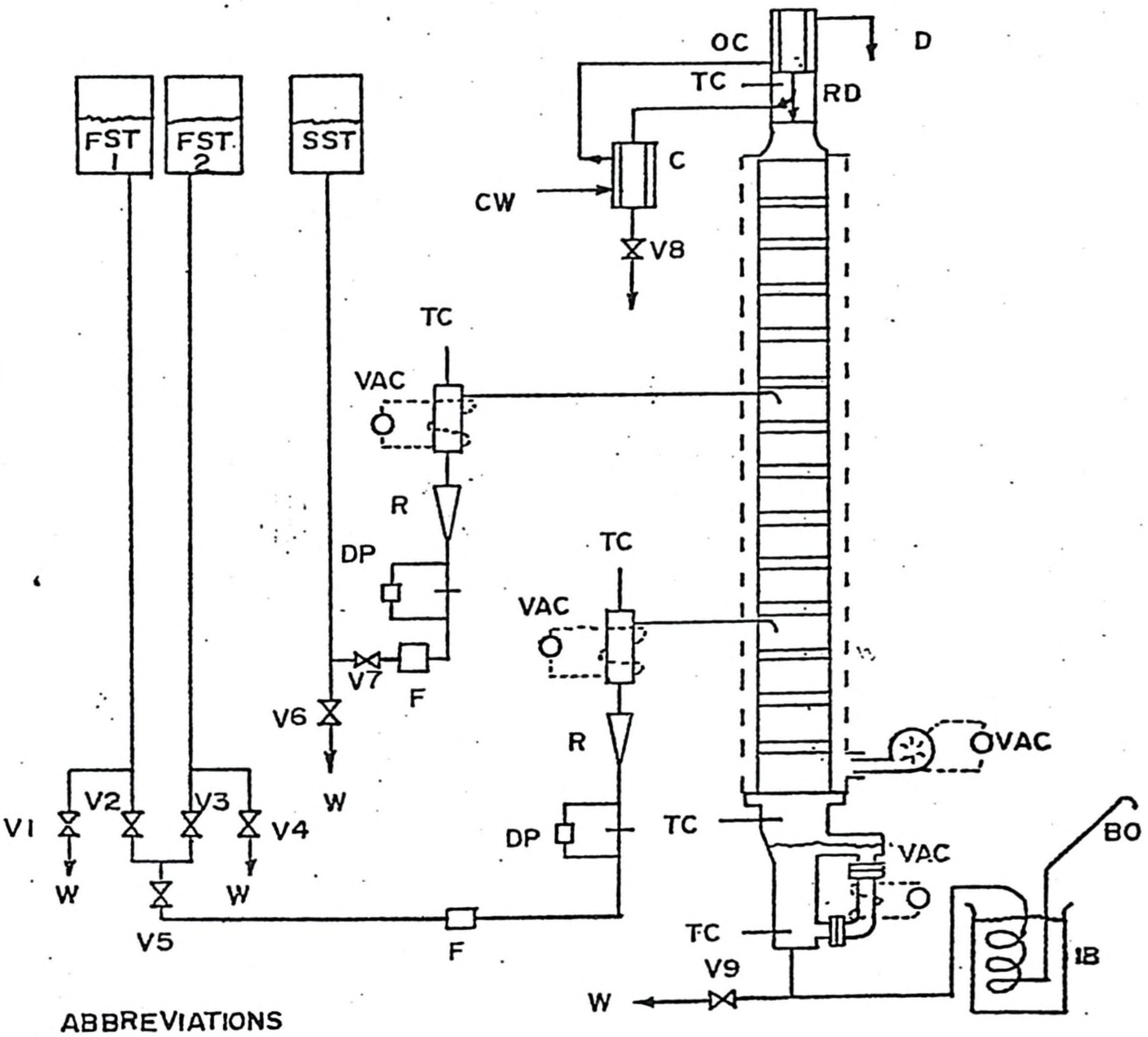
2.1.1 Description

Fig. 2.1 shows the arrangement of the extractive distillation process (10) before the modifications.

Two storage tanks of 10 liters each, made of glass and located on the second floor of the chemical engineering laboratory, were used to supply the feed, at two different proportions of acetone-methanol, to the column through 0.63 cm O.D. stainless steel tubing. A system of on/off valves (V1, V2, V3, V4) allowed the sampling of the feed stream and the switching from one storage tank to the other. The tanks were in contact with the ambient atmosphere, therefore the feed just flowed down by gravity. A needle valve (V5) was used to adjust the flowrate manually. A filter prevented the penetration of foreign material to the glass orifice which was used in conjunction with a Dynascience Corporation differential pressure cell to measure the flowrate; the small signal from the DP cell was sent to a Dynascience Corporation Carrier-Demodulator which converted it into a 0-10 volts DC signal used for control purposes. A F α P precision flowrator indicated the current flowrate. Before entering the column, the feed was electrically heated by a preheater made from a 30.5 cm × 1.27 cm O.D. tube wound with resistance wire. A thermocouple measured the outlet temperature from the preheater for control purposes, and a powerstat transformer could be used to vary the power input to the preheater. The feed line was connected to tray 11 from the top.

The same arrangement was used for the solvent stream, except that only one 10 liter storage tank was present. The tubing from the tank to the preheater input was in Tygon.

The column was made of thirteen sieve plates, a condenser system and a reboiler system. The sieve plates were separated by 15.25 cm × 5.08 cm I.D. Q.V.F. glass pipes. The weir limited the holdup of liquid on a tray; the overflow flowed down the downcomer to the tray below. Each sieve plate fig. 2.1 FLOW DIAGRAM - EXTRACTIVE DISTILLATION COLUMN



ABBREVIATIONS

BO	bottom overflow	OC	overhead condenser
C	cooler	R	rotameter
CW	cooling water	RD	reflux divider
D	drain	SST	solvent storage tank
DP	differential pressure	TC	thermocouple
F	filter	V	valve
FST	feed storage tank	VAC	variable AC power
IB	ice bath	W	waste

was 0.16 cm thick and had 93 holes of 0.89 cm diameter; a feed port, vapor sampler, a thermocouple port and a liquid sampler were provided. A hot jacket surrounding the column to prevent heat loss to the ambient atmosphere was made from two sections of 13.34 cm I.D. perspex plastic pipe. A hot air gun blew the heated air upward in the jacket.

A stainless steel reboiler of 327 ml capacity was used with an overflow arrangement as shown in fig. 2.1; the liquid flowed by gravity or column pressure through the ice bath and went out by the tube exit which was maintained at a certain height that one can adjust to obtain the desired reboiler liquid level. The holdup of liquid from the output of the reboiler to the tube exit was 25 ml.

The condenser was equipped with a water-cooled system and a reflux divider electrically activated by a timing device made by Eagle Signal Corporation.

A more detailed description could be found in the work of Jackson (11) and of O'Neill (10).

2.1.2 Problems in operation and restrictions on dynamic study

The above equipment presented some problems encountered at different stages of the operation.

(1) Before doing any dynamic investigation on the process, one must bring the column to a steady-state level. From that point, perturbations can be put into the desired input variables. A practical question would be "when is the column at steady-state?"

Until the present investigation, temperature readings at the stages and overhead flowrate measurements were used to determine experimentally whether the column is at steady state or not. Very often, when the samples were analyzed later on, it was found that a trend still existed as far as the composition is concerned, showing lack of steady-state reaching. If one wants to study the dynamic of the column due to a certain kind of input disturbance, this fact becomes crucial, since the response depends intimately upon the trend existing at the initial time of perturbation.

(2) It was noticed that during a run, it could happen that the air gun which blows heated air through the jacket, stopped for a significant period of time. This will undoubtedly cause an upset in the negligible heat loss assumption made in the simulation work.

Since each run usually requires at least 10 hours, the presence of electrical devices is very undesirable in the sense that the whole time for the experiment could be wasted if an electrical problem occurs at the last period of the planned run. The feed stream preheater, solvent stream preheater and reboiler heater are locations where electrical devices have been used.

(3) The needle valves on the feed and solvent streams were used to adjust the flowrates to the desired valves. When a flowrate, as measured by the DP cell, exceeded the tolerance limits, the computer sent a signal to the process site and the adjustment had to be carried out manually by checking the readings on the flowrators.

This technique of manual control is very inaccurate. From the dynamic point of view, since the operating range for the flowrates are small, the possible perturbation amplitude is relatively small, therefore, this technique of controlling or varying the flowrate would not be recommended.

(4) Furthermore, oscillatory disturbances like sine wave, square wave or random disturbance, cannot be used as long as the computer is not arranged to do such perturbation. This is due to the inaccuracy of the manual adjustment and lack of time available for it since one must take samples quite often during a dynamic run.

Given the problems pointed out in the last two parts, some modifications were made in order to assist the investigator in obtaining better results as well as to give more versatility to the experimental dynamic work. The two next sections will present modifications on the equipment and on the computer control software.

2.2 Modifications to apparatus

2.2.1 Feed and solvent system

The needle valves V5 on the feed line and V7 on the solvent line were replaced by two control valves, 1/4 in. stainless steel body, trim 0, air to open type with neoprene diaphragm, manufactured by Badger Meter Inc. Fisher electro-pneumatic transducers, type 546, were used to receive a 1-9 VDC input signal from the D/A converter and to convert it into a pressure from 3-15 psi on the control valve diaphragm. An air line at 25 psig supplied the input to the transducers.

One problem occurred when replacing the needle valves by the control valves. The liquid coming from the storage tanks located about 15 feet above, flowed down just by gravity, and it was found that there was not enough head to push the liquid through the control valve orifice at the required rate even if it was fully opened. Two solutions were possible:

(1) One could design an electrical pump, and without altering the storage system, introduce it before the control valve location.

(2) One could replace the glass storage tanks by pressurized tanks and apply a pressure to create additional head on the liquid.

After considering the difficulties encountered while using electrical devices and the economic point of view, the second alternative was chosen.

2.2.2 Storage tank system

Three stainless steel storage tanks were purchased from AC Tanks Co. two for the feed line and one for the solvent line. They could handle a maximum pressure of 200 psi and had a maximum capacity of about 25 liters. A nitrogen cylinder was used to exert the necessary pressure. The hardware connections are presented on fig. 2.2.

Such an arrangement implies that the pressure should be the same in the three tanks. Calibration curves for the feed line and the solvent line are presented in Appendix A. A safety valve (SV) was used for safety purposes and for clearing the pressure at the end of each run.

However, some operational difficulties occurred:

(1) One has to set the gas cylinder valve at the pressure where the calibration was obtained, before each run, which is very difficult with the accuracy given by the actual pressure gauge; therefore calibration

must be checked at every run.

(2) Filling the tanks requires a very substantial amount of time and caution since the mixture acetone-methanol is inflammable and usually a large amount of feed material is used. The feeding should be made from the top hole of the tanks. One must be extremely careful in estimating the quantity of material needed for the run since the tanks can be fed only before the run and not during the run.

The next section will present the modifications made to the computer program developed to assist the experimenter when operating the column.

2.3 Software facilities

2.3.1 Previous control package

D. Jackson (11) has developed a computer program to aid in the manual operation of the column. The software was created within a Real Time Operating System (RTOS) provided by the computer manufacturer. Computer-process interfacing allowed the function of input-output hardware devices according to a criterion which minimized the average time devices must wait for servicing.

Many tasks were created to initiate or terminate the operation, to perform parameter changes, to measure and display process temperatures, and input stream flowrates, to control the input stream temperatures according to an on/off scheme and to create a signal at the process site for manual correction of the input stream flowrates to desired values.

The tasks were scheduled for execution by the operating system depending on their respective priority and the availability of the peripheral devices for input-output tasks.

Complete details of the software package as well as listings of programs and algorithms can be found in the work of D. Jackson (11). Since the software has to be entirely modified, no further description of the previous package is given in this work.

2.3.2 Modifications and new facilities

The actual computer system in the Chemical Engineering department is operating under the Real time Disk Operating System (RDOS) supplied by the computer manufacturer. References 23, 24, 25 are useful to understand how to use commands. All tasks developed to assist the operation of the column have been changed to be used with the Generalized Operating System Executive created by P. Tremblay (22). This executive offers a large number of facilities as well as a lot of flexibility for control purpose programs. Reference (22) must be regarded as the most important source of information on the actual control package for the distillation column.

Several programs were created:

(1) The program INFCE interfaces the user program which consists of a number of independent tasks, to the GOSEX.

(2) Tasks FEETC and SOLTC perform the control of the input feed and solvent temperatures by turning on or off the preheaters, and also

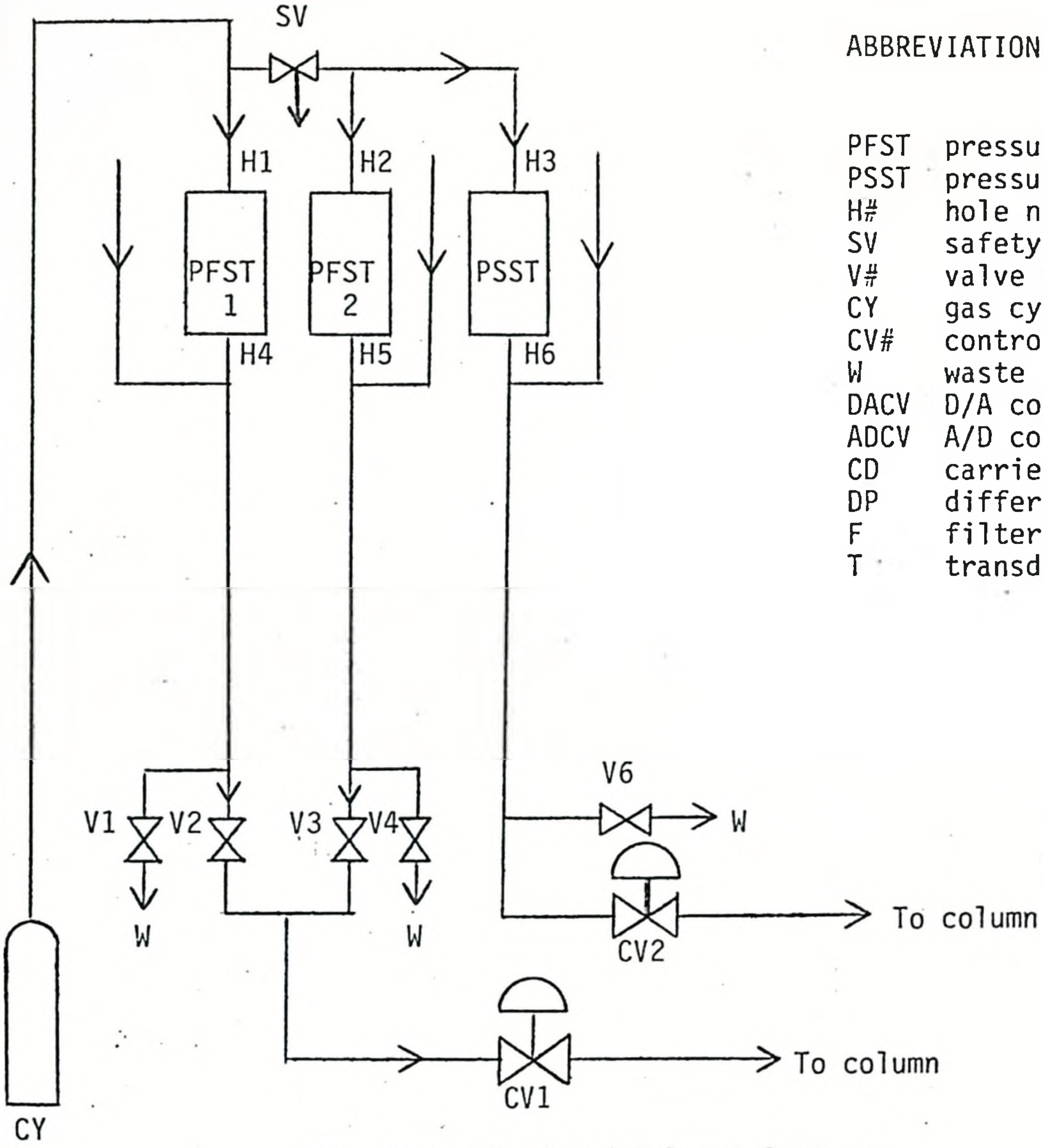
measure up to five temperature locations, convert into engineering units for displaying. There are many possible modes of display. The user should refer to reference (22) to be able to issue correct commands for these display modes.

(3) The introduction of the two control valves in the feed and solvent lines required additional software to actuate the valves via the computer. Two tasks were created, FEEFM and SOLFM, to calculate the control action to the valve based on the flowrate reading from the D.P. cell and based on a PID control algorithm performed by the GOSEX utility routine CONTL. Instructions for using this routine could be found in detail in ref. 22. FEEFM and SOLFM also convert the raw data, i.e. the stream flowrate, and the control action which is the percentage-opening of the valve, into engineering units for display. The simple control loop for the two flowrates on the feed and solvent lines is presented on fig. 2.3.

(4) Task SSAMP was used to acknowledge at the computer site any change occurring at the process site by using the contact senses, for instance the taking of a sample by the operator.

(5) A program to input a square wave to the feed rate variable was also developed. Task FEEFT shifts the feed flowrate set point from a high level (FHL) to a low level (FLL) according to a half period determined by TWAIT. These parameters could be changed on-line by using the GOSEX function PARAM. (see ref. 22)

Algorithms and listings of the programs as well as the loading procedure are presented in Appendix G.



ABBREVIATIONS

PFST	pressurized feed storage tank
PSST	pressurized solvent storage tank
H#	hole number #
SV	safety valve
V#	valve number #
CY	gas cylinder
CV#	control valve number #
W	waste
DACV	D/A converter
ADCV	A/D converter
CD	carrier-demodulator
DP	differential pressure cell
F	filter
T	transducer

fig. 2.2 Feed and Solvent Systems

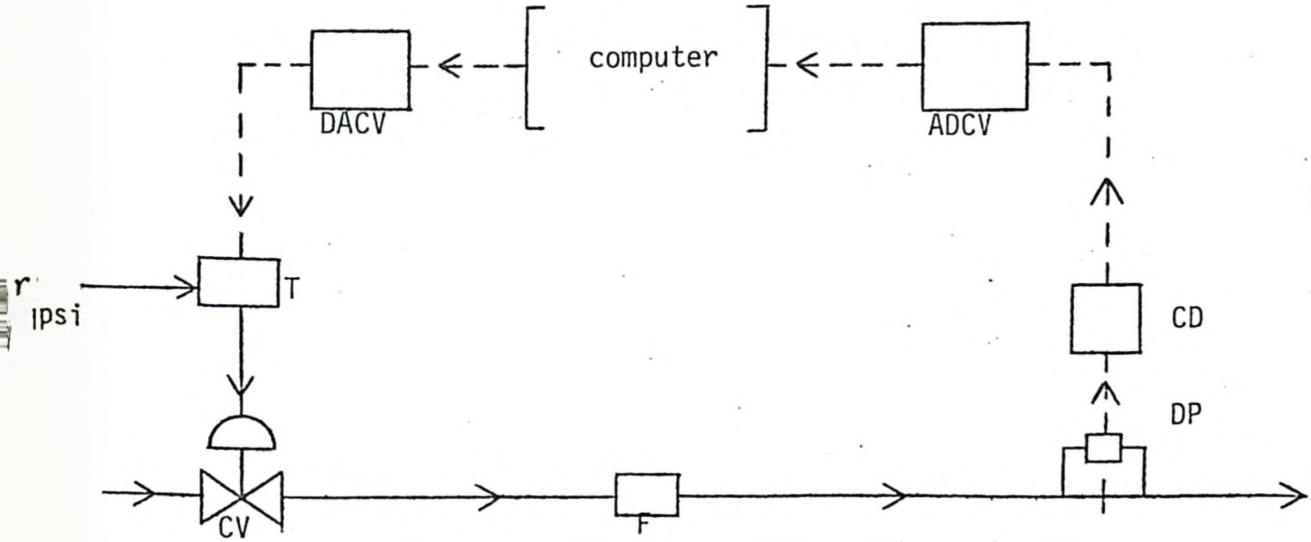


fig. 2.3 Elementary Flowrate Control Loop

2.4 Capabilities of modified apparatus and software

With the present apparatus, the study of the dynamics of the extractive distillation column could be extended to the introduction of oscillatory disturbances such as square wave and sine wave in the feed flowrate variable. The solvent flowrate variable could be investigated in the same way. A good variable to be studied will be the ratio of the feed to the solvent. Random disturbances, whether at two levels or according to a statistical distribution, could also be input into the input stream flowrates. With some modifications on the software, these disturbances could be realized under open-loop or under closed-loop control.

CHAPTER 3

DYNAMIC SIMULATION

As discussed in chapter 1, simulation on a digital computer has become and has been proven to be an efficient and economic tool in the design work of an engineer. Steady state simulation has been used intensively in university research as well as in industrial engineering work, especially to design new equipment, to conceive or improve processes, or to optimize the operating conditions for a better product quality at a minimum cost.

Dynamic simulation finds its importance, for example, in the study of optimal start-up of a process, in the investigation of the effects and the policy to be followed in the case of an equipment failure, and recently, especially in the transient study of process equipment and the design of different control strategies.

Before doing any simulation work, the first step must be to represent the real chemical equipment or process by an adequate mathematical relation. The cost of such model development will obviously increase with the degree of accuracy of the mathematical model; therefore such investment must be related to the objective of the simulation work.

3.1 Literature review on dynamic simulation of distillation

The literature contains many references to the study of the transient behavior of the distillation column which represents a complex and non-linear chemical process.

Marshall and Pigford (7) presented a tentative modelling approach for packed columns. The continuity equation was applied to a differential unit length of the bed to yield differential material balance equations. A linear equilibrium relationship, constant molal holdup per unit length and constant molal overflow were the main assumptions.

Montroll and Newell (8) investigated the exact solution of a set of non-linear differential material balance equations for multistage separation of binary mixtures. On an IBM 701, Jackson and Pigford (9) tried to obtain a numerical solution for the differential material equations by approximating with finite difference equations. Total reflux, a relative volatility equilibrium relationship, a constant Murphree efficiency and a zero condenser hold-up were assumed. Also, the initial liquid compositions on the trays were considered as equal.

Huckaba and Danly (4) have done intensive studies on binary batch distillation which is basically a time dependent process. Besides the differential mass balance equations normally used by previous investigators, differential energy equations were introduced for the first time to take into account the inequality of the liquid flowrates between plates. The relaxation of the equality of flow assumption obviously contributed to a better representation of the real process.

Rosenbrock (5) developed mathematically a method of calculation of the unsteady state behavior of continuous distillation columns. The results for a step change in the feed composition were obtained for binary mixtures, based on total reflux assumption. Several non-linear equilibrium relationships and a Murphree efficiency different from 100% were considered. Rosenbrock's calculated results were compared to the experimental data of Armstrong and Wilkinson (1), and it was found that the method was only acceptable for the initial transient period; fairly substantial divergences in the comparison appeared at later times. However, the use of a digital computer to handle complex and time-consuming calculation was found to be very promising.

An analytical exact solution to a step change in the feed composition was attempted by Voetter (12) using a linear equilibrium relationship. The introduction of the various simpligying assumptions was shown to yield up to 40 percent error in the results.

Acrivos (2) and Katz (3) have extended the dynamic study by introducing randomly fluctuating input conditions in a linear stagewise system with binary mixtures.

Franke (13) has modeled in detail the behaviour of a continuous experimental distillation column using a binary mixture, with a step change in the input feed composition.

The simulation of distillation with more than two components requires a much more powerful method to handle a large number of equations within a reasonable computer time. A typical problem could involve more than 50 components and columns with more than 100 stages.

A new technique of handling differential equations was used in the present study, with a tenary system and a fifteen stage column. Also, a new approach of using the energy balance equations to adjust the liquid and vapor flowrates was investigated.

3.2 General features of the executive package DYNSYS 2.1

In chemical engineering, stiff systems of ordinary differential equations describing the process occur widely in reaction kinetics and in multistage processes. Stiff systems are defined as systems where the time constants differ greatly in magnitude. To be able to study the transient behaviour of such a system, the integration time step size is limited by the smallest time constant. Therefore the prediction of the entire dynamic response of such a system requires a large computer time and cost if conventional numerical techniques are used. An interesting investigation of different existing techniques was presented by Barney (14).

In the past few years, several executive packages have been developed especially for chemical engineers. These are more problem-oriented and use variations of what is known as the modular approach.

In 1970, Brobow, Johnson and Ponton have developed at McMaster University an executive package known as DYNamic SYStems Simulator, using a third order Adams-Moulton-Shell routine to integrate ordinary differential equations. This method proved to be insufficient to handle stiff systems. The integration technique was revised by Barney in 1974 and finally Gear's integration algorithm in conjunction with Key's technique of solving linear algebraic equations was implemented into the new version called DYNSYS 2.1. An option for stiff systems with a tridiagonal Jacobian matrix is also included since this case generally occurs when one deals with counter current stagewise processes.

The principal features of the executive DYNSYS 2.1 used in this project will be outlined in the following sections.

3.2.1 Modularity

A chemical plant is usually made up of processing equipment such as reactors, columns, heat exchangers, etc., with control devices such as analyzers, control valves, etc., connected by material flows and control or measuring signals.

Each equipment item or device can be represented by a mathematical equation or set of equations which form a unit computation or computation module. Then a mathematical model of a plant can be seen as a network of modules in which information will flow in a manner analogous to the material flow or control signals in the real plant. Therefore, the dynamic information flow diagram is very similar to the process flow diagram.

This simulation approach, known as the modular approach, is very convenient and flexible when one wants to study the combinations or configurations of physical equipment or to design control strategies on a digital computer. It is not as efficient computationally as the equation-oriented approach since it involves more information transfer, but it helps the user to visualize his process, and especially it presents a great flexibility and an easy adaptability to different control and operational policies.

As a library of equipment and control modules gradually becomes available, the programming effort for new plant studies is reduced. However, the routines should be made general enough for further use. Another feature is that real variables are used in the modular approach, rather than transformed variables. The modules may be quite non-linear in behavior.

A detailed presentation of the concept can be found in reference 14.

3.2.2 Gear's method of integration

One of the most important features in a dynamic simulation executive is the numerical technique of integration. DYNSYS 2.1 adopted the method developed by C.W. Gear in 1971 to solve ordinary differential equations.

The subroutine is called DIFSUB(30) and uses a variable-order, variable step, linear, predictor-corrector algorithm. To be more flexible in handling a variety of types of chemical processes, the subroutine offers two options, one for stiff systems and one for nonstiff systems.

3.2.2.1 Nonstiff option

When the whole process comprises non-stiff modules, this option must be chosen. Basically, the predictor step is described by the Adams-Bashforth explicit equation:

$$y_{n+1} = y_n + h \sum_{i=1}^{k} \beta_i \, \dot{y}_{n+1-i} \qquad (3.2.a)$$

The corrector step uses the Adams-Moulton equation:

$$y_{n+1} = y_n + h \sum_{i=0}^{k} \beta_i' \, \dot{y}_{n+1-i} \qquad (3.2.b)$$

The coefficients β_i and β_i' may be found in reference 15, and the order may vary from 1 to 7.

3.2.2.2 Stiff option

When the whole process comprises just stiff modules, or both stiff and non-stiff modules, this option must be choosen. Obviously, the degree of stiffness is a matter for the judgement of the investigator.

The following approximations are used:

for the predictor pass:

$$y_{n+1} = h\ \beta_1 \dot{y}_n + \sum_{i=1}^{k} \alpha_i\ y_{n+1-i} \qquad (3.2.c)$$

for the corrector pass:

$$y_{n+1} = h'\beta_0^* \ \dot{y}_{n+1} + \sum_{i=1}^{k} \alpha_i^*\ y_{n+1-i} \qquad (3.2.d)$$

The coefficients in the predictor equation can be found in reference 15. Values of the corrector coefficients are given in reference 16. The order may vary from 1 to 6. The corrector equation has been found stable for stiff equations for order up to 6. Further details concerning accuracy, stability and stiffness coefficient determination are discussed in reference 14.

Since the corrector equation is implicit, a method of iteration is required for convergence of the solutions. Two methods of iteration are available for the stiff option: the direct Jacobi iteration is used for non-stiff modules in a stiff process, and the Newton-Raphson iteration is used for stiff modules. In the latter case, the Jacobian matrix is needed and may be user-supplied or evaluated numerically. Descriptions of the two iterative techniques can be found in reference 14. The application to the distillation column simulation is presented in section 3.3.

3.2.3 Algebraic equation solver

Since the corrector equation is implicit, a set of algebraiec linear equations will be obtained. An important factor in saving computer time cost is the choice of the method to solve these simultaneous equations, since one deals generally with large systems of ordinary differential equations in a plant simulation.

The coefficient matrix of the linear equations was shown to be a function of the Jacobian matrix for the system. In general, in a countercurrent stagewise process such as the distillation column, the Jacobian matrix has many zeros. Such a matrix is called a sparse matrix, and using a technique which just operates on the non-zero elements will reduce greatly the computer time used.

The routine called SIMULT developed by Key (28) in 1973 was adapted and implemented in DYNSYS 2.1. Seven pivoting options are available, but, the option of minimum row-minimum column was chosen for this project: the coefficient matrix is searched to find the row with the least number of non-zero coefficients; then for the non-zero elements in this row, the column with the most zeros is chosen as the pivotal column. If more than one row or column are chosen, the row or column with the smallest index is used.

The storage scheme is illustrated by the following example: let the matrix:

$$\underline{P} = \begin{bmatrix} 0 & 1 & 2 & 0 \\ 0 & 0 & 1 & 1 \\ 0 & 0 & 0 & 1 \\ 1 & 0 & 0 & 3 \end{bmatrix}$$

P is stored as:

$$\underline{A} = \begin{bmatrix} 1 & 2 \\ 1 & 1 \\ 0 & 1 \\ 1 & 3 \end{bmatrix} \qquad \underline{ICOL} = \begin{bmatrix} 2 & 3 \\ 3 & 4 \\ 0 & 5 \\ 1 & 4 \end{bmatrix}$$

One must refer to reference 14 for more details and explanations.

3.2.4 Adaptation of a new module to DYNSYS 2.1

This section will present the main features of how to construct a subroutine representing a unit computation and to adapt it to the main executive. This will be useful for future work when the simulation of the solvent recovery column must be made or when investigation on different control schemes is undertaken.

3.2.4.1 Main components of the DYNSYS 2.1 executive

Basically, the executive consists of the following programs:

-DYN 1:	this subroutine reads in data such as initial values of the variables, parameters of the equipment, characteristic values, etc., and prints out for checking, when called by the main program.
-DYN 2:	this subroutine establishes the logical sequence of execution of the modules, when called by the main program.
-OUTPUT:	this subroutine handles all outputs at each time interval specified by the user, when called by DYN 1 or by the main program.
-PROPS:	this function reads in the properties of each component in the material flow when called by DYN 1.
-DIFSUB:	this subroutine uses the Gear's method to perform the integration and is called by the modules.
-SIMULT:	this subroutine solves the set of linear algebraic equations, simultaneously, when called by DIFSUB.
-TRIDAG:	this subroutine solves the set of linear algebraic equations simultaneously, when called by DIFSUB, if the Jacobian matrix is tridiagonal.
-DYNSYS:	the main program of the executive.

The listings of the above subroutines can be found in reference 14. Fig. 3.1 shows the interactions between the different components involved in the simulation work in this project.

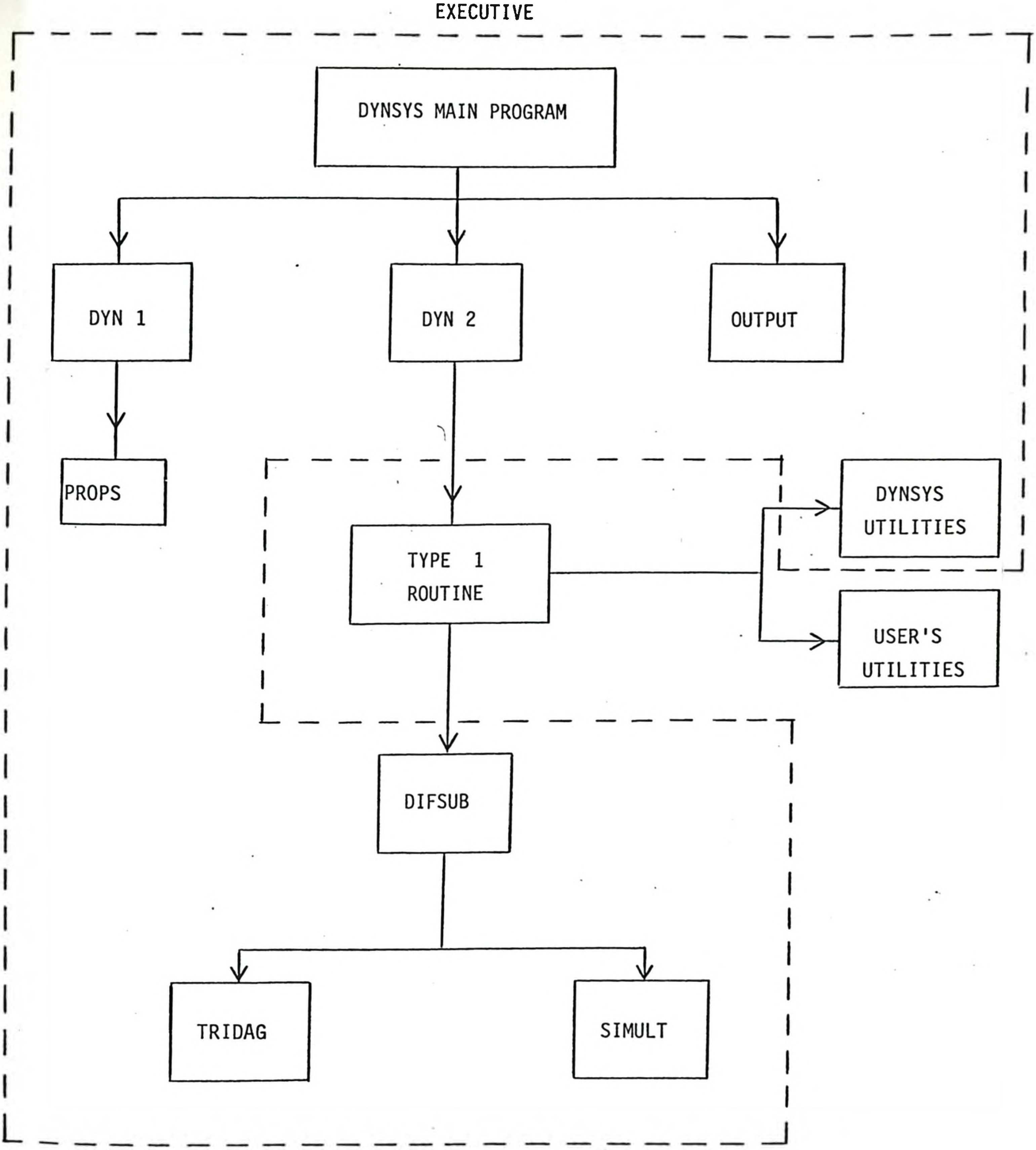


fig. 3.1 INTERACTION BETWEEN DYNSYS COMPONENTS AND USER COMPONENTS

3.2.4.2 Construction of a module

Both the stiff and non-stiff options of Gear's method are available. However, the same option must be used for the entire simulation.

When the simulation diagram contains both stiff and non-stiff modules, the stiff option must be used, which means that the stiff coefficients in the approximative equations will be used in the integration. Then, depending on whether the module is stiff or non-stiff, the Newton-Raphson iteration or Jacobi direct iteration will be chosen.

A detailed skeleton of a module is shown in fig. 3.2.

Up to three iterations are allowed in the corrector pass. If convergence within a user-specified tolerance does not appear after the third iteration, the integration step size is reduced to 1/4 of its current size and the step is repeated for all modules.

The user can access to a number of parameters within the executive. A list is given in Appendix B.

3.3 Distillation column dynamic simulation

A tentative mathematical model has been written to represent the dynamic behavior of the experimental extractive distillation column of fifteen stages with a ternary organic system. The module created is used under the executive DYNSYS 2.1 for further flexibility for future work. The control configuration could be investigated after an appropriate model has been found for the main process.

fig. 3.2 SKELETON OF MODULE FOR DYNSYS 2.1

```
SURROUTINE TYPE2.1

COMMENTS DESCRIBING MODULE

THE FOLLOWING COMMON BLOCKS ARE OR MAY BE REQUIRED:
MAT,CON,PTAB,UNIT,GEAR2,MODULE,ROW,COLUMN,JACOB,SUBDI,DIAG,SUPERD

COMMON/MAT/MP( , ),EP( , ),S( , , ),EX( )
COMMON/CON/IG,NCOMP,NC5,H,NE,NS,NPR,NPOL,TMAX,IORDER,NGRAPH
COMMON/PTAB/IGFLAG,PP( , )
COMMON/UNIT/IM,NMP
COMMON/GEAR2/EPS,TIME,KFLAG,JSTART,NBVMAX,ICONV,ISTIFF
COMMON/MODULE/IDERY,ITER,ITRI,MC,IPIVOT

THE ROW DIMENSION OF JCOL AND XJACOB MUST BE THE SAME AS IN DIFSUB
MC IS MAXIMUM NUMBER OF COLUMNS IN XJACOB AND JCOL

COMMON/COLUMN/JCOL( ,MC)     (NORMAL OPTION)
COMMON/JACOB/XJACOB( ,MC)    (NORMAL OPTION)

N IS NUMBER OF ODES

COMMON/SUBDI/A(N)      (TRIDIAGONAL OPTION)
COMMON/DIAG/B(N)       (TRIDIAGONAL OPTION)
COMMON/SUPERD/C(N)     (TRIDIAGONAL OPTION)

********************************************************************************
                 SECTION #1 : PARAMETER CALCULATIONS
********************************************************************************

CALCULATE MODULE PARAMETERS : STREAM INPUT,INITIAL CONDITIONS,
                              VALUES OF ITER ETC.

INPUT STREAM INFORMATION IS OBTAINED FROM S(IG, , )
VALUE OF INDEPENDENT VARIABLE (Y) NEED ONLY BE SPECIFIED ON FIRST
   INTEGRATION STEP (I.E. INITIAL CONDITIONS)
JSTART=0 ON FIRST INTEGRATION STEP
      =CURRENT ORDER OF INTEGRATION TECHNIQUE ON LATER STEPS

IF ISTIFF=0,NONSTIFF COEFFICIENTS WILL BE USED IN INTEGRATION ALGORITHM
            (DIFSUB) FOR ALL MODULES,
            JACOBIAN MATRIX IS NOT REQUIRED

IF ISTIFF=1,STIFF COEFFICIENTS WILL BE USED FOR ALL MODULES
            THEN,IF ITER=0 DIRECT ITERATION OF CORRECTOR WILL BE USED
                           (NONSTIFF MODULE)
                           JACOBIAN IS NOT REQUIRED
                 IF ITER=1 NEWTON-RAPHSON ITERATION OF CORRECTOR WILL
                           BE USED (STIFF MODULE)
```

```
C                                        JACOBIAN MATRIX MUST BE SUPPLIED
C
C     ITER MUST BE SPECIFIED 0 OR 1
C     IT IS NOT USED UNLESS ISTIFF=1
C
      IF(IG.EQ.2) GO TO 2
C
C**************************************************************************
C                       SECTION #2 : JACOBIAN EVALUATION
C**************************************************************************
C
C     SECTION #2 IS OMITTED FOR NONSTIFF MODULE
C
C     JACOBIAN MATRIX IS REQUIRED ONLY IF MODULE EQUATIONS ARE STIFF
C     CALCULATE JACOBIAN MATRIX ON CORRECTOR PASS ONLY
C     JACOBIAN NEED NOT BE VERY ACCURATE,AS IT IS USED ONLY FOR CONVERGENCE
C        OF CORRECTOR
C     ONLY NONZERO ELEMENTS ARE CALCULATED,BUT ALL DIAGONAL ELEMENTS MUST
C        BE STORED WHETHER OR NOT THEY ARE ZERO (NORMAL OPTION)
C
C              NORMAL OPTION
C
C     IF JACOBIAN MATRIX IS NOT TRIDIAGONAL,KEYS STORAGE SCHEME IS USED
C     COLUMN NUMBERS OF NONZERO ELEMENTS IN ROW I ARE STORED IN
C              (JCOL(I,J),J=1,NUMBER OF NONZEROS IN THAT ROW)
C     SIMILARLY THE VALUES OF THE NONZERO ELEMENTS OF ROW I ARE
C              STORED IN XJACOB(I, )
C     REMEMBER TO STORE ALL DIAGONAL ELEMENTS EVEN IF THEY ARE ZERO
C
C     MC     - MAXIMUM NUMBER OF COLUMNS IN XJACOB MATRIX
C              MC CAN BE LARGER THAN THE MAXIMUM NUMBER OF NONZERO ELEMENTS
C              IN ANY JACOBIAN ROW SINCE ADDITIONAL NONZERO ELEMENTS CAN BE
C              CREATED DURING THE SOLUTION.MC IS CONSTANT FOR ANY SET OF ODES
C              AND MAY HAVE TO BE DETERMINED BY TRIAL AND ERROR
C     IPIVOT - PIVOT OPTION USED IN SIMULT (1-7)
C        IPIVOT=1 : SIMPLE GAUSS-JORDAN ELIMINATION
C        IPIVOT=2 : GAUSS-JORDAN PARTIAL PIVOTING
C        IPIVOT=3 : GAUSS-JORDAN FULL PIVOTING
C        IPIVOT=4 : MINIMUM ROW-MINIMUM COLUMN
C        IPIVOT=5 : MINIMUM COLUMN-MINIMUM ROW
C        IPIVOT=6 : MAXIMUM COLUMN-MINIMUM ROW
C        IPIVOT=7 : MINIMUM OF ROW ENTRIES TIMES COLUMN ENTRIES
C
C     ICOL(I,J)    - COLUMN NUMBER OF JTH NONZERO ELEMENT IN ROW I
C     XJACOB(I,J)  - VALUE OF JTH NONZERO ELEMENT OF ROW I
C
      MC=
      IPIVOT=
      IROW(1,1)=
      XJACOB(1,1)=
         .
         .
```

```
         .
      IROW(N, )=
      XJACOB(N, )=
C
C             TRIDIAGONAL OPTION
C
C     FOR TRIDIAGONAL JACOBIAN MATRIX,THE SUBDIAGONAL,DIAGONAL AND SUPERDIAGO
C     ELEMENT VALUES ARE STORED IN ARRAYS A,B AND C FROM COMMON BLOCKS SUBDI,
C     DIAG AND SUPERD RESPECTIVELY
C
C     N - NUMBER OF ODES
C     A - VALUES OF SUBDIAGONAL ELEMENTS ARE STORED IN A(2)...A(N)
C         A(1) IS NOT USED
C     B - VALUES OF DIAGONAL ELEMENTS
C     C - VALUES OF SUPERDIAGONAL ELEMENTS ARE STORED IN C(1)...C(N-1)
C         C(N) IS NOT USED
C
      B(1)=
      C(1)=
      A(2)=
      B(2)=
      C(2)=
         .
         .
         .
      A(N-1)=
      B(N-1)=
      C(N-1)=
      A(N)=
      B(N)=
    1 CONTINUE
C
C     IF TRIDIAGONAL OPTION IS BEING USED,ITRI MUST BE SET TO 1 HERE.
C     IF NORMAL OPTION IS USED ITRI MAY BE IGNORED
C
C     ITRI=1
C
    2 CONTINUE
C
C*****************************************************************************
C                       SECTION #3 : DERIVATIVE CALCULATION
C*****************************************************************************
C
C     CALCULATE DERIVATIVES
C
      DERY(1)=
         .
         .
         .
      DERY(N)=
C
C*****************************************************************************
```

```
C                    SECTION #4 : CALL DIFSUB
C*******************************************************************************************
C
C     CALL DIFSUB TO SOLVE ODES FOR MODULE
C     DIFSUB MAY BE CALLED MORE THAN ONCE FROM A MODULE FOR EXAMPLE IF THE
C        MODULE CONTAINS A SET OF STIFF O.D.E.S AND ANOTHER SET OF NONSTIFF
C        O.D.E.S
C
C     ARGUMENTS :    N      - NUMBER OF ODES
C                    Y      - INDEPENDENT VARIABLE
C                    DERY   - DERIVATIVES
C
      CALL DIFSUB(N,Y,DERY)
C
C     IF IDERY IS NOT ZERO,THE DERIVATIVES WILL BE RE-EVALUATED AND
C        RETURNED TO DIFSUB
C     ITRI MUST ALSO BE RESET IF IT IS 1
C
      IF(IDERY.NE.0) GO TO 1
C
C*******************************************************************************************
C                    SECTION #5 : STREAM OUTPUT CALCULATION
C*******************************************************************************************
C
C     CALCULATE STREAM OUTPUT (STORED IN S(1, , ))
C
      S(1, , )=
          .
          .
          .
      RETURN
      END
```

3.3.1 Assumptions

The continuous distillation of a liquid mixture in either a stagewise or a continuous-contact column is a very complex phenomenon which involves heat transfer, mass transfer and fluid dynamics. To obtain complete description of such a process, one must formulate the mechanics and interrelationships of these three transport phenomena.

The mechanics of the fluid flow requires the understanding of the column internals, the pipe sizes and arrangement, the condenser system and the reboiler system. The liquid and vapor hold-ups could be determined thereafter using some empirical relations. The next step would be to set the relationship between the vapor and the liquid at these mixing regions. Considering, the liquid holdup on a plate of the column, the composition is a function of time and position within the liquid phase on the plate. The mass transfer rate is basically a function of the difference in composition of the vapor and liquid phases at this point and also of the composition gradient within the liquid phase.

Consequently, it is apparent that a mathematical dynamic model obtained rigorously will be very complex and hard to use.

However, before modeling is attempted for a process, the objectives of such work must be established. The investment in modeling will therefore be determined by the probable return to be gained by using the model.

In the actual project, the main purposes of constructing a dynamic simulation for the distillation column are to predict the transient behavior of the column in response to a variety of types of input disturbances, to investigate the response when the column is disturbed by combinations of possible input variables and to design feasible control strategies as well as to obtain an acceptable design of the control parameters and systems dynamic parameters. If these objectives can be achieved satisfactorily from a relatively simple model, then it is apparent that the pursuit of a completely rigorous model is unwarranted. If the model proves to be insufficient for some of the above objectives, the model could be improved by deleting the least probable assumptions.

The following assumptions have been made in developing the mathematical model of the actual column's transient behavior:

(1) There is no heat loss along the column. This assumption could be shown valid to a certain extent since the column is surrounded by a heated air jacket to minimize the heat transfer from inside the column to outside.

(2) The fluid dynamic response time is negligible compared with that of mass transfer. This assumption is justified when no substantial changes in liquid flowrates are experienced, which is very appropriate for changes in feed composition. Also, the range of input flowrate variation is found to be quite small. Moreover, the work of May (17) indicated that for small towers, the time to reach fluid equilibrium is essentially negligible compared to that for mass transfer. The diameter of the present column is 2 inches.

(3) The molar vapor holdup is small compared with that of the liquid. Huckaba and Danly (4) have shown that this assumption is perfectly valid for atmospheric distillation.

(4) The volumetric liquid holdup at each tray is not a function of time. The initial holdups are estimated by the Francis Weir relationship, using the initial liquid flowrates as the basic information. The equations are given in Appendix C.

(5) The flowing fluids are all single phase streams. The liquid and vapor entrainment are negligible.

(6) The pressure through the column is constant and equals atmospheric pressure. This assumption is valid in this case if one considers the small number of plates and the relatively small height of the liquid layers. This assumption could be deleted to make the pressure a linear function with position along the column. However, no big improvement is expected.

(7) The plate efficiencies are independent of time. The work of O'Neill (10) indicated that the efficiencies could be kept constant within the actual operating range without affecting drastically the results of the steady state simulation.

(8) The volumetric reboiler and condenser hold-ups are estimated experimentally. A 100% mixing is assumed in both holdups. The assumption for the reboiler is questionable.

(9) The liquid composition at each tray is assumed uniform. This could be valid since the diameter of the tray and the weir height are reasonably small.

Based on the above assumptions, the basic equations will be presented in the next section.

3.3.2 Basic equation derivation

3.3.2.1 Material balance equations

The basic principle of conservation has been applied to each tray including the feed tray and the solvent input tray and also to the reboiler and the condenser. All the equations have been derived using the gram-mole as the material unit. The starting point of the derivation is the generalized statement:

Input - output + generation - disappearance = accumulation

Based on the nomenclature in fig. 3.3, one can write for a ternary system:

For a normal tray:

$$\frac{dM_n}{dt} = - V_n - L_n + L_{n-1} + V_{n+1} \tag{3.3.1.a}$$

$$\frac{d(M_n x_{n,i})}{dt} = - V_n y_{n,i} - L_n x_{n,i} + L_{n-1} x_{n-1,i} +$$

$$V_{n+1} y_{n+1,i} \tag{3.3.1.b}$$

$i = 1, 2$

For a feed tray:

$$\frac{d}{dt} M_n = - V_n - L_n + L_{n-1} + V_{n+1} + F \tag{3.3.2.a}$$

$$\frac{d}{dt}(M_n x_{n,i}) = - V_n y_{n,i} - L_n x_{n,i} + L_{n-1} x_{n-1} +$$

$$V_{n+1} y_{n+1,i} + F x_{f,i} \qquad (3.3.2.b)$$

For the condenser (n=1)

$$\frac{d}{dt} M_n = V_{n+1} - D - R \qquad (3.3.3.a)$$

$$\frac{d}{dt}(M_n x_{n,i}) = V_{n+1} y_{n+1,i} - D x_{n,i} - R x_{n,i} \qquad (3.3.3.b)$$

$i = 1, 2$

For the reboiler (n = 15)

$$\frac{d}{dt} M_n = - V_n + L_{n-1} - B \qquad (3.3.4.a)$$

$$\frac{d}{dt}(M_n x_{n,i}) = - V_n\, y_{n,i} + L_{n-1} x_{n-1,i} - B x_{n,i} \qquad (3.3.4.b)$$

$i = 1, 2$

For convenience, the compositions $x_{n,i}$ were used as variables of integration. Therefore, the equations (3.3._.b) were transformed using the equations (3.3._.a) to obtain on the left hand-side the derivative of the composition $x_{n,i}$ as a function of time.

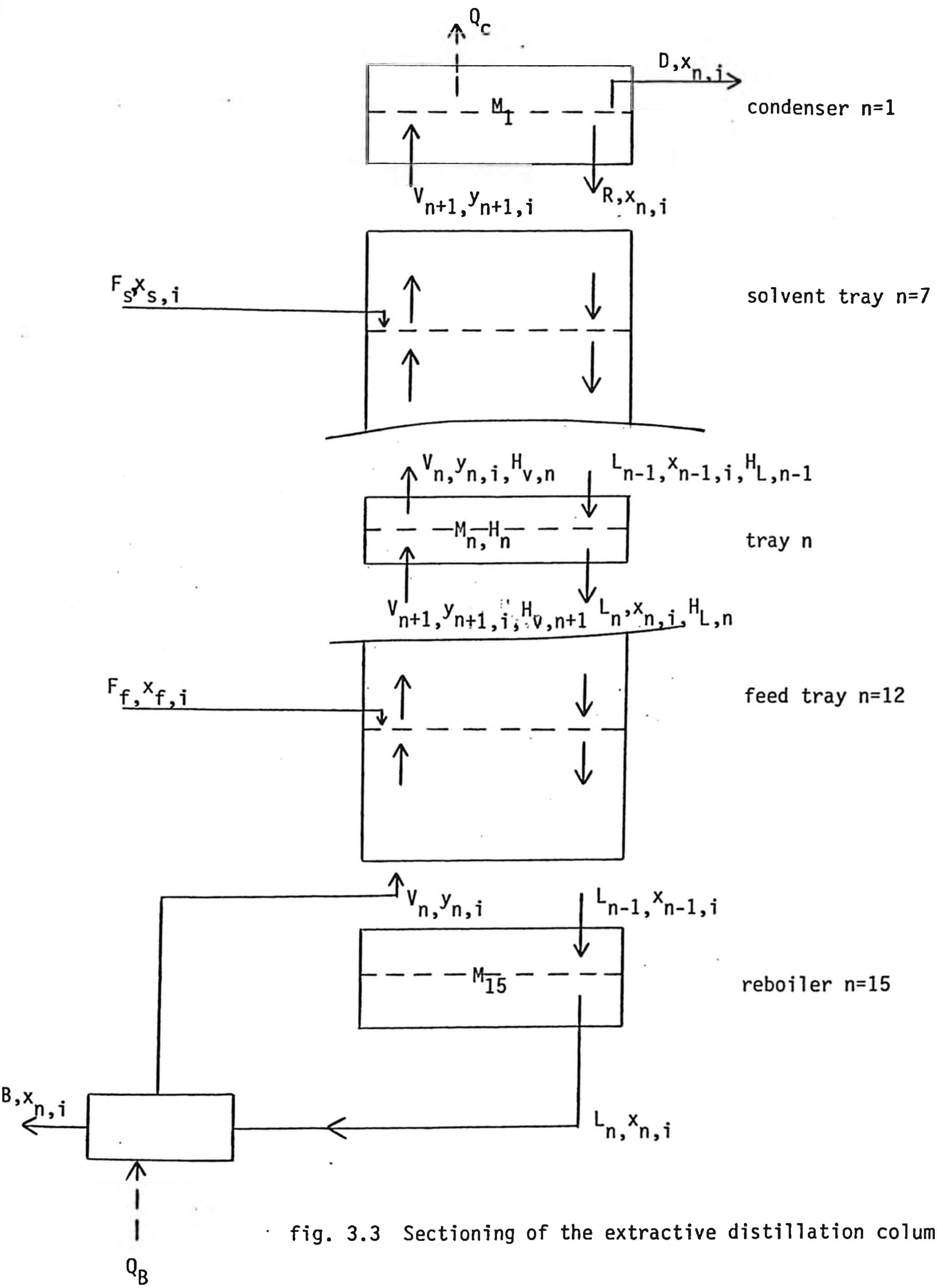


fig. 3.3 Sectioning of the extractive distillation column

After simple manipulations, the following equations were obtained:

For a normal tray:

$$\frac{d}{dt}x_{n,i} = \frac{-V_n y_{n,i} + L_{n-1}x_{n-1,i} + V_{n+1}y_{n+1,i}}{M_n} - \frac{x_{n,i}\,(-V_n + L_{n-1} + V_{n+1})}{M_n}$$

$$i = 1, 2 \qquad (3.3.1.c)$$

For a feed tray:

$$\frac{d}{dt}x_{n,i} = \frac{-V_n y_{n,i} + L_{n-1}x_{n-1,i} + V_{n+1}y_{n+1,i}}{M_n} + \frac{Fx_{f,i} - x_{n,i}\,(-V_n + L_{n-1} + V_{n+1} + F)}{M_n}$$

$$i = 1, 2 \qquad (3.3.2.c)$$

For the condenser (n=1)

$$\frac{d}{dt}x_{n,i} = \frac{(y_{n+1,i} - x_{n,i})\,V_{n+1}}{M_n} \qquad i = 1,2 \qquad (3.3.3.c)$$

For the reboiler (n=15)

$$\frac{d}{dt}x_{n,i} = \frac{-V_n y_{n,i} + L_{n-1}x_{n-1,i} - x_{n,i}\,(-V_n + L_{n-1})}{M_n}$$

$$i = 1, 2 \qquad (3.3.4.c)$$

The 30 equations (3.3._c) were integrated as a function of time, and the third composition (i=3) was calculated according to:

$$\sum_i x_{n,i} = 1 \qquad i = 1,3 \qquad (3.3.5)$$

In this simulation program, the integration was made on the acetone composition (i=1) and the methanol composition (i=2); the water composition (i=3) was obtained by difference.

3.3.2.2 Vapor-liquid equilibrium relations

The activity coefficient of each component in the liquid mixture was calculated using the enthalpic Wilson equation, extended to multicomponent mixtures.

Given a liquid mixture of q components, the activity coefficient of component i in the mixture was calculated from:

$$\ell n\gamma_i = - \sum_{\substack{j=1 \\ j\neq i}}^{q} \frac{x_j \ \ell n\ (\Lambda_{ij}\Lambda_{ji})}{(\sum_p \Lambda_{ip}x_p)(\sum_p \Lambda_{jp}x_p)} \times \alpha_{ij} - $$

$$\sum_{\substack{m,j \\ m\neq j\neq i}} \frac{x_j x_m \ell n(\Lambda_{jm}\Lambda_{mj})}{(\sum_p \Lambda_{jp}x_p)(\sum_p \Lambda_{mp}x_p)} \times \alpha_{ijm} \qquad (3.3.6.a)$$

where

$$\alpha_{ij} = 1 + x_i \; (1 - \frac{1}{\sum_p \Lambda_{ij} x_p} - \frac{\Lambda_{ji}}{\sum_p \Lambda_{jp} x_p}) \qquad (3.3.6.b)$$

$$\alpha_{ijm} = 1 - \Lambda_{ji}/(\sum_p \Lambda_{jp} x_p) - \Lambda_{mi}/(\sum_p \Lambda_{mp} x_p) \qquad (3.3.6.c)$$

$$\Lambda_{ii} = 1 \qquad (3.3.6.d)$$

$$\Lambda_{ij} = \frac{v_j}{v_i} \times \exp(\frac{-\lambda_{ij} - \lambda_{ii}}{RT}) \qquad (3.3.6.e)$$

The parameters λ_{ij} for each pair of components for the present mixture of acetone-methanol-water could be found in Reference 20. Griswold and Buford (21) have tested the above equations for the ternary system mentioned above. The derivation of the enthalpic Wilson equations is presented in reference 20. The v_i's are the molar volume of the components at temperature T in °K.

Once the activity coefficient of each component was calculated, the activity Γ_i could be derived by the following relation:

$$\Gamma_i = \frac{p'_i}{p} \times \gamma_i \quad \text{, p is the solution pressure} \qquad (3.3.7)$$

where

$$p'_i = \exp(a + \frac{b}{T+c}), \quad p'_i \text{ is the vapour pressure} \qquad (3.3.8)$$

The composition of the vapor in equilibrium with the liquid mixture y^*_i

was then calculated from:

$$y^*_i = r_i x_i \qquad (3.3.9)$$

The temperature of the liquid on each tray was used as the parameter to converge the summation of the vapor component compositions to 1.

The Murphree efficiencies $E_{n,i}$ could then be applied to calculate the true composition of the vapor leaving each tray. The following relation was used for tray n:

$$y_{n,i} = y_{n+1,i} + E_{n,i} \times (y^*_{n,i} - y_{n+1,i}) \qquad (3.3.10.a)$$

$$n=1,14$$

As no vapor enters the reboiler (n=15), the Murphree efficiency was defined differently by taking the composition of the liquid leaving the reboiler instead of the composition of the vapor coming to the stage. Therefore, one obtains for the reboiler:

$$y_{n,i} = x_{n,i} + E_{n,i} \times (y^*_{n,i} - x_{n,i}) \qquad (3.3.10.b)$$

$$n=15$$

3.3.2.3. Energy equations

At the end of each time interval, the new compositions of the liquid and the vapor leaving and reaching the stage were obtained as shown in the previous sections. Then, the unsteady-state energy equation at each stage was used to readjust the vapor and liquid flowrates.

The energy imbalance at tray n was obtained according to the following principle:

energy accumulation = energy in - energy out

$$\left(\frac{dH_n}{dt}\right)_t = (H_{L,n-1}L_{n-1} - H_{L,n}L_n + H_{V,n+1}V_{n+1} -$$

$$H_{V,n}V_n + Q_n)_t \qquad (3.3.11)$$

This surplus of energy was assumed to vaporize a certain quantity of liquid from the tray holdup, thus increasing the flowrate of the vapor leaving this tray. The vapor material amount was obtained from:

$$\left(\frac{dV_n}{dt}\right)_t = \frac{(dH_n/dt)_t}{(H_{V,n} - H_{L,n})_t} \qquad (3.3.12)$$

Therefore, the vapor flowrates could be readjusted, starting from the reboiler up the column.

At the top of the column, due to the change in the vapor flowrates, the unbalance of the liquid flow could be computed and flowed down, trying to keep the energy equation instantaneously balanced.

3.3.3. Jacobian matrix derivation

As mentioned previously, the integration procedure requires the

Jacobian matrix of the system of ordinary differential equations, to perform the convergence of the solutions at a certain time t, in the corrector pass.

In general, for the system $\dot{\underline{y}} = f(\underline{y})$, the Jacobian matrix is:

$$\underline{J} = \begin{bmatrix} \frac{\partial f_1}{\partial y_1} & \frac{\partial f_1}{\partial y_2} & \cdots\cdots & \frac{\partial f_1}{\partial y_n} \\ \frac{\partial f_2}{\partial y_1} & \frac{\partial f_2}{\partial y_2} & & \\ \vdots & & & \\ \frac{\partial f_n}{\partial y_1} & \cdots\cdots & & \frac{\partial f_n}{\partial y_n} \end{bmatrix}$$

Taking the partial derivative of the right hand side of equations (3.3.1.c (3.3.2.c), (3.3.3.c) and (3.3.4.c) with respect to $x_{n-1,i}$ and $x_{n,i}$ successively, one could obtain

for a normal tray:

$$\frac{d}{dx_{n-1,i}} \left(\frac{dx_{n,i}}{dt}\right) = \frac{L_{n-1}}{M_n} \qquad i = 1, 2 \qquad (3.3.13.a)$$

$$\frac{d}{dx_{n,i}} \left(\frac{dx_{n,i}}{dt}\right) = - \frac{(-V_n + L_{n-1} + V_{n+1})}{M_n} \qquad i = 1, 2 \qquad (3.3.13.b)$$

for a feed tray:

$$\frac{d}{dx_{n-1,i}}\left(\frac{dx_{n,i}}{dt}\right) = \frac{L_{n-1}}{M_n} \qquad i = 1, 2 \qquad (3.3.14.a)$$

$$\frac{d}{dx_{n,i}}\left(\frac{dx_{n,i}}{dt}\right) = \frac{-(-V_n + L_{n-1} + V_{n+1} + F)}{M_n} \qquad i = 1, 2 \qquad (3.3.14.b)$$

for the condenser (n=1)

$$\frac{d}{dx_{n-1,i}}\left(\frac{dx_{n,i}}{dt}\right) = 0 \qquad i = 1, 2 \qquad (3.3.15.a)$$

$$\frac{d}{dx_{n,i}}\left(\frac{dx_{n,i}}{dt}\right) = \frac{-V_{n+1}}{M_n} \qquad i = 1, 2 \qquad (3.3.15.b)$$

for the reboiler (n=15)

$$\frac{d}{dx_{n-1,i}}\left(\frac{dx_{n,i}}{dt}\right) = \frac{L_{n-1}}{M_n} \qquad i = 1, 2 \qquad (3.3.16.a)$$

$$\frac{d}{dx_{n,i}}\left(\frac{dx_{n,i}}{dt}\right) = \frac{-(-V_n + L_{n-1})}{M_n} \qquad i = 1, 2 \qquad (3.3.16.b)$$

It is apparent to find that the Jacobian matrix has a tridiagonal form.

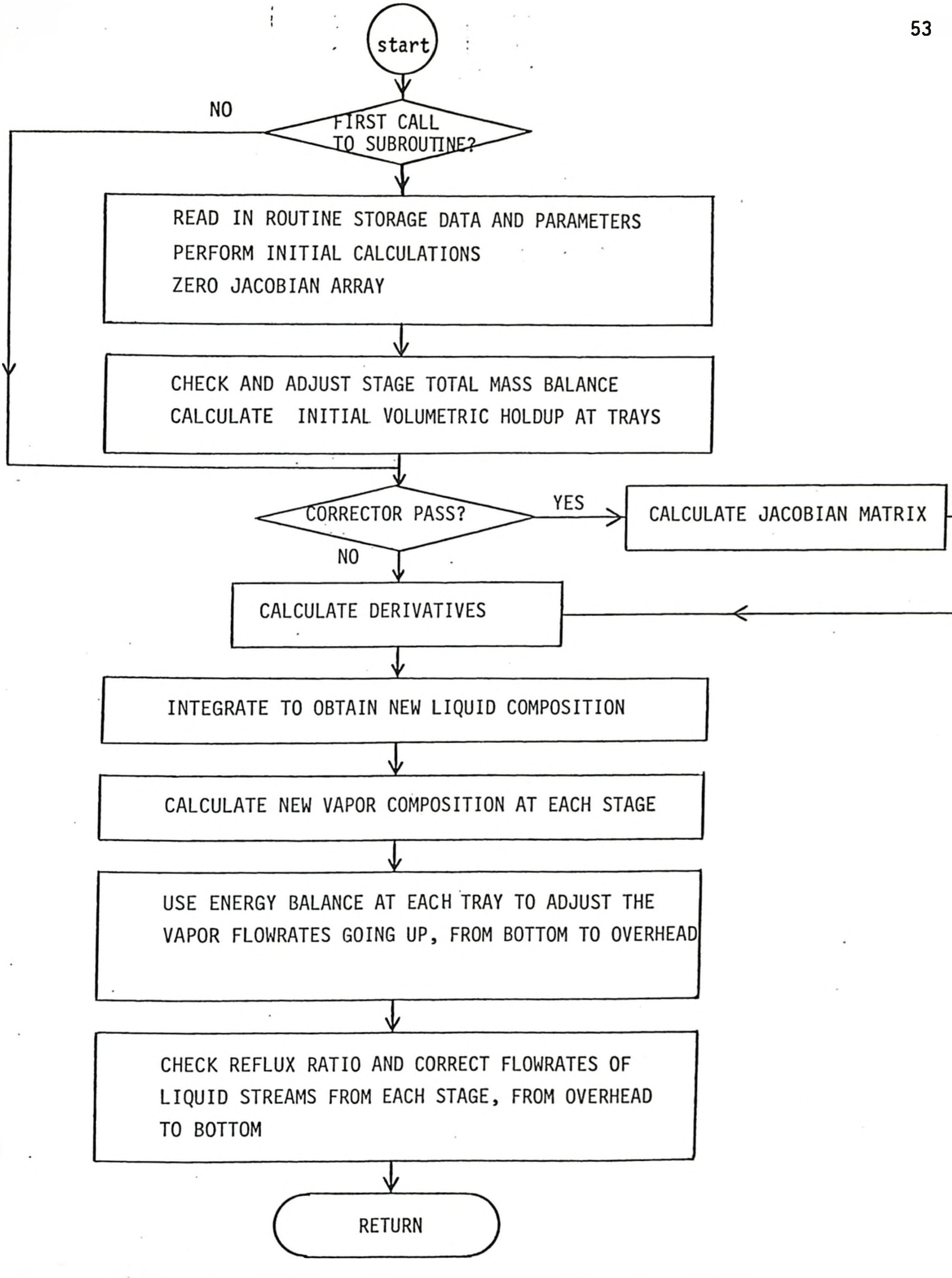


fig. 3.4 GENERAL FLOWCHART OF SUBROUTINE TYPE 1(COLUMN)

3.3.4 Simulation software

A module representing dynamically the experimental distillation column has been created and incorporated to the DYNSYS 2.1 executive. A logic flowchart of the calculation procedure could be found on fig. 3.4.

A stand-alone steady-state program was developed by O'Neill (10) and was used to supply the starting steady-state conditions for the dynamic simulation. Information related to the utilization of the steady-state package was presented in detail in reference 10.

A data transformer program was created to adapt the output of the steady-state program to a suitable format to input into the dynamic simulation program.

The dynamic runs were performed on a CDC 6400 (McMaster University). Listings of the utility subroutines can be found in Appendix F. Listing of the main subroutine is presented in Appendix F.

CHAPTER 4

DYNAMIC RESULTS - STEP AND PULSE DISTURBANCES

The immediate objective of the work was to develop a mathematical model to represent adequately the experimental extractive distillation column. In order to evaluate the present simulation algorithm, conventional disturbances such as step changes and pulse changes, were introduced into input variables such as feed flowrate, feed composition and reflux ratio. The dynamic simulated response has been observed and compared to available experimental results obtained by D. O'Neill (10).

A presentation and analysis of the calculated and experimental results are given in the following sections, for disturbances in feed composition, feed flowrate and reflux ratio respectively.

4.1 Dynamic responses due to feed composition disturbances

In order to observe the dynamic response of the column, especially the transient behavior of the overhead product composition which generally is the controlled variable, the feed acetone composition was disturbed by a negative step change of -5.3% amplitude, a positive step change of +5.6% amplitude and a negative pulse change of -17.4% amplitude. The operating conditions of the three runs could be found on Table 4.1.

Experimentally, the column was brought initially to a certain steady-state level and the disturbance was introduced. A system of on/off valves permits a switch of the feed line from one storage tank to another storage tank with different composition. Samples were taken from the overhead line and from the bottom line, more frequently at the beginning of the change, and were analyzed by a gas chromatograph.

Plots of experimental and simulated results with respect to time were made and presented on fig. 4.1, 4.2, 4.3. The variation of the temperature was found negligible in both experimental and simulated results. The comparison was found more consistent for the case of a pulse change. Bottom product compositions were apparently insensitive to feed composition changes as shown by experimentation and simulation.

4.2 Dynamic responses due to feed flowrate disturbances

The effects of a change in the feed flowrate were investigated. Such a change caused a perturbation in the overall mass balance of the column, therefore affecting the local and component mass balances throughout the stages.

Experimentally, a needle valve allows the manual introduction of a negative step change of -15.4% amplitude, a positive step change of +18.2% amplitude and a positive pulse change of +18.2% amplitude in the feed flowrate. The operating conditions of the three runs are presented in Table 4.2.

In the simulation, the mass change of the feed was supposed to appear instantaneously at the bottom product line, for the total mass balance to be satisfied. Therefore, sudden changes occurred for all liquid flowrates from the feed tray downward to the reboiler. Thus, the transient of the

TABLE 4.1 OPERATING CONDITIONS

FEED COMPOSITION AS DISTURBED VARIABLE

Quantity	Unit	negative step change		positive step change		negative pulse change	
		initial	final	initial	final	initial	final
F	g moles/hr	16.16	16.16	16.16	16.16	16.16	16.16
$x_{F,1}$	mol. fract.	.6076	.5753	.5753	.6076	.6067	.6076
$x_{F,2}$	mol. fract.	.3848	.4171	.4171	.3848	.3848	.3848
T_F	°C	45	45	45	45	45	45
S	g moles/hr	30.75	30.75	30.75	30.75	30.75	30.75
$x_{S,1}$	mol. fract.	0.	0.	0.	0.	0.	0.
$x_{S,2}$	mol. fract.	0.	0.	0.	0.	0.	0.
T_S	°C	70	70	70	70	70	70
RR	dimensionless	1.756	1.756	1.756	1.756	1.756	1.756
REBDT	Kcal/hr	235	235	235	235	235	235
%dist. AMPL.	dimensionless	-5.3 %		+ 5.6 %		- 17.4 %	

fig. 4.1 NEGATIVE STEP CHANGE TO FEED ACETONE COMPOSITION

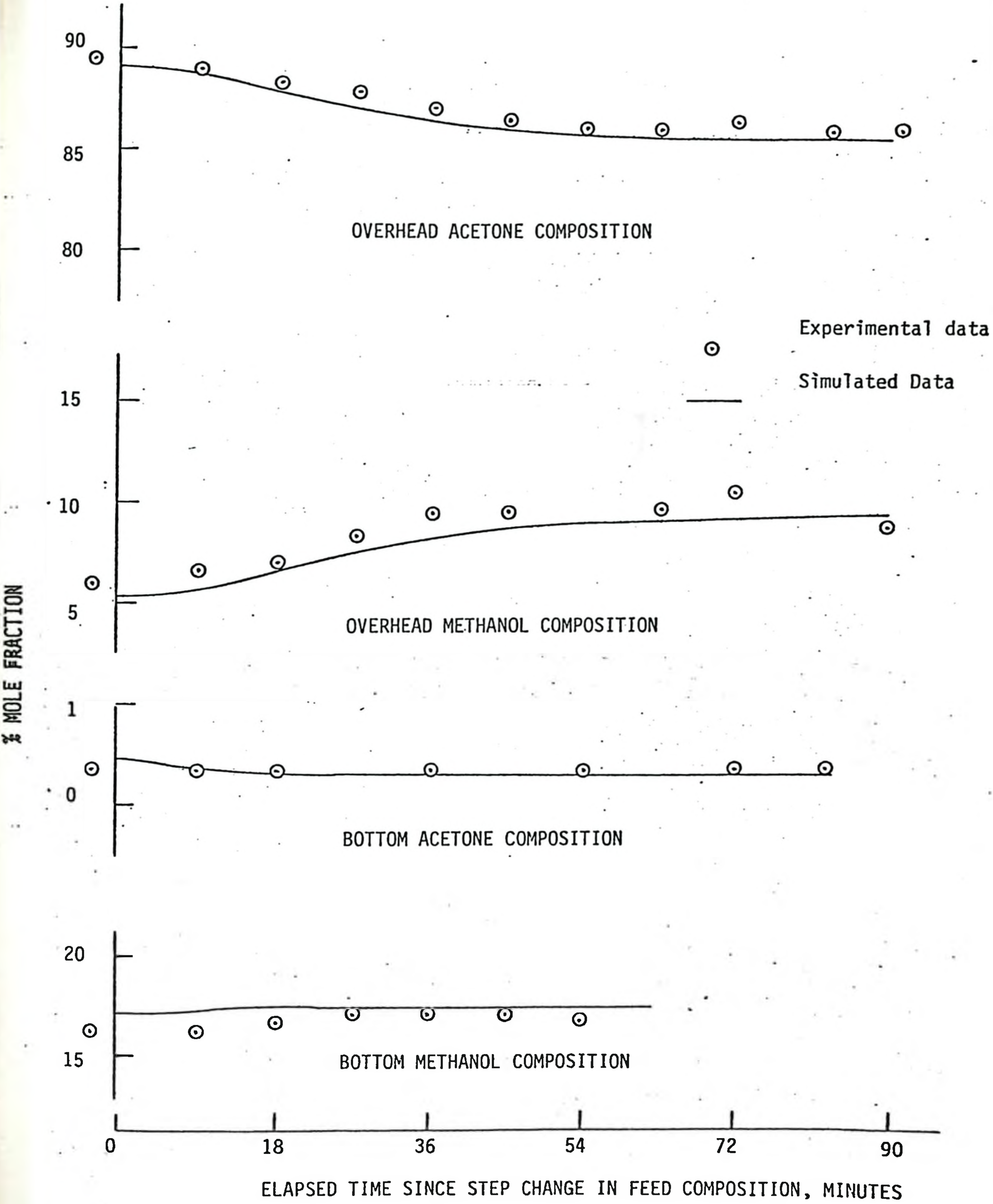


fig. 4.2 POSITIVE STEP CHANGE TO FEED ACETONE COMPOSITION

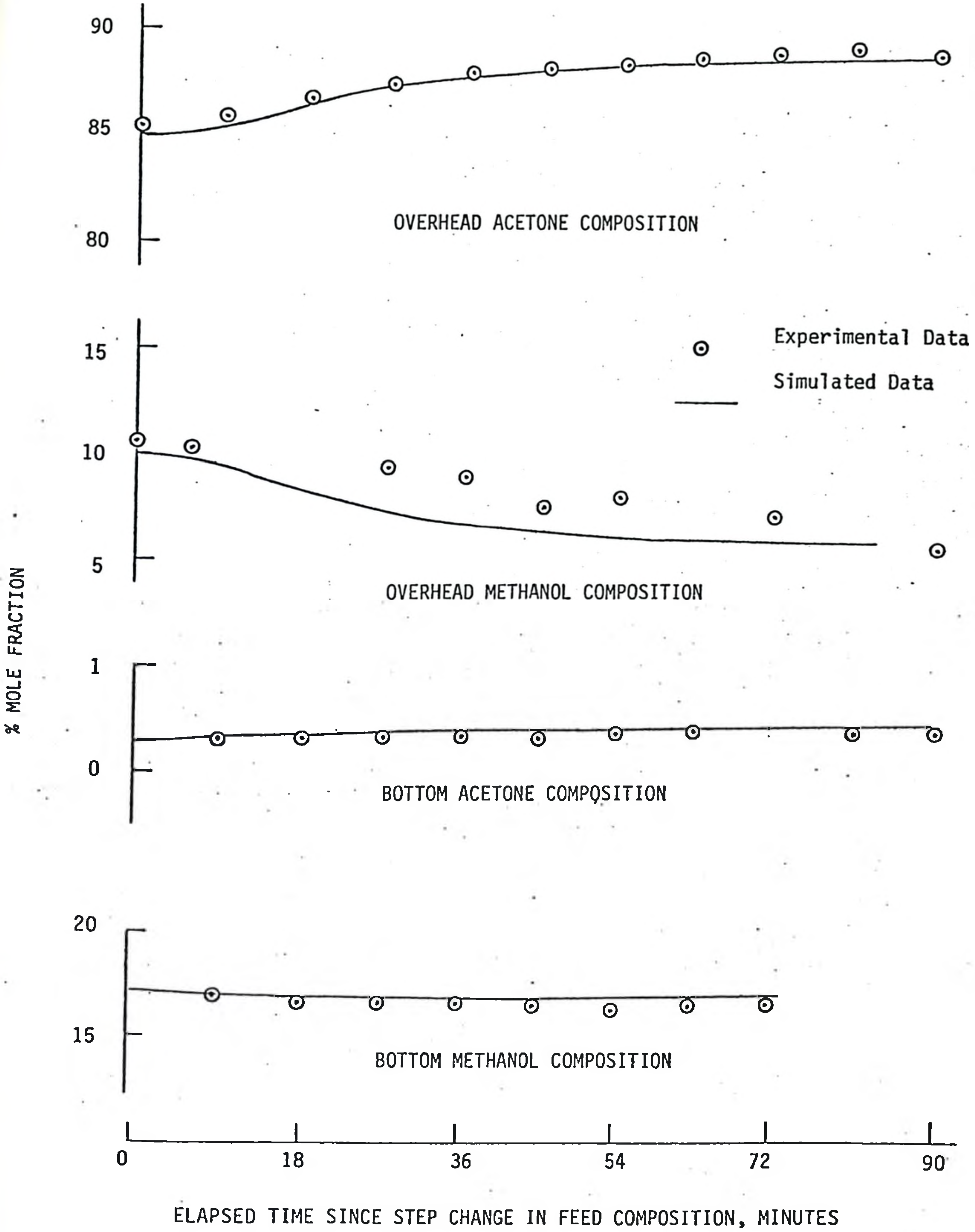
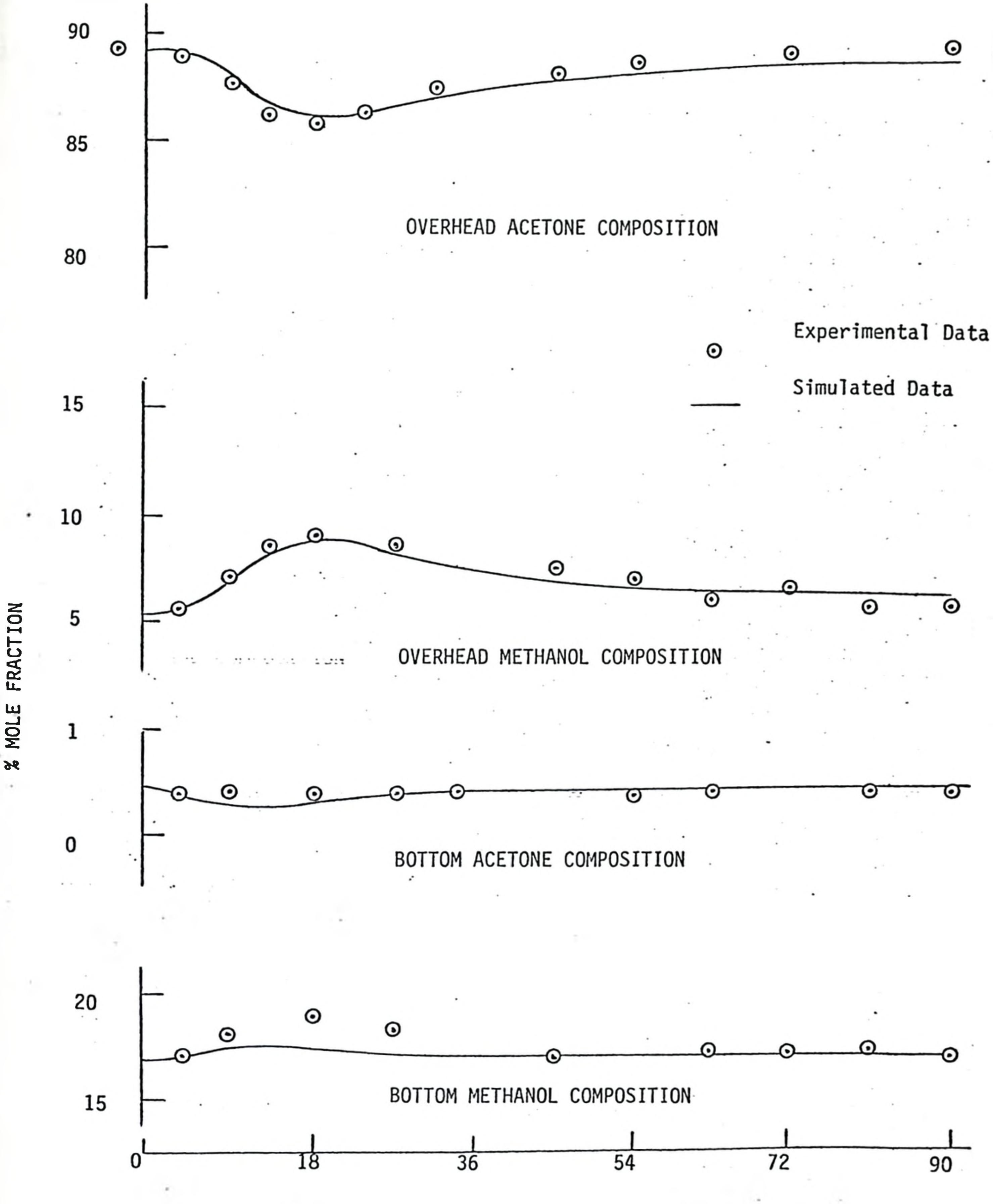


fig. 4.3 NEGATIVE PULSE CHANGE TO FEED ACETONE COMPOSITION



ELAPSED TIME SINCE PULSE CHANGE IN FEED COMPOSITION, MINUTES column resulted from the imbalance of the component mass equation for the trays between the feed and the reboiler. The fluid dynamic equilibrium time was assumed negligible compared to the equilibrium time of the temperature or composition.

The transient responses at the overhead and at the bottom were presented on fig. 4.4 . 4.5, 4.6. An increase in the feed flowrate means an increase in the quantity of acetone and methanol. Since most of the acetone goes up to the condenser, the acetone mole fraction at the overhead should increase relatively to the methanol mole fraction. However, increases of both acetone mole fraction and methanol mole fraction at the reboiler could be explained by the fact that there was more acetone than methanol in the feed line.

4.3 Dynamic responses due to reflux ratio changes

An increase in the reflux ratio at the overhead implies a decrease in the overhead flowrate and vice versa. The results were simulated for the cases of a negative step change of -11.3% amplitude, a positive step change of +12.7% amplitude and a positive pulse change of +75% amplitude.

Once again, the overall mass balance of the column was affected. In the simulation, when a change in the reflux ratio occurred, the total mass balances on each stage were completed by adjusting instantaneously the liquid flowrates leaving each stage. Therefore, an imbalance of the component mass equations resulted and implied a change in the composition of the liquid.

Experimentally, a reflux divider allowed the introduction of such

TABLE 4.2 OPERATING CONDITIONS

FEED FLOWRATE AS DISTURBED VARIABLE

Quantity	unit	negative step change		positive step change		positive pulse change	
		initial	final	initial	final	initial	final
F	g moles/hr	17.53	14.83	14.83	17.53	14.83	14.83
$x_{F,1}$	mol. fract.	.6059	.6059	.6059	.6059	.6059	.6059
$x_{F,2}$	mol. fract.	.3875	.3875	.3875	.3875	.3875	.3875
T_F	°C	45	45	45	45	45	45
S	mol. fract.	30.75	30.75	30.75	30.75	30.75	30.75
$x_{S,1}$	mol. fract.	0.	0.	0.	0.	0.	0.
$x_{S,2}$	mol. fract.	0.	0.	0.	0.	0.	0.
T_S	°C	70	70	70	70	70	70
RR	dimensionless	1.784	1.784	1.784	1.784	1.784	1.784
REBDT	Kcal/hr	234	234	234	234	234	234
% dist.ampl	dimensionless	- 15.8 %		+ 18.2 %		+ 18.2 %	

fig. 4.4 NEGATIVE STEP CHANGE TO FEED FLOW RATE

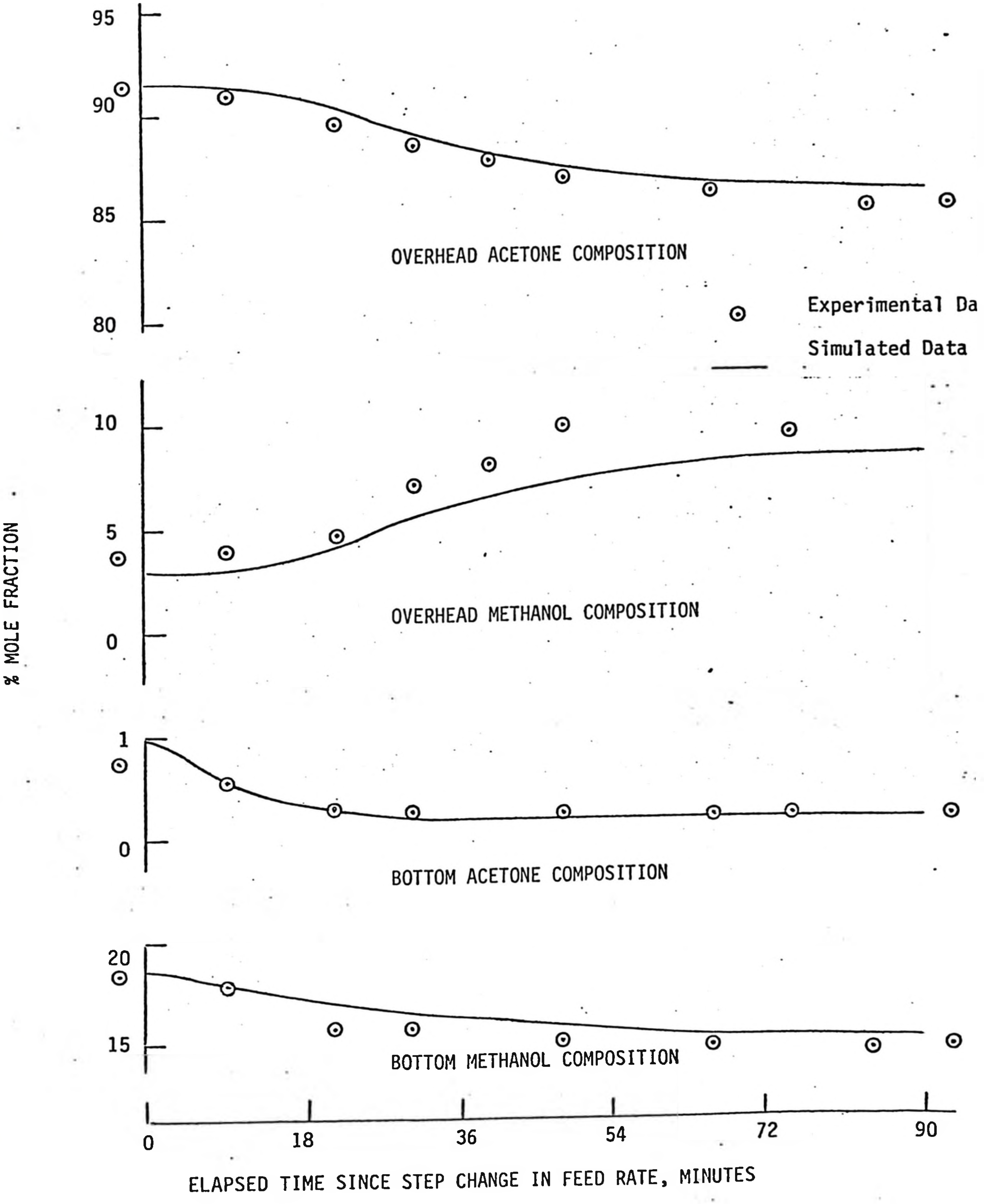


fig. 4.5 POSITIVE STEP CHANGE TO FEED FLOWRATE

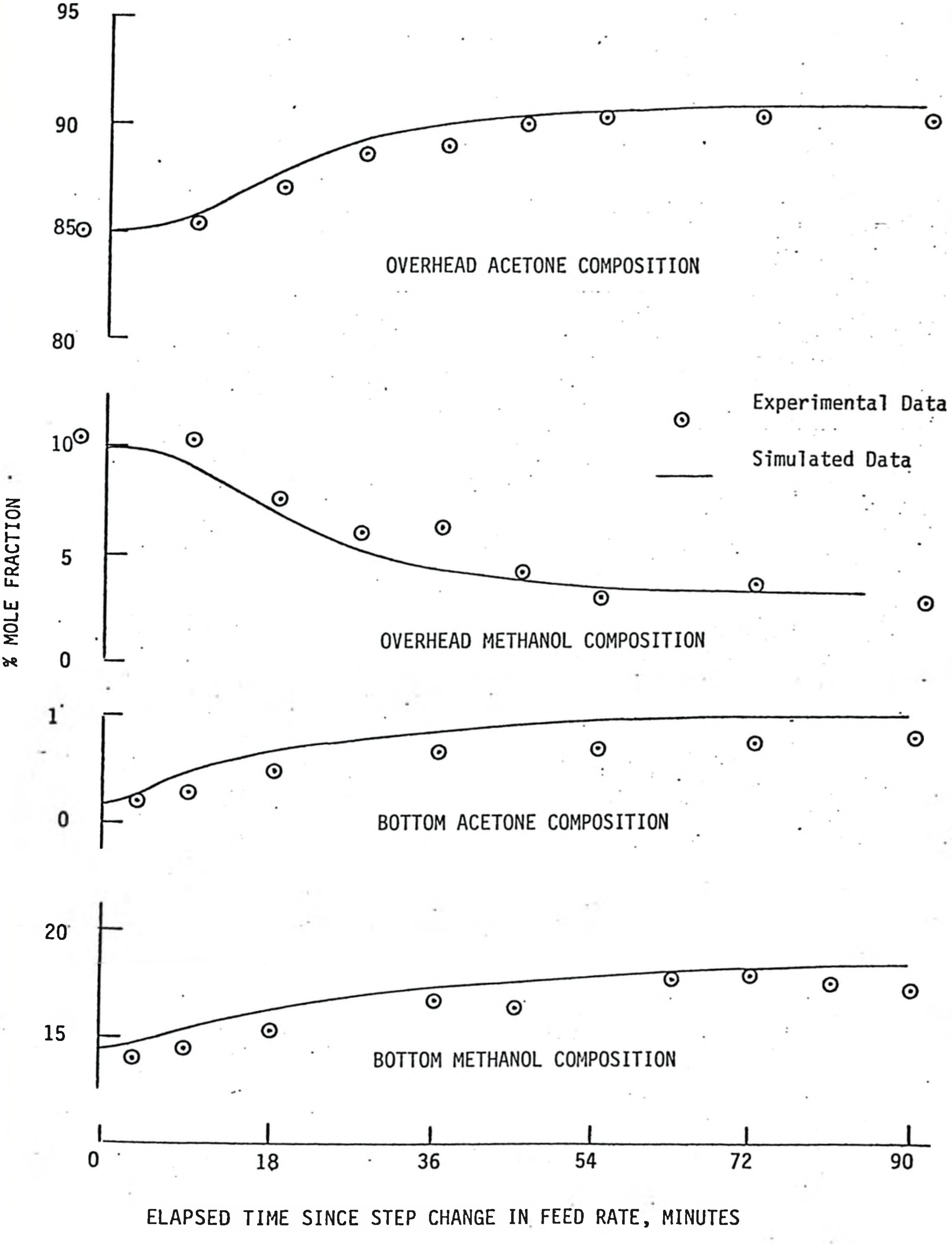


fig. 4.6 POSITIVE PULSE IN FEED FLOW RATE

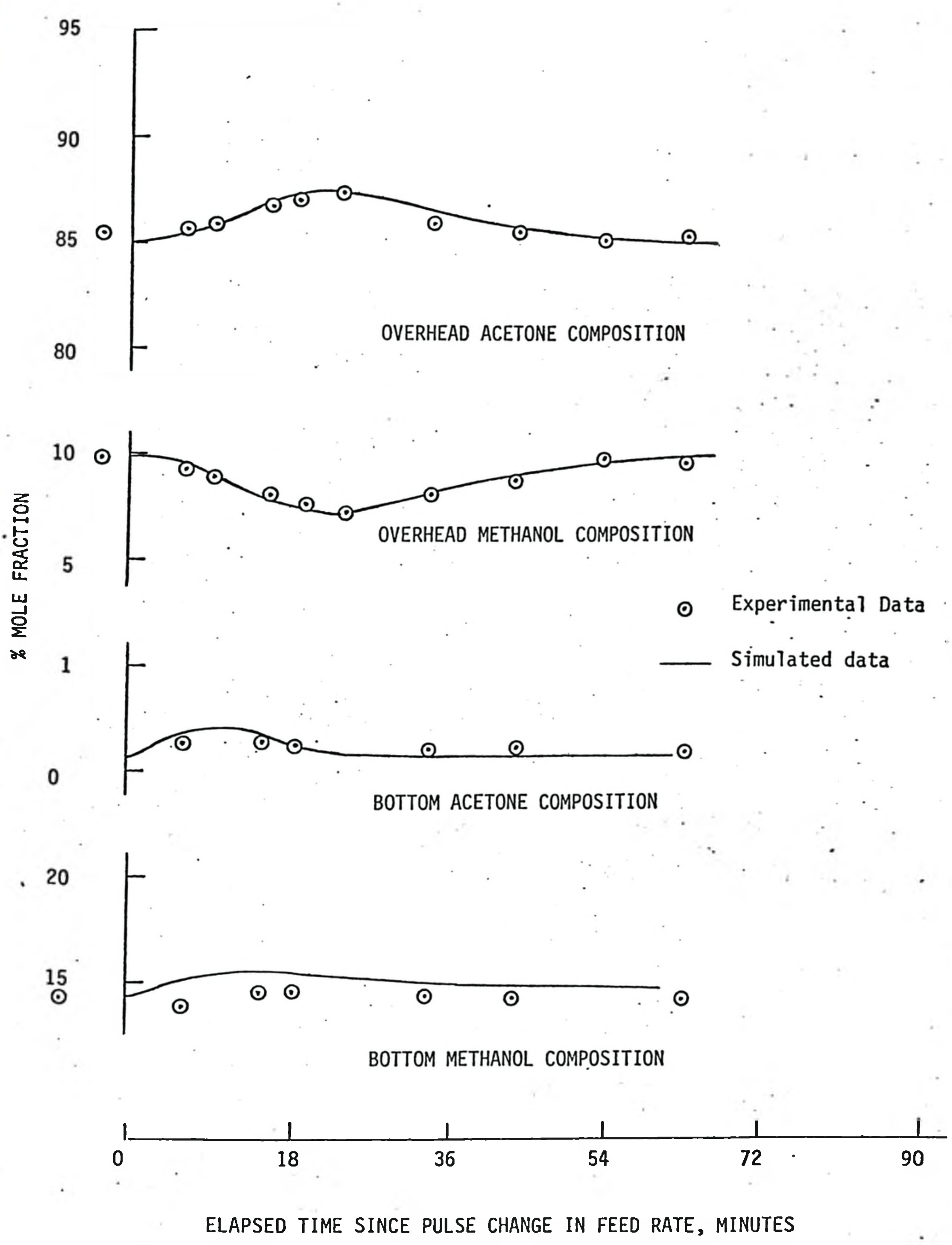
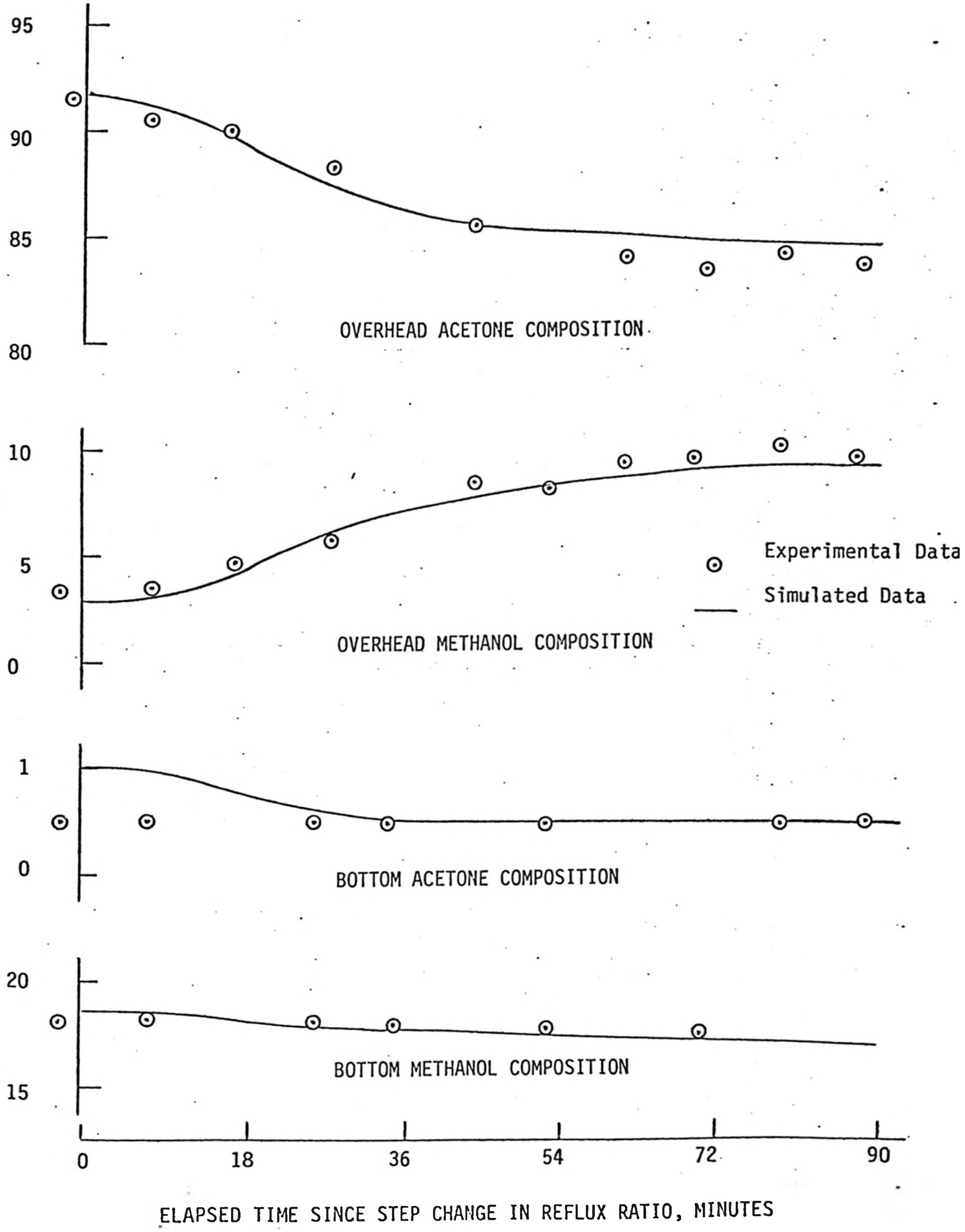


TABLE 4.3 OPERATING CONDITIONS

REFLUX RATIO AS DISTURBED VARIABLE

Quantity	unit	negative step change		positive step change		positive pulse change	
		initial	final	initial	final	initial	final
F	g moles/hr	16.37	16.37	16.37	16.37	16.37	16.37
$x_{F,1}$	mol.fract.	.5886	.5886	.5886	.5886	.5886	.5886
$x_{F,2}$	mol.fract.	.4060	.4060	.4060	.4060	.4060	.4060
T_F	°C	45	45	45	45	45	45
S	g moles/hr	30.75	30.75	30.75	30.75	30.75	30.75
$x_{S,1}$	mol. fract.	0.	0.	0.	0.	0.	0.
$x_{S,2}$	mol. fract.	0.	0.	0.	0.	0.	0.
T_S	°C	70	70	70	70	70	70
RR	dimensionless	1.756	1.558	1.558	1.756	1.756	1.756
REBDT	Kcal/hr	215	215	215	215	215	215
%Dist.ampl.	dimensionless	- 11.3 %		+ 12.7 %		+ 75%	

fig. 4.7 NEGATIVE STEP CHANGE TO REFLUX RATIO



ELAPSED TIME SINCE STEP CHANGE IN REFLUX RATIO, MINUTES

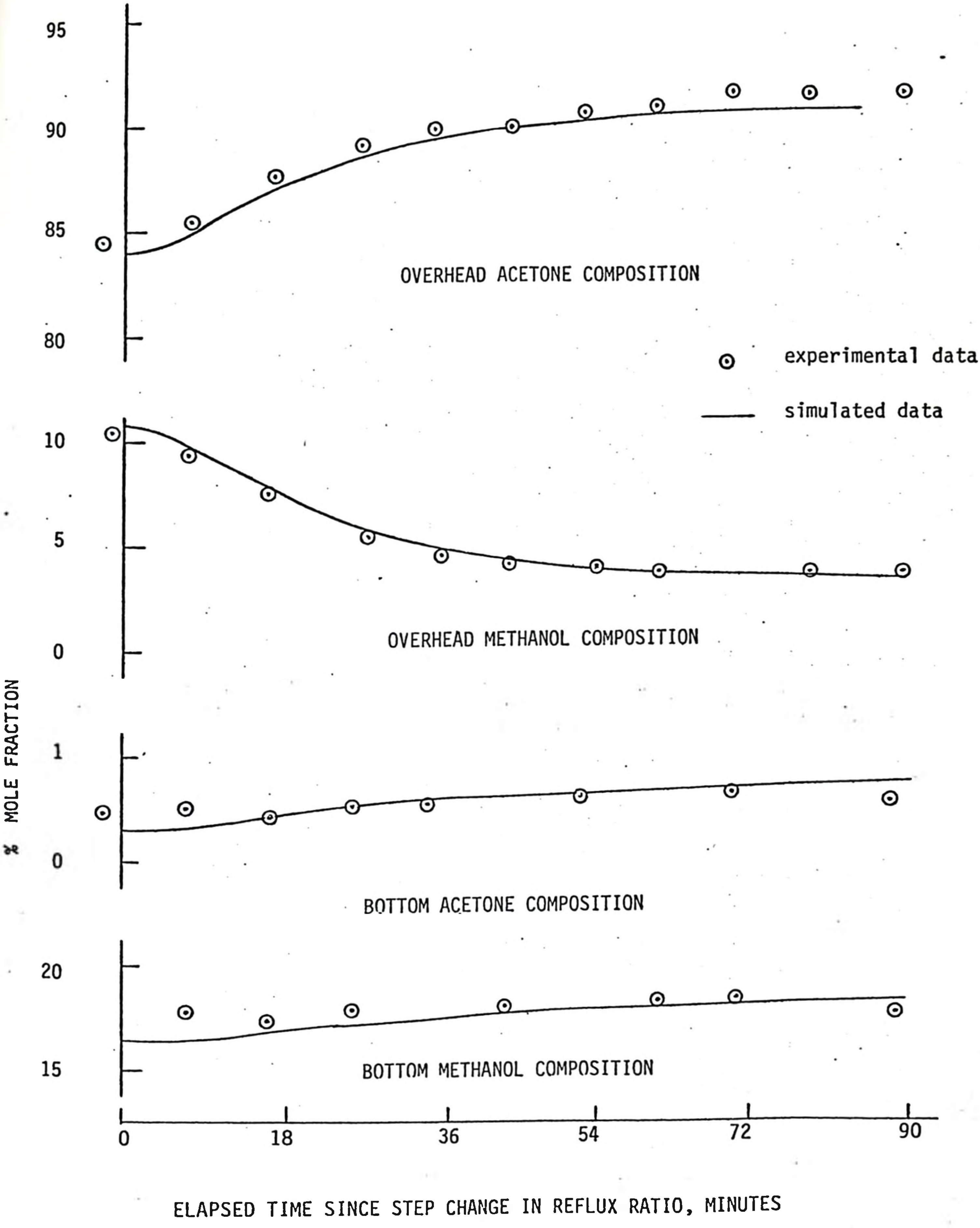
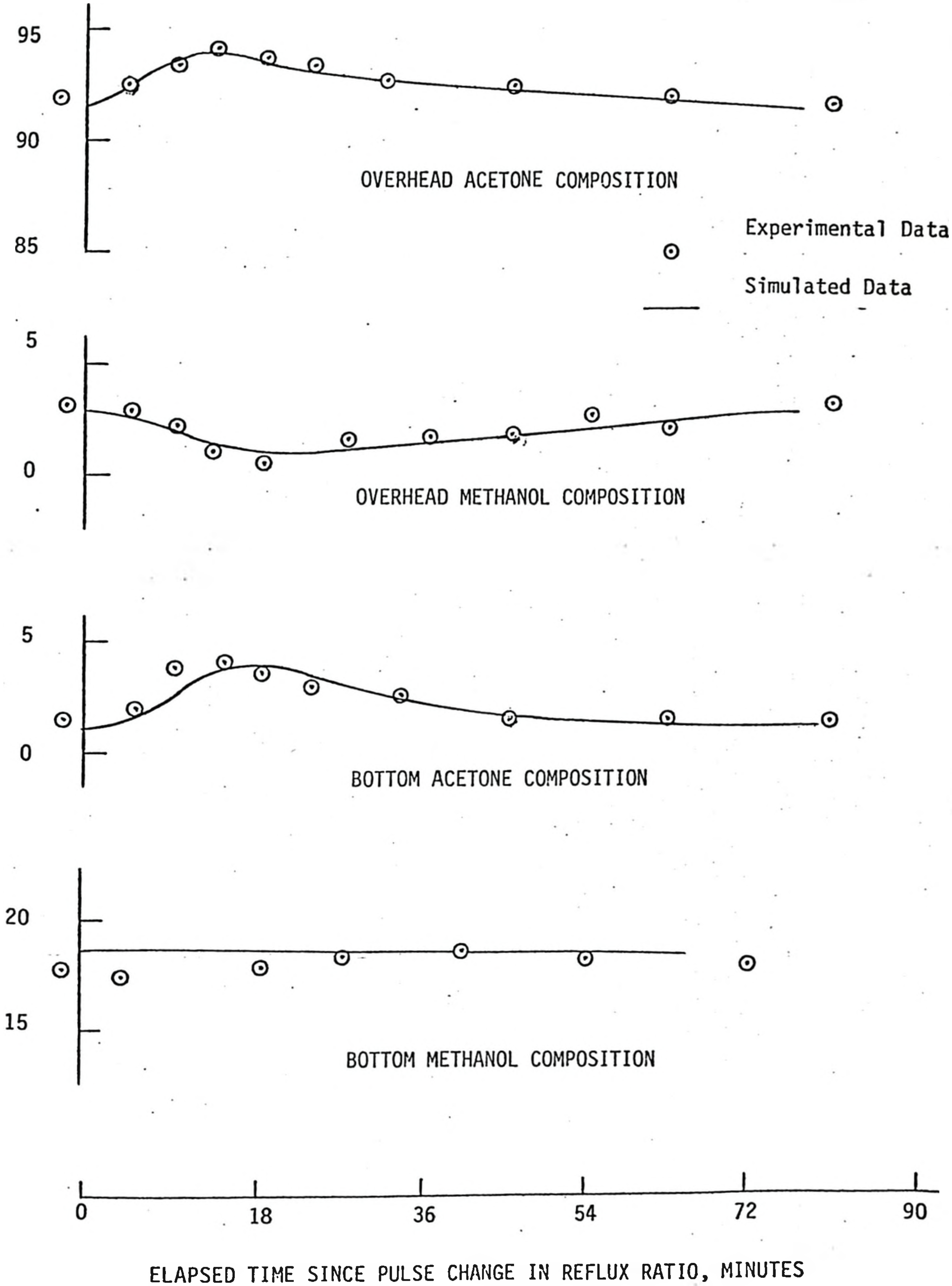


Fig. 4-8 POSITIVE STEP CHANGE IN REFLUX RATIO

fig. 4.9 POSITIVE PULSE CHANGE TO REFLUX RATIO



changes to the reflux ratio variable. The operating conditions are presented in Table 4.3.

The calculated results were compared to the available experimental data and were presented on fig. 4.7, 4.8, 4.9. A fairly good agreement was obtained at the overhead. However, inconsistencies were observed at the bottom of the column. These could be likely explained by the assumption of 100 per cent mixing the reboiler not being satisfied in practice.

4.4 Conclusion

The above study has shown that the dynamic behaviour of a fifteen-stage extractive distillation column can be described by simply laying out the basic mass, energy and equilibrium equations and by choosing an appropriate set of assumptions.The Murphree efficiencies at the stages were kept constant throughout the simulation and came from D. O'Neill's steady-state simulation work (19). The parameters in the enthalpic Wilson equation extended to multicomponent mixtures were taken from Bruin's study (20) about activity coefficient relations in miscible and partially miscible multicomponent systems. Only the parameter in the Francis weir relation was adjusted to obtain a fair consistency in the comparison of simulated data with experimental data. This relation was used to calculate the overflow height as a function of the tray characteristics and the liquid flowrate. A change in the overflow height will cause the total tray holdup to change, and a change in the total tray holdup will affect the model through the values of M_i in equations (3. .1.c), (3. .2.c), (3. .3.c) and (3. .4.c).

The most important feature of the simulation is that it describes the interactions between the fluid dynamic, heat transfer and mass transfer phenomena.

CHAPTER 5

DYNAMIC RESULTS - OSCILLATORY DISTURBANCES

This chapter presents an attempt to study the dynamic behavior of the column subject to oscillatory disturbances. The mathematical model has been found fairly adequate to represent the extractive column with disturbances such as step change and pulse change to feed variables. It was also mentioned that the model depends intimately upon the nature of the disturbance and the disturbed variable.

As a first step toward the identification technique using the black box concept, a test has been made with a square wave disturbance into the feed flowrate variable. Due to a delay in the equipment arrival and the problems encountered when hooking up the control valves as described in chapter 2, a preliminary simulation study of the effect of the frequency and the amplitude which characterize the square wave, was performed on the feed flowrate variable, using the previous simulation program with some minor modifications. The feed flowrate was chosen as the disturbed variable since experimental work could be done later to confirm the simulation study. The reason for choosing a square wave type of disturbance was to overcome the non-linearity problem of control valves in the actual operating range.

The control valves were set up and a reasonable range was determined with the adjustment of the pressure in the feed tank as mentioned in chapter 2. Due to lack of time, only one experimental run was performed with a square wave disturbance into the feed rate variable and the results have been compared with calculated results.

A brief survey about different approaches on unsteady operation of distillation columns is presented in the following section.

5.1 Literature review

Controlled-cycling operation of columns was generally understood as operation in which periods for dense phase to flow down and light phase to flow up are repeated alternately. Schrodt (36) explained qualitatively the improvement as followed: when two phases containing a transferable substance are brought in contact, the driving force for mass transfer is initially very high, than decreases to approach zero, close to equilibrium conditions; controlled-cycling conditions involve more fresh phases brought in contact, therefore, the time-average driving force is higher. McWhirter and Cannon (37) found an efficiency increase as high as 160 per cent on experimental distillation columns operating at total reflux.

A number of investigators such as Belter and Speaker (38) considered a cycle comprising periods for dense phase flow down the column, a coalescence stage with no flow, light phase flow up the column and again a coalescence stage with no flow. Efficiencies were obtained greater than 100 per cent and throughputs were 2 to 10 times greater than the steady-state case. Reference 35 is a useful source of information on controlled-cycling literature.

A more recent approach to unsteady-state operation of distillation process was analytically done by Matsubara et al. (39). Instead of dealing with cases in which a vapor flow period and a liquid flow period are repeated alternately, Matsubara et al. conceive an optimal cycle of operations which, basically, consisted in taking off the top product only when the composition of the more volatile component in the condenser is high and the bottom product only when the composition in the reboiler is low. The numerical study was applied to a binary distillation column with negligible vapor holdup, constant liquid holdup on each stage, identical vapor rate throughout the column, identical liquid rate throughout the enriching or the stripping sections of the column, plates with 100 per cent efficiency and with variable condenser holdup and reboiler holdup. Matsubara et al. concluded that if the maximal admissible vapor rate is large, periodic control required a much smaller number of stages relative to steady control, to attain the same separation efficiency.

Friedly (34) has also mentioned the possible improvement of non-linear processes when operated at unsteady-state conditions rather than at steady-state conditions. Lannus and Kershenbaum (33) noticed an improvement of the conversion in a tubular reactor when the inlet concentration was forced sinusoidally.

5.2 Simulation study

The extractive distillation column was simulated within the possible operating range determined by the work of D. O'Neill (10).

Any change in the feed flowrate involves a change in the fluid dynamics of the column. The same technique was used for a square wave input as was used for a step or pulse change in the feed flowrate. The flow increase was transmitted instantaneously down to the bottom output, changing the liquid flowrates coming out from the trays between the feed location and the bottom, therefore changing the holdup at the trays according to the Francis weir relation.

Fig. 5.1.a. shows the overhead response when a square wave with a period of 0.1 hr., an amplitude of 12% was put into the feed flowrate; the response was observed during a period of time of 2 hr. and gradual stabilization to a higher steady-state level for the acetone overhead composition seemed to occur.

Simulated runs were also performed with three different periods of 0.2 hr., 0.1 hr., 0.04 hr. at a constant amplitude level of 12%, to study the effect of the period on the overhead composition response. Simulation time was limited to 1.4 hr. for economic reasons. From fig. 5.1.b, one can observe three trends:

(1) The acetone composition did not stay at the steady-state level but increased from it substantially. The methanol mole fraction also decreased significantly.

(2) The larger the period is the greater the deviation from the initial steady-state level.

(3) The gradient of the deviation with respect to the period of the wave is not constant. The jump from curve C to curve B is much larger than from curve B to curve A although the increase in the period is almost the same.

The effect of the amplitude of the wave on the overhead response was also investigated by doing runs at the same period value of 0.1 hr. and three different amplitudes of 12, 9, 3 per cent successively. Effects of amplitude on the overhead response is shown in Fig. 5.1.c. It can be seen that in the range of amplitude considered (3% - 12%), the effect of amplitude is very small. This would strongly suggest that most of the improved performance possible by cycling can be obtained with only a very small amplitude (3%) and there is little incentive to go to larger amplitude.

fig. 5.1.a: OVERHEAD COMPOSITION RESPONSES TO SQUARE WAVE DISTURBANCE IN FEED FLOWRATE

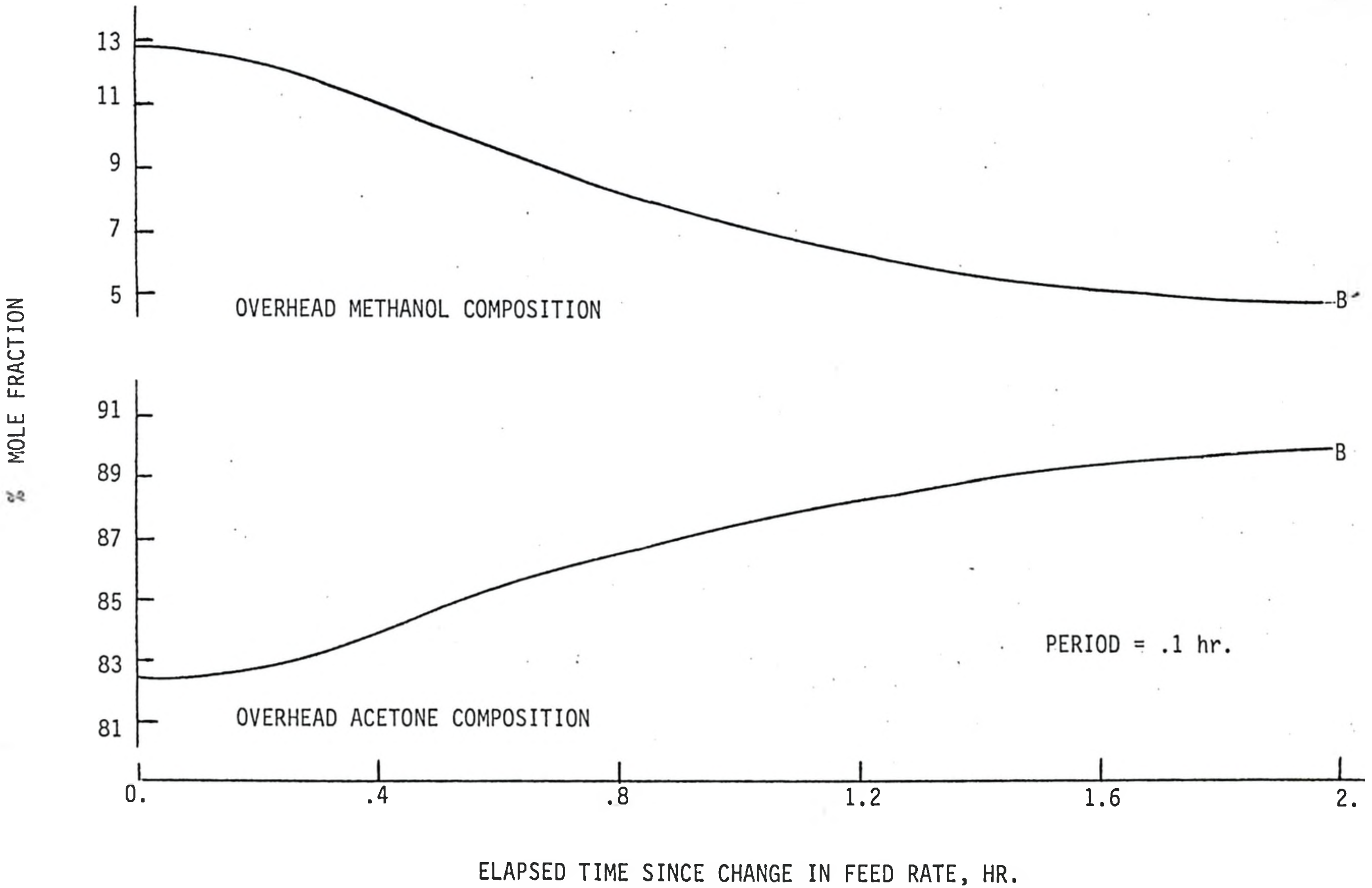


fig.5.1.b EFFECT OF FREQUENCY ON OVERHEAD COMPOSITION

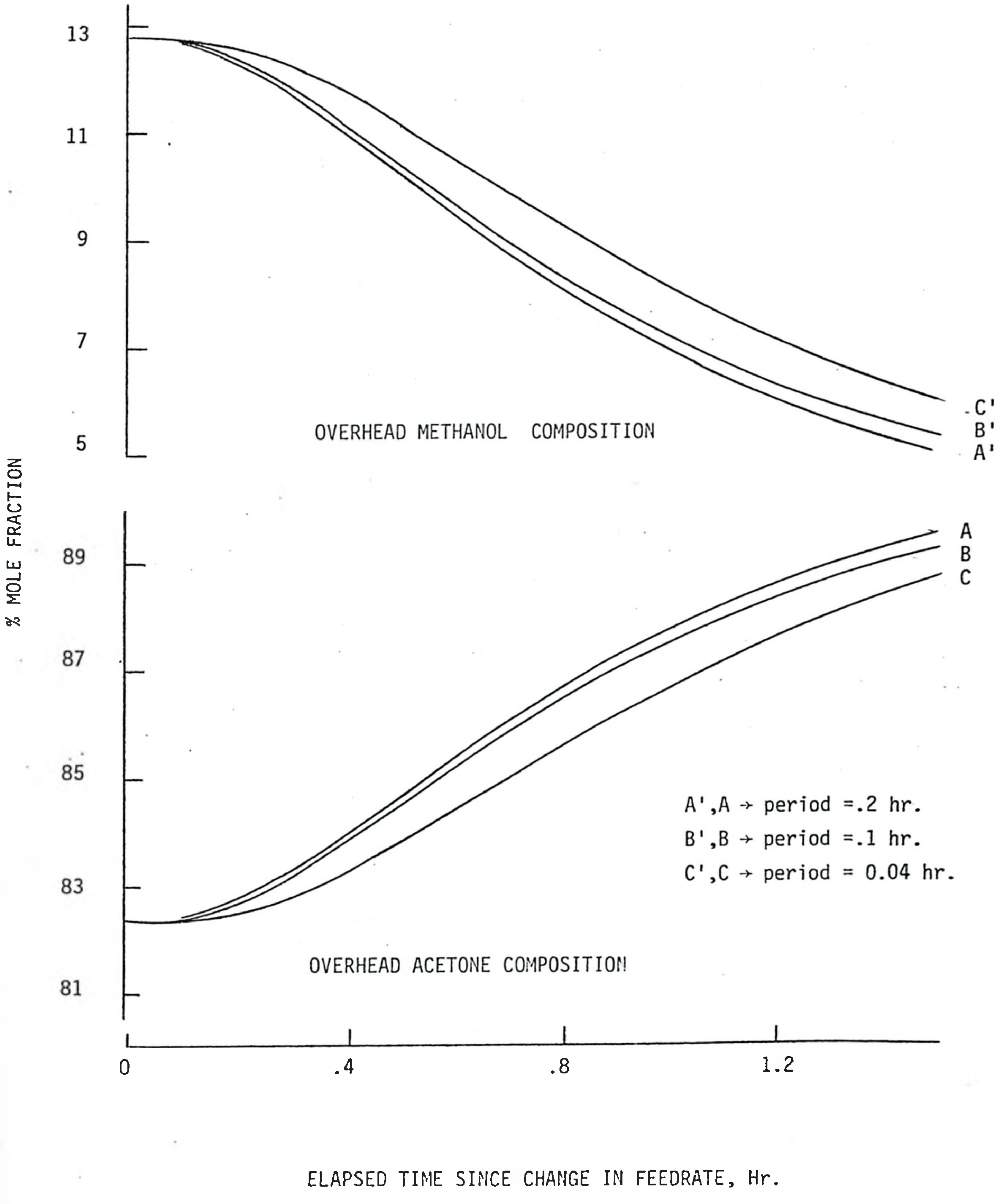
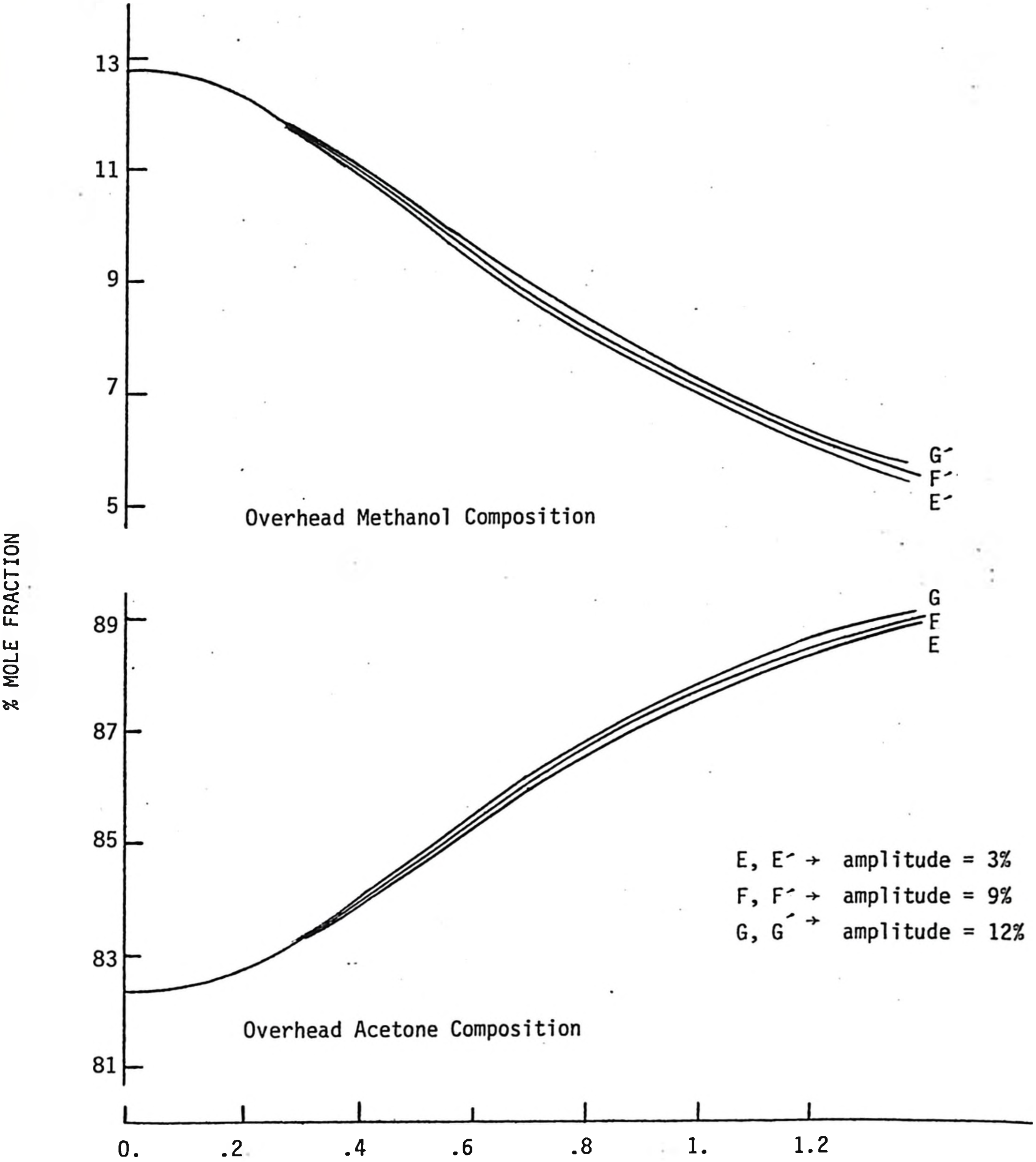


Fig. 5.1.c. Effect of Amplitude on Overhead Composition



ELAPSED TIME SINCE CHANGE IN FEEDRATE, Hr.

5.3 Comparison of experimental and simulated data for the case of a square wave disturbance to the feed flowrate

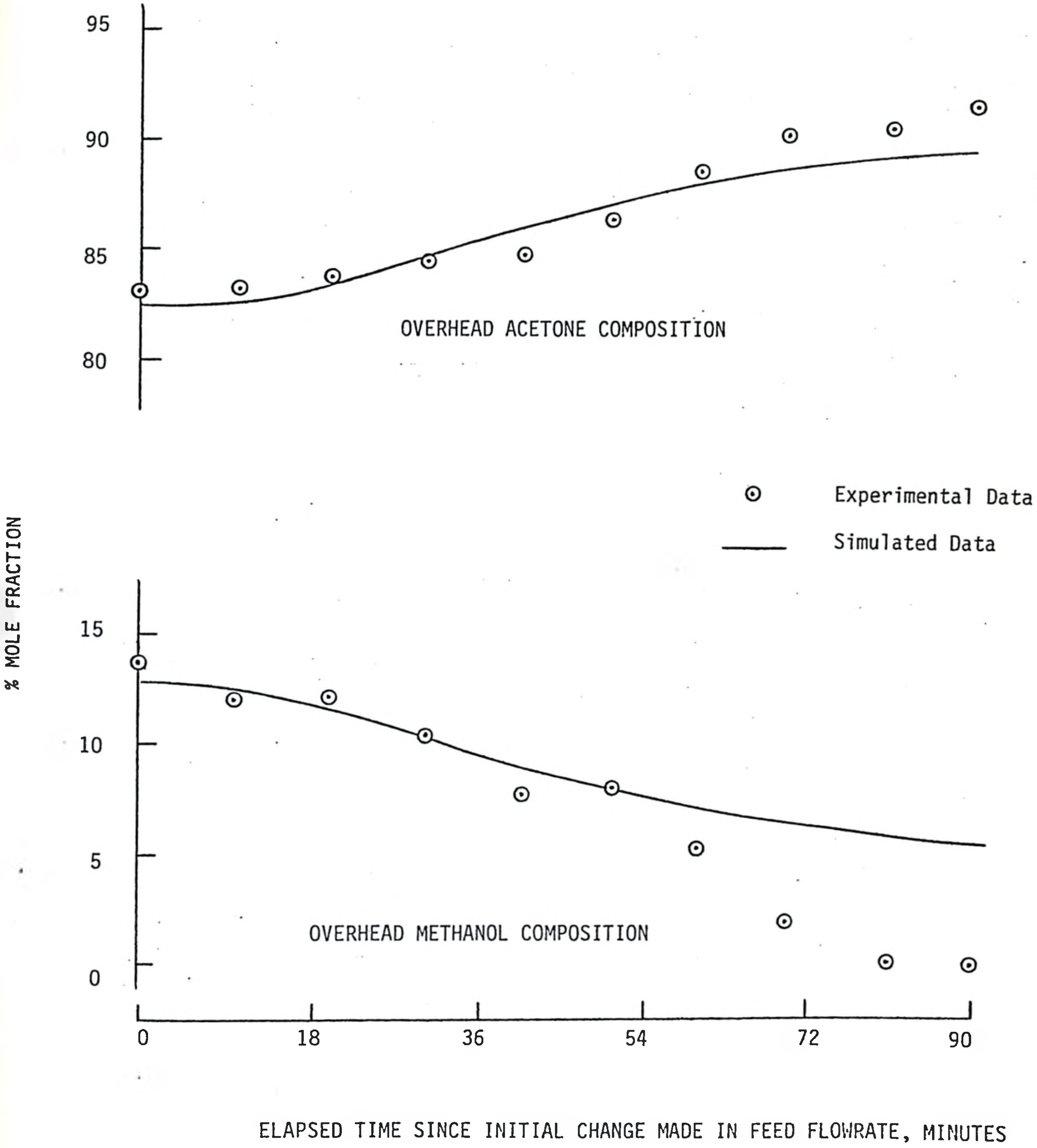
The experimental run was performed by disturbing the feed flowrate by a square wave of 12 per cent amplitude and at a period of .1 hr., through the control valve. As usual, the column was first brought to steady-state and then disturbed. Experimental conditions were presented in Appendix D. The transient response of the overhead composition was observed during a period of 1.5 hr., by taking samples every 10 minutes and analyzing with a gas chromatograph. The settings of the gas chromatograph and the integrator for the analysis can be found in Appendix E.

A comparison of the experimental data and the simulated data is presented in fig. 5.2. The following facts were observed:

(1) Experimentally, the column ceased operating at the initial steady-state level after the disturbance was put into the feed flowrate. The overhead acetone molar fraction increased in fair accordance with the simulated curve. A more detailed explanation of the response behavior by considering heat transfer, mass transfer, equilibrium, and fluid dynamics would be difficult.

(2) The experimental trend of the methanol composition agreed fairly well with the simulation during the initial period of about 1 hr., but the experiment showed that methanol almost disappeared from the overhead product at the end the response. Priming of the top trays was observed during this time. The overhead flowrate was decreasing significantly, and methanol could not reach the top due probably to tray weeping. However, water still appeared in the overhead product, which could be explained by the fact that

fig. 5.2. SQUARE WAVE DISTURBANCE TO FEED FLOWRATE



water was fed very close to the top of the column. Reboiler level increased constantly and has to be readjusted, possibly due to the fact that most of the feed was flowing down.

The simulation did not give the same results in the last section of the response since the constant holdup assumption was used throughout the simulated period of time.

It is suspected that instability of the column would have occurred if the experiment had been continued further, due to the consequences of tray weeping.

5.4 Conclusion

This chapter presented a preliminary study on the response of the column due to controlled-cycling disturbances to feed flowrate variable. The preliminary simulated results were found interesting in the sense that the efficiency seemed to improve substantially when the feed flowrate was cycled with a symmetric square oscillation.

The calculated effects of cycling the feed rate are, thus, at least qualitatively consistent with previous results found in the literature, which have shown considerable improvements in distillation column performance under oscillatory conditions. Furthermore, this work has been done on a multicomponent system as opposed to a binary system, usually mentioned in the literature. Further experimental work is needed on this effect, in particular on possible hydrodynamic instabilities introduced by the controlled-cycling of the feed flowrate (see chapter 6).

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK

A mathematical model was developed to describe the dynamics of a fifteen-stage extractive distillation column using a mixture of acetone methanol and water as the solvent. The unsteady-state component material balance equations were used together with the Enthalpic Wilson equation extended to a multicomponent mixture in the equilibrium calculations. Since the fluid dynamic response of the column was rapid relative to the temperature and composition responses, a quasi-steady state assumption was made in the simulations for the step and pulse changes in feed composition, feed rate and reflux ratio variables. The technique used when the column was subject to a sudden change in the flowrate (e.g. changes in the feed flowrate variable or reflux ratio variable) consisted in satisfying, at the beginning of a change, the total tray material balances by correcting instantaneously the liquid flowrates down the column and therefore, the holdups on the trays. The comparison with available experimental data was found fairly consistent within the possible operating range of the column and showed that a reasonable simulation has been obtained from the assumptions and strategy used.

The simulation of the column was developed inside a DYNSYS 2.1 environment which offered to the user many advantages and flexibilities, e.g. computer time savings, overhead work. Chapter 3 described the desirable features of such an approach.

The real time software developed by Jackson (11) was totally redeveloped to be used with the Generalized Operating System Executive created by J.P. Tremblay (22). The new package consisted of basic data logging tasks, unit conversion tasks, inlet temperature control tasks and inlet flowrate control tasks. A routine to perform a square wave disturbance was also available.

The rearrangement of the feed and solvent system with the addition of the control valves allows a number of possibilities for future work. One of the areas briefly investigated in this work was the study on the dynamic response of the column (overhead composition) when it was subject to controlled-cycling disturbances. A preliminary simulation study showed that the overhead composition in acetone improved significantly when the feed flowrate was disturbed by a square wave. The higher the frequency of the oscillatory disturbance, the smaller was the improvement, in the range of conditions mentioned in chapter 5. This point should be interesting for further future investigation since Lannus and Kershenbaum (33) revealed an improvement in the conversion when the inlet concentration was forced sinusoidally, in the case of a tubular reactor. Also, as stated by Friedly (34), processes which are not strictly linear can sometimes be operated better in the unsteady state than in the steady state; periodic processing takes advantage of the fact that system non linearities can make the average system output different from the steady-state values. A distillation process is one of the many non-linear systems where the non-linearity comes essentially from equilibrium relations. The effect of the amplitude of the periodic wave was also investigated and was found negligible within the operating condition range of the column.

A comparison of the simulated results with the preliminary experimental data was made in chapter 5. A fairly good agreement occurred during the first half of the response, but divergences dominated at the later times. Signs of instability were noticed experimentally, such as top tray weeping, reduction of overhead product flowrate. Since strongly-supported experimental results were not available, the answer to the question "is instability caused by controlled-cycling disturbances in feed flowrate?" will be worth investigating later on.

The following points should be considered for future work.

(1) As discussed in Chapter 2, the feed and solvent arrangement is very inconvenient because of the long time spent in charging the tanks through the small hole at the top. Furthermore, no feed could be added while the column is operating. The storage system should be revamped to rectify the above difficulties. Also, it is desirable that the pressures in the feed tanks and in the solvent tank can be set independently. A method of setting the desired pressure at each run is discussed in Chapter 2.

(2) The software to control the input stream flowrates was developed. The constants of the controller could be determined by trial and error runs. Thereafter, the perturbation could be achieved with closed-loop control, in other words, just by perturbing the controller set point.

(3) A brief review of the literature showed that unsteady-state operation of non-linear processes could improve the performance compared to the steady-state operation. This fact is of great interest for the promising results obtained by the simulation study described in Chapter 5 of this work, when the feed flowrate was disturbed by a square oscillation, and the increase in the overhead acetone composition observed. However, more experimentation must be performed to better explain and analyze the instabilities which occurred during the preliminary investigation in this project.

(4) Since a reflux divider was used, the column, in reality, was subject to an oscillatory change in the reflux flowrate. In the simulation study on controlled-cycling changes in the feed flowrate, the frequency of the oscillation was found to affect significantly the response of the column. Therefore, the

frequency of the reflux divider cycle, which also affected, in a rigorous sense, the fluid dynamics of the column, might have an effect on the response and should be checked in future work (reflux divider period ≈ 5 sec; smallest period investigated ≈ 150 sec) of course the same reflux ratio could be achieved with different frequencies for the reflux divider cycle.

(5) It was mentioned before that the accuracy needed for a simulation model depends upon the objectives of the simulation work. However, another potential direction for further investigation of extractive distillation modeling will be the study of the effects of various assumptions on the model accuracy. As an example, constant volumetric holdup at the stages could be relaxed and the percentage inaccuracy could be analyzed.

(6) A future requirement before any significant control study can be taken will be the implementation of an on-line automatic analyzer.

An outline of the progress of this long-term project is described below:

D. Jackson (11) :
- Basic process instrumentation
- Interfacing with minicomputer (Real Time Operating System, original version)

D. O'Neill (10) :
- steady-state operating conditions determination
- steady-state simulation program
- preliminary study of dynamic behaviour (experimental)

Present work :
- dynamic simulation for step and pulse changes in feed composition, feed flowrate and reflux ratio variables, using DYNSYS 2.1
- Addition of control flexibility for inlet streams
- Modification of real time software to be used with GOSEX

- Preliminary dynamic and experimental study on controlled-cycling effects on column (square oscillation in feed flowrate)

Future work anticipated :

- Effect of reflux-divider cycling on the response of the column
- Further investigation on the effect of a square-wave disturbance in the feed flowrate (experimental and simulation studies)
- Basic assumptions checking in the model development for the extractive distillation column.

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APPENDIX A

CALIBRATION OF THE FEED AND SOLVENT LINES

Calibration curves have been obtained for the feed line and the solvent line. The pressure in the tanks was set at 15.6 psig.

For the feed line, within the possible limits of operation as determined by O'Neill (10), i.e. between 14 ml/min and 18 ml/min, the calibration curve of the feed flowrate vs the voltage read from the D.P cell, expressed in A/D units, could be described approximately a straight line, curve (2), whose equation is:

$$F = 0.1098\ V - 15.7088 \quad \text{with } F \text{ in ml/min}$$
$$V \text{ in A/D units}$$

Curve (1) is the calibration curve of the valve action vs the flowrate, and could be approximated by:

$$P_F = 10.832\ F^2 - 316.492\ F + 2500 \text{ with } F \text{ in ml/min}$$
$$P_F \text{ in D/A units}$$

The same curves were obtained with the solvent line. Solvent flowrate versus the voltage from the DP cell, expressed in A/D units, is represented by curve (4) and the valve action versus the flowrate is represented by curve (3).

for curve (4) : $S = 0.13331\ V - 16.0023$

for curve (3) : $P_S = -5.898\ S^2 + 141.457\ S - 440$

fig. A-1 CALIBRATION CURVES FOR FEED STREAMLINE

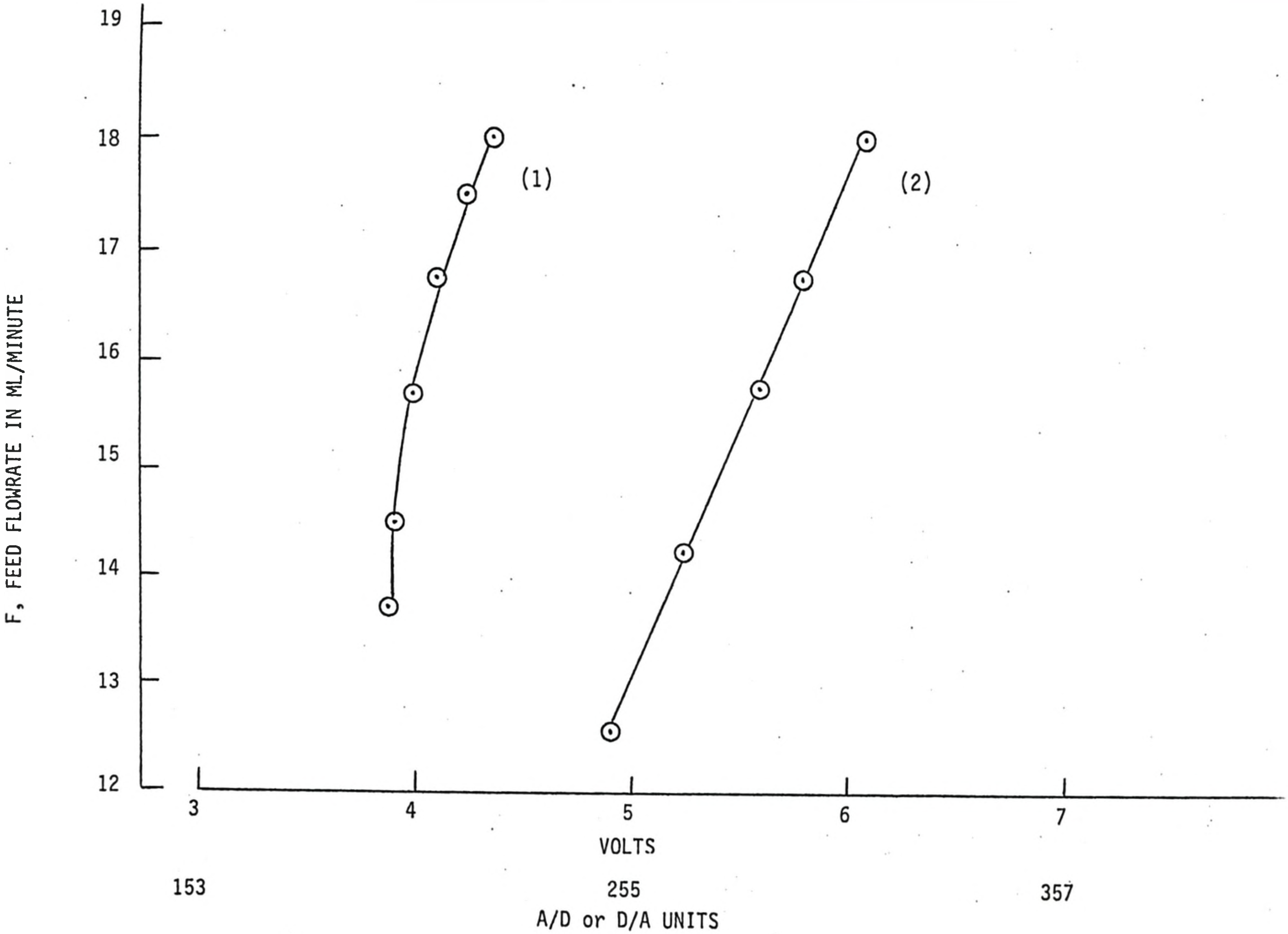
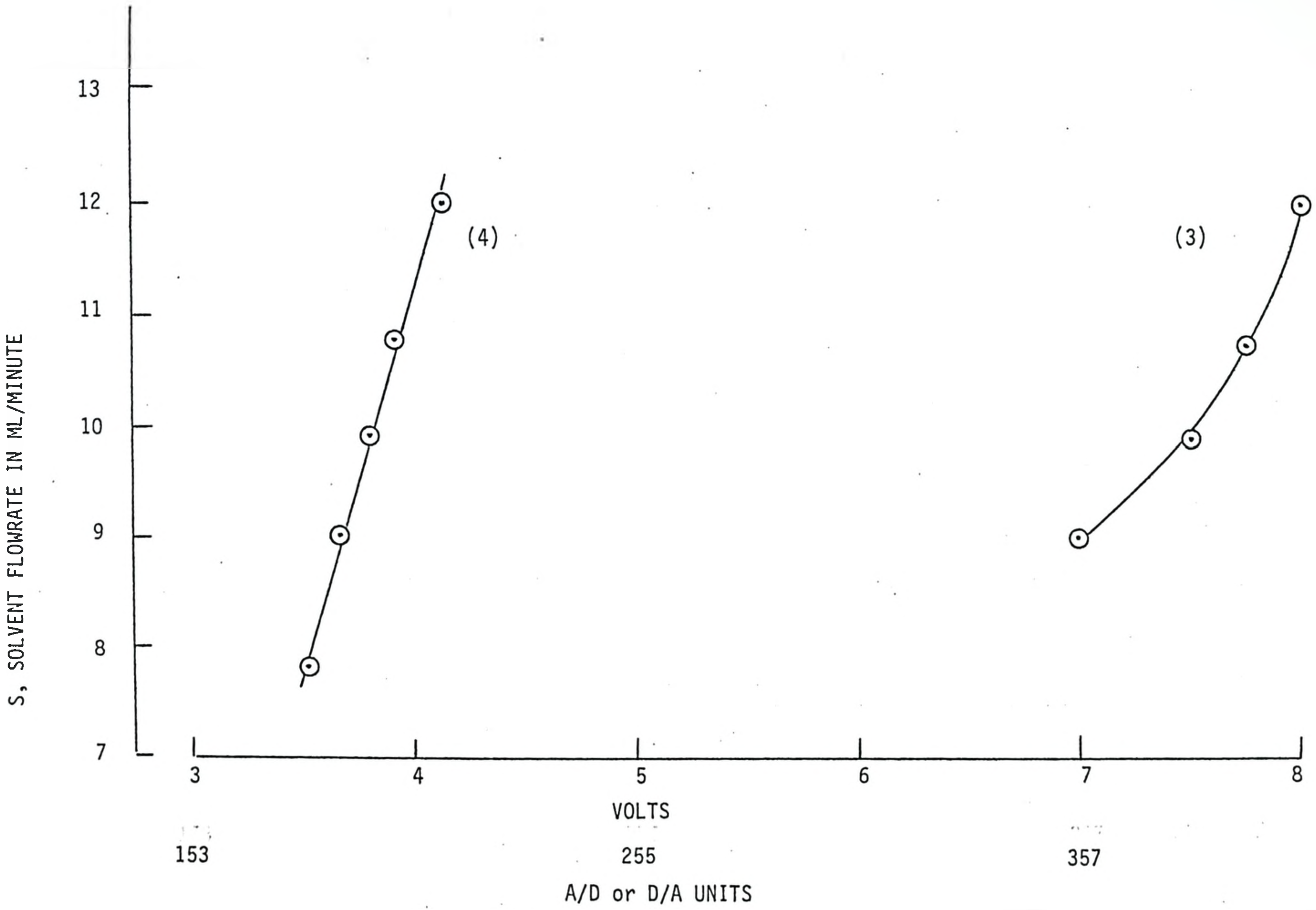


fig. A-2 CALIBRATION CURVES FOR SOLVENT STREAMLINE



- with S in ml/min

P_S in D/A units

V in A/D units

A least squares fitting technique was used.

APPENDIX B

DYNSYS DATA SET

1 Simulation Data

The simulation data begins with the data word "BEGIN". All cards before this are comments and are ignored by DYNSYS.

For example

IN/OUT	3.0	sets the maximum number of streams entering or leaving any module to three (default = 5)
HMAX	0.01	sets the maximum integration set size to 0.01 (default = 0.05)
TIME	3.0	sets the time of simulation to 3.0 (default = 10.0)
LIBRARY	1.0	means that one new module will be defined immediately after this card. Most common modules such as CONT1, STIR1 and VALV1 are already in a library of module names stored inside DYNSYS.
DELAY	9.0	DELAY is the new module being defined. It will be accessed as SUBROUTINE TYPE9

Other possible simulation data words are:

COMPS	X	sets the number of components to X (default = 1)
DELTAT	X	sets the initial integration step size to X (default = 0.00001)
HMIN	X	sets the minimum integration step size to X (default = 0.000001)
TOLERANCE	X	sets the error tolerance for the integration to X (default = 0.001)
ORDER	X	sets the maximum permissible order of integration to X (default = 6)
NONSTIFF		causes the nonstiff option to be used (default = stiff option)
MINPIVOT	X	sets the minimum permissible pivot value for TRGB-TRGB2 to X (default = 0.000001)
PRINTING	X	causes output to be printed every X time steps (default = 1)
LINEPLOT		causes output to be printed as a lineplot (in this case additional data must be given after physical properties data)
GRAPH		causes output to be printed as graph or graphs on CALCOMP plotter as well as Lineplot. LINEPLOT must also be used.

FUNCTION	X	causes X vectors of stream information to be read in as an input function. Stream data follows immediately.

The data word "PROCESS" signifies the beginning of the equipment data and thus the end of the simulation data.

.2 Equipment Data

The data word "PROCESS" signifies the beginning of the equipment data. Here data for the various modules is written in their order of execution.

CONT1	1.0	the first module to be executed is CONT1 , referred to as module #1 in the dynamic information flow diagram

After this, the entering and leaving stream numbers are given. A positive number indicates an input stream while a negative number indicates an output stream.

3.0	-4.0	stream 3.0 enters CONT1 and stream 4 leaves it.

The following card gives the equipment parameters for the module. This is the data which describes the individual units.

```
3.      1500.      1000.     2.     3.
```

For example, the set point for the controller is the third parameter 1000. Up to five parameters may be given. If more are required, the data word "EXTRA" X must follow and the X extra parameters appear on the following data cards.

We continue in this way for all modules. The data word "END" signifies the end of the equipment data.

3 Stream Data

The material flows between equipment are represented by material streams. Information about each stream is stored in the following vector.

position 1 - stream number
2 - flag
3 - total mass flow
4 - temperature
5 - pressure
6 - mass fraction of component 1
7 - mass fraction of component 2, etc.

The stream flag serves the dual purpose of identifying the stream type (flags with absolute value < 10 are material flows, those > 10 are information flows) and, if negative, suppresses printing of the stream vector in the output.

The user may choose his own system of units as long as he is consistent throughout. Usually English units are employed.

STREAMS	5.	means that initial information data will be given for five streams.
EXPLICIT		means that the data will be given explicitly.

The following cards give the data. For the first stream

```
1.     1.     1000.     100.     14.7
1.
```

the stream number is 1. The stream is a material flow (the flag is less than 10). The flow rate, temperature and pressure are 1000., 100. and 14.7 respectively. There is only one component so its mass fraction is 1.0.

Data is given for five streams. The data word "END" signifies the end of the stream data.

4 Physical Properties Data

PROPERTIES -1. means that the physical properties of water are assumed.

The user may provide his own data if desired. See the DYNSYS Manual for details.

The data word "END" signifies the end of the physical properties data and in this case the end of the data set.

APPENDIX C

Francis Weir Relation Used to Calculate the Liquid Tray Holdups

As mentioned in chapter 3, section 3.3.1 the assumption of constant volumetric holdup on the trays was made.

From the steady state simulation program, one can obtain the initial liquid molar flowrates leaving each stage. These flowrate values have been used with the Francis Weir relation to determine the initial volumetric holdups which will be kept constant throughout the dynamic simulation in this work. As described in reference 19 of Luyben, the liquid flowrate F_L is proportional to the weir length WL and to the overflow height h raised to a power of 1.5:

$$F_L = a\,(WL)(h)^{1.5}$$

By rearranging

$$h = \left(\frac{1}{a}\,\frac{F_L}{WL}\right)^{2/3}$$

with a = 11988

F_L in ft³/hr, liquid flowrate leaving the tray.

WL in ft, weir length.

h in ft, overflow height of tray holdup.

The holdup height was then calculated:

$$H = h + WH \qquad \text{WH in ft, weir height}$$

The volumetric holdup could be then obtained.

APPENDIX D

Run with Square Wave as Input Disturbance in Feed Flowrate

Operating conditions

Feed rate:	high level	=	18 ml/min.
	low level	=	16 ml/min.
	period of oscillation	=	6 min.
Feed composition:	60% mole acetone		
	40% mole methanol		
Solvent rate:	9.5 ml/min.		
Solvent composition:	pure water		
Reflux ratio:	1.772		
Solvent temperature:	65°C		
Feed temperature:	40°C		
Feed variac setting:	15		
Solvent variac setting:	27		
Reboiler variac setting:	70		

APPENDIX E

Gas Chromatograph and Integrator Settings

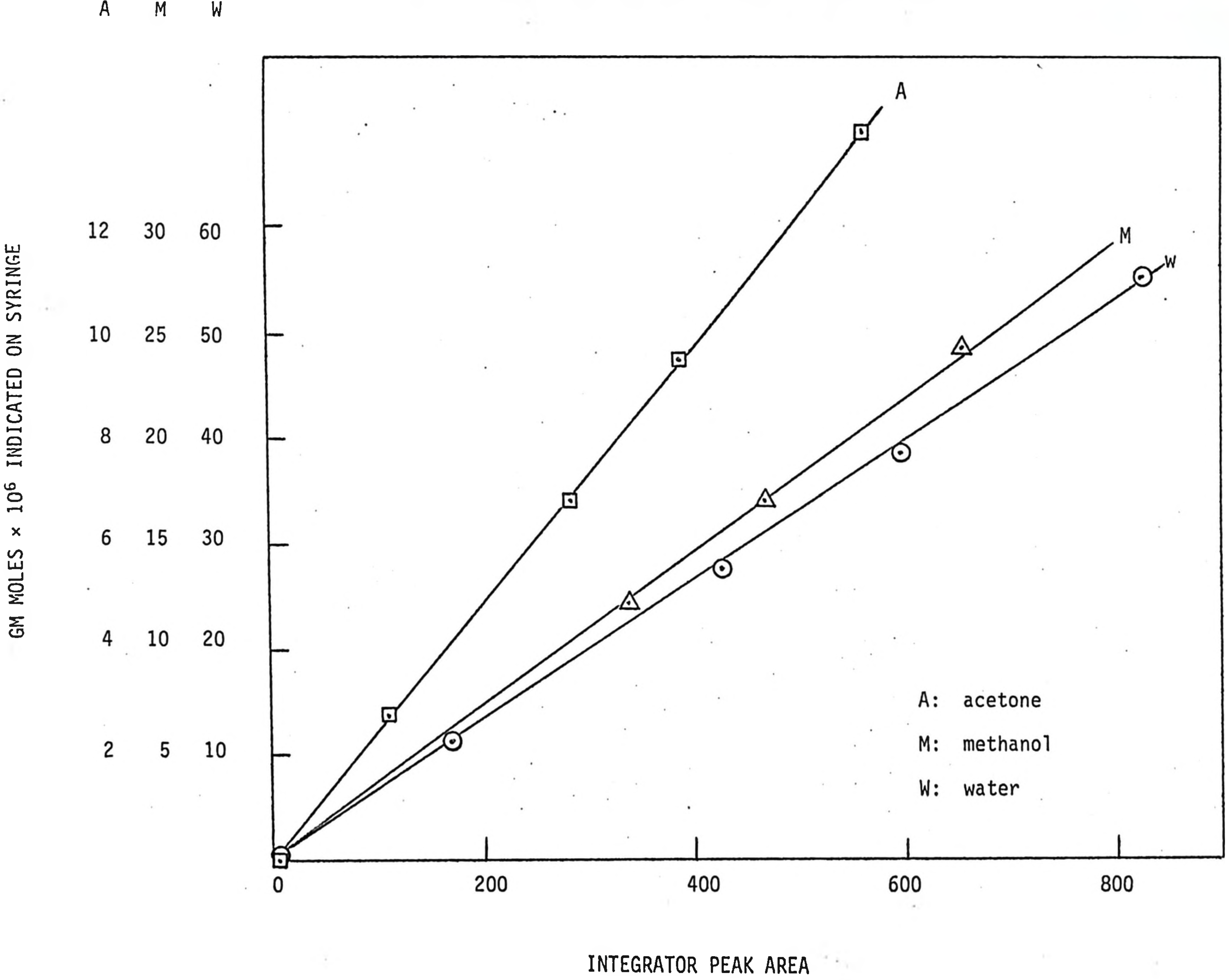
Gas Chromatograph Varian 90-P3

-column - 6' x 1/4" O.D. stainless steel tube
    - Porapak Q, 80 mesh, packing
-carrier gas - Helium @ 65 psig
    - carrier flowrate = 60 ml/min.
    - reference flowrate = 20 ml/min.
-dial settings - column power = 55
    - detecter power = 45
    - injector power = 50
    - attenuatoz = 1
-operating temperatures - column - 175°C
    - detector - 240°C
    - injector - 216°C
    - collector - 212°C

Integrator Hewlett-Packward 3370B

-settings - noise suppresion - maximum
    - slope sensitivity - up = 0.03
    - down = 0.03

GAS CHROMATOGRAPH CALIBRATION USING
PURE COMPONENTS AND VARIABLE SAMPLE SIZE



-B.L. reset delay = 0.5 minutes

-area threshold = 100

-shoulder control - front - on

- rear = 100 mV

APPENDIX F

Dynamic Simulation Programs

A number of utility routines have been developed to assist the dynamic simulation in this work.

ENTHV:	this function calculates the enthalpy of a vapor stream at a specified temperature, pressure and composition
ENTHL:	this function calculates the enthalpy of a liquid stream at a specified temperature, pressure and composition, for the mixture of acetone-methanol-water, taking into account the heat of mixing of the components
ACTY:	this function calculates the coefficient of activity of each component in a mixture of three component using the enthalpic Wilson equation extended to multicomponent mixtures, as described in section 3.3.2.2.
KVAL:	this function calculates the activity of each component in a mixture of components, as described in section 3.3.2.2.

The listings of the above utility programs and the main subroutine Type 1 are presented on the following pages.

```
      SUBROUTINE TYPE1
C     ***************************************************************
C     *                                                             *
C     *                                                             *
C     *                                                             *
C     *              EXTRACTIVE DISTILLATION COLUMN                 *
C     *                                                             *
C     *                                                             *
C     ***************************************************************
C
C
C     BY D.T.TRANG , CHEMICAL ENGINEERING DEPARTMENT
C     MCMASTER UNIVERSITY , OCTOBER 1975
C      THIS SUBROUTINE SIMULATES THE DYNAMICS OF AN EXTRACTIVE
C     DISTILLATION COLUMN WITH THE FOLLOWING CHARACTERISTICS =
C
C
C     13 TRAYS+CONDENSER+REBOILER
C     EXTRACTANT= WATER
C     SOLUTION= ACETONE-METHANOL
C
C
C     THIS SUBROUTINE IS DEVELOPED TO BE USED WITH THE DYNAMIC
C     SIMULATION EXECUTIVE DYNSYS 2.1.
C
C     (REFERENCE=BARNEY,J.R.,PH.D.THESIS,≠DYNAMIC SIMULATION
C     OF LARGE STIFF SYSTEMS IN A MODULAR SIMULATION FRAMEWORK≠,APRIL
C
C
C
C     ALGEBRAIEC EQUATION SOLVER=    SIMULT AND TRIDAG
C
C
C     DYNSYS 2.1 DATA SET
C
C     EQUIPMENT STREAM INPUT,OUTPUT
C          MP(IM,3)   LIQUID SOLVENT FEED STREAM
C          MP(IM,4)   LIQUID FEED STREAM
C          MP(IM,5)   OVERHEAD PRODUCT STREAM
C          MP(IM,6)   BOTTOM PRODUCT STREAM
C
C
C     EQUIPMENT PARAMETERS
C
C          EP(IM,1)   TOTAL NUMBER OF TRAYS(INCLUDING REBOILER AND CON
C          EP(IM,2)   FEED TRAY                                     NSER
C          EP(IM,3)   SOLVENT TRAY
C     EP(IM,4)   REFLUX RATIO
C     EP(I,,5)   REBOILER DUTY KCAL/HR)
C     EX(MP(IM,NMP+1)+I-1)      STAGE I MURPHREE EFFICIENCY FOR METHANOL
C          EX(MP(IM,NMP+1)+I+19)    STAGE I MURPHREE EFFICIENCY FOR WA
C     EX(MP(IM,NMP+1)+40)   CONDENSER HOLD UP (CM3)                   EF
C     EX(MP(IM,NMP+1)+41)   REBOILER HOLD UP (CM3)
C     EX(MP(IM,NMP+1)+42)   WEIR HEIGHT (FT)
C     EX(MP(IM,NMP+1)+43)   WEIR LENGHT (FT)
C     EX(MP(IM,NMP+1)+44)   COLUMN DIAMETER (FT)
C
```

```
C
C     STREAM DATA
C
C        S( , ,1)    STREAM NUMBER
C        S( , ,2)    STREAM FLAG
C        S( , ,3)    TOTAL MOLAR FLOW (GMOLES/HR)
C        S( , ,4)    TEMPERATURE (DC)
C        S( , ,5)    PRESSURE (MM HG)
C        S( , ,6)    MOLE FRACTION ACETONE
C        S( , ,7)    MOLE FRACTION METHANOL
C        S( , ,8)    MOLE FRACTION WATER
C
C     ***********************
C     *********************************
C     ******************************************
C
C     SYSTEM STORAGE SPACE
C
      REAL MW
      COMMON/MAT/MP(5,8),EP(5,5),S(2,50,10),EX(100)
      COMMON/CON/IG,NCOMP,NC5,H,NE,NS,NPR,NPOL,TMAX,IORDER,NGRAPH
      COMMON/UNIT/IM,NMP
      COMMON/GEAR2/EPS,TIME,KFLAG,JSTART,NBVMAX,ICONV,ISTIFF
      COMMON/MODULE/IDERY,ITER,ITRI,MC,IPIVOT
      COMMON/SUBDI/A(50)
      COMMON/DIAG/B(50)
      COMMON/SUPERD/C(50)
C
C     SUBROUTINE STORAGE SPACE
C
      DIMENSION VHUP(20),HUP1(20)
      DIMENSION ETHBAL(20),VDIF(20)
      DIMENSION Y(100),DERY(100)
C
C
C     SECTION 1
C     INITIATION AND PARAMETER CALCULATION
C
C
C     USE NEWTON-RAPHSON ITERATION WITH STIFF OPTION(ITER=1)
C
C
      ITER=1
      IF(JSTART .NE. 0 .OR. IG .EQ. 1) GO TO 200
C
C
C     TRANSFER INFORMATION INTO SUBROUTINE SPACE
C
C     VHUP(15)       REBOILER VOLUMETRIC HOLDUP
C     VHUP(1)        CONDENSER VOLUMETRIC HOLDUP
C     WH             WEIR HEIGHT
C     WL             WEIR LENGTH
C     DIA            TRAY DIAMETER
C     RR             REFLUX RATIO
C     NT             TOTAL NUMBER OF TRAYS
C     NFT            FEED TRAY
C     NST            SOLVENT TRAY
```

```
C
      VHUP(15)=          EX(MP(IM.NMP+1)+41)
      VHUP(1)=         EX(MP(IM,NMP+1)+40)
      WH=EX(MP(IM,NMP+1)+42)
      WL=EX(MP(IM,NMP+1)+43)
      DIA=EX(MP(IM,NMP+1)+44)
      RR=EP(IM,4)
      REBDT=EP(IM,5)
      NT=EP(IM,1)
      NFT=EP(IM,2)
      NST=EP(IM,3)
C
C     ADDITIONAL PARAMETERS
C
      LEVEL=1
      MN=1
      CYCLEF=.1
      TIMP=FLOAT(MN)*CYCLEF
C
C     CALCULATE THE AREA OF ONE TRAY
C
C
      TAREA=3.14/4.*((DIA-WL)*30.48)**2
C
C     NEQ IS THE NUMBER OF ORDINARY DIFFERENTIAL EQUATIONS
C
      NEQ=30
C
      MC=50
C
C
C     CLEAR THE JACOBIAN STORAGE
C
C
      DO 99 I=1,MC
      A(I)=B(I)=C(I)=0.
 99   CONTINUE
100   CONTINUE
C
C
C     CHECK TOTAL MASS BALANCE
C
      NTM1=NT-1
      S(2,NT+20,3)=S(2,NTM1+20,3)-S(2,NT,3)
      DO 17 I=2,NTM1
      J=NT+1-I
      BAL=S(2,J+20-1,3)+S(2,J+1,3)-S(2,J,3)-S(2,J+20,3)
C
C     SPECIAL TERMS FOR FEED AND SOLVENT TRAY
C
      IF(J.EQ.NFT)BAL=BAL+S(2,18,3)
      IF(J.EQ.NST)BAL=BAL+S(2,17,3)
C
C
C     FLOW DOWN TO SATISFY STAGE TOTAL MASS BALANCE
C
C
```

```
        DO 27 IK=J,NT
        S(2,IK+20,3)=S(2,IK+20,3)+BAL
27      CONTINUE
17      CONTINUE
C
C       CHECK FOR CONDENSER
C
        S(2,16,3)=S(2,2,3)/(RR+1.)
        BAL=S(2,2,3)-S(2,16,3)-S(2,21,3)
        DO 37 IK=1,NT
        S(2,IK+20,3)=S(2,IK+20,3)+BAL
37      CONTINUE
C
C
C       CALCULATION OF INITIAL HOLD UP USING THE FRANCIS WEIR RELATION
C       GIVEN THE INITIAL LIQUID FLOW RATE FROM EACH TRAY
C       START CALCULATION FROM THE BOTTOM(EXCLUDING THE REBOILER AND T
C       CONDENSER
C
        NTM2=NT-2
        DO 1 II=1,NTM2
        I=NT-II
        III=I+20
        P=MW(IG,III)
        Q=DENS(IG,III)
C
C       CALCULATE THE OVERHEIGHT(IN FEET)
C
        FLOW    =S(IG,III,3)*P/Q*3.5314*.00001
        H=(FLOW    /11988./WL)**.66667
        HH=H+WH
C
C       CALCULATE THE MOLAR HOLD UP(IN GMOLES)
C
        VHUP(I)=HH*30.48*TAREA
        HUP1(I)=VHUP(I)*Q/P
  1     CONTINUE
C
C
C   TRANSFER THE VARIABLES INTO  WORKI G AREA FOR INTEGRATICNC
C
        DO 5 I=1,NT
        II=I+20
C     MOLE FRACTION ACETONE
        NN=NT+1-I
        Y(NN)=S(IG,II,6)
C     MOLE FRACTION METHANOL
        N=NN+NT
        Y(N)=S(IG,II,7)
  5     CONTINUE
        DO 30 I=3,8
        S(1,17,I)=S(IG,17,I)
        S(1,18,I)=S(IG,18,I)
 30     CONTINUE
C
 200    CONTINUE
C
```

```
C
C     ADDITINAL SECTION FOR SPECIAL DISTURBANCES
C
C
C     PULSE CYCLING TEST SECTION
C
      IF(TIME .GE. TIMP)GO TO 7
      GO TO 6
7     CONTINUE
      IF(LEVEL .EQ. 1) GO TO 9
      S(1,18,3)=S(2,18,3)=16.96
      LEVEL=1
      GO TO 10
9     CONTINUE
      S(1,18,3)=S(2,18,3)=15.04
      LEVEL=0
10    CONTINUE
      MN=MN+1
      TIMP=FLOAT(MN)*CYCLEF
      GO TO 100
6     CONTINUE
C
C     SECTION 2
C  CALCULATION OF JACOBIAN MATRIX ON CORRECTOR PASS
      IF(IG.EQ.2) GO TO 300
C
C
C
C
C     AT NORMAL TRAY
      DO 12 I=2,NTM1
      N=NT+1-I
      IF(N.EQ.NFT) GO TO 13
      IF(N.EQ.NST) GO TO 14
      ADD=0.
      GO TO 15
13    CONTINUE
      ADD=S(IG,18,3)
      GO TO 15
14    CONTINUE
      ADD=S(IG,17,3)
15    CONTINUE
      B(I)=B(15+I)=(S(IG,N,3)-S(IG,N-1+20,3)-S(IG,N+1,3)+
     1ADD)/HUP1(N)
      C(I)=C(15+I)=S(IG,N-1+20,3)/HUP1(N)
12    CONTINUE
C
C     AT REBOILER
C
      B(1)=B(16)= -S(IG,35,3)             /(VHUP(15)*DENS(IG,35)
     1/MW(IG,35))
      C(1)=C(16)=S(IG,34,3)/(VHUP(15)*DENS(IG,35)/MW(IG,35))
C
C     AT CONDENSER
C
      B(15)=B(30)=(S(IG,2,3)-S(IG,15,3))/(VHUP(1)*DENS(IG,21)/MW(IG,21)
16    CONTINUE
```

```
C
C
 300   CONTINUE
       ITRI=1
C
C
C      SECTION 3
C   CALCULATION OF THE DERIVATIVES
C   THIS SEGMENT OF PROGRAM APPLIES FOR THE CASE OF 3 COMPONENT SYSTEM
C
C
C
       DO 20 I=2,NTM1
       N=NT+1-I
       IF(N.EQ.NFT)GO TO 21
       IF(N.EQ.NST)GO TO 22
C      FOR NORMAL TRAY
       ADD4=0.
       ADD5=0.
       ADD6=0.
       GO TO 23
  21   CONTINUE
C   FOR FEED TRAY
       ADD4=S(IG,18,3)
       ADD5= S(IG,18,3)*S(IG,18,6)
       ADD6=S(IG,18,3)*S(IG,18,7)
       GO TO 23
 22    CONTINUE
C    FOR SOLNENT TRAY
       ADD4=S(IG,17,3)
       ADD5=S(IG,17,3)*S(IG,17,6)
       ADD6=S(IG,17,3)*S(IG,17,7)
       GO TO 23
  23   CONTINUE
    C  VARIATION OF ACETONE COMPOSITION
ILL TRANSFER STATEMENT -- TRANSFER IGNORED
       L1=I
       DERY(L1)=(-S(IG,N,3)*S(IG,N,6)+S(IG,N-1+20,3)*Y(L1+1)+
      1S(IG,N+1,3)*S(IG,N+1,5)-Y(L1)*(-S(IG,N,3)+S(IG,N-1+20,3)+S(IG
      1,N+1,3)+ADD4)+ADD5)/HUP1(N)
C   VARIATION OF METHANOL COMPOSITION
       L2=15+I
       DERY(L2)=(-S(IG,N,3)*S(IG,N,7)+S(IG,N-1+20,3)*Y(L2+1)+
      1S(IG,N+1,3)*S(IG,N+1,7)-Y(L2)*(-S(IG,N,3)+S(IG,N-1+20,3)+S(IG
      1,N+1,3)+ADD4)+ADD6)/HUP1(N)
       GO TO 20
  20   CONTINUE
C
C      VARIATION OF THE MOLE COMPOSITION AT THE REBOILER
C
ILL TRANSFER STATEMENT -- TRANSFER IGNORED
       DERY(1 )=(-S(IG,15,3)*S(IG,15,6)+S(IG,34,3)*Y(2 )-S(IG,35,3)*
      1Y(1 ))/ (VHUP(15)*DENS(IG,35)/MW(IG,35))
       DERY(16)=(-S(IG,15,3)*S(IG,15,7)+S(IG,34,3)*Y(17)-S(IG,35,3)*
      1Y(16))/(VHUP(15)*DENS(IG,35)/MW(IG,35))
C
C      VARIATION OF THE MOLE COMPOSITION AT THE CONDENSER
```

```
C
C
      DERY(15)=(S(IG,2,3)*(S(IG,2,6)-Y(15)))/(VHUP(15)*
     1DENS(IG,21)/MW(IG,21))
      DERY(30)=(S(IG,2,3)*(S(IG,2,7)-Y(30)))/(VHUP(15)*
     1DENS(IG,21)/MW(IG,21))
C
C     SECTION 4
C  INTEGRATE
C
      CALL DIFSUB(NEQ,Y,DERY)
      IF(IDERY .NE. 0) GO TO 300
C
C
C     SECTION 5
C     OUTPUT MANIPULATIONS
C
C
C     MANIPULATION 1
C    TRANSFER VARIABLES INTO SYSTEM STRORAGE
997   CONTINUE
      DO 8 I=1,NT
      II=I+20
      NN=NT+1-I
      S(1,II,6)=Y(NN)
      N=NN+NT
      S(1,II,7)=Y(N)
      S(1,II,8)=1.-Y(NN)-Y(N)
  8   CONTINUE
      DO 31 I=1,35
      DO 32 J=4,5
      S(1,I,J)=S(IG,I,J)
32    CONTINUE
31    CONTINUE
C
C     MANIPULATION 2
C     CALCULATION OF THE MOLECULAR HOLD UP
C
      DO 43 II=1,NTM2
      I=NT-II
      III=I+20
      P=MW(1,III)
      Q=DENS(1,III)
      HUP1(I)=VHUP(I)*Q/P
43    CONTINUE
C
C     MANIPULATION 3
C     CALCULATION OF THE COMPOSITION OF THE VAPOR LEAVING EACH TRAY
C     AT THE REBOILER
C
      III=NT
      II=NT+20
      DUM=EQUIL(1,II,III)
C     TRUE SEPARATION AT REBOILER
C
      S(1 ,III,7)=EX(MP(IM,NMP+1)+NT-1)*S(1 ,III,7)+S(1 ,II,7)*(1.-
     1EX(MP(IM,NMP+1)+NT-1))
```

```
      S(1  ,III,8)=EX(MP(IM,NMP+1)+NT+19)*S(1  ,III,8)+S(1  ,II,8)*(1.-
     1EX(MP(IM,NMP+1)+NT+19))
      S(1  ,III,6)=1.-S(1  ,III,7)-S(1  ,III,8)
C
C
C     CALCULATION FOR ALL OTHER TRAYS(INCLUDING THE CONDENSER)
      DO 2 I=2,NT
      III=NT-I+1
      II=III+20
      DUM=EQUIL(1,II,III)
C
C     TRUE SEPARATION AT EACH TRAY
C
      S(1  ,III,8)=EX(MP(IM,NMP+1)+III+19)*S(1  ,III,8)+S(1  ,III+1,8)*
     1(1.-EX(MP(IM,NMP+1)+III+19))
      S(1  ,III,7)=EX(MP(IM,NMP+1)+III-1 )*S(1  ,III,7)+S(1  ,III+1,7)*
     1(1.-EX(MP(IM,NMP+1)+III-1 ))
      S(1  ,III,6)=1.-S(1  ,III,7)-S(1  ,III,8)
    2 CONTINUE
C
C
C     MANIPULATION 4
C     SET SAME COMPOSITION AT OVERHEAD AND REFLUX STREAMS
      DO 26 I=1,NCOMP
      III=I+5
      S(1  ,16,III)=S(1  ,21,III)
   26 CONTINUE
C
C     SET SAME TEMPERATURE
      S(1,16,4)=S(1,21,4)
C
C
C
C     MANIPULATION 5
C     CALCULATE THE VAPOR FLOWRATES AND LIQUID FLOWRATE USING
C     THE ENERGY EQUATION
C
C
C  FOR THE REBOILER
      III=NT
      II=NT+20
      ETHBAL(III)=REBDT+S(IG,II-1,3)*ENTHL(1,II-1)-S(IG,II,3)
     1*ENTHL(1,II)-S(IG,III,3)*ENTHV(1,III)
      VDIF(III)=ETHBAL(III)/(ENTHV(1,III)-ENTHL(1,II))
C
C
C  CALCULATION FOR ALL TRAYS
C   EXCEPT THE CONDENSER
C
C  INITIATE THE CALCULATION
C
      HLOUT=ENTHL(1  ,NT+20-1)
      HVIN=ENTHV(1  ,NT)
      DO 3 I=2,NTM1
      III=NT+1-I
      II=III+20
      HLIN=ENTHL(1  ,II-1)
```

```
              HVOUT=ENTHV(1  ,III)
              IF(III.EQ.NFT)  GO TO 87
              IF(III.EQ.NST)  GO TO 81
        C     NORMAL TRAY
          80  CONTINUE
              ADD=0.
              GO TO 82
        C
        C  FOR FEED AND SOLVENT TRAYS
        C
        87    CONTINUE
              ADD=S(1  ,18,3)*ENTHL(1  ,18)
              GO TO 82
          81  CONTINUE
              ADD=S(1  ,17,3)*ENTHL(1  ,  17)
              GO TO 82
          82  CONTINUE
ULL TRANSFER STATEMENT -- TRANSFER IGNORED
              ETHBAL(III)=S(IG,III+1,3)*HVIN+S(IG,II-1,3)*HLIN-S(IG,II,3)
             1*HLOUT-S(IG,III,3)*HVOUT +ADD
              VDIF(III)=ETHBAL(III)/(HVOUT-HLOUT)
        C  REINITIATE FOR NEXT TRAY CALCULATION
        C
              HVIN=HVOUT
              HLOUT=HLIN
          3   CONTINUE
        C
        C
        C     VAPOUR FLOWRATES
        C
              TVAP=0.
              DO 33 I=1,NTM1
              III=NT+1-I
              TVAP=TVAP+VDIF(III)
              S(1,III,3)=S(IG,III,3)+TVAP
        33    CONTINUE
        C
        C .FOR THE CONDENSER
              S(1  ,16,3)=S(1  ,2,3)/(RR+1.)
              TLIQ=S(1,2,3)-S(1,16,3)-S(IG,21,3)
              S(1,21,3)  =S(1,21,3)+TLIQ
              TVAP=TLIQ
              DO 34 I=2,NT
              TVAP=TVAP-VDIF(I)
              S(1,I+20,3)=S(IG,I+20,3)+TVAP
        34    CONTINUE
        C
        C
        C
              RETURN
         105  CONTINUE
              PRINT 500
         500  FORMAT(10X,≠*****ERROR AT REBOILER*****≠)
              STOP
              END
```

```
      FUNCTION ACTY(IG,IS)
C
C     THIS PROGRAM CALCULATES THE COEFFICIENT OF ACTIVITY OF EACH
C     COMPONENT IN A ACETONE-METHANOL-WATER MIXTURE,
C     USED WITH DYNSYS 2.1
C
      COMMON/MAT/MP(5,8),EP(5,5),S(2,50,10),EX(100)
      COMMON/PTAB/IGFLAG,PP(10,20)
      COMMON /CON/ IU,NCOMP,NC5,H,NE,NS,NPR,NPOL,TMAX,IORDER,NGRAPH
      COMMON/KVALU/HKK(10)
      DIMENSION A(3),B(3),C(3),AA(3),BB(3),CC(3),X(3),GAM(3,3)
     1 ,ACT(3)  ,LAM(3,3),Y(3),V(3),VV(3),Y2(3)
      DIMENSION ALP(3,3),ALP1(3,3,3),SUMGX(3)
      REAL LAM
      DATA AA/56.8622,21.795,22.8876/,BB/8.4555E-3,4.9507E-2,-3.642E-2/,
     1 CC/1.65E-4,5.68E-5,6.856E-5/
      LAM(1,2)=-216.966
      LAM(2,1)=506.093
      LAM(1,3)=-436.468
      LAM(3,1)=1185.436
      LAM(2,3)=-208.7681
      LAM(3,2)=571.3078
C     ENTHALPIC WILSON EQUATION TO PREDICT VAPOUR MOL FRACTION IN EQUIBRIUM
C     WITH GIVEN LIQUID CONCENTRATION
C     BINARY COEFFICIENTS FOR WILSON EQUATION,VAPOUR PRESSURE ANTOINE
C     COEFFICIENTS AND LIQUID MOLAR VOLUMES FROM
C     ACETONE -1, METHANOL -2,WATER -3
      TK=S(IG,IS,4)+PP(1,20)
      ACTY=0.
C
C     CALCULATE THE GAMMA≠S BY (15-B)
C
      DO  20 J=1,3
      GAM(J,J)=1.
      DO  10 I=1,3
      IF(I.EQ.J) GO TO 10
      FC=EXP(-LAM(I,J)/(1.987*TK))
      V(J)=AA(J)+(BB(J)+CC(J)*TK)*TK
      VV(I)=AA(I)+(BB(I)+CC(I)*TK)*TK
      GAM(I,J)=V(J)/VV(I)*FC
   10 CONTINUE
   20 CONTINUE
C
C     READ IN THE PROPERTIES OF THE STREAMS
C
      DO 70 I=6,NC5
      N=I-5
      X(N)=S(IG,IS,I)
  70  CONTINUE
      DO 107 I=1,3
      SUMGX(I)=0.0
      DO 101 IP=1,3
  101 SUMGX(I)=SUMGX(I)+GAM(I,IP)*X(IP)
1.07  CONTINUE
      DO 102 I=1,3
      DO 102 J=1,3
  102 ALP(I,J)=1.0+X(I)*(1.0-1.0/SUMGX(I)-GAM(J,I)/SUMGX(J))
      DO 106 I=1,3
      DO 105 J=1,3
      IF(I.EQ.J) GO TO 105
      DO 103 M=1,3
      IF(M.EQ.I .OR. M.EQ.J) GO TO 103
      ALP1(I,J,M)=1.0-GAM(J,I)/SUMGX(J)-GAM(M,I)/SUMGX(M)
 103  CONTINUE
 105  CONTINUE
 106  CONTINUE
      DO 60 I=1,3
      SUM1=0.0
```

```
      DO 30  J=1,3
      SUM1=SUM1+X(J)*ALOG(GAM(I,J)*GAM(J,I))*ALP(I,J)/(SUMGX(I)*SUMGX
     1 (J))
30    CONTINUE
      SUM2=0.0
      DO 50  M=1,3
      IF(M.EQ.I)  GO TO 50
      DO 55  J=1,3
      IF(J.EQ.I  .OR.J.EQ.M)  GO TO 55
      SUM2=SUM2+X( J)*X(M)*ALOG(GAM(J,M)*GAM(M,J))*ALP1(I,J,M)/
     1 (SUMGX(J)*SUMGX(M))
55    CONTINUE
 50   CONTINUE
      PP(I,9)=EXP(-SUM1-SUM2)
60    CONTINUE
      RETURN
      END
```

```
      REAL FUNCTION KVAL(IG,IL,IV)
C
C     THIS PROGRAM CALCULATES THE ACTIVITY OF EACH COMPONENT IN
C     A MIXTURE
C
      COMMON/MAT/MP(5,8),EP(5,5),S(2,50,10),EX(100)
      COMMON /CON/ IU,NCOMP,NC5,H,NE,NS,NPR,NPOL,TMAX,IORDER,NGRAPH
      COMMON/PTAB/IGFLAG,PP(10,20)
      COMMON/KVALU/HKK(10)
C     DUMMY STATEMENT
      KVAL=0.0
      T=S(IG,IL,4)+PP(1,20)
      P=S(IG,IL,5)
      DUM=ACTY(IG,IL)
      DO 1  I=1,NCOMP
      HKK(I)=10.**(PP(I,6)-PP(I,7)/(PP(I,8)+T-273.))*PP(I,9)/P
 1    CONTINUE
      RETURN
      END
```

```
      FUNCTION ENTHV(IG,IS)                                          ETV
C
C     THIS PROGRAM CALCULATES THE ENTHALPY OF A VAPOR MIXTURE,
C     USED WITH DYNSYS 2.1
C
      COMMON/MAT/MP(5,8),EP(5,5),S(2,50,10),EX(100)
      COMMON/CON/IU,NCOMP,NC5,H,NE,NS,NPR,NPOL,TMAX,IORDER,NGRAPH
      COMMON/PTAB/IGFLAG,PP(10,20)
      T=S(IG,IS,4)
      HV=0.
      DO 10  I=1,NCOMP
110     HV=HV+S(IG,IS,I+5)
     K               *((((((PP(I,15)*T+PP(I,14))*T+PP(I,13))*T
     K               +PP(I,12))*T+PP(I,11))*T+PP(I,10))
      ENTHV=HV
      RETURN
      END
```

```
      FUNCTION ENTHL(IG,IS)

      THIS PROGRAM CALCULATES THE ENTHALPY OF A LIQUID MIXTURE
      OF ACETONE-METHANOL-WATER,USED WITH DYNSYS 2.1

      COMMON/MAT/MP(5,8),EP(5,5),S(2,50,10),EX(100)
      COMMON/CON/IU,NCOMP,NCS,H,NE,NS,NPR,NPOL,TMAX,IORDER,NGRAPH
      COMMON/PTAB/IGFLAG,PP(10,20)
      DIMENSION X(3)
      HACEL(T)=30.6E-3*T
      HMETL(T)=19.74E-3*T
      HWATL(T)=18.016E-3*T

      FORMULATION OF HEAT OF MIXING
      THE PARAMETERS ARE GIVEN BY A LEAST SQUARE FITTING

      HMIXWA(XC)=-6.00506E-3+0.368503*XC-2.1499*XC**2+5.30655*XC**3
     1-3.50732*XC**4
      HMIXMA(XC)=2.01772E-3-0.753418*XC+0.94408*XC**2-0.192395*XC**3
      HMIXWM(XC)=6.05874E-3+0.171893*XC+0.646148*XC**2-0.813034*XC**3


      SET THE NECESSARY VARIABLES FROM THE STREAMS


      T=S(IG,IS,4)
      DO 10 I=1,3
      II=I+5
      X(I)=S(IG,IS,II)
10    CONTINUE
      ENTHL=HACEL(T)*X(1)+HMETL(T)*X(2)+HWATL(T)*X(3)
      IF(X(1).EQ.1. .OR.X(2).EQ.1..OR.X(3).EQ.1.) GO TO 1
      X1B=X(1)+X(2)
      X1A=X(2)/X1B
      X2B=X(3)+X(2)
      X2A=X(3)/X2B
      X3B=X(3)+X(1)
      X3A=X(3)/X3B
      ENTHL=ENTHL-X1B*HMIXMA(X1A)-X2B*HMIXWM(X2A)-X3B*HMIXWA(X3A)
      CONTINUE
      RETURN
      END
```

APPENDIX G

Distillation Control Programs

The control software of the extractive distillation column has been developed under the GOSEX created by J.P. Tremblay (22). The minicomputer NOVA/2 in the Chemical Engineering computer room was equipped with necessary peripheral devices (A/D converter, D/A converter, relays, contact sense, one operator console and one data logging line printer) and could be used to run the column.

The user must provide:

(1) a symbol table to facilitate duties as parameter change, variable logging... (SYMBOL OF TAPE #2 OF 2)

(2) a disk file containing user messages (MESSG.TR)

(3) user control programs (INFCE, FSCON, SWAVE, SSAMP)

The listings are presented on the following pages with explanations.

The listings of GOSEX parameter tape, symbol table and executive message file (see reference 22) are also presented.

A loading map could be found at the end of Appendix G, and the loading procedure is detailed in reference 22.

```
 INFCF
                       .TITL INFCE
                       ;INTERFACE TO THE GOSEX EXECUTIVE
                       ;DECEMBER, 1975
                       ;MCMASTER UNIVERSITY
                       ;CHEMICAL ENGINEERING DEPARTMENT
                       ;
                       ;
                       .ENT OVLAY,ALARM,DATUM,RELAY,SENSE
                       .ENT SCAN,CLOOP,SUPRS,TASKS,UDEF
                       .EXTN FEEPT,SOLPT
                       .EXTN SOLFM,SOLVB,FEEFM,FEEFB
                       .EXTN FEETC,SOLTC
                       .EXTN TRAP,SSAMP
                       .EXTD TIMER
   000001             ..TXTM 1
                       .NREL
            ;OVERLAY FILENAME DEFINITION
000'000002"OVLAY:   .+1*2
                       .TXT @DISTIL.OL@
011'042111
022'051524
033'044514
044'027117
055'046000
            ;ALARM TABLE DEFINITION
066'000010 ALARM:   10                            ;# OF ALARMS IN TABLE
   000010           .BLK 10
   000012 DATUM:    .BLK ATODS                    ;A/D CHANNELS
   000012 RELAY:    .BLK RLAYS                    ;RELAYS
433'177777 SENSE:   TRAP
444'000043'         TRAP
455'000044'         TRAP
466'000045'         TRAP
477'000046'         TRAP
500'000047'         TRAP
511'000050'         TRAP
522'000051'         TRAP
533'000052'         TRAP
544'000053'         TRAP
            ;MULTIPLEXER SCAN RATE
555'000002 SCAN:    2                             ;2 TIMES PER SECOND
            ;PID CONTROL LOOP DEFINITION
56'177777 CLOOP:    SOLFM                         ;SOLVENT LOOP
57'177777           SOLVB
60'177777           FEEFM                         ;FEED LOOP
61'177777           FEEFB
62'100000           1B0
            ;USER SHUTDOWN ROUTINE
63'054407 SUPRS:    STA 3 .+7
64'126420           SUBZ 1 1                      ;ZERO AC1
65'044752           STA 1 RELAY+6                 ;TURN OFF FEED HEATER
66'044750           STA 1 RELAY+5                 ;TURN OFF SOLVENT HEATER
67'024001$          LDA 1 TIMER                   ;WAIT FOR RELAYS TO BO SET
70'006017            .SYSTM
71'010400            .DELAY
72'000000           0
73'002777           JMP @ -1
            ;USER TASK DEFINITION TABLE
74'177777 TASKS:    FEETC                         ;FEED TEMP. MONITOR
```

```
 TINFCE
5'000110          110                              ;PRIORITY
6'000100          100                              ;IDENTIFICATION
7'100000          1B0
0'177777          SOLTC                            ;SOLVENT TEMP  MONITOR
1'000110          110
2'000110          110
3'100000          1B0
4'000056'         SOLFM                            ;SOLVENT RATE MONITOR
5'000110          110
6'000120          120
7'100000          1B0
0'000060'         FEEFM                            ;FEED RATE MONITOR
1'000110          110
2'000130          130
3'100000          1B0
4'177777          SSAMP                            ;SAMPLING MONITOR(OVERHEAD AND BOTTOM)
5'000110          110
6'000140          140
7'100000          1B0
0'177777          FEEFT                            ;SQUARE WAVE IN FEED RATE ROUTINE
21'000110         110
22'000150         150
23'100000         1B0
24'177777         SOLFT                            ;INPUT FLOWRATES SETTING ROUTINE
25'000110         110
26'000160         160
27'100000         1B0
30'100000         1B0

311'000000 UDEF:  0
                  .END
```

```
                  TITL FSCON
                 ;CONTROL PROGRAM HAS 4 TASKS
                 ;FEETC           ON/OFF CONTROL OF FEED TEMPERATURE
                 ;                AND CONVERT DATA TO ENG  UNITS
                 ;                FNOFF   CONTROL PERIOD
                 ;                FTSET   SET POINT
                 ;
                 ;SOLTC           ON/OFF CONTROL OF SOLVENT TEMPER
                 ;                AND CONVERT DATA TO ENG  UNITS,
                 ;                ALSO CONVERT 3 OTHER TEMP READ-
                 ;                INGS TO ENG  UNITS
                 ;                SNOFF   CONTROL PERIOD
                 ;                STSET   SET POINT
                 ;
                 ;SOLFM           PID CONTROL FOR SOLVENT RATE
                 ;                AND CONVERSION OF DATA INTO ENG UNITS
                 ;
                 ;FEEFM           PID CONTROL FOR FEED RATE AND
                 ;                CONVERSION OF DATA INTO ENG UNITS
                 ;
                 ;DECEMBER, 1975
                 ;MCMASTER UNIVERSITY
                 ;CHEMICAL ENGINEERING DEPARTMENT
                 ;
                 ;
                  ENT SOLFM,SOLVB,FEEFB,FEEFM
                  ENT FEETC,SOLTC
                  ENT FNOFF,FTSET,ADFTE,FTEMP
                 .ENT SNOFF,STSET,ADSTE,STEMP
                  ENT FFLOW,OFERF,FFCON
                 .ENT SFLOW,OFERS,SFCON
                 .EXTN CONTL,QUAD,QNVRS,MTPLY
                 .ENT CONVB,ADT
                 .EXTN LNEAR,LNVRS
                 .EXTD TIMER
                 .ENT FEETB,SOLTB
                 .ENT OUTBS,OUTBF
                 .NREL
          ;FEED TEMPERATURE CONTROL ROUTINE

00'034434 FEETC:  LDA 3 FEETB-1
01'054016         STA 3 USP
02'026427         LDA 1.@DTUM4             ;GET RAW FEED TEMP  READING
03'022477         LDA 0 @DTUM5             ;GET REFERENCE READING
04'106400         SUB 0 1                  ;A/D-A/D REFERENCE
05'044420         STA 1 ADFTE              ;STORE FEED TEMP (IN A/D UNITS)
06'020447         LDA 0 ACONB
07'006424         JSR @AQUAD
10'030437         LDA 2 AENT
11'045004         STA 1 4 2
12'020414         LDA 0 FTSET              ;GET SET POINT TEMP
13'030414         LDA 2 FHSW               ;GET RELAY SWITCH WORD
14'024411         LDA 1 ADFTE              ;GET FEED TEMP
15'122433         SUBZ# 1 0 SNC            ;SKIP IF TEMP <=SET POINT TEMP
16'152400         SUB 2 2
17'052411         STA 2 @RLWF             ,INSTALL RELAY WORD
20'024403         LDA 1 FNOFF              ;GET WAIT PERIOD
21'006017          SYSTM
22 010400          DELAY
```

```
  FSCON
23'000764 FNOFF:    500
24'000756           JMP FEETC+2


25'000000 ADFTE     0
26'000000 FTSET.    0
27'001000 FHSW:     1000                              ;BIT 6
   000003 RLWF:      GADD RELAY, 3
   000004 DTUM4:    .GADD DATUM, 4
   000005 DTUM5:    .GADD DATUM, 5
33'177777 AQUAD     QUAD
34'000035'          FEETB
   000012 FEETB:    .BLK 10.
47'000050'AENT:     ENTAB
50'000000 ENTAB:    0                               ;TABLE IN ENG. UNITS
51'000000           0
52'000000           0
53'000000 STEMP:    0
54'000000 FTEMP:    0
55'000056'ACONB:    CONVB
56'175710 CONVB:    -1080.
57'076272           +31930.
60'000000           +0

          ;SOLVENT TEMPERATURE ROUTINE

61'034470 SOLTC:    LDA 3 SOLTB-1
62'054016           STA 3 USP
63'026447           LDA 1 @DTUM5
64'022746           LDA 0 @DTUM5
65'106400           SUB 0 1
66'044445           STA 1 ADSTE
67'020766           LDA 0 ACONB                     ;CONVERSION
70'006743           JSR @AQUAD
71'030756           LDA 2 AENT
72'045003           STA 1 3 2
73'020441           LDA 0 STSET                      ;CONTROL ACTION
74'030442           LDA 2 SHSW
75'024436           LDA 1 ADSTE
76'122433           SUBZ# 1 0 SNC
77'152400           SUB 2 2
00'052435           STA 2 @RLWS                    ;INSTALL RELAY
01'024436           LDA 1 C3                       ;CONVERSION OF OTHER TEMP
02'044436           STA 1 CNT
03'024441           LDA 1 AADT
04'044441           STA 1 ADT1
05'024742           LDA 1 AENT
06'044435           STA 1 AENT1
07'024432           LDA 1 DTUM0
0'044432            STA 1 CNT1
1'026431            LDA 1 @CNT1
2'022720            LDA 0 @DTUM5
3'106400            SUB 0 1                        ;A/D-A/D REFERENCE
4'046431            STA 1 @ADT1                    ;STORE RESULT IN A/D TABLE
5'020740            LDA 0 ACONB
6'006715            JSR @AQUAD
7'046424            STA 1 @AENT1
0'010435            ISZ AENT1
```

```
 FACON
21'010424            ISZ ADT1
22'010420            ISZ CNT1
23'014415            DSZ CNT
24'000765            JMF  -13
25'024403            LDA 1 SNOFF                   ;GET WAIT FERIOD
26'006017             SYSTM
27'010400            .DELAY
30'000764 SNOFF      500
31'000732            JMF SOLTC+2
   000003 DTUM3:     .GADD DATUM, 3
33'000000 ADSTE:     0
34'000000 STSET:     0
   000002 RLWS:      .GADD RELAY, 2
36'002000 SHSW:      2000                          ;BIT 5
37'000003 C3:        3
40'000000 CNT:       0
   000000 DTUMO:     .GADD DATUM, 0
42'000000 CNT1:      0
43'000000 AENT1:     0
44'000146'AADT:      ADT
45'000000 ADT1:      0
46'000000 ADT:       0                             ;TABLE OF TEMP. IN A/D UNITS
47'000000            0
50'000000            0
51'000152'           SOLTB
   000012 SOLTB:     .BLK 10

          ;FID CONTROLLER FOR SOLVENT FLOWRATE
64'034435 SOLFM:     LDA 3 SOLVB-1                ;GET CONTROLLER BLOCK ADDRESS
65'054016            STA 3 USF
66'023427            LDA 0 @VABLE 3               ;GET RAW DATA
67'041423            STA 0 SMOTH 3                ;STORE
70'177777            CONTL                        ;CALL CONTROLLER
          ;CONVERSION TO ENG. UNITS FOR OUTFUT.
71'021450            LDA 0 SFCON 3                ;CONVERT INFUT VARIABLE
72'025423            LDA 1 SMOTH 3
73'007451            JSR @SFAZE 3
74'044416            STA 1 SFLOW
75'021454            LDA 0 MNCON 3                ;CONVERT OUTFUT ACTION
76'025422            LDA 1 OUTFT 3
77'007455            JSR @MNAZE 3
00'044413            STA 1 OFERS
01'025442            LDA 1 DELAY 3
02'030015            LDA 2 TIMER
03'177777            MTFLY
04'101004            MOV 0 0 SZR
05'126000            ADC 1 1
06'006017            .SYSTM
07'010400             DELAY
10'000000            0
11'000757            JMF SOLFM+4
12'000000 SFLOW      0
13'000000 OFERS      0
14'177502 SFCON      -190                           ,A*10**5
15'003136            +1630                         ,B*10**5
16'175210            -1400                         ,C*10
17'000000 OUTBS      0                  ;TRUE A *10 **3
20'000000            0                  ;TRUE B*10
          ;CONTROLLER DATA BLOCK FOR SOLVENT RATE MONITORING
```

1803

```
   FSCON
21'000222'           SOLVB
   000026 SOLVB:      BLK 22
50'000000             0                           ;FILTER
   000006              GADD DATUM, 6              ;ADDRESS OF INPUT DATA
52'000000             0                           ;INPUT HIGH ALARM LIMIT
53'000000             0                           ;INPUT LOW ALARM LIMIT
54'000000             0                           ;HI ALARM MESSAGE NUMBER
55'000001             1                           ;LO ALARM MESSAGE NUMBER
56'000002             2                           ;TIME INTER  FOR SUCC ALARM DISPL
57'000000             0                           ;SET POINT IN A/D UNITS
60'000000             0                           ;KP
61'000000             0                           ;KPD
62'000000             0                           ;KPI
63'000000             0                           ;SCALE
64'000001             1                           ;DELAY
65'000000             0                           ;DIRECT OR CASCADE CONTROL
66'000000             0                           ;D/A CHANNEL NUMBER
67'000000             0                           ;UPPER LIMIT ON OUTPUT ACTION
70'000000             0                           ;LOWER LIMIT ON OUTPUT ACTION
71'000000             0                           ;MEAN VALUE OF OUTPUT ACTION
72'000214'           SFCON                        ;ADDRESS OF CONSTANT BLOCK FOR CONV
73'000033'           QUAD                         ;ADDRESS OF CONV  ROUTINE
74'177777            QNVRS                        ;INVERSE CONVERSION
75'000144            100
76'000217'           OUTBS
77'177777            LNEAR
00'177777            LNVRS
01'000145            101
02'000146            102
         ;PID CONTROLLER FOR FEED FLOWRATE
03'034435 FEEFM:     LDA 3 FEEFB-1                   ;GET CONTROLLER BLOCK ADDRESS
04'054016            STA 3 USP
05'023427            LDA 0 @VRBLE 3                  ;GET RAW DATA
06'041423            STA 0 SMOTH 3
07'000170'           CONTL                           ;CALL CONTROLLER
         ;CONVERSION TO ENG  UNITS FOR OUTPUT
10'021450            LDA 0 SPCON 3               ;CONVERT INPUT VARIABLE
11'025423            LDA 1 SMOTH 3
12'007451            JSR @SPAZE 3
13'044416            STA 1 FFLOW
14'021454            LDA 0 MNCON 3               ;CONVERT OUTPUT ACTION
15'025422            LDA 1 OUTPT 3
16'007455            JSR @MNAZE 3
17'044413            STA 1 OPERF
20'025442            LDA 1 DELAY 3
21'030001$           LDA 2 TIMER
22'000203'           MTPLY
23'101004            MOV 0 0 SZR
24'126000            ADC 1 1
25'006017             SYSTM
26'010400             DELAY
27'000000            0
30'000757            JMP FEEFM+4
31'000000 FFLOW:     0
32'000000 OPERF:     0
33'176650 FFCON:     -600                       ;A*10**5
34'005132            2650                       ;B*10*3
35'176042            -990                       ;C*10
36'000000 OUTBF:     0                          ;TRUE A *10 **5
```

```
 FFACON
7 '000000           0                          ;TRUE B *10
        ;CONTROLLER DATA BLOCK FOR FEEDRATE
        ;MONITORING
0 '000341'          FEEFB
   000026 FEEFB:     BLK 22
7 '000000           0                          ;FILTER
   000007            GADD DATUM, 7             ;ADDRESS OF INPUT DATA
1 '000000           0                          ;INPUT HIGH ALARM LIMIT
2 '000000           0                          ;INPUT LOW ALARM LIMIT
3 '000002           2                          ;HI ALARM MESSAGE NUMBER
4 '000003           3                          ;LO ALARM MESSAGE NUMBER
5 '000002           2                          ;TIME INTER  FOR SUCC  ALARM DISP
6 '000000           0                          ;SET POINT IN A/D UNITS
7 '000000           0                          ;KP
00'000000           0                          ;KPD
01'000000           0                          ;KPI
02'000000           0                          ;SCALE
03'000001           1                          ;DELAY
04'000000           0                          ;DIRECT OR CASCADE CONTROL
05'000001           1                          ;D/A CHANNEL NUMBER
06'000000           0                          ;UPPER LIMIT ON OUTPUT ACTION
07'000000           0                          ;LOWER LIMIT ON OUTPUT ACTION
10'000000           0                          ;MEAN VALUE OF OUTPUT ACTION
11'000333'          FFCON                      ;ADDRESS OF CONSTANT BLOCK FOR CONV
12'000273'          QUAD                       ;ADDRESS OF CONV  ROUTINE
13'000274'          QNVRS                      ;INVERSE CONV
14'000144           100.
15'000336'          OUTBF
16'000277'          LNEAR
17'000300'          LNVRS
20'000145           101.
21'000147           103.
                     END
```

```
                   TITL SSAMP
                  ;ROUTINE TO ACKNOWLEDGE CHANGES OCCURRED AT PROCESS
                  ;SITE AND TO PRINT OUT AN APPROPRIATE MESSAGE
                  ;RELAY BIT 0AND CONTACT SENSE BIT 0 FOR OVERHEAD CHANGE
                  ;RELAY BIT 1 AND CONTACT SENSE BIT 1 FOR BOTTOM CHANGE
                  ;DECEMBER,1975
                  ;MCMASTER UNIVERSITY
                  ;CHEMICAL ENGINEERING DEPARTMENT
                  ;
                  ;
                   EXTN  REC
                   ENT SSAMP,TRAP
                  .NREL
0'034443 SSAMP:   LDA 3 SSAMB-1
1'054016          STA 3 USP
2'102400          SUB 0 0
3'040454          STA 0 TRAP
4'020452          LDA 0 TRAP-1
5'177777          .REC
6'020452          LDA 0 MASK
7'123420          ANDZ 1 0                 ;MASK UNWANTED BITS
0'152520          SUBZL 2 2                ;PUT 1 IN AC2
1'112414          SUB# 0 2 SZR             ;SKIP IF BOTTOM SIGNAL OCCURS
2'000415          JMP OVERH
3'125113          MOVL# 1 1 SNC            ;SKIP IF SIGNAL ON
4'000406          JMP BTOPE                ;GO TO SIGNAL OFF
5'052444          STA 2 @BTRLA             ;SET RELAY BIT ON
6'052444          STA 2 @BTON              ;PRINT MESSAGE
7'152400          SUB 2 2                  ;PUT 0 IN AC2
0'052443          STA 2 @BTOF
1'000763          JMP SSAMP+4
2'176400 BTOPE:   SUB 3 3
3'056436          STA 3 @BTRLA             ;CLEAR RELAY BIT
4'052437          STA 2 @BTOF              ;PRINT MESSAGE
5'056435          STA 3 @BTON
6'000756          JMP SSAMP+4
7'125113 OVERH:   MOVL# 1 1 SNC            ;SKIP IF SIGNAL ON
0'000406          JMP OVOPE
1'052433          STA 2 @OVRLA             ;SET RELAY BIT
2'052433          STA 2 @OVEON             ;PRINT MESSAGE
3'152400          SUB 2 2
4'052432          STA 2 @OVEOF
5'000747          JMP SSAMP+4
6'176400 OVOPE:   SUB 3 3
7'056425          STA 3 @OVRLA             ;CLEAR RELAY BIT
0'052426          STA 2 @OVEOF             ;PRINT MESSAGE
1'056424          STA 3 @OVEON
2'000742          JMP SSAMP+4


3'000044          SSAMB
  000012 SSAMB:   .BLK 10.
6'000057          TRAP
7'000000 TRAP:    0
0'000017 MASK:    17
  000001 BTRLA:    GADD RELAY,1
  000006 BTON:     GADD ALARM,6
  000007 BTOF:     GADD ALARM,7
  000000 OVRLA:    GADD RELAY,0
```

```
SSAMP
 000004 OVEON:     GADD ALARM,4
 000005 OVEOF:     GADD ALARM,5
                   END
```

```
                   TITL SWAVE
                  ;DECEMBER, 1975
                  ;MCMASTER UNIVERSITY
                  ;CHEMICAL ENGINEERING DEPARTMENT
                  ;
                  ;
                  ;
                   ENT SOLFT,FFLO1,SFLO1,WAIT,TWAIT
                  .ENT FEEFT,FHL,FLL,FLG
                  .EXTN LNEAR,MTPLY
                  .EXTD TIMER
                   NREL
                  ;PROGRAM TO SET FEED AND SOLVENT FLOWRATES
                  ;AND TO READ FLOWRATES FROM DP CELLS,AND
                  ;CONVERT THE READINGS INTO ML/MIN DATA FOR
                  ;DISPLAY
                  ;
                  ;CHANS          D/A CHANNEL FOR SOLVENT VALVE
                  ;CHANF          D/A CHANNEL FOR FEED VALVE
                  ;SOLFR          STTING OF SOLVENT RATE IN D/A UNITS
                  ;FEEFR          SETTING OF FEED RATE IN D/A UNITS
                  ;SFLO1          CURRENT SOLVENT RATE IN ML/MIN
                  ;FFLO1          CURRENT FEED RATE IN ML/MIN
                  ;
                  ;
0'034431 SOLFT:   LDA 3 SOLF-1
1'054016          STA 3 USP
2'020442          LDA 0 CHANS               ;SET POINT TO SOLVENT RATE
3'024442          LDA 1 SOLFR
4'062023          DOB 0 DACV
5'065023          DOA 1 DACV
6'020440          LDA 0 CHANF               ;SET POINT TO FEED RATE
7'024440          LDA 1 FEEFR
0'062023          DOB 0 DACV
1'065023          DOA 1 DACV
2'026436 READ:    LDA 1 @SOLAD              ;READ SOLVENT RATE IN A/D UNITS
3'020436          LDA 0 SCON-1
4'006444          JSR @LINEA                          ;CONVERSION
5'044444          STA 1 SFLO1
6'026436          LDA 1 @FEEAD              ;READ FEED RATE IN A/D UNITS
7'020436          LDA 0 FCON-1
0'006440          JSR @LINEA                          ;CONVERSION
1'044441          STA 1 FFLO1
2'024405          LDA 1 WAIT
3'030001$         LDA 2 TIMER
4'177777          MTPLY
5'006017           SYSTM
6'010400          .DELAY
7'000005 WAIT:    5
0'000762          JMP READ
1'000032'         SOLF
  000012 SOLF:     BLK 10
4'000000 CHANS:   0
5'000600 SOLFR:   384.
6'000001 CHANF:   1
7'000321 FEEFR:   209
  000006 SOLAD:    GADD DATUM,6
1'000052'         SCON
2'000202 SCON:    130
```

```
 SWAVF
53'177540          -160
  000007 FEEAD:     GADD DATUM,7
55'000056          FCON
56'000156 FCON:    110
57'177543          -157
60'177777 LINEA:   LNEAR
61'000000 SFLO1:   0
62'000000 FFLO1:   0

                   ;PROGRAM TO SHIFT THE FEED FLOWRATE TO A HIGH
                   ;LEVEL (FHL) AND TO A LOW LEVEL (FLL) ALTERNA-
                   ;TIVELY.
                   ;TWAIT         WAITING HALF PERIOD IN SEC.
                   ;FHL AND FLL ARE IN D/A UNITS
                   ;
                   ;
73'034421 FEEFT:   LDA 3 FEEVB-1
74'054016          STA 3 USP
75'020433          LDA 0 FHL                     ;GET HIGH LEVEL
76'024431          LDA 1 FLG                     ;GET FLAG VALUE
77'124005          COM 1 1 SNR           ;HI-LO?
700'020431         LDA 0 FLL             ;IF FLAG=0,GET LOW LEVEL
701'030755         LDA 2 CHANF
702'072023         DOB 2 DACV            ;SET D/A CHANNEL
703'061023         DOA 0 DACV
704'044423         STA 1 FLG             ;SAVE FLAG VALUE
705'024405         LDA 1 TWAIT
706'030001$        LDA 2 TIMER
707'000024'        MTPLY
000'006017         .SYSTM
001'010400          DELAY
002'000264 TWAIT:  180.
003'000762         JMP FEEFT+2
004'000105'        FEEVB
  000012 FEEVB:    .BLK 10
17'000000 FLG:     0
20'000426 FHL:     278.
21'000274 FLL:     188
                   .END
```

```
GOSEX PARAMETER TAPE

USERS OF THE GOSEX EXECUTIVE MUST CONFIGURE ALL OF
THE EXECUTIVE PROGRAMS WITH THIS PARAMETERS TAPE
THE USER MAY SPECIFY THE VALUE OF A CERTAIN NUMBER
OF THE PARAMETERS HEREIN.   THESE PARAMETERS ARE:
        NVARS
        LPTS
        TTYS
        PTPS
        CASS
        ATODS
        PTRS
        CNTCT
        RLAYS
        DTOAS
        FORT4

ON NO ACCOUNT SHOULD OTHER PARAMETER VALUES BE CHANGED

DISPLACEMENTS IN THE GENERAL WORK SPACE

DUSR    GENRL=  10.     ;SIZE OF GENERAL WORK SPACE
DUSR    RETUR=  0       ;RETURN ADDRESS STORAGE
DUSR    RETAD=  1       ;RETURN ADDR STORAGE
DUSR    BYTEP=  2       ;BYTE POINTER STORAGE
DUSR    COUNT=  3       ;COUNTER
DUSR    KOUNT=  4       ;COUNTER
DUSR    WORD1=  5       ;TEMPORARY STORAGE
DUSR    WORD2=  6       ;TEMPORARY STORAGE
DUSR    WORD3=  7       ;TEMPORARY STORAGE
DUSR    WORD4=  8.      ;TEMPORARY STORAGE
DUSR    WORD5=  9.      ;TEMPORARY STORAGE

DISPLACEMENTS USED FOR THE LOG AND PLOT ROUTINES

DUSR    LOOP1=  GENRL   ;LOOP COUNTER
DUSR    LOOP2=  GENRL+1 ;LOOP COUNTER
DUSR    SPOT1=  GENRL+2 ;POINTER
DUSR    SPOT2=  GENRL+3 ;POINTER
DUSR    NDATA=  GENRL+4 ;# DATA TO LIST
DUSR    INTVL=  GENRL+5 ;# SECS BETWEEN RECORDS
DUSR    CHANL=  GENRL+6 ;CHANNEL # USED FOR OUTPUT
DUSR    LNCNT=  GENRL+7 ;PAGE LINE COUNT
USR     HOUR=   GENRL+8.        ;STARTING HOUR OF RECORD
USR     MIN=    GENRL+9.        ;STARTING MINUTE OF RECORD
USR     SEC=    GENRL+10.       ;STARTING SECOND OF RECORD
USR     TMSHO=  GENRL+11.       ;COUNTER FOR TIME OF DAY DISPLAY
USR     MNSIZ=  GENRL+12.       ;LOWER RANGE VALUE OF PLOT
USR     MXSIZ=  GENRL+13.       ;UPPER RANGE VALUE OF PLOT
USR     DNAME=  GENRL+14.       ;BYTE POINTER OF DEVICE FILENAME
USR     UNIT=   GENRL+15.       ;DEVICE UNIT #
USR     ADSYM=  GENRL+16.       ;ADDR OF POINTER TO SYMBOLS LIST
USR     ADVAL=  GENRL+17.       ;ADDR OF POINTER TO VALUES TABLE
USR     ADBUF=  GENRL+18.       ;BYTE POINTER TO WRITTING BUFFER
USR     NVARS=  20.     ;# VARIABLES TO BE LOGGED OR PLOTTED
USR     IDCOD=  0       ;OUTPUT TASK DEFINITION CODE
USR     TABSA=  1       ;DEVICE STORAGE BLOCK S.A
USR     IDENT=  2       ;TASK IDENTIFICATION NUMBER
USR     PRIOR=  3       ;TASK PRIORITY

UTPUT TASK DEFINITION CODES
```

```
WSR   LOGTT=  1B0+321
WSR   LOGPN=  1B0+331
WSR   LOGDK=  1B0+341
WSR   LOGCS=  1B0+351
WSR   PLTCL=  1B0+302
WSR   PLTLP=  1B0+312
WSR   PLTTT=  1B0+322
WSR   PLTPN=  1B0+332
WSR   PLTDK=  1B0+342
WSR   PLTCS=  1B0+352

ONSTANTS WHICH ARE USED TO DEFINE DISPLACEMENTS
OR THE DISK MESSAGE ROUTINE

WSR   TMESS=  GENRL   ;TOTAL NUMBER OF MESSAGES IN DISK FILE
WSR   MXLGT=  GENRL+1 ;MAXIMUM MESSAGE LENGTH PERMITTED
WSR   DSKCL=  GENRL+2 ;CHANNEL # TO DISK FILE
WSR   DSKNM=  GENRL+3 ;BYTE POINTER TO DISK FILE NAME
WSR   DEVCL=  GENRL+4 ;CHANNEL # TO OUTPUT DEVICE
WSR   DEVNM=  GENRL+5 ;BYTE POINTER TO OUTPUT DEVICE NAME
WSR   BUFFR=  GENRL+6 ;BYTE POINTER TO BUFFER AREA OF LENGTH MXLGT
WSR   HIBYP=  0       ;HIGH ORDER BYTE POINTER TO MESSAGE
WSR   LOBYP=  1       ;LOW ORDER BYTE POINTER TO MESSAGE
WSR   LNGTH=  2       ;MESSAGE LENGTH IN BYTES

ISPLACEMENTS IN THE USER TASKS DEFINITION BLOCKS.
ACH BLOCK IS FOUR WORDS LONG AND EACH USER TASK HAS
NE BLOCK DEFINING ITS NECESSARY ATTRIBUTES.

WSR   TSKSA=  0       ;S.A. OF THE TASK
WSR   TSKPR=  1       ;PRIORITY OF THE TASK(1 TO 377-NEVER 0)
WSR   TSKID=  2       ;TASK IDENTIFICATION CODE-1 TO 300,
                      ;NEVER 0 OR ABOVE 300
WSR   TSKIL=  3       ;TASK'S KILL FLAG
      THE TASK DEFINITION BLOCK IS EXTENDED BY ONE WORD
      IF THE FORT4 PARAMETER, DEFINED BELOW IS NON-ZERO
WSR   TSKMD=  4       ;TASK MODE-IF FORTRAN IV TASK, ZERO
                      ;         -IF ASSEMBLER ONLY, NON-ZERO

HE FOLLOWING CONSTANTS DEFINE THE SYSTEM CONFIGURATION
ERO:          DEVICE NOT IN THE SYSTEM
ON-ZERO:      NUMBER OF DEVICES CONFIGURED IN SYSTEM

USR   LPTS=   1       ;LINE PRINTER
USR   TTYS=   1       ;# TELETYPES ON CPU
USR   PTPS=   0       ;PAPER TAPE PUNCH
USR   CASS=   0       ;CASSETTE FILE
USR   ATODS=  10.     ;ANALOG TO DIGITAL CONVERTER
USR   PTRS=   0       ;PAPER TAPE READER
USR   CNTCT=  10.     ;CONTACT SENSE
USR   RLAYS=  10.     ;RELAYS
USR   DTOAS=  10.     ;DIGITAL TO ANALOG CONVERTER

DOS DISPLACEMENTS USED BY THE GOSEX PACKAGE

USR   USTP=   12      ;POINTER TO CURRENT PROGRAM'S UST
USR   TID=    12      ;TASK ID IN TASK'S TCB
USR   USTCT=  14      ;CURRENT TCB POINTER TO PROGRAM'S UST
USR   USTIN=  17      ;START OF USER NREL CODE IN UST

ORTRAN IV PARAMETER
HEN ZERO, ALL TASKS MUST BE IN ASSEMBLER
HEN NON-ZERO, THIS PARAMETER SPECIFIES THE MAXIMUM NUMBER OF
ARALLEL FORTRAN IV TASKS THAT MAY RUN AT ONE TIME   THE
```

```
DUSR FORT4=     0

CONTROLLER BLOCK DISPLACEMENTS

DUSR     WRKSP=   GENRL    ;WORK SPACE-2 WORDS
DUSR     SIGN=    GENRL+2 ;SIGN STORAGE
DUSR     ERORO=   GENRL+3 ;ERROR AT T
DUSR     EROR1=   GENRL+4 ;ERROR AT T-1
DUSR     EROR2=   GENRL+5 ;ERROR AT T-2
DUSR     EROR3=   GENRL+6 ;ERROR AT T-3
DUSR     ITSUM=   GENRL+7 ;INTEGRAL SUM STORAGE
DUSR     OUTPT=   GENRL+8. ;CURRENT OUTPUT ACTION
DUSR     SMOTH=   GENRL+9. ;;FILTERED DATA STORAGE
DUSR     CNTHI=   GENRL+10. ;HIGH LIMIT ALARM COUNTER
DUSR     CNTLO=   GENRL+11. ;LOW LIMIT ALARM COUNTER
DUSR     FILTR=   GENRL+12. ;FILTER CONSTANT
DUSR     VRBLE=   GENRL+13. ;ADDR OF RAW DATA
DUSR     INHI=    GENRL+14. ;INPUT HIGH LIMIT
DUSR     INLO=    GENRL+15. ;INPUT LOW LIMIT
DUSR     MESHI=   GENRL+16. ;MESSAGE # FOR HIGH LIMIT ALARM IN MESSG.MG
DUSR     MESLO=   GENRL+17. ;MESSAGE # FOR LOW LIMIT ALARM IN MESSG.MG
DUSR     AWAIT=   GENRL+18. ;INTERVAL BETWEEN SUCCESSIVE ALARMS
DUSR     SETPT=   GENRL+19. ;SET POINT
DUSR     KP=      GENRL+20. ;PROPORTIONAL GAIN
DUSR     KPD=     GENRL+21. ;PROPORTIONAL DERIVATIVE GAIN
DUSR     KPI=     GENRL+22. ;PROPORTIONAL INTEGRAL GAIN
DUSR     SCALE=   GENRL+23. ;SCALE FACTOR WITH ACTION DIRECTION
DUSR     DELAY=   GENRL+24. ;TIME INTERVAL BETWEEN CONTROL ACTIONS
DUSR     CASCD=   GENRL+25. ;CASCADE FLAG
DUSR     DESTN=   GENRL+26. ;ADDR OF DESTINATION DATA BLOCK OR
                            ;D/A CHANNEL #
DUSR     OUTHI=   GENRL+27. ;HIGH LIMIT ON OUTPUT
DUSR     OUTLO=   GENRL+28. ;LOW LIMIT ON OUTPUT
DUSR     MEAN=    GENRL+29. ;MEAN OUTPUT VALUE
DUSR     SPCON=   GENRL+30. ;ADDR OF CONSTANT BLOCK FOR SETPT CONVERSION
DUSR     SPAZE=   GENRL+31. ;ADDR OF ROUTINE TO CONVERT FROM
                            ;A/D TO ENGINEERING UNITS
DUSR     SPEZA=   GENRL+32 ;ADDR OF ROUTINE TO CONVERT FROM
                            ;ENGINEERING TO A/D UNITS
DUSR     SPCAP=   GENRL+33. ;MESSAGE # FOR SETPT CAPTION IN OPCOM.MG
DUSR     MNCON=   GENRL+34. ;ADDR OF CONSTANT BLOCK FOR MEAN CONVERSION
DUSR     MNAZE=   GENRL+35. ;ADDR OF ROUTINE TO CONVERT FROM
                            ;D/A (OR A/D) TO ENGINEERING UNITS
DUSR     MNEZA=   GENRL+36. ;ADDR OF ROUTINE TO CONVERT FROM
                            ;ENGINEERING TO D/A (OR A/D) UNITS
DUSR     MNCAP=   GENRL+37. ;MESSAGE # FOR MEAN CAPTION IN OPCOM.MG
DUSR     LPMES=   GENRL+38. ;LOOP'S DESCRIPTIVE MESSAGE # IN OPCOM.MG

         .EOT               ;END OF GOSEX PARAMETER TAPE
```

```
        .TITL SYMBL

;SYMBL TAPE # 1 OF 2
;USER BUILT SYMBOL TABLE
;THIS TAPE-GOSEX SYMBOLS

        .ENT USST,UEST

;EXECUTIVE FUNCTION SYMBOLS
        .EXTN START,CEASE,STATS,ENDIO,ENDAL,CLEAR,CHNGE,PARAM
        .EXTN TUNER,LOG,PLOT,READY,RELSE,RETRN,CLOCK,PROFL

;DEVICE SYMBOLS
        .EXTN CONSL,PRINT,TTYPE,PUNCH,DISK,CASET,READR

        .NREL

;USER MUST DECLARE THE FOLLOWING AT THE END OF TAPE # 2
;       SST:    LAST SYMBOL-CODE WORD # 1
;
;       UEST=   EST
;       USST=   SST

;ALSO, THE USER MUST INCLUDE THE APPROPRIATE  EXTN
;DECLARATIONS FOR TAPE # 2   THIS MAY BE DONE IN TAPE # 2 ITSELF
;IN THIS INSTANCE, IT IS NECESSARY THAT THE EXTERNAL
;DECLARATIONS PRECEED ANY OTHER STATEMENTS

;CODING TABLE FOR RADIX 50 SYMBOLS

;       CHARACTER       VALUE(DECIMAL)
;       NULL            0
;       0               1.
;       1               2
;       2               3.
;       3               4.
;       4               5.
;       5               6.
;       6               7.
;       7               8.
;       8               9.
;       9               10.
;       A               11.
;       B               12.
;       C               13.
;       D               14.
;       E               15.
;       F               16.
;       G               17.
;       H               18.
;       I               19.
;       J               20.
;       K               21.
;       L               22
;       M               23
;       N               24.
;       O               25.
;       P               26.
;       Q               27.
```

```
;       R               28.
;       S               29.
;       T               30.
;       U               31.
;       V               32.
;       W               33.
;       X               34.
;       Y               35.
;       Z               36.
;       .               37.

;EXAMPLE OF CODING:
;SYMBOL "ABCDE" MAY BE ENCODED BY THE USER IN THE FOLLOWING WAY
;       [D]*40.+[E]*32.         ;SYMBOL CODE WORD # 2
;       [A]*40.+[B]*40.+[C]     ;SYMBOL CODE WORD # 1

;WHERE [#] DECOTES THE CODING VALUE OF THE CHARACTER AS
;GIVEN IN THE TABLE ABOVE

;WRITTEN OCTOBER, 1974 BY
;       JEAN-PIERRE TREMBLAY
;       DEPARTMENT OF CHEMICAL ENGINEERING
;       MCMASTER UNIVERSITY
;       HAMILTON, ONTARIO, CANADA
```

```
0003   SYMBI
00000'000000'EST:


                   ;EXECUTIVE FUNCTIONS LIST
00001'177777             START          ;SYMBOL VALUE-USUALLY ESTABLISHED AT LOAD
00002'107700             36800.         ;SYMBOL CODE WORD # 2
00003'134773             47611          ;SYMBOL CODE WORD # 1
00004'177777             CEASE
00005'111340             37600.
00006'051643             21411.
00007'177777             STATS
00010'114640             39328.
00011'134773             47611.
00012'177777             ENDIO
00013'061040             25120.
00014'060616             24974.
00015'177777             ENDAL
00016'034700             14784.
00017'060616             24974.
00020'177777             CLEAR
00021'035200             14976.
00022'052277             21695.
00023'177777             CHNGE
00024'053340             22240.
00025'052050             21544.
00026'177777             PARAM
00027'034740             14816.
00030'122124             42068.
00031'177777             TUNER
00032'047200             20096.
00033'140160             49264.
00034'177777             LOG
00035'000000             0
00036'106571             36217.
00037'177777             PLOT
00040'113000             38400.
00041'123011             42505.
00042'177777             READY
00043'045140             19040.
00044'130543             45411.
00045'177777             RELSE
00046'111340             37600.
00047'130556             45422.
00050'177777             RETRN
00051'107400             36608.
00052'130566             45430.
00053'177777             CLOCK
00054'041640             17312
00055'052311             21705.
00056'177777             PROFL
00057'051300             21184.
00060'123371             42745.

                   ;DEVICES LIST
00061'177777             CONSL
00062'111700             37824.
00063'052500             21824.
00064'177777             PRINT
00065'075700             31680.
00066'123363             42739.
```

```
00004  SYMBL
 00067'177777          TTYPE
 00070'101740          33760.
 00071'140123          49235.
 00072'177777          PUNCH
 00073'041500          17216.
 00074'123560          42864.
 00075'177777          DISK
 00076'064400          26880.
 00077'055225          23189.
 00100'177777          CASET
 00101'047300          20160.
 00102'051425          21269.
 00103'177777          READR
 00104'044600          18816.
 00105'130543          45411.

           ;CONTROLLER PARAMETERS LIST-VALUE SPECIFIED IN PARAMETER TAPE
000106'000026          FILTR
000107'114600          39296.
000110'063416          26382.
000111'000030          INHI
000112'057400          24320.
000113'075222          31378.
000114'000031          INLO
000115'076400          32000.
000116'075226          31382.
000117'000035          SETPT
000120'102700          34240.
000121'133666          47030.
000122'000036          KP
000123'000000          0
000124'103520          34640.
000125'000037          KPD
000126'000000          0
000127'103536          34654.
000130'000040          KPI
000131'000000          0
000132'103543          34659.
000133'000041          SCALE
000134'067740          28640.
000135'133523          46931.
000136'000042          DELAY
000137'035540          15200.
000140'054756          23022.
000141'000045          OUTHI
000142'056140          23648.
000143'120466          41270.
000144'000046          OUTLO
000145'070440          28960.
000146'120466          41270.
000147'000047          MEAN
000150'074000          30720.
000151'111043          37411.

                        .EOT          ;END OF SYMBL TAPE # 1 OF 2
```

```
0005   SYMBI
                          ;SYMBOL OF TAPE #2 OF 2
                           . EXTN SOLFT, FFLO1, SFLO1, WAIT, TWAIT
                           . EXTN FEEFT, FHL, FLL, FLG
                           . EXTN FEETC, FNOFF, FTSET, ADFTE
                           . EXTN SOLTC, SNOFF, STSET, ADSTE
                           . EXTN FFLOW, OPERF, FFCON, FTEMP
                           . EXTN SFLOW, OPERS, SFCON, STEMP
                           . EXTN SOLFM, FEEFM, CONVB, ADT
                           . EXTN DATUM, SCAN
                           . EXTN OUTBS, OUTBF
                           . EXTN SSAMP

                  ;FEED TEMPERATURE CONTROL LOOP PARAMETERS
00152'177777               FEETC
00153'113640               30. *40. +13. *32.
00154'063147               16. *40. +15. *40. +15.
00155'177777               FNOFF
00156'051500               16. *40. +26. *32
00157'063731               16. *40. +24. *40. +25.
00160'177777               FTSET
00161'047300               15. *40. +30. *32.
00162'064315               16. *40. +30. *40. +29.
00163'177777               ADFTE
00164'113740               30. *40. +15. *32
00165'043400               11. *40. +14. *40. +16
00166'177777               FTEMP
00167'073100               23. *40. +26. *32
00170'064277               16. *40. +30. *40. +15.
                  ;SOLVENT TEMPERATURE CONTROL LOOP PARAMETERS
00171'177777               SOLTC
00172'113640               30. *40. +13. *32
00173'134476               29. *40. +25. *40. +22.
00174'177777               SNOFF
00175'051500               16. *40. +26. *32
00176'134431               29. *40. +24 *40. +25.
00177'177777               STSET
00200'047300               15. *40. +30. *32
00201'135015               29. *40. +30. *40. +29
00202'177777               ADSTE
00203'113740               30. *40. +15. *32
00204'043415               11. *40. +14. *40. +29
00205'177777               STEMP
00206'073100               23. *40. +26. *32.
00207'134777               29. *40. +30. *40. +15.
00210'177777               ADT
00211'000000               0. *40. +0. *32
00212'043416               11. *40. +14. *40. +30
                  ;FLOWRATES(FEED AND SOLVENT)CONTROL LOOP PARAMETERS
00213'177777               SFLOW
00214'100440               25. *40. +33. *32
00215'133726               29. *40. +16. *40. +22.
00216'177777               OPERS
00217'107640               28. *40. +29. *32
00220'120137               25. *40. +26. *40. +15.
00221'177777               SFCON
00222'100000               25. *40. +24. *32
00223'133715               29. *40. +16. *40. +13.
00224'177777               FFLOW
00225'100440               25. *40. +33. *32.
```

```
0006   SYMBI
 00226'063226          16. *40 +16. *40. +22
 00227'177777          OPERF
 00230'107000          28. *40 +16. *32
 00231'120137          25 *40. +26. *40. +15
 00232'177777          FFCON
 00233'100000          25. *40. +24. *32
 00234'063215          16. *40 +16 *40. +13
 00235'177777          SOLFM
 00236'051340          16. *40 +23. *32
 00237'134476          29. *40 +25. *40 +22
 00240'177777          FEEFM
 00241'051340          16. *40. +23 *32
 00242'063147          16. *40. +15. *40 +15
 00243'177777          OUTBS
 00244'037640          12 *40 +29. *32
 00245'120466          25 *40. +31. *40. +30.
 00246'177777          OUTBF
 00247'037000          12 *40. +16. *32
 00250'120466          25. *40. +31 *40. +30
 00251'177777          DATUM
 00252'116740          31. *40. +23. *32
 00253'054526          14. *40 +11. *40 +30
 00254'177777          SCAN
 00255'074000          24. *40 +0. *32
 00256'133523          29. *40 +13. *40. +11
 00257'177777          CONVB
 00260'120600          32 *40. +12 *32
 00261'052500          13. *40. +25. *40. +24
                       ;SQUARE WAVE TO FEED RATE ROUTINE
 00262'177777          FEEFT
 00263'051700          16. *40 +30. *32.
 00264'063147          16. *40. +15 *40 +15.
 00265'177777          FHL
 00266'000000          0. *40. +0. *32.
 00267'063346          16. *40. +18. *40 +22
 00270'177777          FLL
 00271'000000          0. *40. +0. *32.
 00272'063606          16. *40. +22. *40. +22.
 00273'177777          FLG
 00274'000000          0. *40. +0. *32.
 00275'063601          16 *40. +22 *40. +17.
 00276'177777          SOLFT
 00277'051700          16 *40 +30 *32
 00300'134476          29. *40. +25 *40. +22
 00301'177777          FFLO1
 00302'076500          25 *40. +2 *32
 00303'063226          16. *40 +16. *40. +22
 00304'177777          SFLO1
 00305'076500          25 *40. +2 *32
 00306'133726          29. *40 +16 *40 +22
 00307'177777          WAIT
 00310'113000          30. *40 +0 *32
 00311'150013          33. *40 +11. *40 +19.
 00312'177777          TWAIT
 00313'061300          19. *40. +30 *32
 00314'140263          30 *40. +33. *40. +11
                ;SAMPLING MONITOR ROUTINE
 00315'177777          SSAMP
 00316'073100          23 *40 +26 *32
```

```
0007   SYMBL
 00317'134723 SST:       29.*40.+29.*40.+11.
      000000'UEST=EST
      000317'USST=SST
                     .END
```

```
NE #   LINE AS READ FROM SOURCE FILE
0000   <15><12>
0001   <12>NOT AN EXECUTIVE FUNCTION
0002   <12>EXECUTIVE FUNCTION NOT AVAILABLE
0003   <12>TASK NAME:
0004   <12>NOT A USER TASK<15>
0005   <12>USER TASK CREATED
0006   <12>TASK IS ALREADY ACTIVE
0007   <12>USER TASK KILLED
0008   <12>TASK IS ALREADY INACTIVE
0009   <12>TASK IS INACTIVE
0010   <12>TASK IS ACTIVE
0011   <12>DEVICE NAME:
0012   <12>NOT A SYSTEM DEVICE<15>
0013   <12>DEVICE NOT CONFIGURED FOR USE
0014   <12>SPECIFIED DEVICE DEACTIVATED
0015   <12>DEVICE NOT IN USE
0016   <12>ALL CONFIGURED DEVICES DEACTIVATED
0017   <12>DISK DIRECTORY & FILENAME:
0018   <12>DELETION CONFIRMED
0019   <12>FILENAME ERROR<15>
0020   <12>CURRENT DISPLAY INTERVAL(SECONDS):
0021   <12>RESET DISPLAY INTERVAL(SECONDS):
0022   <12>CHANGE CONFIRMED
0023   <12>INTERVAL IMPROPERLY SPECIFIED<15>
0024   <12>INTERVAL TOO LONG FOR CURRENT RTC FREQUENCY<15>
0025   <15><12><40>*?*.
0026   <15><12><40>
0027   <15><12>LOOP NAME:
0028   <12>NOT A CONTROLLER LOOP<15>
0029   FILTER CONSTANT(0 TO +1000.)/<40>
0030   INPUT HIGH ALARM LIMIT-
0031   INPUT LOW ALARM LIMIT-
0032   CURRENT SET POINT-
0033   PROPORTIONAL GAIN(+VE)/<40>
0034   PROPORTIONAL-DERIVATIVE GAIN(+VE)/<40>
0035   PROPORTIONAL-INTEGRAL GAIN(+VE)/<40>
0036   SCALE FACTOR & DIRECTION OF ACTION/<40>
0037   CONTROL INTERVAL(SECONDS)/<40>
0038   OUTPUT HIGH LIMIT-
0039   OUTPUT LOW LIMIT-
0040   OUTPUT MEAN-
0041   <15><12>CONVERSION ERROR
0042   <15><12><40>*?*<15><12>
0043   <12>FILE IN USE
0044   <12>DELETION NOT POSSIBLE, CODE<40>
0045   <12>VARIABLES LIST-MAX:
0046   <12>DEVICE IN USE
0047   <12>DEVICE NOT EQUIPPED FOR LOGGING
0048   <12>DISPLAY MODE:
0049   <12>NO LIST<15>
0050   <12>END OF LIST<15>
0051   <12>LOGGING INTERVAL(SECONDS):
0052   <12>OUTPUT TASK CREATED
0053   <12>UNABLE TO INITIATE OUTPUT TASK
0054   <12>CASSETTE UNIT & FILE #:
0055   <12>DEVICE NOT EQUIPPED FOR PLOTTING
0056   <12>SCALE FACTOR:
0057   <12>PLOTTING INTERVAL(SECONDS):
0058   <12>MIN RANGE OF PLOT:
```

```
0059    <12>MAX RANGE OF PLOT:
0060    <15><12>USER IN BACKGROUND
0061    <15><12>USER IN FOREGROUND
0062    <15><12>USER BUILT SYMBOL TABLE NOT DEFINED<15><12>
0063    <15><12>RELAY BIT STATUS TABLE NOT DEFINED<15><12>
0064    <15><12>A/D READ DATA STORAGE NOT DEFINED<15><12>
0065    <15><12>USER CLOCK SET TO RTC FREQUENCY<15><12>
0066    <15><12>USER CLOCK NOT DEFINED<15><12>
0067    <15><12>CONTACT SENSE DISPATCH TABLE NOT DEFINED<15><12>
0068    <15><12>USER SHUTDOWN ROUTINE NOT DEFINED<15><12>
0069    <15><12>USER TASK DEFINITION TABLE NOT DEFINED<15><12>
0070    <15><12>ERROR DETECTED IN USER TASK DEFINITION TABLE<15><12>
0071    <15><12>USER ALARM DEFINITION TABLE NOT DEFINED<15><12>
0072    <15><12>USER MESSAGE FILE NOT DEFINED<15><12>
0073    <15><12>EXECUTION ABORTED<15><12>
0074    <15><12>AOK-TO RUN, STRIKE ANY KEY<15><12>
0075    <12>DIRECTORY OR DEVICE NAME:
0076    <12>DIRECTORY OR DEVICE INITIALIZED
0077    <12>DIRECTORY OR DEVICE ALREADY INITIALIZED
0078    <12>INITIALIZATION NOT POSSIBLE, CODE<40>
0079    <12>DIRECTORY OR DEVICE RELEASED
0080    <12>DIRECTORY OR DEVICE IN USE
0081    <12>RELEASE NOT POSSIBLE, CODE<40>
0082    <15><12>USER SPACE RELEASED<15><12>
0083    <15><12>CONTROL LOOP DEFINITION TABLE NOT DEFINED<15><12>
0084    <15><12>ERROR DETECTED IN CONTROL LOOP DEFINITION TABLE<15><12>
0085    <15><12>ERROR DETECTED IN CONTACT SENSE DISPATCH TABLE<15><12>
0086    <12>AXIAL POSITION:
0087    <15><12>UDEF EXECUTIVE FUNCTION LIST EXTENSION NOT DEFINED<15><12>
0088    UNUSED
0089    UNUSED
0090    UNUSED
0091    UNUSED.
0092    UNUSED
0093    UNUSED
0094    UNUSED
0095    UNUSED
0096    UNUSED
0097    UNUSED
0098    UNUSED
0099    UNUSED
0100    MILLIMETERS*10./MINUTE<40>
0101    %VALVE OPEN<40>
0102    <12>SOLVENT FLOWRATE CONTROL LOOP<15>
0103    <12>FEED FLOWRATE CONTROL LOOP<15>
```

MESSG.

```
INE #   LINE AS READ FROM SOURCE FILE
00000   <12>SOLVENT RATE NEAR MAXIMUM<15>
00001   <12>SOLVENT RATE NEAR MINIMUM<15>
00002   <12>FEED RATE NEAR MAXIMUM<15>
00003   <12>FEED RATE NEAR MINIMUM<15>
00004   <12>OVERHEAD SIGNAL ON <15>
00005   <12>OVERHEAD SIGNAL OFF <15>
00006   <12>BOTTOM SIGNAL ON <15>
00007   <12>BOTTOM SIGNAL OFF <15>
```

```
TIL SV        LOADED BY RLDR REV 03 AT 12:29:51 10/12/75
NFCE
SCON
SAMP
WAVE
YMBL
NTRP
EYED
IDER
INVT
       006002
       000,000   START
                 CEASE
                 STATS 000216
       000,001   ENDIO
                 ENDAL
                 CLEAR
                 CHNGE 000502
       000,002   PARAM 000560
       000,003   TUNER 000613
       000,004   LOG   000455
       000,005   HEADL 000320
       000,006   PLOT  000502
       000,007   HEADP 000314
       000,010   PREP  000536
       000,011   READY
                 RELSE
                 RETRN
                 CLOCK 000263
       000,012   PROFL 000506
       000,013   PROHD 000274
       007002
OGER
LTER
FILE
OUNS
VLY
ERMO
ACAL
XMT
PRI
IDC
WMOD
 CONTL 001274
 QTCK  011071


  NMAX 012076
  ZMAX 000123
  CSZE 000000
   EST 000000
   SST 000000

 TIMER 000056
 PCTMP 000101
  TSAV 000102
 .TMN1 000103
 CTCB  000104
 OVLAY 000633
 ALARM 000641
 DATUM 000652
 RELAY 000664
 SENSE 000676
```

```
SCAN   000710
CLOOP  000711
SUPRS  000716
TASKS  000727
UDEF   000764
FEETC  000765
FNOFF  001010
ADFTE  001012
FTSET  001013
FEETB  001022
STEMP  001040
FTEMP  001041
CONVB  001043
SOLTC  001046
SNOFF  001115
ADSTE  001120
STSET  001121
ADT    001133
SOLTB  001137
SOLFM  001151
SFLOW  001177
OPERS  001200
SFCON  001201
OUTBS  001204
SOLVB  001207
FEEFM  001270
FFLOW  001316
OPERF  001317
FFCON  001320
OUTBF  001323
FEEFB  001326
.UIEX  001400
.UPEX  001400
.UCEX  001400
SSAMP  001407
TRAP   001466
SOLFT  001476
WAIT   001525
SFLO1  001557
FFLO1  001560
FEEFT  001561
TWAIT  001600
FLG    001615
FHL    001616
FLL    001617
UEST   001620
USST   002137
INTRP  002140
MODE   002244
START  002270
CEASE  002273
STATS  002276
ENDIO  002301
ENDAL  002304
CLEAR  002307
CHNGE  002312
PARAM  002315
TUNER  002320
LOG    002323
PLOT   002326
READY  002331
RELSE  002334
RETRN  002337
CLOCK  002342
```

```
MESS   002355
MNITR  002412
MTFLX  003154
ADDCT  003221
CTDCT  003260
DEVCE  003335
CONSL  004207
PRINT  004213
TTYPE  004217
PUNCH  004220
DISK   004221
CASET  004233
READR  004234
OPCOM  004235
MESSG  004243
.. BDEC 004750
.. BOCT 005001
.. DBIN 005102
.. OBIN 005150
.. BFLT 005217
.. FBIN 005412
EXC01  006002
EXC04  006002
EXC08  006002
EXC09  006002
 EXC10 006002
 EXC12 006002
 EXC14 006002
 EXC15 006002
 EXC17 006002
 EXC11 006004
 EXC13 006004
 EXC20 006004
 ERROR 006050
 DSKMG 006051
 MTPLY 006052
 MTPLA 006053
 DIVID 006054
 EXC16 006054
 ROUND 006055
 KEYED 006057
 DECOD 006060
 ENCOD 006061
 LOCAT 006062
 RETRV 006063
 RANGE 006064
 BDEC  006065
 EXC02 006065
 BOCT  006066
 DBIN  006067
 OBIN  006070
 BFLT  006071
 FBIN  006072
 .OVKI 006073
 .TOVL 006074
 .OVRE 006075
 .OVEX 006076
 .KILL 006077
 .TASK 006100
  AKIL 006105
 .ASUS 006106
 .AFEN 006106
 .ARDY 006107
 .AUNP 006107
 .IXMT 006110
```

```
 XMT  006111
 XMTW 006112
 REC  006113
 PRI  006114
 IDST 006115
 TIDS 006116
 TIDR 006117
 TIDK 006120
 TIDP 006121
 SMSK 006122
EXC17 006125
EXC05 006127
EXC03 006150
EXC18 006155
EXC06 006232
EXC07 006341
LOGER 007002
PAUZE 007203
TFORM 007301
DATER 007327
PLTER 007401
PTEXT 007414
CLEAN 007633
NUMBS 007652
PFILE 007733
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LNEAR 010246
QNVRS 010317
LNVRS 010320
OVWRT 010650
TMAX  010763
TNXT1 011007
QTCNT 011067
LQTSC 011070
SYST2 011077
SYST1 011102
TLINK 011115
LNKPR 011155
TREL  011162
INHB  011170
TUNLK 011200
TSKER 011331
TERCM 011332
TIDSR 011355
XSRH  011622
TPRO1 011675
CONTL 177777
QTCK  177777
OVY01 000,000
OVY02 000,000
OVY03 000,000
OVY04 000,001
OVY05 000,001
OVY06 000,001
OVY07 000,001
OVY08 000,002
OVY09 000,003
OVY10 000,004
OVY11 000,005
OVY12 000,006
OVY13 000,007
OVY14 000,010
OVY15 000,011
OVY16 000,011
OVY17 000,011
OVY18 000,011
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