DYNAMIC SIMULATION

OF A

MULTIPRODUCT CHEMICAL PLANT

by

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SCOPE AND CONTENTS:

This report deals with the use of the "Modular Approach", in the simulation of change-over operations, for a continuous multiproduct chemical plant.

A dynamic model of the continuous fat hydrolysis process was built, within the DYNSYS framework, with the purpose of evaluating the usefulness of this programme in simulating different change-over operation policies.

Operator's activities were simulated making use of an interactive version of DYNSYS, demonstrating its value for studying alternate changeover policies and possibly as a training device.

Supervisory control was implemented in order to automate the change-over operations, simulating the functions of a small computer controlling the process.

Improved operating policies for change-over are proposed and the validity of the simulation is discussed from both a practical and a theoretical point of view.

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1. INTRODUCTION

1.1 Incentive for Multiproduct Plants

In the early days of the modern chemical industry, batch operation, processing different materials in the same equipments, was predominant and can be considered as the first manifestation of multiproduct plants. Explosive expansion in the markets for chemicals and technological advances including the development of the concept of unit operations and process control, marked the decline of batch processing and the advent of continuous processes featuring a smaller labor requirement, easier control of equipment, improved grade of final product and larger capacities. However, profitable operation of continuous processes may require large production volumes; therefore in countries with small local markets, the operation of "small" continuous plants is not advantageous. As a result of this situation, large plants in highly developed countires started getting larger and absorbing international markets, making virtually impossible the profitable operation of small plants.

A possible solution to this problem, which may enable small local firms to compete with the large ones is to operate economically-viable plants as multiproduct plants. This may require greater technological skill, in view of the fact that the operation of such plants will require frequent changes in conditions, while changing

from one product to another. Technological advances in the field of computerized process control and better understanding and use of chemical engineering principles, may provide the means for designing and operating such multiproduct plants competitively.

Multiproduct plants can be classified as:

Complete batch plants: In this type of plant usually several units, each producing one or more products, operate in parallel. Operational problems in this case are materials handling and scheduling of the units; since more than one operation can take place in the same equipment,solution of these kinds of problems belongs to the province of Operations Research and no further mention will be made here.

Continuous plants with changes in raw materials: This type of plant operates in a continuous way but processes different materials; it can be considered as a continous process subjected to disturbances. The interest here is to perform the changes at the lowest cost possible and to achieve this goal, a series of proper manipulations in the controlling variables must be carried out; these require a good understanding of the dynamic behaviour of the plant under such circumstances.

1.2 Why Simulation?

As pointed out before, understanding of the dynamic behaviour of the process is required if we want to operate continuous multi-product chemical plants. Today an engineer can obtain that understanding of the process by diverse methods ranging from pilot plant studies, to collection of operating data in actual industrial plants, to the development and use of computer simulations. The first has been employed many times in the past with significant successes, but suffers from some drawbacks. First, if time is short, a pilot plant study may not be possible. Second, there is always uncertainty about the successful performance of commercial sized equipment that has been scaled-up from pilot plant data. Third, and most important, of all development cost, the largest item is usually for pilot plant construction and operation.

The second method, data acquisition in actual plants, can be used for improving control operations through techniques such as system identification and adaptive control, but its usefulness at the design step is limited for all practical purposes to updating the design of a process that is being replicated; this is a severe restriction from a general point of view.

The third method is simulation, which consists of developing a model of the process and then operating the model to observe its performance. This kind of study can avoid most of the major difficulties of the other two methods examined here. Simulations can be done in a relatively short time, at the design stage or later, and at lower cost. The amount of time and the effort required may depend on the questions to be answered and the degree of reliability required of the answer.

The term model can have various connotations including physical models, analogs, drawings, and mathematical representations. Himmelblau gives a complete classification of the models used in chemical engineering [6].

1.3 Guidelines for Digital Simulation of Processes

In preparing a simulation study two major stages can be considered; the representation of the process and the solution of the equations involved in the model.

The first is concerned with the construction of a mathematical model to represent the process. An important thing to bear in mind when using models for any purpose is that a model can never give a complete description of the actual events, first because of necessary simplifications, second because of inaccurate knowledge of model parameters, and third because of intentional omission of a number of factors. A perfect model can only be an identical replica of the process itself.

The second stage is the solution of the set of equations which constitute the simulation model. In general, for chemical processes, the models will be lumped or distributed parameter models involving ordinary differential equations, partial differential equations, nonlinear algebraic equations, etc. whose solution requires, in most of the cases, use of numerical techniques. An excellent review of these is given by Carnahan et.al. [2].

1.4 Modular Approach to Dynamic Simulation

Literature concerning steady state simulation is rather large, but only a few works have been published dealing with dynamic behaviour of complete chemical plants [1, 5, 12, 23].

Different approaches to dynamic simulation have been used by different authors and they can be classified under two major groups:

1) Equation oriented systems.

2) Equipment or modular systems.

The first considers the problem of solving simultaneously a set of differential equations representing the process, in some cases Laplace transforms are used and analytical solutions can be obtained; unfortunately, this is restricted to linear or easily linearized systems. In the more general case numerical solutions are used.

The second, modular approach, considers that as a chemical plant is made up of processing units such as heat exchangers, reactors, etc., with connecting lines of material flows and controllers activating controlling devices - each unit can be described by mathematical relationships relating outputs to inputs. The assembly of mathematical equations for simulation of one of these process units, including controllers and control devices, in a computer subroutine is defined as a unit computation or computation module.

Just as the chemical plant is made up of assemblies of physical units, a mathematical model of the plant can be a "network" of unit computations, among which information will flow in a manner analogous to the material flow or control signals in the real plant; a graphical representation of this network is called the dynamic information flow diagram. Figure 1 shows the instrumented flow diagram of a simple chemical process and its corresponding dynamic information flow diagram; it is seen that there is a very close similarity between them.

The modular approach is believed to have three advantages. Firstly, the transformation of the physical plant to a dynamic information flow diagram, in preparation for the simulation, is facilitated by its close correspondence with the instrumented process flow diagram. This is of particular value for the chemical engineer who is accustomed to consider the process as a collection of unit operations, reactors and control devices. Furthermore, study of alternative plant and control configurations is greatly facilitated

Secondly, as a library of equipment and control modules becomes available it may be readily used for simulation of new plant, provided that the modules have a reasonable level of generality. Thus, the programming effort for new plants is appreciably reduced.

Finally, for dynamic studies, the modular approach, as used in this work, deals with the real variables of the process, rather than with transformed variables (i.e. Fourier transform uses frequencies instead of time) and the modules may be quite non-linear. This should encourage the use of modular approach by process and design engineers, even if they are not familiar with control theory and terminology.

The major disadvantage of the modular approach, compared with equation oriented systems, is that usually the computational time required for the execution of a case in the latter method will be less.

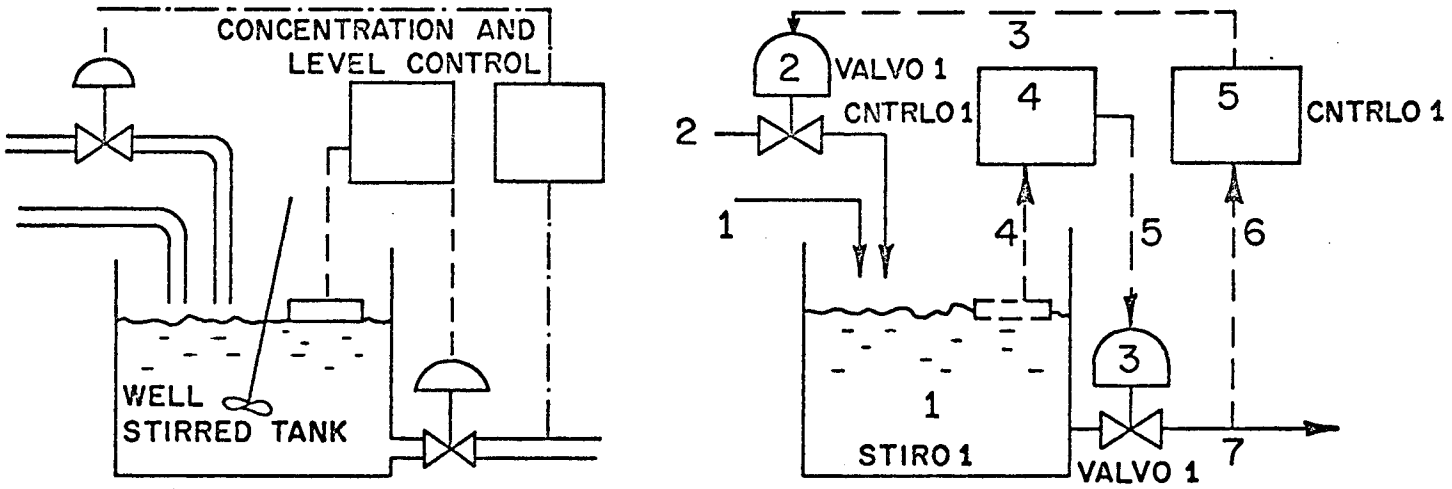


FIGURE I.

INSTRUMENTED PROCESS

DYNAMIC INFORMATION
FLOW DIAGRAM

1.5 DYNSYS - A Dynamic Systems Simulation Programme

DYNSYS is the name given to a system which has been developed by Professor Johnson and his students [1] at McMaster University, for the simulation of the transient behaviour of manufacturing processes. Its approach is modular, that is, based on computing blocks, and equipment oriented in that its basic blocks are chemical plant equipments rather than differential equations as with equation oriented systems like MIMIC.

DYNSYS is similar to a number of steady state simulation systems, such as PACER [3], GEMCS [7], CHESS [22], in its handling of plant equipment information, and like these programmes it calls subroutines for the "unit computations" in a predetermined order, and then repeats the sequence. However, as these programmes are concerned with iterating towards a steady state value, with no relation whatsoever to the time, as independent variable , in DYNSYS each sequence of computations is associated with a given time.

Two general classes of unit computations are handled by the DYNSYS programme:

(i) The Differential Unit Computations contain capacity as parameter; their mathematical description includes sets of differential equations that must be solved by the numerical integration routine which forms part of the executive system. The STIRϕI of Figure 1 is an example of this kind of module.

(ii) Algebraic Modules do not contain capacity and their output streams depend entirely on the current values of the input streams. A valve or tee in a pipe line is an example of an algebraic module.

In addition the consideration of process and instrument lags has become important for the simulation of the dynamic behaviour of chemical processes.

Details of the information handling system, and the integration routine procedure used by DYNSYS have been given by Bobrow, Ponton and Johnson [1], so no further reference will be done here.

DYNSYS can operate either batchwise or interactively. In the latter mode the engineer using it is able to make changes through the keyboard of a teletype or C.R.T.* terminal while observing the behaviour of the systems on the printed output or screen of the unit. This method has been demonstrated to be of great value in "getting acquainted" with the plant, and may be useful in training operators and testing new control policies.

*Cathode Ray Tube.

2: The Continuous Fat Hydrolysis Plant

2.1 General Description of the Process

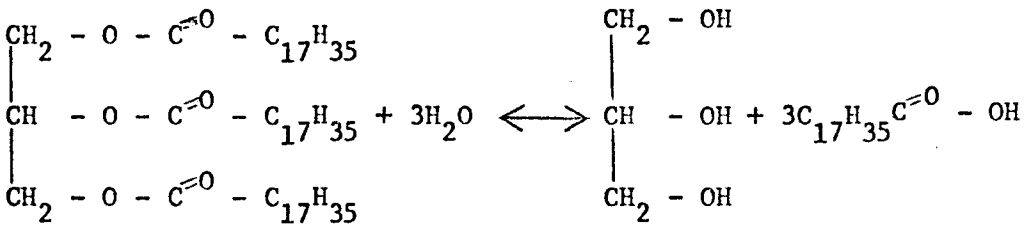
The hydrolysis of animal and vegetable fats to produce glycerine and fatty acids is a long standing chemical process. Originally this process was carried out batchwise by saponifying animal fat with caustic alkali thereby obtaining soap and glycerine. Fatty acids were obtained by acidifying the soap.

Nowadays, soap is seldom manufactured by a strictly batchwise process; continuous high pressure hydrolysis of fats by means of hot water in a spray tower has become the conventional process for the production of glycerine and fatty acids.

Figure 2 is an instrumented process flow diagram of a fat hydrolysis plant similar to that operated by Procter and Gamble of Canada Ltd. in Hamilton; the purpose of this unit is to carry out the conversion of oil and fats fed to it to a distilled fatty acid, suitable for soap manufacturing and a glycerine solution for further concentration.

Briefly, the continuous hydrolysis of fats is performed by feeding water at 450° - 500°F and 600 - 700 psia pressure through sprays in the top of the column and the fat mixed with live steam at

the bottom. Droplets of water pass down the column and come in contact with the partially hydrolyzed fat ascending from the base of the column. A small amount of water dissolves in the fat phase and there reacts chemically producing glycerine and the corresponding fatty acids. Using tristearin (one of the main components of tallow) as an example the basic reaction carried out in the column is:



TRISTEARIN GLYCERINE STEARIC ACID

This chemical reaction takes place in the presence of zinc oxide catalyst; the glycerine so formed distributes itself between the fat and water phases but is mainly in the water, with the result that glycerine is extracted from the fat and leaves the base of the column as dilute aqueous solution called "sweetwater". The hydrolysis of the fat is 97 - 99.5% complete by the time the product leaves the column. Therefore, the continuous phase leaving is essentially free fatty acid containing small amounts of undesirable impurities which would impart an objectionable colour. Then it is passed through a flash tank to eliminate all the water carried by the fatty acid. The fatty acid free of water is passed to the fatty acid still, where a mixture of

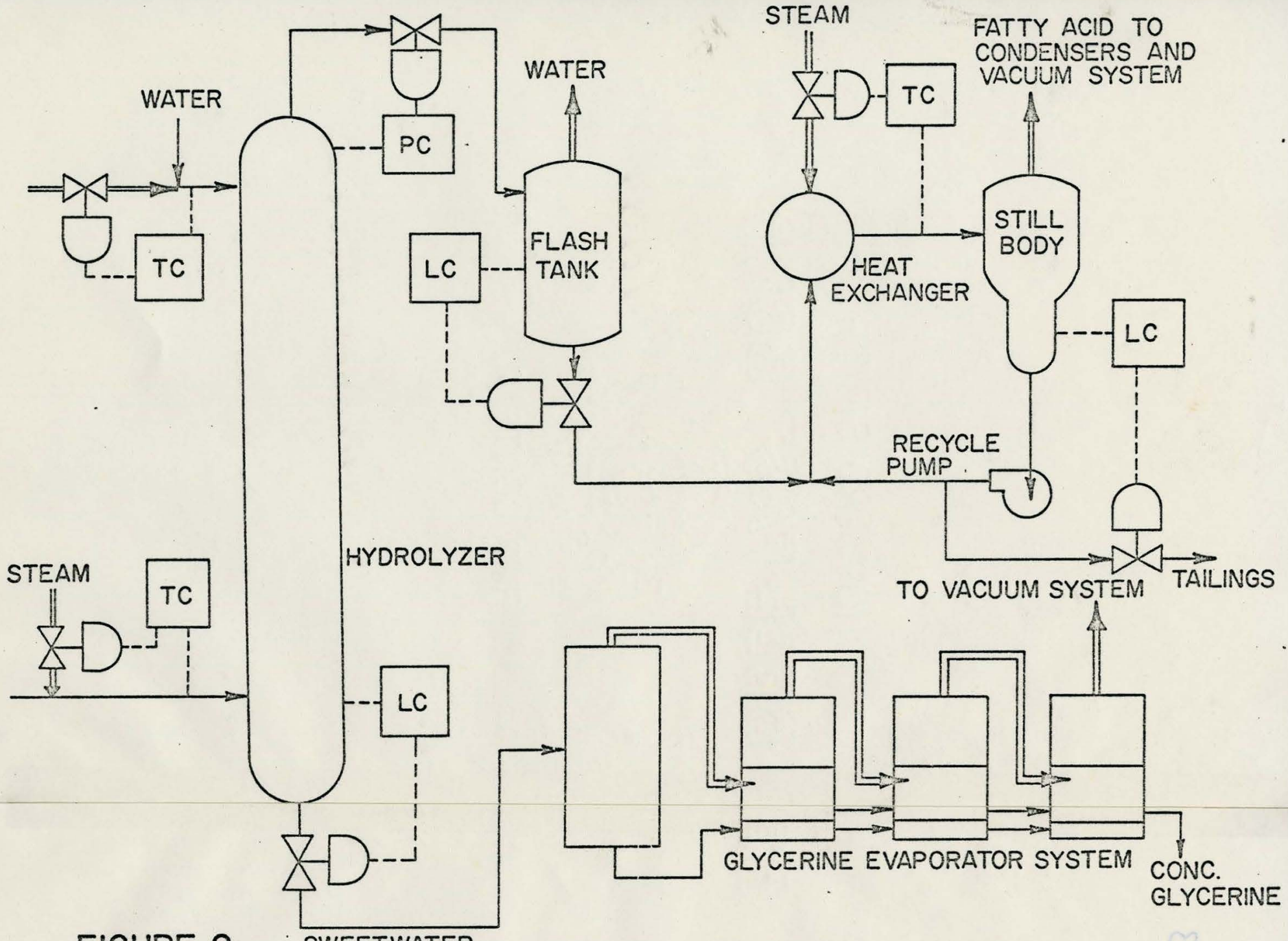


FIGURE 2.
INSTRUMENTED PROCESS FLOW DIAGRAM OF A FAT HYDROLYSIS PLANT

fatty acid and non-volatile material is recirculated through heat exchangers where it is heated. The hot liquid enters the still body where the fatty acids flash off and flow to the condensers. A liquid containing all the non-volatiles and part of the fatty acid accummulates in the still body, whence it is recirculated by the recycle pump. The heat exchangers are steam heated and the outlet temperature is controlled by regulating the flow of steam.

From the recirculation pump a tailings stream is drained in order to maintain the amount of non-volatile in the recycle constant. The flow of this stream is controlled according to the liquid level in the still body. For higher grade fats and oils, the tailings can be recirculated to the blending tanks for further hydrolysis; in poorer grade fats the tailings must be removed and burned.

The sweetwater leaving the bottom of the column, under interface level control, is fed to the glycerine evaporator system. Flashing occurs as pressure is reduced on entering the flash tank; the vapour produced provides the heating medium for the first effect in which the solution is further concentrated and vapour for the second effect is generated. The evaporator system is a conventional forward feed unit, vapour from the third effect going to the steam jet vacuum system.

The glycerine solution from the evaporator has a concentration of 75-80% and is the second major product of the unit.

Conventional proportional plus integral (PI) controllers act upon outlet streams in the flash tank in order to maintain the liquid levels and in the column to control the pressure.

2.2 Operation of the Plant

The plant which has been described is operated under conditions of frequent change in the oil or fat used as feedstock. For this reason, this plant has been selected as an example of a multiproduct plant.

The change in feed can be considered as a series of one or more step changes in the input, followed by corresponding manipulations of the control variables in order to conduct the plant in the best way possible to the steady state conditions. For each new feed there are steady state operating conditions which must be achieved by having the operator make some adjustments in the plant. Steady state simulation [8] of this plant has provided such conditions.

It is important to realize that a change in feed will produce a disruption in the operation of the plant, so there may be some control policies for bringing about the change in the minimum possible time, and with the least amount of off grade material which, in the best case, may be used as lower grade products, and in the opposite, disposed as waste, with consequent economic implications.

2.2.1 Change-over Operations

There are two types of change-over procedures which will be examined using the modular approach:

The Running Break - in which the oil feed is changed from the old to the new by step change. There is a gradual displacement of the fatty acid produced from the old feed to the new fatty acid. The composition of the fatty acid leaving the still body will change in a similar way, as shown in Figure 3A.

The Clean Break - in this the old fat feed is stopped, water is then pumped in and used to displace the oil phase out of the column by a process of "floating-off". When the oil phase has been essentially displaced from the unit, the new oil feed to the column is started and the complete plant is brought up to steady state for the new feed. Figure 3B illustrates the changes on fatty acid leaving the still body through this clean break operation.

FIGURE 3a.

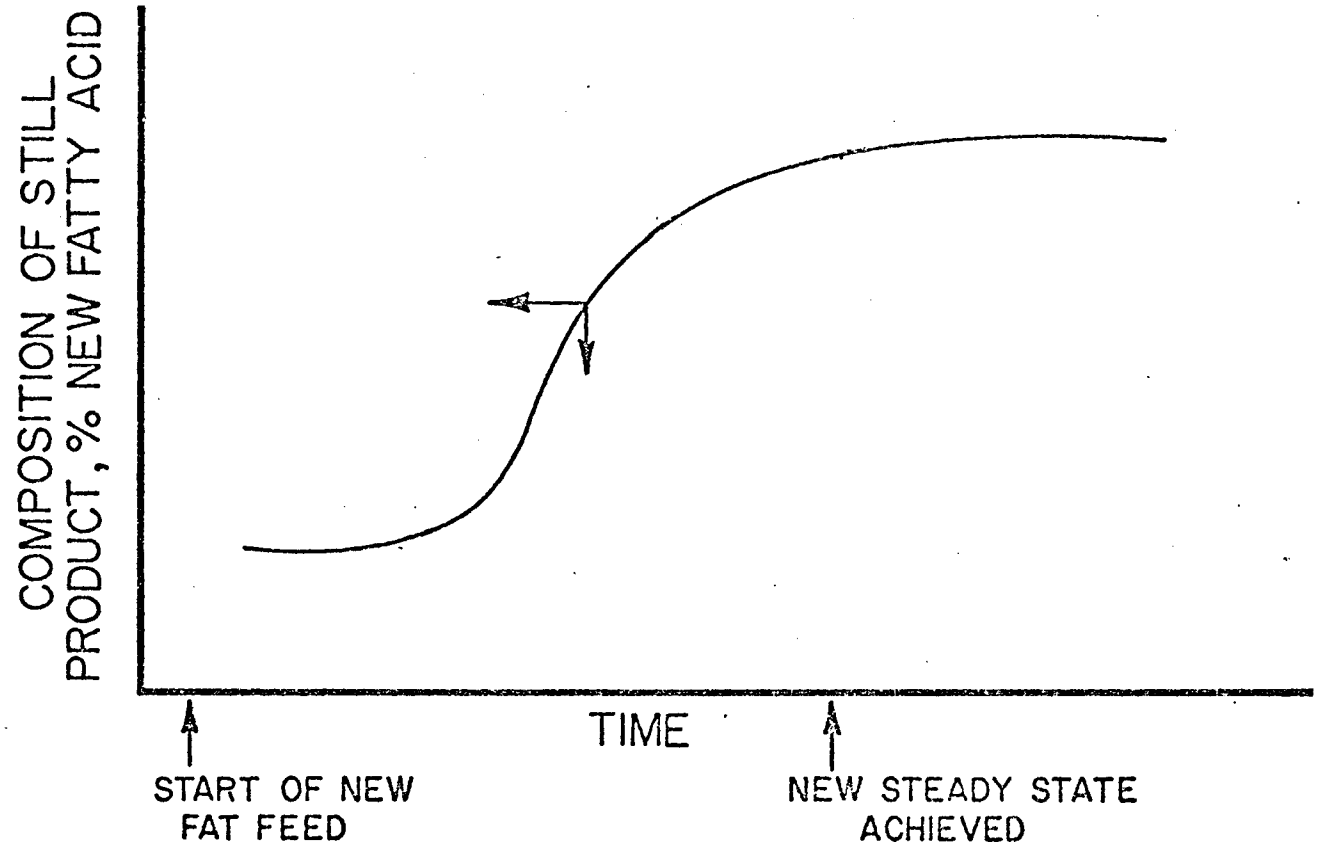
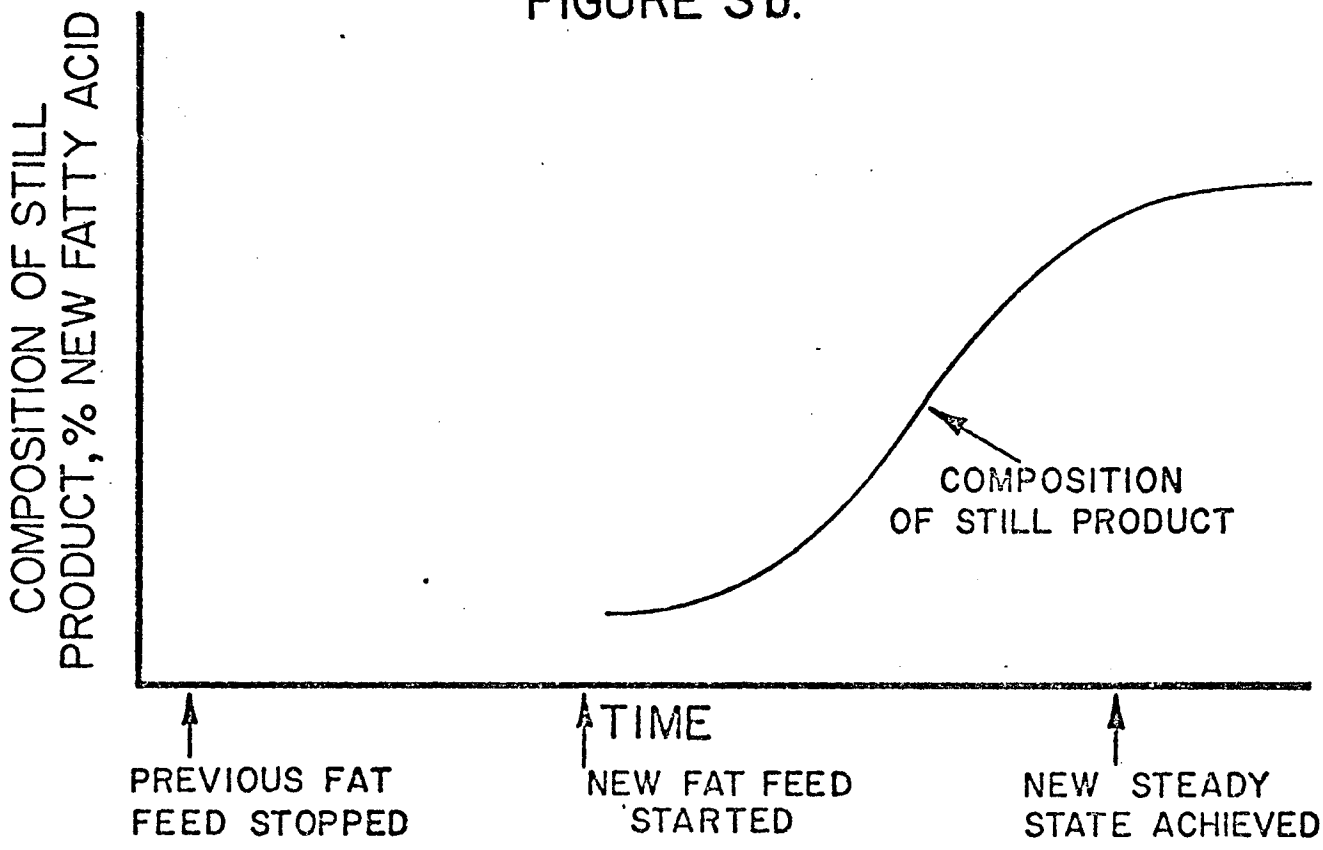


FIGURE 3b.



3. Modeling Strategy

3.1 The Information Flow Diagram

The most difficult part of a simulation study is usually the development of a suitable mathematical representation of the process. It should be realized that there are several levels of sophistication in modeling the same unit. Usually, approximate models are used in the initial stages of setting up the plant simulation to make preliminary studies, and then, if necessary, these modules can be easily replaced by more comprehensive ones. Very often it is found that only some of the models need to be comprehensive and the remainder may be approximate.

Following this philosophy, the first step was to establish the information flow diagram for the plant; this is shown in its final version in Figure 4.

It was realized that a rigorous dynamic model of the hydrolysis column may involve partial differential equations since we are dealing with a distributed parameter system; to avoid these difficulties, it has been represented by an assembly of lumped parameter modules, the number of which may be determined by comparison of the model with plant data.

No major simplifications were required in creating modules for representing the other pieces of equipment and controllers.

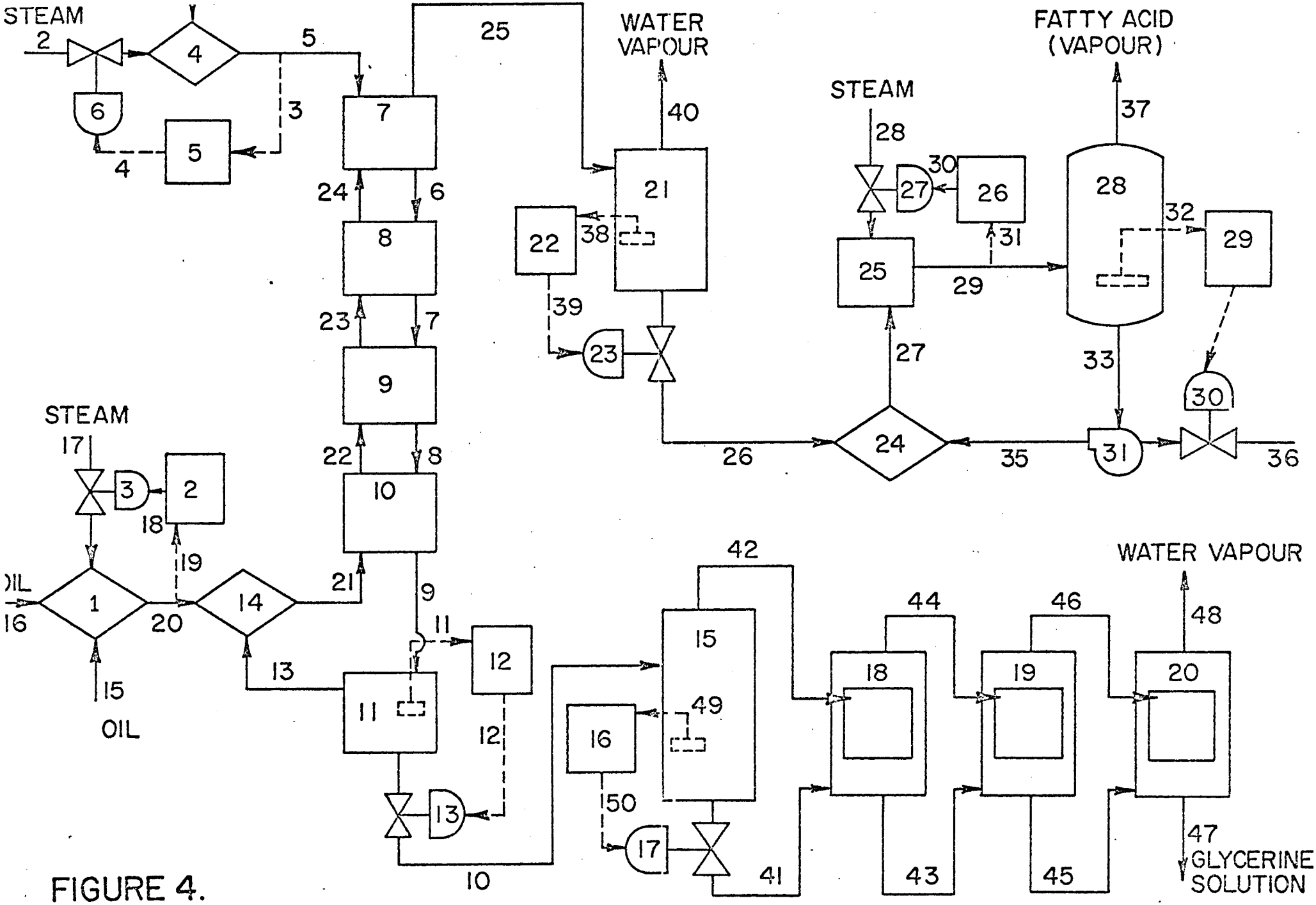


FIGURE 4.

DYNAMIC INFORMATION FLOW DIAGRAM OF THE FAT HYDROLYSIS UNIT SHOWN

3.2 Stream Variables

As, for this study, a dynamic model was sought to test alternative change-over policies characteristic of multiproduct plants, it was, therefore, desirable to define the information flows to include at least two different types of oil and their corresponding fatty acids. The final choice of stream information is as shown in Table 3.2

TABLE 3.2.1

STREAM INFORMATION
FOR THE DYNAMIC SIMULATION OF THE PLANT

Stream Element	Variable	Units
1	Stream number	
2	Stream flag	
3	Total flow	Lbs/hr.
4	Temperature	^{o}F
5	Pressure	Psia
6	Fat (1) (Tallow)	Weight Fractions
7	Fat (2) (Coconut oil)	
8	Steam	
9	Water	
10	Glycerine	
11	Unsaponifiable	
12	Fatty acid (1) (Tallow based)	
13	Fatty acid (2) (Coconut oil based)	

3.3 Unit Computations

After the information flow diagram was created, there was a consideration of the available unit computations, and the new ones to be created.

Inspection of the DYNSYS library [1] revealed that the following modules could be used with no modification:-

CNTRLϕ1 - Proportional plus integral control

VALVϕ1 - Parabolic type control valve

The mixed splitter module MXPLTϕ1 was slightly modified to account for latent heat of condensation of steam, but in essence, remained the same.

Then the following modules have to be created:-

EXTCϕ1 - To represent each one of the two-phase contactor-reactors into which the column was divided.

FLASϕ1 - To represent the flash tanks in which water is separated.

EXCHϕ1 - To represent the heat exchanger in the still.

STILLϕ1 - To represent the adiabatic flash tank in which fatty acid is separated.

PUMPϕ1 - To represent the recycle pump in the still.

SETLϕ1 - To represent the separator of aqueous and oil phase in the bottom of the column.

EVAPϕ1 - To represent a stage of the evaporator train.

3.4 Redimensioning of DYNSYS arrays

Numbering of streams and equipments gave information necessary to dimension arrays in DYNSYS. The following values were used:-

NEQ	-	Number of unit computations	40
NSTR	-	Number of streams	50
NC	-	Number of components	8
NPP	-	Number of property entries per component	10

The complete programme used 37000 (octal) words in the CDC 6400.

3.5 Physical properties

The DYNSYS physical properties handling scheme [1] was used for:- molecular weight, specific heat and density, with experimental values reported by Jeffreys, Jenson and Edwards [10]. Water and steam data were taken from Perry [22].

With the exception of the still which required vapour pressure and latent heats for fatty acids (available from the previous steady state simulation [8]) no other physical properties were required.

3.6 Modeling of the hydrolysis column

3.6.1 Hydrolysis of Fats

The hydrolysis of fats has been studied for about 100 years when Tilghman patented a high pressure hydrolysis. Later, studies showed that hydrolysis of triglycerides in animal fats and vegetable oils takes place step wise. Lascary [20, 21] studied the rate of hydrolysis and showed that the reaction proceeds principally in the oil phase of the two-phase system, and when little fatty acid is present, the reaction is confined to the interface, between the oil and water.

Struzingger and Sturm [26] assumed that the oil phase was always saturated with water and suggested a pseudo-first order reaction in terms of fat concentration. The reverse reaction was ignored although they estimated the equilibrium constant. Jenson, Jeffreys and Edwards (op. cit.) performed experimental work to determine the rate constant assuming a second order reversible reaction, and pointed out that in commercial continuous hydrolysis columns, as steam is injected with the fat at the bottom, an ample supply of water exists and may justify the assumption of a first order reaction mechanism as proposed by Struzingger. Values for the rate constants of 10.2 $hr.^{-1}$ for tallow and 9.2 $hr.^{-1}$ for coconut oil and a catalyst concentration of 0.25% (zinc oxide) were used in the present study.

3.6.2 Extractor Module EXTCϕ1

This module represents a section of the countercurrent hydrolysis column in which water is brought into contact with oil flowing countercurrentlyand a hydrolysis reaction and mass transfer takes place.

A pseudo-first order chemical reaction is assumed to occur in the oil phase, the reaction is

$$C_3H_5 - (COOR)_3 + 3\ H_2O \Leftrightarrow C_3H_5(OH)_2 + 3RCOOH$$

Two different fats were accounted for as reactants, and also the corresponding fatty acids and glycerine as products.

The heat of reaction in hydrolysis of fats was experimentally determined and computed by Jeffreys, Jenson and Edwards (op. cit.) to be of the order of 2000 cal/g; this is small and validates the assumption made for this module of isothermal operation.

The glycerine, formed in the reaction, was assumed to be completely transferred to the aqueous phase, and the oil phase was assumed to be saturated with water so the first order mechanism holds.

The total hold-up is assumed to be constant.

Making use of the previous assumptions the mass balance can be written as:

For fat

$$\frac{d\, m_{fi}}{dt} = FI*x_{fi} - FO*z_{fi} - R_{fi} \qquad i = 1,2 \qquad (1)$$

$$\text{with } R_{fi} = k_i\, m_{fi} \qquad (2)$$

For water in fat phase

$$\frac{d\, m_w}{dt} = FI*x_w - FO*z_w - 3*M_w \left(\frac{R_{f1}}{M_{f1}} + \frac{R_{f2}}{M_{f2}}\right) \qquad (3)$$

For fatty acids

$$\frac{d\, m_{fai}}{dt} = FI*x_{fai} - FO*z_{fai} + 3*\frac{M_{fai}}{M_{fi}}\, R_{fi} \qquad i = 1,2 \qquad (4)$$

In the aqueous phase

For water

$$\frac{d\, A_w}{dt} = \psi I*y_w - \psi O*\zeta_w - S_1\, R_{f1} - S_2\, R_{f2} \qquad (5)$$

For glycerine

$$\frac{d\, A_g}{dt} = \psi I*y_g - \psi O*\zeta_g + M_g \left(\frac{R_{f1}}{M_{f1}} + \frac{R_{f2}}{M_{f2}}\right) \qquad (6)$$

where

m_i = hold-up of component i in oil phase (lb)

A_i = hold-up of component i in aqueous phase (lb)

FI = input flow rate of oil phase (lb/hr)

ψI = input flow rate of aqueous phase (lb/hr)

FO = output flow rate of oil phase (lb/hr)

ψO = output flow rate of aqueous phase (lb/hr)

x_i = mass fraction of i in oil input

z_i = mass fraction of i in oil outlet

y_i = mass fraction of i in aqueous input

ζ_i = mass fraction of i in aqueous outlet

M_i = molecular weight of component i

S_1 = solubility of water in fatty acid (1)

S_2 = solubility of water in fatty acid (2)

k_i = rate constant for hydrolysis of fat "i"

Subscripts

fi = fat (i = 1 tallow, i = 2 coconut oil)

fai = fatty acid (i = 1 tallow based, i = 2 coconut based)

w = water

g = glycerine

3.6.3 Settler Module SETLϕI

The stream flowing down from the bottom of the last stage in the hydrolysis column contains two inmiscible phases. This module acts like a splitter, separating the incoming mixture in two distinct layers. Part of the oil phase is recycled and the level of the aqueous phase is given as output for use in a controller. In view of the fact that no reaction or mass transfer is considered to occur between the phases in the settler, the outlet concentrations of all the components are calculated during the splitting of the inlet stream, and simple mass balances are used to determine the rate of change of the hold-up in the aqueous phase, in order to give the signal for the controller.

3.7 Modeling the Fatty Acid Still

3.7.1 General Description of the Modules

The fatty acid still group is represented by three new modules:

Heat-exhanger EXCHϕ1 - where the feed to the still body is heated up to a preset temperature,

The Still Body - where the fatty acid contained in the input stream is flashed-off,

The recirculation pump - which circulates the liquid from the still body.

Due to the large difference between the vapour pressure of the fatty acids and the unsaponifiable residues and zinc soap, it was assumed that, under the operating pressure of the still body, 8 mm Hg, only fatty acid can exist in the vapour, and Raoult's law can be used to calculate the composition of such vapour. Furthermore, fatty acid vapour pressure dependence upon temperature was assumed to follow an Antoine equation.

The pressure in the still is considered constant and no reference will be made to the vacuum system.

3.7.2 Heat exchanger, module EXCHϕ1

This is a very simple module which calculates the temperature of the outlet stream, given the input flow rate and temperature. The heating medium is saturated vapour condensing at constant temperature. Values of the area and overall heat transfer coefficient have to be given as input parameters for the module; the heat transfer coefficient is considered as a constant.

Heat balance for the fatty acid stream is as follows:-

$$q = FI * C_p * (TI - TO) \tag{7}$$

and for the steam

$$q = FS * h \tag{8}$$

the heat transfer equation is:

$$q = UA\ (\Delta T)_{AV} \tag{9}$$

where

q = heat transferred (BTU/hr)

FI = flowrate of fatty acid (lb/hr)

C_p = specific heat of fatty acid (BTU/lb ^{o}F)

FS = flow rate of steam (lb/hr)

h = latent heat of condensation (BTU/lb)

U = overall heat transfer coefficient (BTU/hr ^{o}F sq.ft.)

$(\Delta T)_{AV}$ = arithmetic mean temperature difference (^{o}F)

3.7.3 Recycle Pump module PUMPϕ1

This module sets the output flow, for the still body, to a certain given value and calculates the recycle flow by subtracting tailings flow. It accounts, also, for the increase in pressure from the still to the pressure in the recirculation loop.

The flow and pressure change (PSIA) are given by the user as parameters for this module.

3.7.4 Still Body Module STILϕ1

The main module of the fatty acid still group was modeled as an adiabatic flash, which calculates the amount and composition of the vapour and liquid streams resulting from the flashing of the feed. Liquid hold-up is calculated from component mass balances and the temperature from a total energy balance in the liquid phase.

The present module uses the results of the steady state model [8] as "split factors" (ALPHA)defined as:-

ALPHA (1,IG) - ratio of flash temperature to feed temperature

ALPHA (i,IG) - i=2,9, ratio of component"i" mass flow in the liquid stream resulting from the flash to its flow in the feed.

The liquid hold-up is calculated from the component mass balances:

$$\frac{d\,m_i}{dt} = FI^* ALPHA(i+1,IG)^*x_i - FO^*z_1 \qquad i=1,NCOMP \qquad (10)$$

The heat balance is

$$\frac{d\varepsilon}{dt} = FI^* ALPHA\,(1,IG) * C_p * TI - FO^*C_p * TO \qquad (11)$$

$$TO = \frac{\varepsilon}{\Sigma m_i} \qquad (12)$$

where

TO = temperature of the liquid in the still (^oF)

TI = temperature of the feed (^oF)

C_p = specific heat (BTU/lb ^oF)

3.8 Flash Tank Module FLASϕ1

This unit performs a separation of excess water carried along with the input stream coming from the hydrolyzer. The module calculates the composition and flow of the liquid and vapour resulting from the flashing operation. The amount of vapour is calculated by a heat balance, considering that part of the heat carried by the input stream is used to evaporate water as the pressure reduces (in this case from 700 psia in the column to the pressure in the evaporator system or in the still cycle).

The level of the liquid is determined performing mass balances in the liquid in much the same way as for the still body.

The following assumptions were used: the pressure is fixed, only water is vaporized and no appreciable hold-up exists in the vapour phase.

3.9 Evaporator Module EVAPϕ1

In view of the fact that the main interest was placed in simulating change-overs on the plant, we were fundamentally concerned with modeling the hydrolysis column and the still, so only a very simple model of the evaporator train was incorporated, in order to achieve completeness in the study.

This routine is very similar to the one written for the flash tank, it allows the calculation of the outlet vapour and liquid flows and compositions once the input streams are given. The amount of heat transferred is calculated using a constant heat transfer coefficient and the temperature difference between the liquid in the evaporator and the steam in the chest; both the heat transfer and the area must be given as parameters for the module.

The outlet liquid stream carries heat, part of which is used to evaporate water. As a result of the lowering in pressure there is also a decrease in the boiling temperature so the temperature difference between the liquid and the steam in the chest allows heat to be transferred to the liquid contributing to the evaporation of more water; this results in a more concentrated liquid stream at the outlet and higher vapour flow.

Assumptions made include: no glycerine is evaporated; no compensation for the elevation in boiling point as the liquid becomes more concentrated was introduced; as in the flash, constant pressure and no vapour hold-up were assumed.

Appendix A1 shows complete listings of these routines.

4. A Comment on the Solution Technique

As pointed out in section 1.2 simulation problems present two aspects, one related to the representation of the system, and the other with the solution of the proposed mathematical relationships. It is apparent, from the previous section, that the mathematical representation for the dynamic behaviour of the fat hydrolysis plant involved non-linear ordinary differential equations.

DYNSYS uses a predictor-corrector (3rd order Adams-Moulton-Shell, "AMOS", [4]) technique to perform the numerical integration of ordinary differential equations. The integration step is adjusted automatically by the DYNSYS executive according to a truncation error calculation performed as part of the AMOS routine. However, values for the initial integration step and the upper and lower limits for the error can be provided by the user in order to make efficient use of this feature.

In solving systems of ordinary differential equations, initial and/or boundary conditions must be given. The DYNSYS integration routine requires a complete set of initial conditions, and as it stands now, is unable to cope with boundary value problems.

Initial conditions for all the streams in the plant were obtained from previous steady-state simulation studies [8] and used to calculate the corresponding control information streams.

5. Simulation and Control Studies

5.1 Preliminary Considerations

With the information flow diagram, depicted in figure 4, and the corresponding unit computations discussed in section 3, the following plan for the dynamic simulation of the change-over operations in the plant was adopted.

Firstly, consider the hydrolysis column. The fatty acid still and the evaporator train were to be assembled and tested as separate units. This provided an opportunity for becoming acquainted with the DYNSYS information handling structure and data set organization while working with a simpler problem than the complete plant. At this stage different integration steps were tested and the number of units necessary for proper representation of the hydrolysis column was determined. In this early stage, the validity of the models was examined and it was found that both the hydrolysis column and the fatty acid still behaved in the manner expected; for example, the fact that the reaction takes place mainly in the bottom of the column and that conversion of fat is almost complete (reported by Jeffreys [11]) were confirmed from the results of the model.

Secondly, with the experience gained in running separate units, limits for errors and integration time step were set, facilitating the assembly of the complete plant and the first runs.

With the first results of the complete plant, it was decided to make use of the interactive version of the DYNSYS program, a complete description of which is beyond the scope of this work and can be found elsewhere in [1].

Thirdly, the response of the plant to step changes in the feed was

simulated, in the interactive mode , which allows the user to interact with his dynamic model from the keyboard of a teletype or CRT unit while watching the printed output or the screen displaying the results. What follows is a teletype output indicating the commands used to simulate a step change in the oil feed stream and a plot showing the changes in composition for the column overhead, still vapour, and an intermediate stream on the column, as functions of time. This operation was carried out using the CDC 6400 under INTERCOM 3.

i) Once the user has established contact with the computer, the "LOGIN" operation, consisting of typing the user's name and password is:

```
MCMASTER UNIV INTERCOM 3.Ø
DATE  Ø6/3Ø/71
TIME  11.19.33.
HPJL
PLEASE LOGIN
LOGIN.
ENTER USER NAME-HPJL
■■■■■■■■■■ ENTER PASSWORD-
```

If the "LOGIN" has been successful, the computer displays the following message

```
Ø6/3Ø/71   LOGGED IN AT  11.2Ø.19.
           WITH USER-ID AF
           EQUIP/PORT 7Ø/Ø6
```

ii) SCOPE INTERCOM commands [25] to access the files containing the input data (DATAØ1)* and the compiled program (DYNS) are

```
COMMAND- BATCH.
TYPE FILE NAME-DATAØ1
TYPE DISPOSITION-PRIVATE
TYPE FILE NAME-DYNS
TYPE DISPOSITION-PRIVATE
TYPE FILE NAME-END
```

(*)Note all stream and equipment numbers in this section and following sections refer to the information flow diagram shown as figure 4.

iii) To start the execution of the program, type "DYNS" after the COMMAND -, and the program will start.

To each enquiry the user should reply with a Y for yes or N for no.

```
COMMAND- ETL,20.
COMMAND- EFL,55000.
COMMAND- DYNS.

         WATCHING CRT...N
         WANT TO SEE YOUR DATA SET...  N
         WANT TO USE EXISTING GRAPH FORM ... Y
  FAT HYDROLYSIS PLANT SIMULATION
BEGIN

PROCESS VARIABLES AT TIME =     .00100
```

STM	FLOW	T	P	FAT(1)	FAT(2)	STAM	H2O	GLY	USAP.	F/A(1)	F/A(2)
2	3466.2	515.00	777.96	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.00
5	5807.5	429.21	776.70	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.00
10	4645.6	478.00	776.70	.000	.000	.000	.799	.201	.000	0.000	.00
11	500.0	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00
17	2139.1	515.00	777.96	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.00
21	10551.9	414.40	776.70	.635	.159	0.000	.173	0.000	.018	.013	.00
25	14359.1	478.00	776.70	.020	.003	.000	.248	.000	.010	.567	.15
26	12677.2	289.29	55.45	.022	.006	.000	.101	.000	.014	.671	.18
29	62082.7	466.27	55.45	.019	.294	0.000	.021	.000	.397	.200	.04
32	1014.5	407.07	.10	.017	.368	.000	.000	.000	.486	.120	.00
33	50999.7	407.07	.10	.017	.368	.000	.000	.000	.486	.120	.00
36	1633.7	407.07	55.45	.017	.368	.000	0.000	0.000	.496	.079	.00
37	12506.3	433.43	0.00	0.000	0.000	0.000	.102	.000	0.000	.583	.18
38	998.8	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00
45	1671.8	228.10	20.29	.000	.000	.000	.579	.421	.000	0.000	.00
46	890.8	228.10	20.29	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.00
47	740.1	178.00	7.18	.000	.000	.000	.436	.564	.000	0.000	.00
49	1000.2	0.00	0.00	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.00

This table shows some of the most important streams in the plant (as selected by the user).

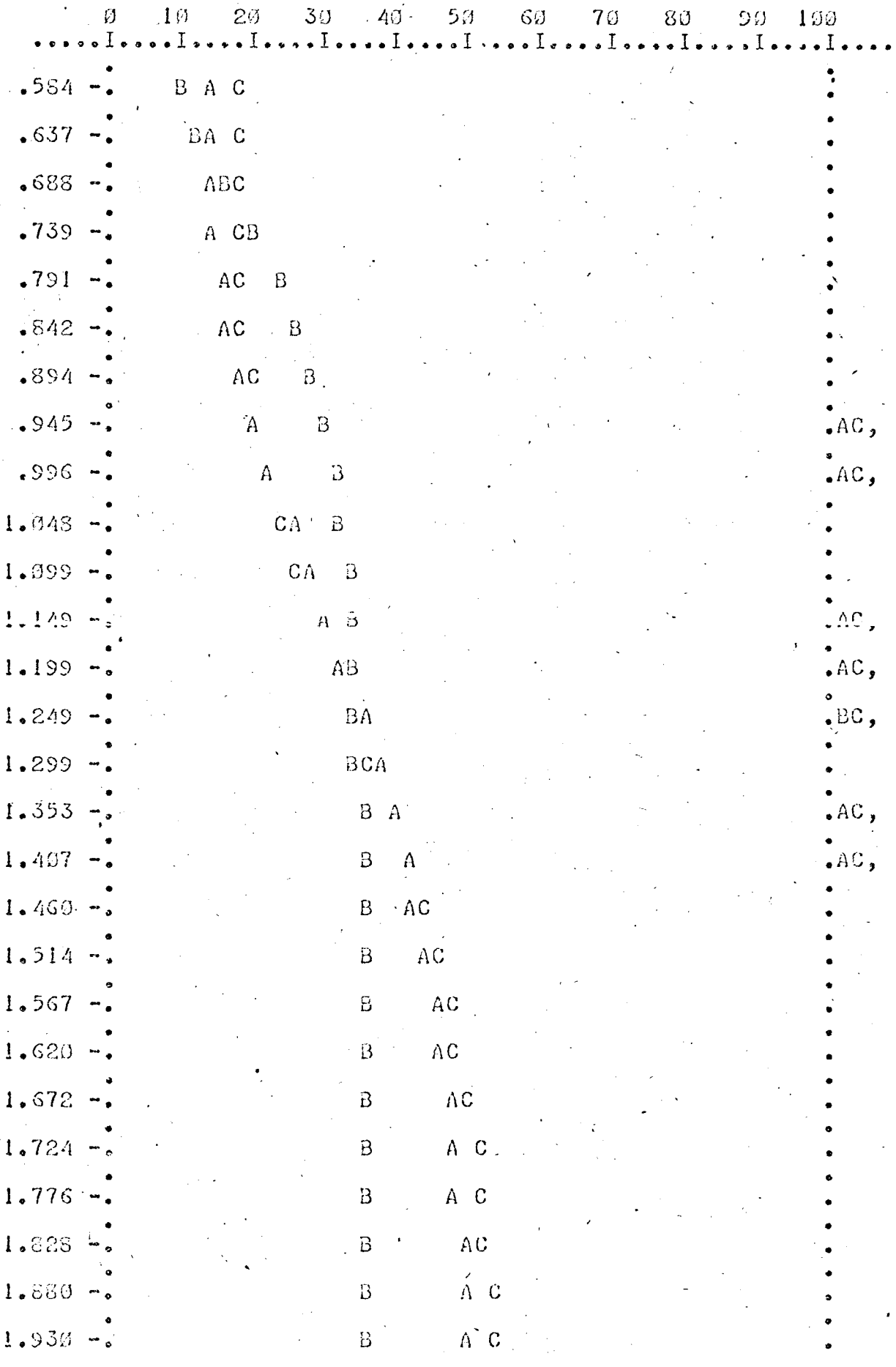
iv) To simulate a step change in the oil feed composition (for example from tallow 63% to coconut 70%) both streams 15 and 16 are changed using the following commands

```
          TYPE IN N,A,J,S,E,G OR F ... S
          NUMBER...15.
          CHANGE STREAM  15
          ALL OR ELEMENT ...E
          ELEMENT NO. ...3.
          OLD =    6983.5400      NEW = 1745.
          ANY MORE ...N
  NEW4VALUES
     15   1745.00 220.00  14.70  .9600 0.0000 0.0000 0.0000 0.0000  .021
Ø  .0190 0.0000

     IF MORE THEN

          TYPE IN N,A,J,S,E,G OR F ... S
          NUMBER...16.
          CHANGE STREAM  16
          ALL OR ELEMENT ...E
          ELEMENT NO. ...3.
          OLD =    1745.9100      NEW = 6983.
          ANY MORE ...N
  NEW4VALUES
     16   6983.00 220.00  14.70 0.0000  .9600 0.0000 0.0000 0.0000  .02
Ø 0.0000  .0180
```

v) The setting up of a graph from the teletype is illustrated below. Note that the plot will show the concentration of coconut based fatty acid in the stream leaving the top of the hydrolysis column (25), in the vapour leaving the still body (37) and in a stream "inside" the column (22).

```
          TYPE IN N,A,J,S,E,G OR F ... G
          NEW CURVES...Y
          POINTS YOU WANT ...10.
          NO. OF CURVES REQUIRED ...3.
          INTERVAL YOU WANT...0.1
          STREAM NO.  ...25.
          ELEMENT NO. ...13.
          LOWER LIMIT ...0.
          UPPER LIMIT ...1.
          STREAM NO.  ...22.
          ELEMENT NO. ...13.
          LOWER LIMIT ...0.
          UPPER LIMIT ...1.
          STREAM NO.  ...37.
          ELEMENT NO. ...13.
          LOWER LIMIT ...0.
          UPPER LIMIT ...1.
          VERY WELL
```

The following is the changeover plot



IF MORE THEN

TYPE IN N,A,J,S,E,G OR F ... N

The following ranges and stream variable definitions are provided

SYMBOL	STREAM	VARIABLE	RANGES	
A	25	F/A(2)	0.000	1.000
B	22	F/A(2)	0.000	1.000
C	37	F/A(2)	0.000	1.000

It is apparent from the graph that the system approaches closely to the new steady-state 1.45 hour after the change.

The first goal of the project has been successfully achieved. We are in possession of a dynamic model of the process to a step change on the oil feed. Moreover, this model can be used in an interactive mode enabling the user to operate in much the same way that the operator does in the real plant.

The change-over simulation here shown required about 90 sec. of CDC 6400 computation time.

5.2 Simulation of Different Operation Policies During the Change-over

5.2.1 Running Break

Recalling the definition of running break given in section 2.2.1, it is evident that the operation described in the previous section is a running break. However, this was done keeping all the controllers with fixed set points and

making changes only in the feed; in the real plant, the operator should adjust the set points in the different controllers, in order to reduce the amount of off-grade material produced, that is the fatty acid mixture formed during the grade change. This can be achieved by reducing the hold up in the still body and flash tank. To test this policy the example of the previous section was run, but now prior to the change in the feed, the level in the still body was lowered by resetting the set point in the corresponding level controller (equipment 29) from 1000 to 600 lbs. Teletype commands for this operation follow.

```
IF MORE THEN

     TYPE IN N,A,J,S,E,G OR F ... E
     NUMBER...29.
     CHANGE EQUIPMENT  29
     ALL OR ELEMENT ...E
     ELEMENT NO. ...3.
     OLD =     1000.0000      NEW = 600.
     ANY MORE ...N
NEW4VALUES
   3 0. 600.00  10.00     .50
```

After this, the running break was performed in the same way shown before; the resulting plot is reproduced here.

SYMBOL	STREAM	VARIABLE	RANGES	
A	25	F/A(2)	Ø.ØØØ	1.ØØØ
B	22	F/A(2)	Ø.ØØØ	1.ØØØ
C	37	F/A(2)	Ø.ØØØ	1.ØØØ

```
      Ø   10   2Ø   3Ø   4Ø   5Ø   6Ø   7Ø   8Ø   9Ø  1ØØ
     I....I....I....I....I....I....I....I....I....I....I....
.001 -.     B  AC                                          .
.052 -.      B AC                                          .
.1Ø3 -.        AC                                          .AB,
.154 -.        ACB                                         .
.2Ø5 -.        AC  B                                       .
.256 -.        AC   B                                      .
.3Ø7 -.         A    B                                     .AC,
.357 -.        AC    B                                     .
.41Ø -.         AC    B                                    .
```

TYPE IN N,A,J,S,E,G OR F ... N

```
.463 -.           CA    B                                  .
.515 -.             A   B                                  .AC,
.567 -.              A  B                                  .AC,
.62Ø -.               CAB                                  .
.672 -.                CA                                  .AB,
.724 -.                  A                                 .AB,AC,
.776 -.                  BA                                .AC,
.829 -.                  B A                               .AC,
```

TYPE IN N,A,J,S,E,G OR F ...
AC, DROP IN TEN MIN.
N

```
  .881 -.                  B  A                                  .AC,
  .933 -.                  B   A                                 .AC,
  .986 -.                  B   AC
 1.038 -.                  B    AC
 1.090 -.                  B     AC
 1.142 -.                  B     AC
 1.195 -.                  B     A C
 1.247 -.                  B      AC

        TYPE IN N,A,J,S,E,G OR F ... N

 1.299 -.                  B      A C
 1.352 -.                  B      A C
 1.404 -.                  B      A C
 1.456 -.                  B      A C
 1.509 -.                  B      A C
 1.561 -.                  B      A  C
 1.613 -.                  B      A  C
 1.665 -.                 B        A C

        TYPE IN N,A,J,S,E,G OR F ... N

 1.718 -.                  B        A C
 1.770 -.                  B        A C
 1.822 -.                  B        A C
 1.875 -.                  B        A C
TIME LIMIT
COMMAND- LOGOUT.
CP TIME    128.210
PP TIME    226.911
 05/01/71  LOGGED OUT AT 11.56.41.<
```

The time required to reach the same values as in the previous case was 1.35 (11 min. less) and therefore the amount of off-grade material produced is less, demonstrating the incentive for performing this simple control operation and encouraging the investigation of other control policies.

5.2.2 Clean Break

The alternative method used in the plant for change-over is the clean break. This operation keeps the amount of off-grade material to a minimum, but in general is more costly in view of the fact that it is necessary to stop the production for a longer time than is involved in the transition from one steady-state to another in a running break. In plants like this, where all the production can be placed on the market, the time that the unit is down is of extreme importance. However, there are certain fatty acids which may pass narrow tolerances, and to be safe clean breaks are carried out to produce them.

What follows is a line printer output for the clean break operation.

i) First fat feed is replaced by water, this has been done by setting the composition of oil feeds (streams 15 and 16) as pure water.

```
TYPE IN N,A,J,S,E,G OR F ... S
NUMBER...  15.
CHANGE STREAM  15
ALL OR ELEMENT ... E
ELEMENT NO. ... 6.
CLD =          .9600      NEW = 0.
ANY MORE ... Y
ELEMENT NO. ...12.
CLD =          .0210      NEW = 0.
ANY MORE ... Y
ELEMENT NO. ...11.
CLD =          .0190      NEW = 0.
ANY MORE ... Y
ELEMENT NO. ... 9.
CLD =         0.0000      NEW = 1.
ANY MORE ... N
NEW VALUES
   15 -16983.54 220.00  14.70 0.0000 0.0000 0.0000 1.0000 0.0000
                              0.0000 0.0000 0.0000
```

```
IF MORE THEN
     TYPE IN N,A,J,S,E,G OR F ... S
     NUMBER... 16.
     CHANGE STREAM  16
     ALL OR ELEMENT ...E
     ELEMENT NO. ... 9.
     CLD =        0.0000      NEW = 1.
     ANY MORE ... Y
     ELEMENT NO. ... 7.
     CLD =         .9600      NEW = 0.
     ANY MORE ... Y
     ELEMENT NO. ...13.
     CLD =         .0220      NEW = 0.
     ANY MORE ... Y
     ELEMENT NO. ...11.
     CLD =         .0180      NEW = 0.
     ANY MORE ...  N
NEW VALUES
  16 -11745.91 220.00  14.70 0.0000 0.0000 0.0000 1.0000 0.0000
                             0.0000 0.0000 0.0000
```

ii) The compositions in the column were observed and when the top of the column, in this case stream 24, shows a high water content (98%) it is an indication that the oil phase in the column has been "floated off", see figure 5.

iii) The next step was to reduce the hold up of the tanks and still body to avoid mixing of products.

iv) Once the column is completely filled with water, the new oil is fed and water is displaced by the oil phase, levels in still and tanks are reset once the system approaches the steady-state. This is shown in figure 6.

The complete clean break operation was performed in 4.5 hrs. when the new steady-state was approached.

5.3 Comments on the Change-over Operations

Results of the previous two sections showed that it was possible to simulate operator's activities during the change-over and demonstrate the usefulness of the modular approach in simulating the operation of continuous multi-product plants; the main objective of this project.

Several policies were tested for both clean and running break, and ways of improving the operation were found.

Figure 5 "Float off" Operation During a Clean Break

SYMBOL	STREAM	VARIABLE	RANGES	
A	21	F/A(1)	0.00000	1.00000
B	24	F/A(1)	0.00000	1.00000
C	37	F/A(1)	0.00000	1.00000
D	37	F/A(2)	0.00000	1.00000
E	32	TOTFLOW	1000.00000	12000.00000
F	38	TOTFLOW	500.00000	1500.00000

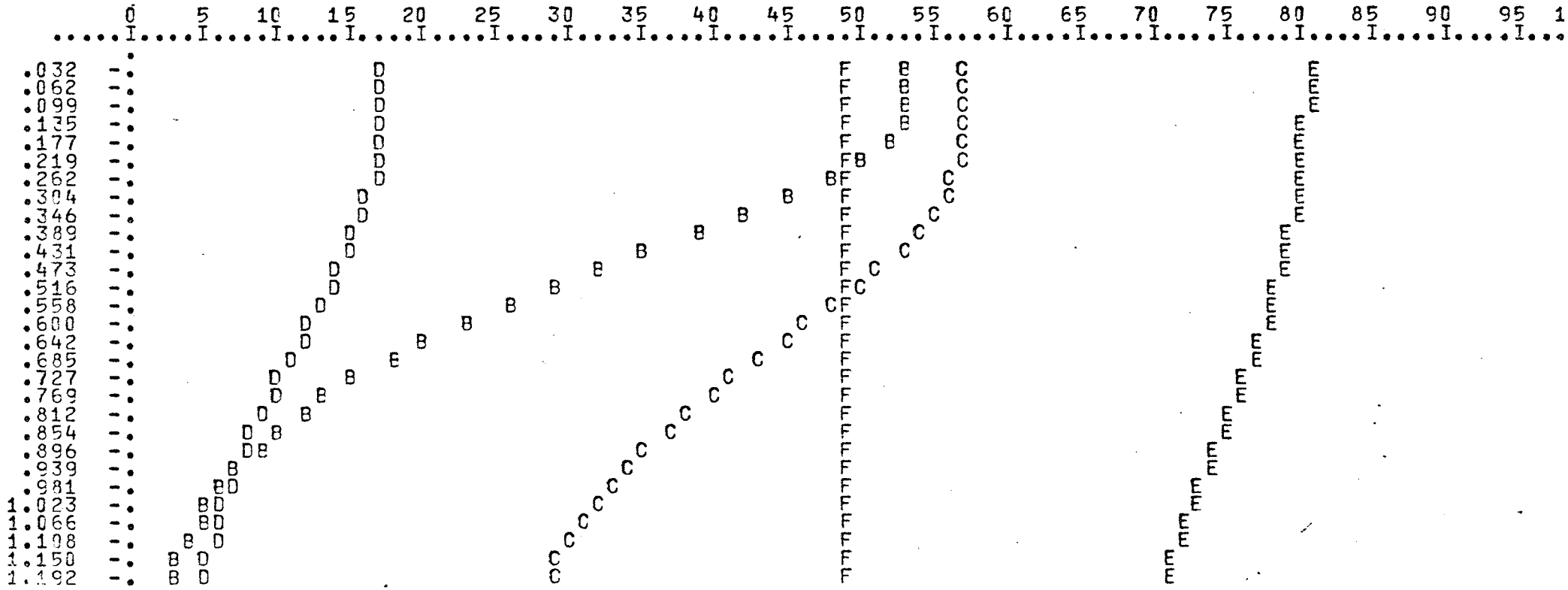


FIGURE 6

TANK DRAINING AND NEW FEED BUILD-UP DURING CLEAN BREAK

(Scale as in Fig. 5)

It is suggested that the leveling down of the flash tank and still body as part of any change-over operation be implemented.

The use of high flow rates in the feed stream at the bottom of the column did reduce the time required for change-over, however, it must be realized that flows in the column are constrained by hydrodynamic and mass transfer considerations. Exceedingly high flow rates in oil phase will increase the entrainment and correspondingly a lower yield in glycerin and fatty acid. It is desirable to include these limitations (like entrainment, flooding, etc.) on to the hydrolysis column model.

6. Automatic Change-over

The availability of low priced mini-computers for on line control of processes has opened new opportunities for implementing control policies like those discussed here. The implementation, within DYNSYS, of a new module able to simulate a computer controlling the process was a challenging goal to achieve.

Two routines were developed for such purpose; CLBR and RUBK. They perform supervisory control of the plant during the change-over operations, as well as some routine operations and alarm monitoring during the normal operation of the plant in much the same way that a mini-computer would do in the real plant.

Experience and understanding of the plant acquired during the modeling, simulation and change-over studies, was used throughout the developing of the "automatic change-over routines".

Both routines are called upon by the DYNSYS executive after the corrector step of the integration routine (detailed information about the predictor-corrector manipulations of DYNSYS can be found in [1]); however, the sampling operations are performed according to a given sampling time, while alarm scanning may be performed in every call.

Operations carried out by these routines may be classified into four categories:

i) Stream sampling and set point reading

ii) Supervisory control operations

iii) Alarm monitoring and emergency procedures

iv) Others; like report generation, mass balancing for inventory, and production control, etc.

The change-over is performed by a sequence of tasks, controlled by the routine according to certain pre-specified parameters or conditions; i.e. percent of water in the column overhead. A system of flags is used to indicate if a task is DONE or queuing for execution.

The running break has been represented as a set of four tasks:

1) Adjust set point of level controllers in flash and still body to decrease their levels.

2) Switch off the "old" oil feed and start the new; sample the concentration of new feed if it is not given by the operator.

3) Monitor composition of fatty acid in the vapour leaving the still, until a prefixed value is reached (% old fatty acid).

4) Reset set points in level controllers to steady state operation values.

Whenever a task is complete the DONE flag is set enabling the execution of the next task in the queue.

```
      SUBROUTINE RUBK(IFIRST)
*****************************************************************************
C       THIS SUBROUTINE SIMULATES THE OPERATIONS
C       PERFORMED BY A COMPUTER CONTROL
C       DURING THE RUNNING-BREAK OPERATION
*****************************************************************************

      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1  EMAX,NPOL,TMAX,NCOUNT,JM,KJ,MPR,NFORM
      DIMENSION ONEW(13)
      INTEGER FLAG(5),AL(4)
      DATA AL/4*8H        /
      NA=4
V       ALARM MONITORING
      IF(S(1,11,3).LT.10.) AL(1)=8HSETLER
      IF(S(1,38,3).LT.10.)AL(2)=8HFA/FLASH
      IF(S(1,32,3).LT.10.)AL(3)=8HSTILL
      IF(S(1,49,3).LT.10.)AL(4)=8HSW/FLASH
      DO 10 I=1,NA
      IF(AL(I).EQ.8H          )GO TO 10
      PRINT 11,AL(I)
      AL(I)=8H
   10 CONTINUE
   11 FORMAT(10X,15H LOW LEVEL IN  ,A8,/)
      IF(IFIRST.EQ.1) GO TO 40
C       SET INITIAL VAUES
      IFIRST=1
C       SAMPLING TIME IN HRS.
      TSAM=0.08
C       COMPOSITION OF NEW FEED
      DO 20 I=1,13
      ONEW(I)=0.
   20 CONTINUE
      ONEW(7)=0.96
      ONEW(11)=0.021
      ONEW(12)=0.019
      ONEW(3)=8790.
      TS=TIME(1)
C       INITIALIZE FLAGS
      DO 30 I=1,5
      FLAG(I)=4H
   30 CONTINUE
C       WATER TO OIL RATIO
      RATIO=8./7.
   40 CONTINUE
C       CHEK FOR SAMPLING TIME
  100 IF(TIME(1).LT.TS) RETURN
      TS=TIME(1)+TSAM
C       FIRST OPRATION STARTS
      IF(FLAG(1).EQ.4HDONE) GO TO 100
      DO 50 J=1,2
      DO 51 I=6,NC5
   51 S(J,15,I)=ONEW(I)
      S(J,15,3)=ONEW(3)
   50 CONTINUE
C     ADJUST SET POINTS TO EMPTY STILL AND FLASH
      EP(22,3)=100.
      EP(29,3)=100.
```

```
      FLAG(1)=4HDONE
      IF(FLAG(2)=EQ.4HDONE) GO TO 200
C       CHECK COMPOSITION OF F/A VAPOUR
      IF(S(1,37,12).GT.0.02)GO TO 200
      EP(22,3)=1000.
      EP(25,3)=2000.
      FLAG(2)=4HDONE
  200 AUX=RATIO*(S(1,21,6)+S(1,21,7))*S(1,21,3)-S(1,21,9)*S(1,21,3)
      S(1,1,3)=AUX-S(1,2,3)
      IF(S(1,1,3).LT.0.)S(1,1,3)=0.
      S(2,1,3)=S(1,1,3)
      RETURN
      END
```

For the clean-break the following tasks were programmed:

1) Switch off the oil feed and start feeding water, to float off the column.

2) Adjust set points in level controllers in flash tank and still body to drain the old product.

3) Monitor water content in the oil phase at the top of the column, until the oil phase disappears.

4) Start feeding the new oil to the column.

5) Monitor fat and fatty acid in top of the column, until the fat phase appears again.

6) Reset the set points in level controllers to steady-state operation values.

Besides the supervisory control, all the levels are sampled each time interval and compared with safety values. If there is any level out of the safety range an alarm message will be issued to inform the operator. It is proposed to have an emergency routine to attend to alarms and simulate the interlock systems for emergencies.

```
      SUBROUTINE CLBK(IFIRST)
*****************************************************************************
C       THIS SUBROUTINE SIMULATES THE OPERATIONS
C       PERFORMED BY A COMPUTER CONTROL
C       DURING THE CLEAN BREAK OPERATION
*****************************************************************************
C
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1 EMAX,NPOL
      INTEGER FLAG(5),AL(4)
      DIMENSION ONEW(13)
      DATA AL/4*8H        /
C     SET ALARMS
      IF(S(1,11,3).LT.10.) AL(1)=8HSETLER
      IF(S(1,38,3).LT.10.)AL(2)=8HFA/FLASH
      IF(S(1,32,3).LT.10.)AL(3)=8HSTILL
      IF(S(1,49,3).LT.10.)AL(4)=8HSW/FLASH
      DO 1 I=1,4
      IF(AL(I).EQ.8H          )GO TO 12
      PRINT 11,AL(I)
      AL(I)=8H
    1 CONTINUE
   11 FORMAT(10X,15H LOW LEVEL IN  ,A8,/)
   12 CONTINUE
      IF(IFIRST.EQ.1)GO TO 10
      IFIRST=1
C     INITIALIZE PARAMETERS ON FIRST CALL
C     SAMPLING TIME
      TSAM=0.08
C     COMPOSITION OF THE NEW OIL
      DO 2 I=1,13
      ONEW(I)=0.
    2 CONTINUE
      ONEW(11)=0.1
      ONEW(7)=0.9
C     FLOW RATE OF NEW OIL
      ONEW(3)=7000.
C     WATER FLOW DURING COLUMN 'FLOATING'
      WF=8000.
      TS=TIME(1)
      DO 20 I=1,5
      FLAG(I)=4H
   20 CONTINUE
C     WATER/OIL RATIO
      RATIO=8./7.
   10 CONTINUE
      IF(TIME(1).LT.TS) RETURN
      TS=TIME(1)+TSAM
C     FIRST OPERATION
C     SWITCH OFF OIL FEED
      IF(FLAG(1).EQ.4HDONE) GO TO 100
      DO 30 I=6,NC5
      S(1,15,I)=0.
      S(2,15,I)=0.
      S(1,16,I)=0.
      S(2,16,I)=0.
   30 CONTINUE
      S(1,16,3)=0.
```

```
      S(2,16,3)=0.
      S(1,15,9)=1.
      S(2,15,9)=1.
      S(1,15,3)=WF
      S(2,15,3)=WF
      FLAG(1)=4HDONE
      RETURN
  100 CONTINUE
      IF(FLAG(2).EQ.4HDONE)GO TO 200
C      ADJUST SET POINTS IN FLASH AND STILL
      EP(22,3)=100.
      EP(29,3)=100.
      FLAG(2)=4HDONE
      RETURN
  200 CONTINUE
      IF(FLAG(3).EQ.4HDONE) GO TO 300
C     MONITOR WATER CONTENT IN COLUMN OVERHEAD
      IF(S(1,25,9).LT.0.9) RETURN
      IF(S(1,32,3).LE.100.)GO TO 40
      IF(S(1,38,3).LE.100.)GO TO 40
C     EMPTY TANKS
      S(1,34,3)=100.
      S(2,34,3)=100.
      S(1,39,3)=100.
      S(2,39,3)=100.
   40 CONTINUE
C     SET UP THE NEW FEED COMPOSITION AND FLOW
      DO 50 I=6,NC5
      S(1,15,I)=ONEW(I)
      S(2,15,I)=ONEW(I)
   50 CONTINUE
      S(1,15,3)=ONEW(3)
      S(2,15,3)=ONEW(3)
      FLAG(3)=4HDONE
      RETURN
  300 IF(FLAG(4).EQ.4HDONE) GO TO 400
      EP(22,3)=1000.
      EP(29,3)=2000.
      FLAG(4)=4HDONE
  400 CONTINUE
C     SUPERVISORY CONTROL ON W/O RATIO
      AUX=RATIO*(S(1,21,6)+S(1,21,7))*S(1,21,3)-S(1,21,9)*S(1,21,3)
      S(1,1,3)=AUX-S(1,2,3)
      IF(S(1,1,3).LT.0.)S(1,1,3)=0.
      S(2,1,3)=S(1,1,3)
      RETURN
      END
```

Testing of the automatic change-over routines was carried out by performing both running and clean break from a feed of 100% tallow to one of 100% coconut oil, which is considered to be an extreme situation and the most difficult that can be simulated.

Figures 7 and 8 show the results of these automated operations. Variables plotted are:

A mass fraction of tallow based fatty acid in the combined feed to the column (21)

B mass fraction of coconut oil in the combined feed to the column (21)

C mass fraction of tallow based fatty acid in an intermediate stream in the column (23)

D mass fraction of tallow based fatty acid in the outlet at the top of the column (25)

E mass fraction of coconut based fatty acid in the outlet at the top of the column (25)

F mass fraction of tallow based fatty acid in the vapour leaving the still body (37)

G mass fraction of coconut based fatty acid in the vapour leaving the still body (37)

H liquid level in the flash tank (38)

I liquid level in the still body (32)

Examination of these results demonstrated that improvements were achieved over manual operation through a better precision in control actions and elimination of human negligence by automation of the decisions.

SYMBOL	STREAM	VARIABLE	RANGES	
A	21	F/A(1)	0.00000	1.00000
B	21	FAT(2)	0.00000	1.00000
C	23	FAT(1)	0.00000	1.00000
D	25	F/A(1)	0.00000	1.00000
E	25	F/A(2)	0.00000	1.00000
F	37	F/A(1)	0.00000	1.00000
G	37	F/A(2)	0.00000	1.00000
H	38	TOTFLOW	0.00000	1100.00000
I	32	TOTFLOW	0.00000	2000.00000

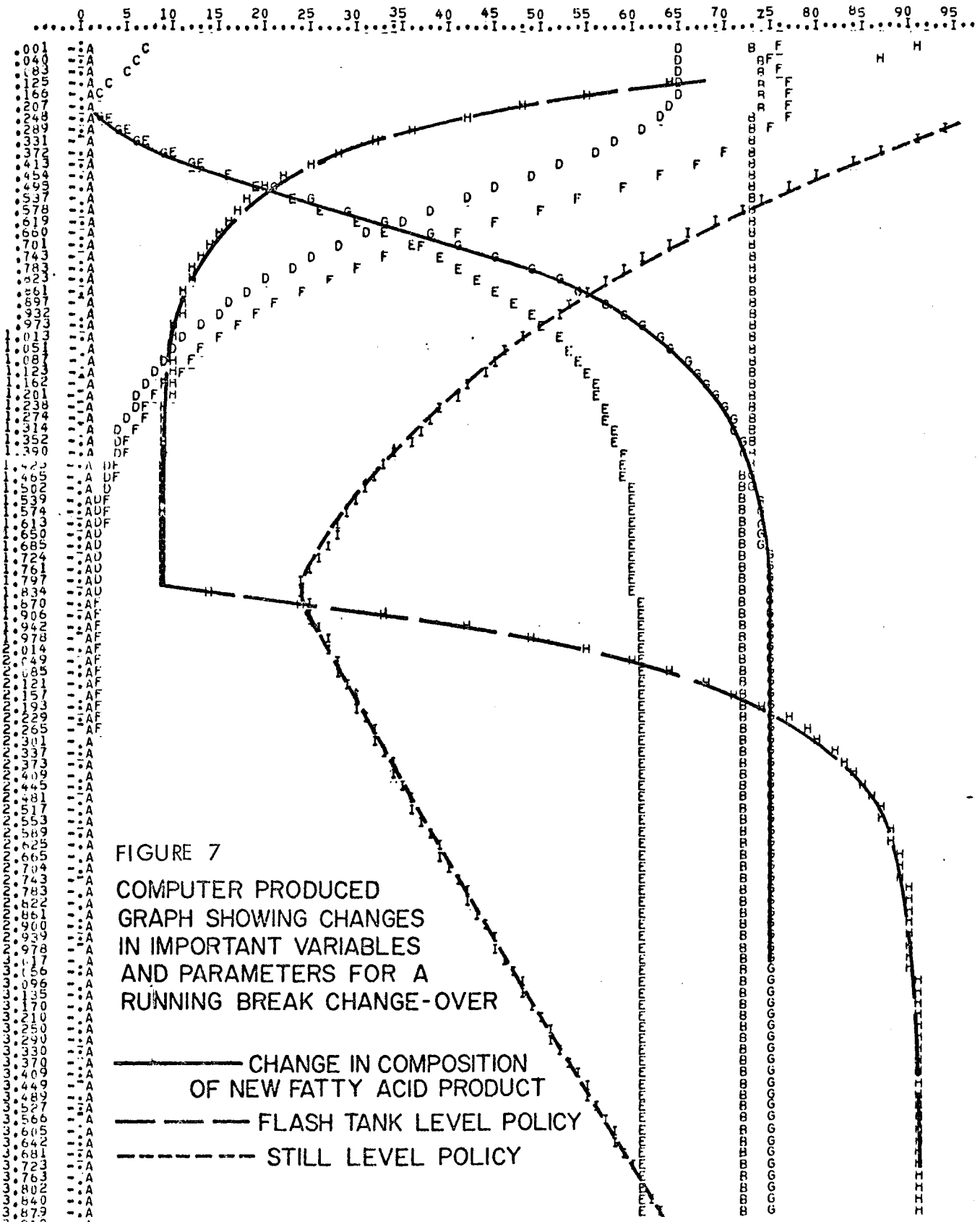


FIGURE 7

COMPUTER PRODUCED GRAPH SHOWING CHANGES IN IMPORTANT VARIABLES AND PARAMETERS FOR A RUNNING BREAK CHANGE-OVER

——— CHANGE IN COMPOSITION OF NEW FATTY ACID PRODUCT

— — — FLASH TANK LEVEL POLICY

- - - - - - - STILL LEVEL POLICY

SYMBOL	STREAM	VARIABLE	RANGES	
A	21	F/A(1)	0.00000	1.00000
B	21	FAT(2)	0.00000	1.00000
C	23	FAT(1)	0.00000	1.00000
D	25	F/A(1)	0.00000	1.00000
E	25	F/A(2)	0.00000	1.00000
F	37	F/A(1)	0.00000	1.00000
G	37	F/A(2)	0.00000	1.00000
H	38	TOTFLOW	0.00000	1100.00000
I	32	TOTFLOW	0.00000	2000.00000

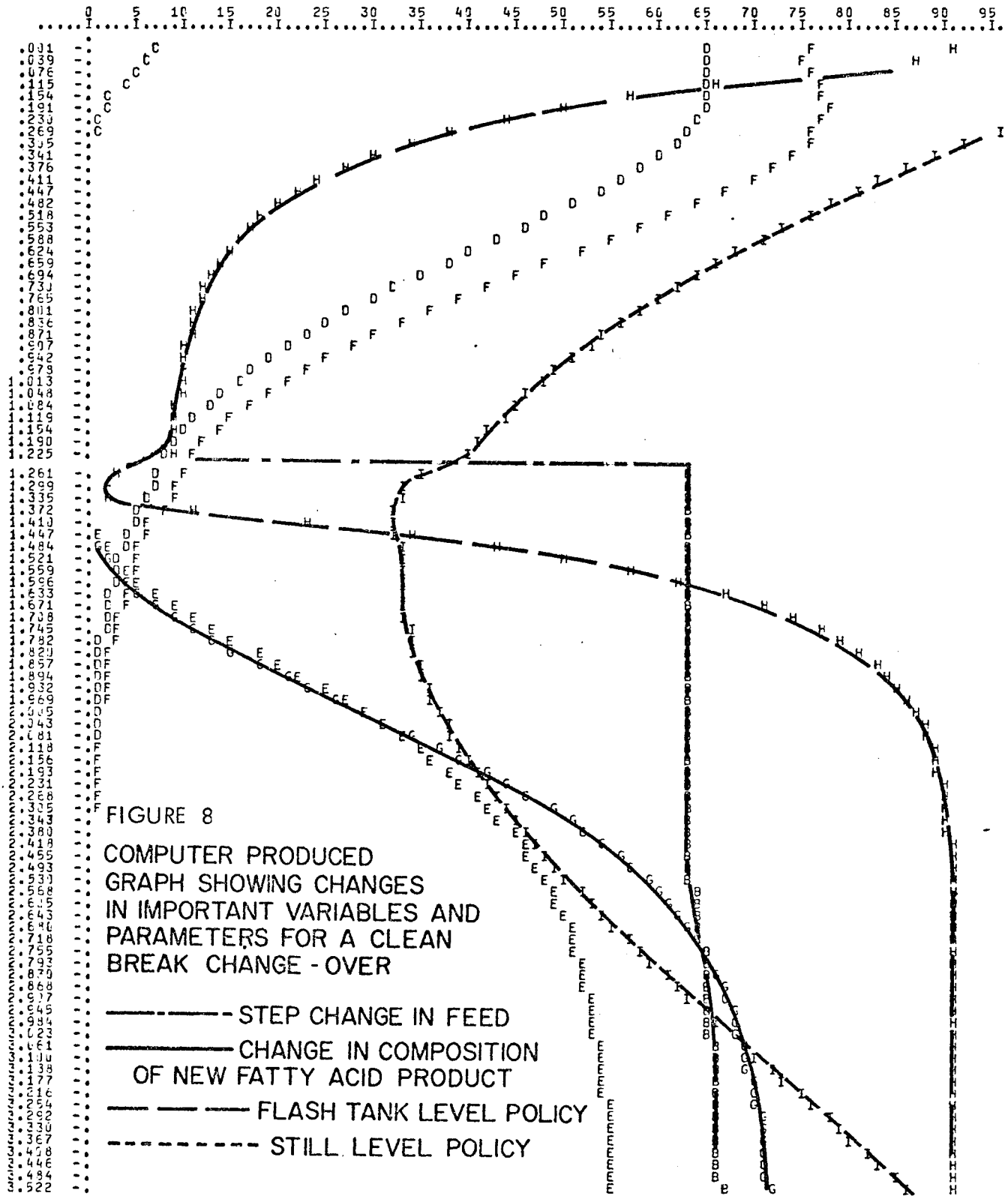


FIGURE 8

COMPUTER PRODUCED GRAPH SHOWING CHANGES IN IMPORTANT VARIABLES AND PARAMETERS FOR A CLEAN BREAK CHANGE - OVER

Computation time for these type of studies was of the order of 120 seconds in the CDC 6400 and the simulated operation time was 3.5 hrs.

7. Evaluation Of The Complete Study

7.1 Practical Considerations

Recalling the objectives of this study, some discussion with plant personnel was carried out to determine whether or not the results had practical value.

There was a close agreement with plant behaviour. The possibility of replacing some of the clean breaks by controlled running breaks (which will give a good indication of when the new product is virtually free of old residuals) was demonstrated; this may reduce significantly the time that the unit is out of normal production due to clean breaks, with a corresponding increase in productivity.

Some recommendations are in order, especially related to assumptions made during the modeling:

i) Consideration of entrainment problems in the hydrolysis column and its fluid mechanical behaviour.

ii) Possible addition of a module to account for a high vacuum flash which exists in the plant between the flash tank and the still, which will complete the elimination of all the water carried by the fatty acid. In fact, this can be easily done by adding another FLASHØ1 module operating under high vaccum.

7.2 Computational Considerations

From the computational point of view, the following considerations may help future users of the modular approach for dynamic simulation studies and

particularly DYNSYS users.

1) It was observed that when chemical reaction or mass transfer takes place in a module, the concentrations of one or more of the components may reach extremely small values. This may cause numerical instabilities to the integration routine and to the module itself. It is advisable to skip the integration routine whenever the derivative and the dependent variable have reached very small values.

2) Alternate modeling of the adiabatic flash operation, as a non-equilibrium module, must be considered since it has been recently reported by Kleinspeter and Weaver [13] that this model has extraordinatory stability; however, information related to mass transfer coefficients for fatty acids was not available so it was not possible to make use of it.

3) Of special interest, in order to perform optimization, optimal and adaptive control studies, is to have transfer functions of the process or of its units, which will enable a very fast execution. Koppel [14] has indicated that dynamical behaviour of many chemical process operations may be well reproduced by the following transfer function

$$G_p(s) = \frac{K_p\, e^{-a\tau s}}{(\tau s + 1)(b\tau s + 1)}$$

if there is sufficient information to evaluate the parameters a, b, τ and the process gain K_p. Examples including heat exchanges, distillation columns and chemical reactors have been published [14]. In this case, information generated in dynamic studies with the time domain model may be used to "identify" the

system with a transfer function, and then use this for optimization and control work. The validity of this will be strongly determined by the validity of the model and may be used only in points close to those used for the "identification".

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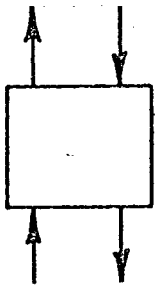
APPENDIX A1

This appendix contains the new additions to the DYNSYS library of unit computations used in modeling this plant, and fully described in sec. 3 of this work. A listing of the complete input data is also included.

Symbol



DYNSYS Subroutine Name - EXTCΦI - Type 1

Written by - P.K.J. Weng

Reviewed by - A. Lozada, August, 1971.

Purpose

This module represents a section of the countercurrent hydrolysis column in which tallow or coconut oil is contacted with water, and a hydrolysis reaction occurs.

Description

In this module a pseudo first order reaction is assumed to take place in the fat phase. The reaction is expressed as

$$C_3H_5(COOR)_3 + 3H_2O \xrightarrow{k_R} C_3H_5(OH)_3 + 3RCOOH$$

The module considers two types of reactant:

Tallow (k_R^* = 10.2 hr^{-1}) and coconut oil (k_R^* = 9.0 hr^{-1}).

These reaction rate constants are for 490°F and 650 psia and for 0.25% ZnO catalyst.

The temperature and pressure are assumed to be constant through the reactor. The glycerine produced in the reaction is assumed to be transferred completely to the aqueous phase.

Initial mass holdups of fat phase and water phase are provided; these are adjusted with time taking into account the chemical reaction and the transfer of material.

Stream Designations

MP(IM,3) - inlet stream of water phase
MP(IM,4) - inlet stream of oil phase
MP(IM,5) - outlet stream of water phase
MP(IM,6) - outlet stream of oil phase

---

*from Jenson, V.G., Jeffreys, G.V., (op.cit.)

Parameters

EP(IM,1) - mass holdup of water phase at TIME (1)

EP(IM,2) - mass holdup of oil phase at TIME (1)

```
      SUBROUTINE TYPE  1
      COMMON/UNIT/IM,NMP
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1  EMAX,NPOL,TMAX
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      COMMON/PTAB/  IGFLAG,PP(10,10)
      COMMON/BKV/NBV,BV(2,200)
C    EXTRAC01 --- TYPE 1
C    LIQUID-LIQUID EXTRACTOR TAKING ACCOUNT OF CHEMICAL REACTION
C    PARAMETERS  ---  EP TABLE
C  1  --  WATER PHASE HOLD-UP  2  --  OIL PHASE HOLD-UP
C  3  AND  4  ARE USED AS STORES
      DIMENSION  RATE(20),REQ(20)
      MWI=MP(IM,3)
      MOI=MP(IM,4)
      MWO=-MP(IM,5)
      MOO=-MP(IM,6)
      IF(TIME(2).NE.0 .AND. IG.NE.2) GO TO 3
      EP(IM,3)=EP(IM,1)
      EP(IM,4)=EP(IM,2)
    3 T=S(IG,MWO,4)
      TALOW=0.8
      AK1=10.2
      AK2=9.
      SUMW=SUMO=0.
      TMASS=EP(IM,1)+EP(IM,2)
      IF(IG.EQ.2) GO TO 10
      WO=S(1,MWI,3)+S(1,MOI,3)
      S(1,MWO,3)=WO*EP(IM,1)/TMASS
      S(1,MOO,3)=WO*EP(IM,2)/TMASS
   10 CONTINUE
      DO 20 I=1,NCOMP
   20 RATE(I)=0.
      REQ(3)=REQ(6)=0.
      REQ(1)=AK1*EP(IM,4)*S(2,MOO,6)/PP(1,1)
      IF(IG.EQ.1) REQ(1)=AK1*EP(IM,2)*S(1,MOO,6)/PP(1,1)
      REQ(2)=AK2*EP(IM,4)*S(2,MOO,7)/PP(2,1)
      IF(IG.EQ.1) REQ(2)=AK2*EP(IM,2)*S(1,MOO,7)/PP(2,1)
      REQ(4)=3.*(REQ(1)+REQ(2))
      REQ(5)=-(REQ(1)+REQ(2))
      RATE(5)=ABS(REQ(5))
      REQ(7)=-REQ(1)*3.
      REQ(8)=-REQ(2)*3.
      RATE(4)=-REQ(4)+1.97*REQ(7)+3.88*REQ(8)
C  CHEMICAL REACTION IS ASSUMED TO TAKE PLACE IN OIL-PHASE ONLY
      DO 40 I=1,NCOMP
      I6=I+5
      FO=(S(IG,MOI,3)*S(IG,MOI,I6)-S(IG,MOO,3)*S(IG,MOO,I6))/PP(I,1)
     2  -RATE(I)-REQ(I)
      Y2=EP(IM,4)*S(2,MOO,I6)/PP(I,1)
      IF(IG.EQ.1) X=EP(IM,2)*S(1,MOO,I6)/PP(I,1)
      S(1,MOO,I6)=Y1(1,Y2,FO,X)*PP(I,1)
      SUMO=SUMO+S(1,MOO,I6)
      FW=(S(IG,MWI,3)*S(IG,MWI,I6)-S(IG,MWO,3)*S(IG,MWO,I6))/PP(I,1)
     2  +RATE(I)
      Y2=EP(IM,3)*S(2,MWO,I6)/PP(I,1)
      IF(IG.EQ.1) X=EP(IM,1)*S(1,MWO,I6)/PP(I,1)
      S(1,MWO,I6)=Y1(1,Y2,FW,X)*PP(I,1)
```

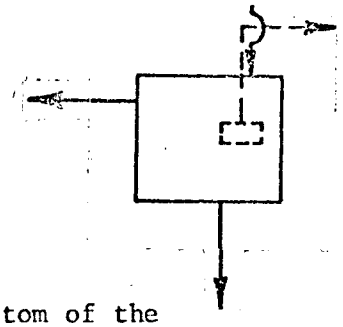
```
   40  SUMW=SUMW+S(1,MWO,I6)
C   NORMALIZATION
       FACT=TMASS/(SUMW+SUMO)
       EP(IM,1)=SUMW*FACT
       EP(IM,2)=SUMO*FACT
C   OUTLET CONCENTRATION
       DO 60 I=1,NCOMP
       I6=I+5
       S(1,MWO,I6)=S(1,MWO,I6)/SUMW
   60  S(1,MOO,I6)=S(1,MOO,I6)/SUMO
C    ADJUST TOTAL OUTLET FLOW
       TMASS=EP(IM,1)+EP(IM,2)
       WO=S(1,MWI,3)+S(1,MOI,3)
       S(1,MWO,3)=WO*EP(IM,1)/TMASS
       S(1,MOO,3)=WO*EP(IM,2)/TMASS
       S(1,MWO,4)=T
       S(1,MOO,4)=S(1,MWO,4)
       IF(IG.EQ.2) RETURN
       EP(IM,3)=EP(IM,1)
       EP(IM,4)=EP(IM,2)
       RETURN
       END
```

DYNSYS Subroutine Name - SETLϕI - Type 3

Written by - P.K.J. Weng.

Revised by - A. Lozada, August, 1971.

Symbol



Purpose

A settler module is attached to the bottom of the hydrolyzer for separating the light phase from the heavy phase.

Description

The stream flowing down from the bottom stage of the hydrolyzer contains two immissible phases. This module acts like a splitter separating the incoming mixture into two distinct layers and the level of the heavy phase is maintained by a level controller and a valve for the outlet stream of the heavy phase. A simple differential equation is written to determine the rate of change of the hold up of heavy phase in order to give signals of the level position to the controller in each time stage.

Stream Designations

MP(IM,3) - inlet stream, mixture of aqueous solution and fatty material.

MP(IM,4) - leaving aqueous phase.

MP(IM,5) - leaving fatty phase.

MP(IM,6) - level signal

Parameters

EP(IM,1) - holdup of aqueous phase at predictor step

EP(IM,2) - holdup of aqueous phase at corrector step

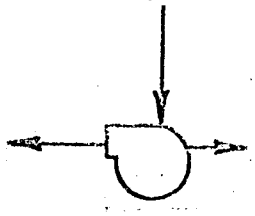
```
      SUBROUTINE TYPE 3
C  SETTLER MODULE FOR P AND G   SETL01 -- TYPE 3
C  PARAMETERS  1 --PREDICTED  2 -- CORRECTED VALUES OF HOLDUP OF WATER PHASE
      COMMON/UNIT/IM,NMP
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1 EMAX,NPOL
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      COMMON/PTAB/ IGFLAG,PP(10,10)
      DIMENSION FCT(20)
      IN=MP(IM,3)
      IWO=-MP(IM,4)
      IOO=-MP(IM,5)
      ISIG=-MP(IM,6)
      IF(TIME(2).EQ.0. .AND. IG.EQ.2) EP(IM,2)=EP(IM,1)
      SU=SV=SX=0.
C  CONCENTRATION OF THE OUTLET OIL PHASE
      DO 10 I=6,NC5
      FCT(I)=0.
   10 S(IG,IOO,I)=0.
      DO 11 I=11,13
      FCT(I)=S(IG,IN,3)*S(IG,IN,I)
   11 SU=SU+FCT(I)
      IF(SU.LT.1.0E-10) GO TO 13
      DO 12 I=11,13
      FCT(I)=FCT(I)/SU
      IF(IG.EQ.1) S(1,IOO,I)=FCT(I)
   12 CONTINUE
      S(1,IOO,3)=SU
   17 CONTINUE
C ACTUALL FLOW AND COMPOSITION OF INLET WATER STREAM
      DO 14 I=6,NC5
      FCT(I)=S(I7,IN,3)*S(IG,IN,I )-SU*FCT(I)
   14 SV=SV+FCT(I)
C  MASS BALANCE FOR THE WATER PHASE
      DO 15 I=6,NC5
      FCT(I)=FCT(I)/SV
      IF(IG.EQ.1)S(1,IWO,I)=FCT(I)
   15 CONTINUE
C    NEW HOLDUP
      EP(IM,1)=Y1(1,EP(IM,2),(SV-S(IG,IWO,3)),EP(IM,1))
      S(1,ISIG,3)=EP(IM,1)
      IF(IG.EQ.1)EP(IM,2)=EP(IM,1)
      RETURN
   13 CONTINUE
      S(1,IOO,3)=0.
      S(1,IOO,11)=0.
      S(1,IOO,12)=0.
      S(1,IOO,12)=0.
      S(1,IOO,13)=0.
      GO TO 17
      END
```

DYNSYS Subroutine Name - PUMPϕI - Type 7

Written by - A. Lozada

Reviewed by - A. Lozada, August, 1971.

Symbol



Purpose

This module sets the flow of tailings and pressure of the liquid streams leaving the pump.

Description

This module sets the liquid outlet flow from the still body and calculates the recycle flow by subtracting the tailings flow.

Stream Designations

MP(IM,3) - feed stream

MP(IM,4) - tailings stream

MP(IM,5) - recycle stream

Parameters

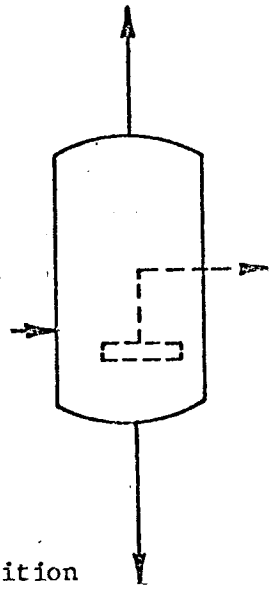
EP(IM,1) - tailings flow in lb/hr

EP(IM,2) - increase in pressure on passing through the pump.

```
      SUBROUTINE TYPE7
**********************************************************************
C       PUMP MODULE
C       EP VECTOR
C        EP(1)  FLOW HANDLED BY THE PUMP IN LB/HR
C       EP(2) INCREMENT IN PRESSURE PRODUCED BY THE PUMP

CNOTE THAT THE FIRST STREAM IN IS THE INPUT TO THE PUMP
C               SECOND STREAM IS THE TAILINGS
C               THIRD STREAM IS THE OUTPUT OF THE OUMP AFTER REMOVING OF TAI
**********************************************************************
      COMMON/UNIT/IM,NMP
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1  EMAX,NPOL,TMAX
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      IN=MP(IM,3)
      FLOW=EP(IM,1)
      IN=MP(IM,3)
      IT=IABS(MP(IM,4))
      IO=IABS(MP(IM,5))
      DO 10 K=5,NC5
      S(1,IT,K)=S(1,IN,K)
      S(1,IO,K)=S(1,IN,K)
   10 CONTINUE
      S(1,IT,4)=S(1,IN,4)
      S(1,IO,4)=S(1,IN,4)
      S(1,IT,5)=S(1,IN,5)+EP(IM,2)
      S(1,IO,5)=S(1,IN,5)+FP(IM,2)
      S(1,IN,3)=FLOW
      AUX=FLOW-S(1,IT,3)
      IF(AUX.LT.0.) GO TO 20
      S(1,IO,3)=AUX
      RETURN
   20 CONTINUE
      S(1,IT,3)=FLOW
      S(1,IO,3)=0.
      RETURN
      END
```

Symbol



DYNSYS Subroutine Name - STILLϕI - Type 8

Written by - A. Lozada,

Revised by - A. Lozada, August, 1971.

Purpose

To simulate the dynamic behaviour of the adiabatic flash in the fatty acid still body.

Description

This module calculates the amount and composition of the liquid stream formed after flashing of the liquid phase. The present model uses the results of the steady state model as "split factors".

ALPHA (1, IG) - ratio of the flash temperature to the feed temperature.

ALPHA (K, IG) - for K = 2, ..., 9 - ratios of the component mass flows in the liquid phase to those in the feed stream.

ALPHA (10,IG) - total mass flow of liquid to mass flow of liquid in the feed.

The liquid phase holdup is calculated from the component mass balances for the liquid phase:

$$\frac{dm_i}{dt} = FL * w_i - FO * w_s$$

where FL - liquid produced in the flashing of the feed.

w_i - the weight fraction of component i in this stream produced by flashing

FO - output liquid flow from still

w_s - the weight fraction of component i in the still

The temperature of the liquid in the still is calculated by the heat balance

$$\frac{dHT_s}{dt} = FL * C_P * TF - FO * C_P * TS$$

where TF - the temperature of the feed after flashing.

TS - the temperature in the still

Parameters

EP(IM, 1) - initial holdup

EP(IM, 2) - total still volume

EP(IM, 3) - not used

EP(IM, 4) - last holdup, used internally.

Stream Designations

MP(IM, 3) - inlet stream mixture of fatty acids and non volatiles.

MP(IM, 4) - leaving liquid phase

MP(IM, 5) - leaving vapor phase

MP(IM, 6) - level signal

```
      SUBROUTINE TYPE 8
C     ******************************************************************
C      STILL MODULE
C     PARAMETERS......
C     -1-     INITIAL HOLDUP
C     -2-     VOLUME
C     -3-     NOT USED
C     -4-     LAST HOLUP, USED INTERNALLY
C     -5-     1ST. OUPUT STREAM, USED INTERNALLY
C     4 AND 5 ARE NOT SPECIFIED BY USER.
C     ******************************************************************
      COMMON/UNIT/IM,NMP
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1  EMAX,NPOL,TMAX
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      COMMON/PTAB/IGFLAG,PP(10,10)
      DIMENSION CMPT(10)
      DIMENSION ALPHA(10,2)
      ALPHA(1,1)=ALPHA(1,2)=.92957
      ALPHA(2,1)=1LPHA(2,2)=1.
      ALPHA(3,1)=ALPHA(3,2)=1.
      ALPHA(4,1)=ALPHA(4,2)=0.
      ALPHA(5,1)=ALPHA(5,2)=0.
      ALPHA(6,1)=ALPHA(6,2)=0.
      ALPHA(7,1)=ALPHA(7,2)=1.
      ALPHA(8,1)=ALPHA(8,2)=.412319
      ALPHA(9,1)=ALPHA(9,2)=.13154648
      ALPHA(10,1)=ALPHA(10,2)=0.86944
      NC1=NCOMP+1
      VOL=EP(IM,2)
      IS=MP(IM,3)
      ISO=IABS(MP(IM,6))
      IFT=IABS(MP(IM,5))
      IOUT=IABS(MP(IM,4))
      IF(KLOOP.GT.1) GO TO 40
C     CALC.  INITIAL CONDITIONS IF NOT SPECIFIED.
      IF(VOL.GT.0. .AND.EP(IM,1).LE.0.) EP(IM,1)=VOL*SG(1,IS)
      EP(IM,4)=EP(IM,1)
   40 CONTINUE
C     CALCULATION OF DERIVATIVES
      DO 50 J=1,NC1
      CMPT(J)=0
   50 CONTINUE
C     FIRST INPUT IS FLASHED
      AUX=0.
      DO 71 K=2,NC1
      KK=K+4
      CMPT(K)=CMPT(K)+S(IG,IS,KK)*S(IG,IS,3)*ALPHA(K,IG)
      AUX=AUX+CMPT(K)
   71 CONTINUE
      CMPT(1)=CMPT(1)+S(IG,IS,4)*CP(IG,IS)*ALPHA(1,IG)*AUX-
     1S(IG,IOUT,4)*CP(IG,IOUT)*S(IG,IOUT,3)
      S(IG,IFT,3)=S(IG,IS,3)-AUX
      DO 72 K=2,NC1
      KK=K+4
      S(IG,IFT,KK)=(S(IG,IS,KK)*S(IG,IS,3)-CMPT(K))/S(IG,IFT,3)
      IF(KK.EQ.9)S(IG,IFT,KK-1)=S(IG,IFT,KK)
   72 CONTINUE
      S(IG,IFT,9)=0.
```

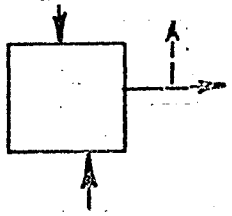
```
C       CALCULATION FOR NOO-FLASHING INPUTS AND OUTPUTS
C       MASS BALANCES
        DO 75 K=2,NC1
        KK=K+4
 75     CMPT(K)=CMPT(K)-S(IG,IOUT,KK)*S(IG,IOUT,3)
C       PREDICT/CORRECT NEW VALUES
C       MASSES
        SUM=0.
        DO 150 K=6,NC5
        KK=K-4
        Y2=EP(IM,4)*S(2,IOUT,K)
        X=EP(IM,1)*S(1,IOUT,K)
        S(1,IOUT,K)=Y1(1,Y2,CMPT(KK),X)
 150    SUM=SUM+S(1,IOUT,K)
        DO 175 K=6,NC5
        S(1,IOUT,K)=S(1,IOUT,K)/SUM
 175    CONTINUE
C       HEAT
        Y2=EP(IM,4)*S(2,IOUT,4)*CP(2,IOUT)
        X=EP(IM,1)*S(1,IOUT,4)*CP(1,IOUT)
        HEAT=Y1(1,Y2,CMPT(1),X)
C       THE NEW TEMPERATURE
 200    S(1,IOUT,4)=HEAT/(SUM*CP(1,IOUT))
C       INSERT THE NEW VALUES IN THE OUTPUT STREAMS
        S(1,IFT,4)=S(1,IOUT,4)
C       FOR CONTROLLER OUTPUT,HOLD-UP STORED   IN PLACE OF FLOW
        S(1,ISO,3)=SUM
        EP(IM,1)=SUM
        IF(IG.EQ.1) EP(IM,4)=SUM
        RETURN
        END
```

DYNSYS Subroutine Name - EXCHϕI - Type 9

Written by - A. Lozada

Revised by - A. Lozada, August, 1971.

Symbol



Purpose

This module calculates the outlet temperature of the liquid leaving the heat exchanger, given steam flow, heat of condensation, exchanger area and overall heat transfer coefficient.

Stream Designation

MP(IM, 3) - cool liquid inlet
MP(IM, 4) - steam
MP(IM, 5) - hot liquid outlet
MP(IM, 6) - outlet temperature signal.

Parameters

EP(IM, 1) - exchanger area ft^2
EP(IM, 2) - overall heat transfer coefficient BTU/ft^2 ^oF
EP(IM, 3) - steam condensation heat BTU/lb

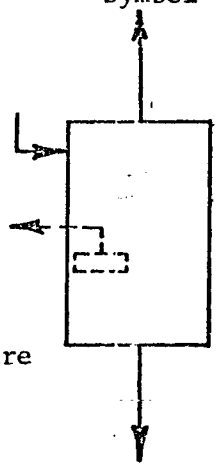
```
      SUBROUTINE TYPE 9
*****************************************************************
C       HEAT EXCHANGER WITH CONDENSING VAPOUR IN SHELL
C     EP VECTOR
C       EP( ,1) HEAT EXCHANGER AREA (SQFT)
C       EP( ,2) OVERALL HEAT TRANSFER COEFICIENT(BTU/HR SQFT F)
C       EP( ,3) VAPOUR CONDENSATION HEAT(BTU/LB)
*****************************************************************
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      COMMON/UNIT/IM,NMP
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1 EMAX,NPOL
      REAL LAT
      A=EP(IM,1)
      H=EP(IM,2)
      LAT=EP(IM,3)
      INP=MP(IM,3)
      IST=MP(IM,4)
      IO=IABS(MP(IM,5))
      ISI=IABS(MP(IM,6))
      F=S(1,INP,3)
      TI=S(1,INP,4)
      STM=S(1,IST,3)
      P=S(1,IST,5)
      TS=S(1,IST,4)
      S(1,IO,3)=S(1,INP,3)
      FO=S(1,IO,3)
      DO 20 I=1,NCOMP
      K=I+5
      S(1,IO,K)=S(1,INP,K)
   20 CONTINUE
      TO=S(1,IO,4)
      CPI=CP(1,INP)
      CPO=CP(1,IO)
      Q=S(1,IST,3)*LAT
      TO=(F*CPI*TI+Q)/(FO*CPO)
      IF(TO.GT.TS)TO=TS
   30 S(1,IO,4)=TO
      S(1,ISI,4)=TO
      RETURN
      END
```

DYNSYS Subroutine Name - FLASϕI - Type 10

Written by - D.D. Dang

Revised by - A. Lozada, August, 1971.

Symbol



Purpose

To flash off the excess water of the inlet liquid stream as a result of the lowering in pressure in flash tank.

Description

This subroutine allows the calculation of the outlet vapour and liquid flow and its compositions knowing the inlet liquid flow and composition as well as the equilibrium temperature and pressure in the tank. Part of the heat carried by the input liquid stream is used to transfer some of the water into vapour as a result of the lowering pressure in flash. The level of liquid in flash tank is adjusted by a controller and a controlled valve located at the outlet liquid stream.

Assumptions

The following assumptions are made on the module:

1) temperature and pressure in flash tank are fixed
2) only water is vaporized
3) there is no holdup of vapour phase in tank.

Stream Designations

MP(IM, 3) - inlet stream of wet fatty acid
MP(IM, 4) - outlet water vapour stream
MP(IM, 5) - outlet dried fatty acid
MP(IM, 6) - level signal

Parameters

To execute the module the following parameters are supplied by the user:

EP(IM, 1) - initial liquid hold-up

EP(IM, 2) - hold-up at corrector step

EP(IM, 3) - latent heat of vaporization for water at flash pressure BTU/lb

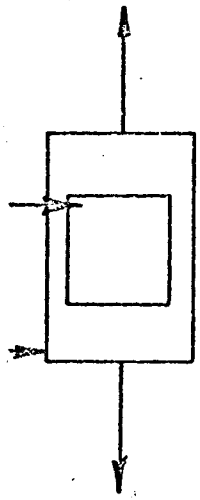
EP(IM, 4) - flash temperature ^{o}F

EP(IM, 5) - flash pressure psia.

```
      SUBROUTINE  TYPE 10
*****************************************************************************
C       FLASH TANK MODULE
C       EP VECTOR
C       EP( ,1) INITIAL HOLDUP
C       EP( ,2) HOLD UP FOR PREVIOUS STEP
C       EP( ,3) LATENT HEAT OF VAPORIZATION AT FLASH PRESSURE AND TEMP
C       EP( ,4)  FLASH TEMPERATURE (D.F)
C       EP( ,5) OPERATION PESSURE
*****************************************************************************
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1  EMAX,NPOL,TMAX,NCOUNT,JM,KJ,MPR
      COMMON/PTAB/ IGFLAG,PP(10,10)
      COMMON/UNIT/IM,NMP
      DIMENSION  SL(10),HUV(2)
      IN=MP(IM,3)
      IVO=-MP(IM,4)
      ILO=-MP(IM,5)
      ISIG=-MP(IM,6)
      IF(TIME(2).EQ.0. .AND. IG.EQ.2) EP(IM,2)=EP(IM,1)
C  COMPOSITION OF THE INLET STREAM AFTER SEPARATING FROM THE VAPOR
      SUM=0.
      SY=0.
      S(1,IVO,4)=EP(IM,4)
      S(1,ILO,4)=EP(IM,4)
      S(1,IVO,5)=EP(IM,5)
      S(1,ILO,5)=EP(IM,5)
      WR=S(IG,IN,3)*S(IG,IN,9)
      HVP=EP(IM,3)
      QL=S(IG,IN,3)*CP(IG,IN)*(S(IG,IN,4)-S(IG,ILO,4))
      S(IG,IVO,3)=QL/HVP
      IF(S(IG,IVO,3).GT.WR) S(IG,IVO,3)=WR
      DO 11 I=1,NCOMP
      I6=I+5
      SL(I)=S(IG,IN,3)*S(IG,IN,I6)
      IF(I.EQ.4) SL(I)=SL(I)-S(IG,IVO,3)
   11 SUM=SUM+SL(I)
      DO 12 I=1,NCOMP
   12 SL(I)=SL(I)/SUM
C  MASS BALANCE FOR THE L-PHASE
      DO 14 I=1,NCOMP
      I6=I+5
      DL=SUM*SL(I)-S(IG,ILO,3)*S(IG,ILO,I6)
      Y2=EP(IM,2)*S(2,ILO,I6)
      IF(IG.EQ.1) X=EP(IM,1)*S(1,ILO,I6)
      S(1,ILO,I6)=Y1(0,Y2,DL,X)
   14 SY=SY+S(1,ILO,I6)
      EP(IM,1)=SY
      S(1,ISIG,3)=EP(IM,1)
C  CONCENTRATION OF OUTLET LIQUID STREAM
      DO 15 I=6,NC5
   15 S(1,ILO,I)=S(1,ILO,I)/SY
C  CONCENTRATION OF OUTLET VAPOR STREAM
      DO 10 I=6,NC5
   10 S(1,IVO,I)=0.
      S(1,IVO,8)=1.
      IF(IG.EQ.1) EP(IM,2)=EP(IM,1)
      RETURN
```

END

Symbol



DYNSYS Subroutine Name - EVAPϕ1 - Type 11

Written by - D.D. Dang

Revised by - A. Lozada, August, 1971.

Purpose

To concentrate the glycerine solution by heat transfer from steam chest, using natural circulation boiling.

Description

This subroutine is very similar to that of the flash tank. It allows the calculation of the outlet vapour and liquid flow and its compositions knowing the inlet liquid stream and compositions. The module also calculates the amount of heat transferred from the steam chest, knowing the temperature in evaporator and steam chest, the heat exchange area and heat transfer coefficient of the steam chest.

Assumptions

The following assumptions are made on the module:

1) temperature and pressure are fixed.
2) no glycerine vapour in vapour stream.
3) no vapour hold-up in evaporator.

Stream Designations

MP(IM, 3) - inlet glycerine solution to be concentrated
MP(IM, 4) - inlet vapour stream to steam chest
MP(IM, 5) - outlet vapour stream
MP(IM, 6) - outlet concentrated glycerine solution

Parameters

EP(IM, 1) - initial liquid hold-up

EP(IM, 2) - hold-up at corrector step

EP(IM, 3) - heat transfer coefficient BTU/hr ft^2 ^oF

EP(IM, 4) - latent heat of vaporization for the liquid in the evaporator BTU/lb

EP(IM, 5) - latent heat of condensation for the steam in the chest BTU/lb

```
      SUBROUTINE TYPE11
C       THIS SUBROUTINE SIMULATES THE DYNAMICS OF
C       AN EVAPORATOR UNIT

C       EN(IM, ) PARAMETERS
C       1- HOLDUP OF LIQUID IN THE P'EDICTOR
C       2- HOLDUP OF LIQUID IN THE CORRECTOR
C       3- OVERALL HEAT TRANSFER COEF
C       4- LATENT HEAT OF VAPORIZATION
C       5- LATENT HEAT OF CONDENSATION FOR STEAM IN THE CHEST
      COMMON/UNIT/IM,NMP
      COMMON/CON/IG,KLOOP,NCOMP,NC5,DELT,NE,TIME(30),NS,NPR,TOLL,TOLU,
     1  EMAX,NPOL,TMAX,NCOUNT,JM,KJ,MPR
      COMMON  MP(40,8),EP(40,5),S(2,50,20),EX(200)
      DIMENSION FCT(8)
      AREA=378.
      ILI=MP(IM,3)
      IVI=MP(IM,4)
      ILO=-MP(IM,5)
      IVO=-MP(IM,6)
      IF(TIME(2).EQ.0..AND.IG.EQ.2)EP(IM,2)=EP(IM,1)
C       SET COMPOSITION OF THE VAPOUR AS PURE STEAM
      DO 10 K=6,NC5
      S(IG,IVO,K)=0.
   10 CONTINUE
      S(IG,IVO,8)=1.
      SL=0.
      SX=0.
C       HEAT TRANSFERED TO THE LIQUID
      QT=EP(IM,3)*AREA*(S(IG,IVI,4)-S(IG,ILO,4))
C       HEAT AVAILABLE FROM STEAM CONDENSATION
      QV=S(IG,IVI,3)*EP(IM,5)
      IF(QV.GT.QT)QV=QT
C       HEAT GAINED BY THE LIQUID
      QL=S(IG,ILI,3)*CP(IG,ILI)*(S(IG,IVI,4)-S(IG,ILO,4))
      S(IG,IVO,3)=(QV+QL)/EP(IM,4)
      AUX=S(IG,ILI,3)*S(IG,ILI,9)
      IF(S(IG,IVO,3).GT.AUX)S(IG,IVO,3)=AUX
      DO 11 I=1,NCOMP
      I6=I+5
      FCT(I)=S(IG,ILI,3)*S(IG,ILI,I6)
      IF(I.EQ.4)FCT(I)=FCT(I)-S(IG,IVO,3)
   11 SL=SL+FCT(I)
C       MAS BALANCE FOR COMPONENTS IN LIQUID PHASE
      DO 12 I=1,NCOMP
      I6=I+5
      DL=FCT(I)-S(IG,ILO,3)*S(IG,ILO,I6)
      Y2=EP(IM,2)*S(2,ILO,I6)
      IF(IG.EQ.1)X=EP(IM,1)*S(1,ILO,I6)
      S(1,ILO,I6)=Y1(0,Y2,DL,X)
      SX=SX+S(1,ILO,I6)
   12 CONTINUE
      DO 13 I=6,NC5
   13 S(1,ILO,I)=S(1,ILO,I)/SX
      S(1,IVO,4)=S(1,ILO,4)
      IF(IG.EQ.1) EP(IM,2)=EP(IM,1)
      RETURN
      END
```

Data Set for the Dynamic Simulation

A data set for studying the dynamic behaviour of the fat hydrolysis plant, based on the dynamic information flow diagram shown as Figure 4, follows.

This data set may be stored as a permanent file on the computer and used for both interactive or batch runs. While interactive mode is used SCOPE INTERCOM commands allow the user to make alterations in the data set previous to the execution of the programme.

In total ten different modules were used and 50 streams. A listing of the complete data set follows.

```
**********************************************************************

            DYNAMIC SIMULATION
                  OF A
            FAT HYDROLYSYS UNIT

**********************************************************************

BEGIN
COMPS       8.
DELT        0.01
PROCESS
MXPLT01     1.
            15.         16.         17.         -20.        -19.
            12.         1.
CNTRL01     2.
            19.         -18.
            4.          1000.       478.        3.0         4.0
VALV01      3.
            18.         -17.
            3.          0.3         -1.
MXPLT01     4.
            2.          1.          -5.         -3.
            13.         1.
CNTRL01     5.
            3.          -4.
            4.          1000.       478.        3.0         1.5
VALV01      6.
            4.          -2.
            6.          0.5         -1.
EXTC01      7.
            5.          24.         -6.         -25.
            830.        2390.
EXTC01      8.
            6.          23.         -7.         -24.
            830.        2390.
EXTC01      9.
            7.          22.         -8.         -23.
            830.        2390.
EXTC01      10.
            8.          21.         -9.         -22.
            830.        2390.
SETL01      11.
            9.          -10.        -13.        -11.
            500.
CNTRL01     12.
            11.         -12.
            3.          700.        500.        3.          3.5
VALV01      13.
            12.         -10.
            13.         1.0         1.          0.          0.
MXPLT01     14.
            20.         13.         -21.
             9.         1.
FLAH01      15.
            10.         -42.        -41.        -49.
            1000.       0.          918.        0.          0.
CNTRL01     16.
```

```
                49.             -50.
                3.              2000.           1000.           5.              4.0
VALV01          17.
                50.             -41.
                17.             1.0             1.              0.              0.
EVAP01          18.
                41.             42.             -43.            -44.
                312.            0.              200.            937.4           918.
EVAP01          19.
                43.             44.             -45.            -46.
                312.            0.              140.            959.6           937.4
EVAP01          20.
                45.             46.             -47.            -48.
                312.            0.              70.              991.4          959.6
FLAH01          21.
                25.             -40.            -26.            -38.
                1000.           0.              918.            0.              0.
CNTRL01         22.
                38.             -39.
                3.              2000.           1000.           10.             4.
VALV01          23.
                39.             -26.
                23.             1.5             1.              0.              0.
STIL01          28.
                29.             -33.            -37.            -32.
                2000.
CNTRL01         29.
                32.             -34.
                3.              500.            2000.           10.             1.
VALV01          30.
                34.             -36.
                30.              .8             1.
PUMP01          31.
                33.             -36.            -35.
                50999.67        55.35
MXPLT01         24.
                26.             35.             -27.
                                1.              1.
EXCH01          25.
                27.             28.             -29.            -31.
                551.88
CNTRL01         26.
                31.             -30.
                4.              50.             460.            0.5             1.
VALV01          27.
                30.             -28.
                                0.5             -1.0
END
```

STREAMS 50.
EXPLICIT

1.00	-1.00	4000.00	200.00	14.70
0.00000	0.00000	0.00000	1.00000	0.00000
0.00000	0.00000	0.00000		
2.00	-1.00	2438.68	515.00	777.96
0.00000	0.00000	1.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
3.00	-11.00	0.00	478.94	0.00
0.00000	0.00000	0.00000	1.00000	0.00000
0.00000	0.00000	0.00000		
4.00	-11.00	30.16	0.00	0.00
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
5.00	-1.00	6438.68	478.94	776.70
0.00000	0.00000	0.00000	1.00000	0.00000
0.00000	0.00000	0.00000		
6.00	-1.00	6390.25	478.00	776.70
.00000	.00000	.00000	.99722	.00278
.00000	.00000	.00000		
7.00	-1.00	6308.60	478.00	776.70
.00000	.00000	.00000	.98877	.01123
.00000	.00000	.00000		
8.00	-1.00	6127.89	478.00	776.70
.00000	.00000	.00000	.96230	.03770
.00000	.00000	.00000		
9.00	-1.00	5633.54	478.00	776.70
.00000	.00000	.00000	.87079	.12921
.00000	.00000	.00000		
10.00	1.00	5630.41	478.00	776.70
.00000	.00000	.00000	.87079	.12921
0.00000	0.00000	.00000		
11.00	11.00	499.84	0.00	0.00
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
12.00	-11.00	75.04	0.00	0.00
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
13.00	-1.00	.00	478.00	776.70
0.00000	0.00000	0.00000	0.00000	0.00000
.00475	1.76143	.23383		
14.00	-1.00	667.28	478.00	776.70
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	.80265	.19735		
15.00	-1.00	6983.54	220.00	14.70
1.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
16.00	-1.00	0.00	220.00	14.70
0.00000	0.00000	0.00000	1.00000	0.00000
0.00000	0.00000	0.00000		
17.00	-1.00	2374.30	515.00	777.96
0.00000	0.00000	1.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
18.00	-11.00	11.04	0.00	0.00
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
19.00	-11.00	0.00	493.70	0.00
.74627	0.00000	0.00000	.25373	0.00000
0.00000	0.00000	0.00000		

```
  20.00      -1.00     9357.93     493.70      776.70
 .74627    0.00000     0.00000     .25373     0.00000
0.00000    0.00000     0.00000
  21.00       1.00     9357.93     493.70      776.70
 .74627    0.00000     0.00000     .25373     0.00000
 .00000     .00000      .00000
  22.00      -1.00     9852.13     478.00      776.70
 .22600     .00000      .00000     .30399      .00000
 .00000     .47001      .00000
  23.00      -1.00    10032.88     478.00      776.70
 .07311     .00000      .00000     .32386      .00000
 .00000     .60303      .00000
  24.00      -1.00    10114.47     478.00      776.70
 .02419     .00000      .00000     .33521      .00000
 .00000     .64060      .00000
  25.00       1.00    10162.89     478.00      776.70
 .00809     .00000      .00000     .34334      .00000
 .00000     .64857      .00000
  26.00       1.00     8581.67     289.29       55.45
 .00961     .00000      .00000     .22666      .00000
 .00000     .76373      .00000
  27.00      -1.00    59581.34     405.08       55.45
 .03884     .32303     0.00000     .03265      .00000
 .42423     .17753      .00489
  28.00      -1.00     2580.88     515.00      776.70
0.00000    0.00000     1.00000    0.00000     0.00000
0.00000    0.00000     0.00000
  29.00       1.00    59581.34     451.05       55.45
 .03884     .32303     0.00000     .03265      .00000
 .42423     .17753      .00489
  30.00     -11.00       28.15       0.00        0.00
0.00000    0.00000     0.00000    0.00000     0.00000
0.00000    0.00000     0.00000
  31.00     -11.00        0.00     451.05        0.00
0.00000    0.00000     0.00000    0.00000     0.00000
0.00000    0.00000     0.00000
  32.00      11.00     2000.       399. 9         .10
 .01240     .36610     0.00000    0.00000     0.00000
 .49030     .12160      .00960
  33.00      -1.00    50999.67     426.67         .10
 .04376     .37739      .00000     .00000      .00000
 .49305     .08505      .00075
  34.00     -11.00     -411.58       0.00        0.00
0.00000    0.00000     0.00000    0.00000     0.00000
0.00000    0.00000     0.00000
  35.00      -1.00    50999.67     426.67       55.45
 .04376     .37739      .00000    0.00000     0.00000
 .49561     .07889      .00571
  36.00       1.00        0.00     426.67       55.45
 .04376     .37739      .00000    0.00000     0.00000
 .49561     .07889      .00570
  37.00       1.00     8344.75     426.67        0.00
0.00000    0.00000      .23309    0.00000      .00000
0.00000     .74492      .03031
  38.00      11.00      998.36       0.00        0.00
0.00000    0.00000     0.00000    0.00000     0.00000
0.00000    0.00000     0.00000
```

39.00	-11.00	75.64	0.00	0.00
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
40.00	-1.00	1580.21	289.29	55.45
0.00000	0.00000	1.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
41.00	-1.00	4531.40	289.29	55.45
.00000	.00000	.00000	.84005	.15995
.00000	.00000	.00000		
42.00	-1.00	1094.01	289.29	55.45
0.00000	0.00000	1.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
43.00	-1.00	3348.32	264.50	38.45
.00000	.00000	.00000	.78387	.21613
.00000	.00000	.00000		
44.00	-1.00	1183.08	264.50	38.45
0.00000	0.00000	1.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
45.00	-1.00	2077.24	228.10	20.29
.00000	.00000	.00000	.65294	.34706
.00000	.00000	.00000		
46.00	-1.00	1271.08	228.10	20.29
0.00000	0.00000	1.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
47.00	1.00	757.40	178.00	7.18
.00000	.00000	.00000	.07642	.92358
.00000	.00000	.00000		
48.00	-1.00	1319.84	178.00	7.18
0.00000	0.00000	1.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
49.00	11.00	999.97	0.00	0.00
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		
50.00	-11.00	67.32	0.00	0.00
0.00000	0.00000	0.00000	0.00000	0.00000
0.00000	0.00000	0.00000		

END

```
PROPERTIES   5.
850.4        0.          0.          .55
756.95       0.          0.          .58
18.          0.          0.          .435        .000071
18.          0.          0.          1.          0.
92.11        0.          0.          .576        0.
500.         0.          0.          .7          0.
284.         0.          0.          .63         0.
270.         0.          0.          .69
9.
0.1
21.          12.         0.          1.
21.          7.          0.          1.
23.          6.          0.          1.
25.          12.         0.          1.
25.          13.         0.          1.
37.          12.         0.          1.
37.          13.         0.          1.
38.          3.          0.          1100.
32.          3.          0.          3000.
```