THE DETERMINATION OF PLATE EFFICIENCIES IN UNSTEADY-STATE PLATE COLUMN MODELS

by

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1. ABSTRACT

1. Abstract

Other investigations have shown the relationship between various plate efficiencies and the relationship between the Murphree plate efficiency and the point efficiency for steady-state conditions. So far there has been no work done on the relationship between the Murphree plate efficiency and the point efficiency for unsteady-state conditions. In unsteady-state distillation simulation, the Murphree plate efficiency has been held constant and it is thought that this may be the reason for the differences between experimental and theoretical liquid composition responses in some cases. Further, the liquid mixing models used to represent the mixing occurring on a distillation plate, do not include downcomer sections and the need for experimental investigation of the downcomer effect is required to produce a realistic model.

Experimental conductivity impulse responses, using potassium chloride tracer, are obtained on a 7ft. x 1.5 ft. sieve plate using the system air-water. The sieve trays, weir height and downcomer segmental area are variable, and the responses are compared with those obtained by the diffusion model. Intermediate responses along the tray are also compared with those of the diffusion model and with

those of the theoretical model to be used in the unsteady-state simulation. Empirical equations for estimating the eddy diffusivity are given and the significance of the downcomer area is shown.

Steady-state experiments using a continuous helium injection in the vapour stream are used to show that the general assumption of perfect vapour mixing between the sieve trays is untrue. An equation for estimating the eddy diffusivity in the vapour phase is given and the assumption of no vapour mixing is shown to be more realistic.

From the experimental results, a generalised model of a sieve tray is given, taking into account the liquid and vapour mixing characteristics. This generalised model consists of a series of perfectly mixed liquid pools with recycle between each pool and the vapour leaving each pool rises to the pool immediately above on the next tray.

Using the theoretical model, the unsteady-state equations for a 5-plate distillation column are formulated. Step changes in the feed composition, feed flow rate, boilup and reflux are investigated for binary systems and the effect of these changes on the Murphree vapour plate efficiency and liquid compositions are shown.

The generalised model was used for various investigations. The liquid composition responses of the generalised model were compared with those obtained by simplified models using a constant Murphree plate efficiency which included the liquid mixing effect and for those obtained where the liquid was perfectly mixed.

The unsteady-state equations of a linearised model are also formulated using constant Murphree vapour plate efficiencies and the comparison of the liquid composition responses is made with those of the generalised model where the plate efficiency is variable. From the comparison, simplifying assumptions about the significance of the Murphree plate efficiency during unsteady-state operations are given. The relationship between the extent of plate efficiency changes during unsteady-state and the equilibrium relationship is shown.

The generalised model is also used in representing a gas absorption column and the unsteady-state equations for the trays are the same as those for the distillation column. Step changes in the liquid feed composition are made and the unsteady-state composition responses are investigated and simplifying assumptions are again given.

The generalised model is also used for investigating ternary systems in distillation. Step changes in the feed composition are made and constant relative volatilities are used. It is shown that large differences between the initial and final steady-state values of the Murphree plate efficiency for the middle phase component occur and thus when using simplified models with constant plate efficiencies, the light and heavy phase components should be those operated on.

A general discussion on the type and size of disturbance introduced and its effect on the Murphree vapour plate efficiency is given. The limiting values of the equilibrium relationships and the simplifications that can be made from this investigation are given. A discussion on the results of this investigation for simplifying distillation simulation and suggestions of its use in future work are given.

2. INTRODUCTION

2. Introduction

Since the late 1950's, the unsteady-state behaviour of plate columns has been extensively investigated. Chemical engineers have for a long time been interested in the column dynamics, but until the introduction of computers, little work had been done. With the arrival of computers came an increased interest in the unsteady-state simulation of plate columns, but due to the inadequate numerical integration routines and the actual computation time, assumptions of perfect liquid mixing on the plates or at any rate constant plate efficiencies were made. Making the assumption of perfect liquid mixing gives the Murphree plate efficiency equal to that of the point efficiency which hardly changes for composition disturbances introduced.

Work done using linearised models and matrix methods for solving the unsteady-state behaviour also assumed constant plate efficiency. There were no grounds for this assumption and the assumption of perfect liquid mixing is only valid for small diameter columns, but for large diameter columns (i.e. greater than 0.5 metre diameter) there is a concentration gradient across the tray and the liquid mixing must be accounted for.

The aim of this work is to investigate the liquid mixing characteristics of large sieve tray columns in order to derive a theoretical model which can be easily used in unsteady-state simulation and still give accurate transient responses. The extent of vapour mixing between the trays and its effect on the numerical solution are investigated. The theoretical model used in the unsteady-state simulation is based on the experimental results.

The theoretical model is used to investigate the changes in the liquid compositions and the Murphree plate efficiency for unsteady-state operations. The disturbances investigated for binary systems are step changes in:-

1) feed composition
2) feed flow with vapour and liquid reflux flow held constant
3) vapour flow with liquid reflux flow held constant
4) liquid reflux flow with vapour flow held constant

The magnitude and transient behaviour of the plate efficiency responses and the difference in their initial and final steady-state values are investigated.

The liquid transient responses of the theoretical model for step changes in feed composition are compared with those obtained from a simplified model with perfect liquid mixing and with those of a linearised model with constant plate efficiency to see if any simplifying assumptions can be made.

As most, but not all, distillation systems are gas phase limited, a gas absorption column which was liquid-phase-limited was simulated so that small values of the plate efficiency could be investigated.

The theoretical model is further used to investigate the composition and plate efficiency responses of a ternary distillation system for step changes in the feed composition. The conclusions drawn from the binary results are reviewed to see if they apply to multicomponent systems.

In all the cases where composition changes only occur, the assumption of constant molal flow is made. This assumption is also made for the initial and end steady-state simulation periods when flow changes occur.

The numerical integration routine used to solve the unsteady-state equations is relatively new. The integration routine used in this problem must be able to solve stiff equations as unstiff methods increase the actual computation time by an order of 20 or more and are less accurate. Thus it is only the increased speed of modern computers and the greater efficiency of modern integration routines that have made this work possible.

3. LITERATURE SURVEY ON PLATE EFFICIENCY:- GENERAL CONSIDERATIONS

3. Literature Survey on Plate Efficiency:

General Considerations

3.1 The Function of Efficiency

Equilibrium-stage efficiencies mentioned in this work can be applied to all stagewise separation processes. Separation efficiencies for plate columns were introduced as a means of presenting the extent of mass and heat transfer between the phases in contact on an actual plate. If an ideal column can be defined as one where the two phases in contact on a plate leave the plate in perfect equilibrium and where its behaviour can be calculated using material and enthalpy balances, the efficiency can be used to estimate true compositions on plates compared to what they would have been if the plate had been ideal. The equilibrium value used for defining efficiency is that of the streams leaving the plate.

3.2 Column or Overall Efficiency

There are three main efficiencies to be considered in distillation processes; point efficiency, plate efficiency and column or overall efficiency. The column or overall efficiency is calculated from:-

$$\text{Column efficiency} \quad E_O = \frac{\text{Number of ideal stages}}{\text{Actual number of stages}} \qquad \ldots\ldots \; 3.1$$

The plate efficiency estimated for specified operating conditions has generally been the basis of estimating the column efficiency.

The overall column efficiency can be calculated for the case where the equilibrium and operating lines are essentially straight but not necessarily parallel, by the equation (1):- where $\lambda = \frac{mV}{L}$

$$E_O = \frac{\ln(1 + E_a(\lambda - 1)}{\ln \lambda} \qquad \text{.........} \; 3.2$$

where E_a is the tray efficiency if the entrainment is considered. If the column has two different values of λ for the rectifying section and the stripping section then equations 3.1 and 3.2 are to be applied for both sections.

A different approach using empirical equations for calculating the overall column efficiency was presented by O'Connell (73). For standard operating conditions of vapour and liquid flows and for standard tray design, he derived some simple relationships between column parameters and the overall efficiency for absorption and distillation bubble-cap tray columns.

3.3 Point Efficiency

Fig 3.1 shows a schematic representation of one tray 'n' in a multitray column. This tray n delivers liquid to the tray below of average composition X_n. If a specific point

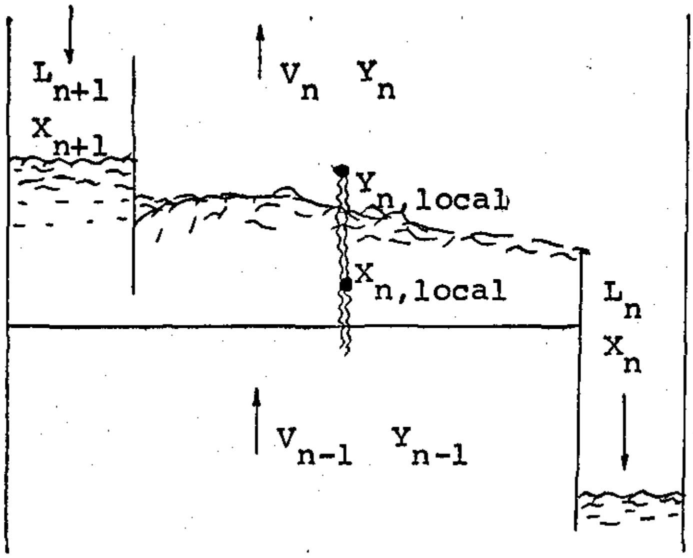


Fig 3.1 Schematic representation of one tray 'n' in a multitray column.

on that tray was investigated, it would have an element of vapour of composition $Y_{n-1,local}$ which enters the liquid of concentration $X_{n,local}$ and leaves with a concentration of $Y_{n,local}$. The efficiency for this point can then be defined by:-

$$E_{OG} = \frac{Y_{n,local} - Y_{n-1,local}}{Y^{*}_{n,local} - Y_{n-1,local}} \quad \text{.........} \; 3.3$$

where $Y^{*}_{n,local}$ is the vapour composition in equilibrium with the liquid $X_{n,local}$ .The point efficiency as defined above, plays a great part in distillation simulation if the liquid on a plate is not perfectly mixed. The means of calculating the point efficiency has been given in many papers (1,93,104, 107) but the main relationships are:-

$$E_{OG} = 1 - \exp(-N_{OG}) \qquad \text{........... 3.4}$$

where $$\frac{1}{N_{OG}} = \frac{1}{N_G} + \frac{mV}{L}\cdot\frac{1}{N_L} \qquad \text{........... 3.5}$$

In equation 3.5 the terms on the right represent the gas and liquid resistances to mass transfer, respectively. The theory for estimating N_{OG} is given in the A.I.Ch.E Manual (1).

3.4 Murphree Efficiency

If the average compositions of all the local elements of vapour across the tray (Fig 3.1) are considered, then the Murphree efficiency (69)can be defined as:-

$$E_{MV} = \frac{Y_n - Y_{n-1}}{Y_n^* - Y_{n-1}} \qquad \text{......... 3.6}$$

where Y_n^* is the vapour in equilibrium with the liquid leaving the plate. The Murphree efficiency based on the liquid phase can be defined as:-

$$E_{ML} = \frac{X_{n+1} - X_n}{X_{n+1} - X_n^*} \qquad \text{.......... 3.7}$$

If the liquid on a tray is perfectly mixed giving a uniform concentration across it and the vapour entering is of uniform concentration, then:-

$$E_{MV} = E_{OG} \qquad \text{............ 3.8}$$

If the liquid travels across the tray in 'plug flow', then by integrating across the tray the following relationship is obtained (64):-

$$E_{MV} = \frac{1}{\lambda}\left\{e^{\lambda E_{OG}} - 1\right\} \qquad \text{........ 3.9}$$

If the liquid on the tray is partially mixed then for steady-state conditions the vapour plate efficiency can be estimated using (1):-

$$E_{MV} = \frac{1 - e^{-(\eta+Pe)}}{(\eta+Pe)\left(1+\frac{\eta+Pe}{\eta}\right)} + \frac{e^{n} - 1}{\eta\left(1+\frac{\eta}{\eta+Pe}\right)} \qquad \text{........ 3.10}$$

where $\eta = \frac{Pe}{2}\left\{\sqrt{1 + \frac{4\lambda E_{OG}}{Pe}} - 1\right\}$

For steady-state conditions and for constant λ across the tray, relationships between the liquid and vapour plate efficiencies can be derived (1,35).

$$E_{MV} = \frac{E_{ML}}{E_{ML} + \lambda(1 - E_{ML})} \qquad \text{....... 3.11}$$

Standart (95) pointed out that the vapour-phase plate efficiency is useful for plate-to-plate calculations up the column while the liquid-phase plate efficiency is convenient for plate calculations down the column.

For unsteady-state systems and where the significant holdup is that of the liquid phase, Lees (60) showed that the vapour and not the liquid-phase plate efficiency should be used in principle. However, Lees (59) also showed that for some particular gas absorption columns, the difference between the computed frequency responses of the models using the liquid-phase and those using the vapour-phase plate efficiency was small although the principle of using the liquid-phase plate efficiency is wrong.

3.5 Generalised Plate Efficiency and Hausen Efficiency

In 1953 Hausen (42) introduced his definition of efficiency based on the assumption of constant molal flow rates. Hausen considers only material efficiency and ignores the requirement of phase saturation. The Hausen plate efficiency will be defined after the generalised plate efficiency has been derived.

To include heat transfer between liquid and vapour phases, Standart (95) in 1965 introduced the generalised plate efficiency which is a generalisation of the Hausen efficiency. The analysis was later extended by Standart and Kastanek (97) to include the effect of liquid weeping and entrainment.

The idea of the generalised plate efficiency is based on keeping the inflowing streams constant when comparing the actual and ideal plates. Ho and Prince (45) related the Hausen plate efficiency to the mass transfer coefficient and discussed its use in design calculations. For an actual and ideal plate, Standart set down:-

1) overall material balance

$$V_{n-1} + L_{n+1} = V_n + L_n = V'_n + L'_n \qquad \ldots\ldots\ldots\ 3.12$$

2) material balance for the ith component

$$V_{n-1}Y_{n-1,i}+L_{n+1}X_{n+1,i}= V_nY_{n,i}+L_nX_{n,i}= V'_nY'_{n,i}+L'_nX'_{n,i} \qquad \ldots\ 3.13$$

3) overall enthalpy balance

$$V_{n-1}H_{v\ n-1}+L_{n+1}H_{l\ n+1}-Q_n = V_nH_{v\ n}+L_nH_{l\ n}$$

$$=V'_nH'_{v\ n}+L'_nH'_{l\ n} \qquad \ldots\ldots\ldots\ 3.14$$

where Q_n is the rate of heat lost from the plate n to the surroundings. Thus the definition of efficiency is:-

$$E_G = \frac{\text{Total change in property across the actual plate}}{\text{Total change in property across the ideal plate}} \qquad \ldots\ 3.15$$

1) overall material efficiency:-

$$E_V = \frac{V_n - V_{n-1}}{V'_n - V_{n-1}} \quad ; \quad E_L = \frac{L_n - L_{n+1}}{L'_n - L_{n+1}} \qquad \ldots\ldots\ 3.16$$

2) material efficiency for component i:-

$$E_{Vi}= \frac{V_nY_{n,i}-V_{n-1}Y_{n-1,i}}{V'_nY'_{n,i}-V_{n-1}Y_{n-1,i}} \quad ; \quad E_{Li}= \frac{L_nX_{n,i}-L_{n+1}X_{n+1}}{L'_nX'_{n,i}-L_{n+1}X_{n+1}} \qquad \ldots\ldots\ 3.17$$

3) overall thermal efficieny:-

$$E_{HV} = \frac{V_nH_{v\ n}-V_{n-1}H_{v\ n-1} + r_nQ_n}{V'_nH'_{v\ n}-V_{n-1}H_{v\ n-1} +r_nQ_n} \quad ;$$

$$E_{HL} = \frac{L_nH_{l\ n} -L_{n+1}H_{l\ n+1}+ (1-r_n)Q_n}{L'_nH'_{l\ n} -L_{n+1}H_{l\ n+1}+ (1-r_n)Q_n} \qquad \ldots\ldots\ 3.18$$

where r_n is the fraction of the heat lost by the vapour phase.

From equations 3.12 to 3.14 it can be seen that:-

$$E_V =E_L=E \quad ; \quad E_{Vi}=E_{Li}=E_i \quad ; \quad E_{HV}=E_{HL}=E_H$$

Unlike the Murphree plate efficiency, the generalised plate efficiency is the same for both phases. For multicomponent systems of k components then there are k+1 independent generalised efficiencies including the overall thermal efficiency.

If only binary systems are considered with constant molal flow then:-

$$V_n = V'_n = V_{n-1} \quad ; \quad L_n = L_{n+1} = L'_n$$

and from the equation 3.17

$$E_{HG} = E_1 = E_2 = \frac{Y_{n,1} - Y_{n-1,1}}{Y'_{n,1} - Y_{n-1,1}} \quad \text{.........} \; 3.19$$

which by definition (42) is the Hausen plate efficiency and is of the same form as the Murphree plate efficiency:-

$$E_{MV} = \frac{Y_n - Y_{n-1}}{Y^*_n - Y_{n-1}} \quad \text{.........} \; 3.20$$

but the value of Y'_n does not equal Y^*_n as can be shown by comparing the operating lines. Y'_n is the vapour composition in equilibrium with the liquid X'_n which has been shifted from X_n in order to satisfy both the material balance and the thermal equilibrium.

The Murphree ideal plate for the vapour phase is based on:-

$$L^*_n = L_n \quad ; \quad V^*_{n-1} = V_{n-1} \quad ; \quad X^*_n = X_n \quad ; \quad Y^*_{n-1} = Y_{n-1} \quad \text{.....} 3.21$$

The Hausen ideal plate is based on:-

$$L'_{n+1} = L_{n+1} \quad ; \quad V'_{n-1} = V_{n-1} \quad ; \quad X'_{n+1} = X_{n+1}$$

$$Y'_{n-1} = Y_{n-1} \qquad \text{.........} \; 3.22$$

For the component balance:-

$$V_{n-1}Y^{*}_{n-1} + L_{n+1}X^{*}_{n+1} = V_nY^{*}_n + L_nX^{*}_n \qquad \text{........} \; 3.23$$

The Murphree operating line for an ideal plate is obtained using equation 3.21 :-

$$Y^{*}_n - Y_{n-1} = \frac{L_n}{V_n} (X^{*}_{n+1} - X_n) \qquad \text{.........} \; 3.24$$

which is a straight line passing through the point (X_n, Y_{n-1}) with a slope of L/V and is identical to the operating line of the real plate.

The Hausen operating line is obtained using equation 3.22

$$Y'_n - Y_{n-1} = \frac{L_n}{V_n} (X_{n+1} - X'_n) \qquad \text{..........} \; 3.25$$

This is a straight line passing through the points (X_{n+1}, Y'_n) and (X'_n, Y_{n-1}) with a slope of L/V. (see Fig 3.2). This is parallel to the actual operating line, but not the same. The liquid concentration leaving the plate X'_n with which Y'_n is in equilibrium, shifts from X_n to satisfy the material balance.

For steady- state conditions and for linear equilibrium and operating lines, a relationship between the Murphree and Hausen plate efficiencies can be derived from the definitions of these efficiencies (45).

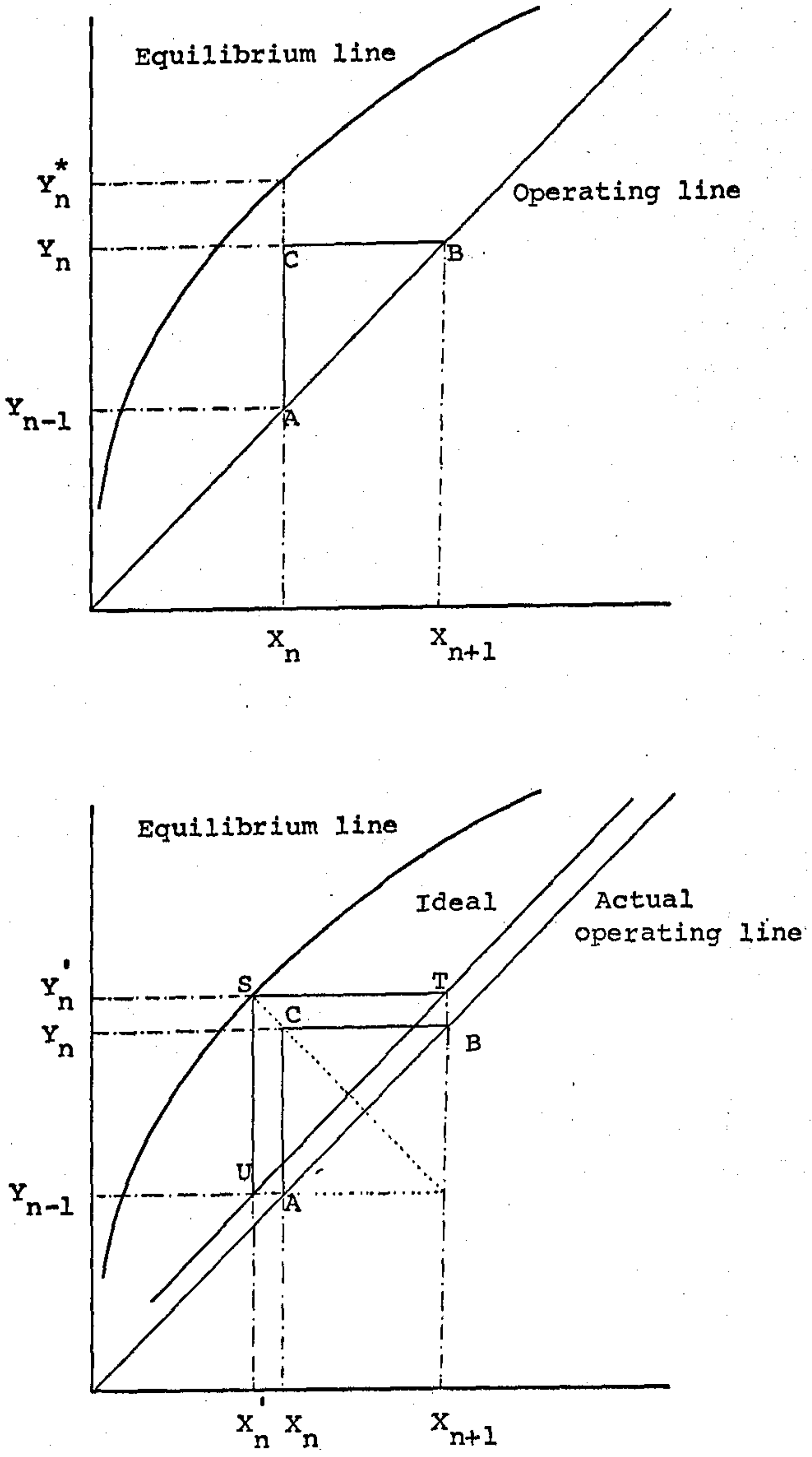


Fig 3.2 Graphic representation of the Murphree and Hausen plate efficiencies.

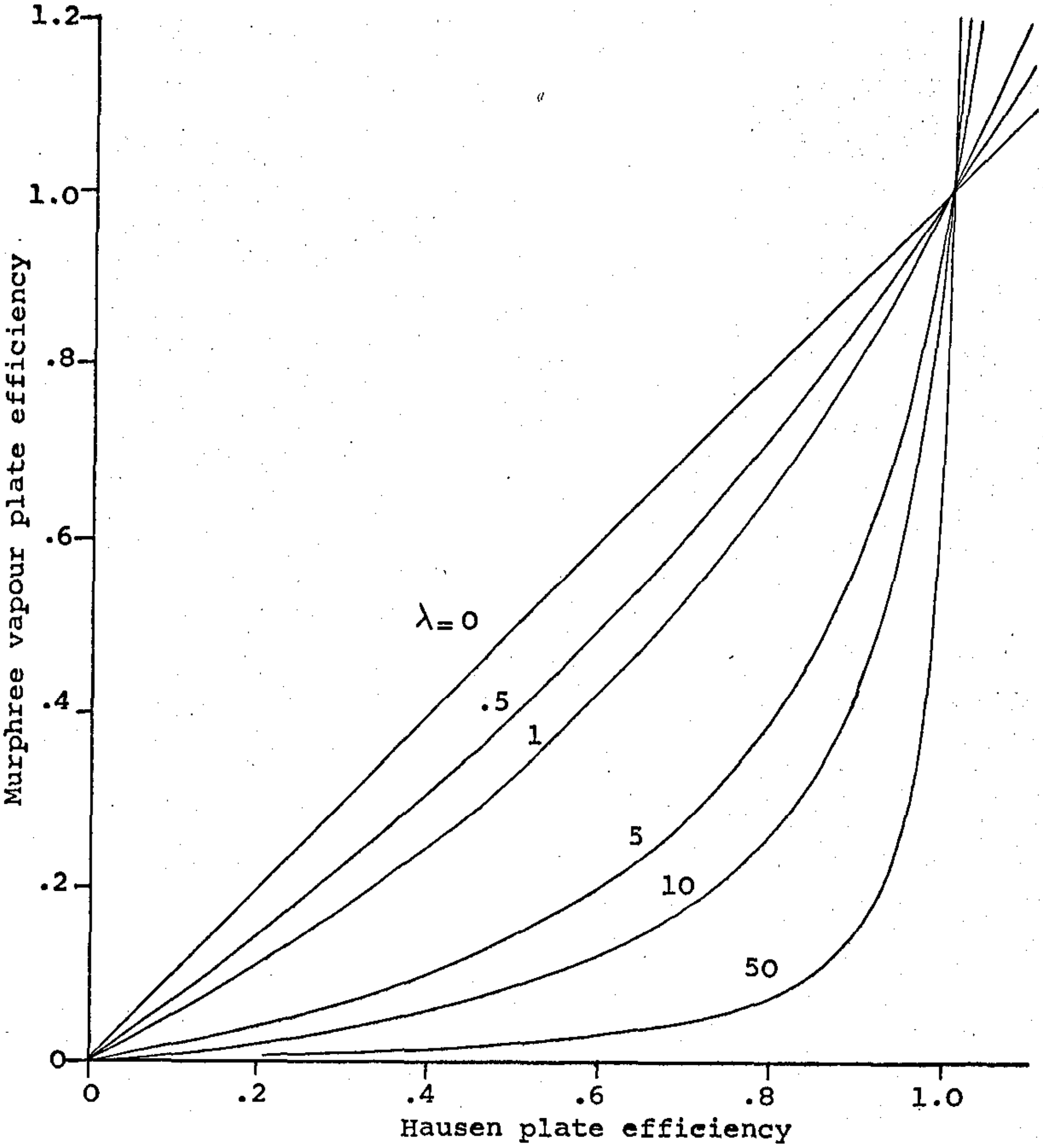


Fig 3.3 Comparison of the Murphree vapour plate efficiency with the Hausen plate efficiency.

$$E_{HG} = \frac{\lambda + 1}{\lambda + \frac{1}{E_{MV}}} \qquad = \frac{\lambda + 1}{1 + \frac{\lambda}{E_{ML}}} \qquad \text{........ 3.26}$$

A comparison of the Hausen and Murphree efficiency is shown in Fig 3.3 from which it can be seen that if the vapour phase is controlling and λ is small then the Murphree efficiency tends to that of the Hausen plate efficiency.

Similarly for large λ and liquid phase controlling, the Murphree liquid-phase efficiency tends to that of Hausen. The need to differentiate between the two phases is *eliminated if the Hausen or generalised efficiency is used.*

3.6 Vaporisation Efficiency

Vaporisation efficiency was initially defined by McAdams (70) in batch steam distillation for a system containing one volatile component. Carey (15) derived a relationship between the efficiency and the characteristics of the substance being distilled.

$$E = 1 - \exp\left(-\frac{KL_h}{D}\right) \qquad \ldots\ldots\ldots 3.27$$

(where L_h is the depth of the liquid through which the steam rises

D is the diameter of the steam bubbles

K is the constant characteristic of the volatile)

The vaporisation efficiency was extended to apply to multicomponent systems by Holland and Welch (46). Holland (47) defines an equilibrium vapour composition for a component i on plate n as:-

$$Y'_{n,i} = K_{n,i}X_{n,i} \qquad \ldots\ldots\ldots 3.28$$

(Y' *in the vaporisation efficiency is not the same as that used* by Standart, but has been used again to comply with the notation of Holland). The equilibrium constant K is calculated at the same temperature and pressure as that of the liquid leaving plate n with composition $X_{n,i}$. Assuming the liquid to leave at its bubble point, then the sum of vapour compositions must be unity and the vaporisation efficiency is defined as (47):-

$$E^{o}_{n,i} = \frac{Y_{n,i}}{Y'_{n,i}} \quad \text{.........} \; 3.29$$

where $Y_{n,i}$ is the average vapour leaving the tray.

Substituting in equation 3.28 gives:-

$$E^{o}_{n,i} K_{n,i} X_{n,i} = Y_{n,i} \quad \text{........} \; 3.30$$

For a multicomponent system of k components and for a set of known values of $E^{o}_{n,i}$ and $X_{n,i}$, the temperature of the plate is the positive value of T_n which makes $f_n=0$

where:- $$f_n = \sum_{i=1}^{k} E^{o}_{n,i} K_{n,i} X_{n,i} - 1 \quad \text{.......} \; 3.31$$

The definition of the multicomponent Murphree efficiency is:-

$$E_{Mn,i} = \frac{Y_{n,i} - Y_{n-1,i}}{Y^{*}_{n,i} - Y_{n-1,i}} \quad \text{.......} \; 3.32$$

(where $Y^{*}_{n,i} = K_{Mn,i} X_{n,i}$; $\sum Y^{*}_{n,i} = 1$; K_M is calculated at the bubble point temperature.)

Holland (48) defined a modified Murphree efficiency where he substituted $Y'_{n,i}$ for $Y^{*}_{n,i}$ giving:-

$$E^{M}_{n,i} = \frac{Y_{n,i} - Y_{n-1,i}}{Y'_{n,i} - Y_{n-1,i}} \quad \text{.........} \; 3.33$$

In a recent paper, Holland and McMahon (49) showed the relationship between these efficiencies and the vaporisation efficiency.

$$E^{o}_{n,i} = \frac{Y_{n-1,i}}{Y'_{n,i}} + E_{Mn,i}\left\{\frac{K_{Mn,i}}{K_{n,i}} - \frac{Y_{n-1,i}}{Y'_{n,i}}\right\} \qquad \text{........ } 3.34$$

$$E^{o}_{n,i} = \frac{Y_{n-1,i}}{Y'_{n,i}} + E^{M}_{n,i}\left\{1 - \frac{Y_{n-1,i}}{Y'_{n,i}}\right\} \qquad \text{........ } 3.35$$

They showed the vaporisation efficiency to be non-zero, finite and positive, provided the equilibrium constants are non-zero and positive for the component to which the efficiency applies. From examples taken in the paper (49), the Murphree efficiency values were in the range of $-\infty$ to $+\infty$ and the conclusion was drawn that the vaporisation efficiency was superior to that of the Murphree efficiency for multicomponent systems.

3.7 Factors Affecting Efficiency

3.7.1 General Discussion

Due to the large amount of work done on factors affecting efficiency, only a brief review will be carried out in this section. It can be seen from the definitions of various efficiencies, that they are dependent on a large number of system variables. Other variables not treated in this section which also affect efficiency can be found in the reviews given by Treybal (104) and Van Winkle (107).

In the case of physical properties, their effect on plate efficiency is mainly due to their effect on point efficiency. This is also true of changes in the vapour and reflux flows. (Table 3.1)

3.7.2 Physical Properties of the System

The physical properties that affect plate efficiency include relative volatility, liquid density, viscosity and surface tension. These parameters really affect the point efficiency which in turn affects the plate efficiency in the same way.

From the correlations derived by O'Connell (73), it can be seen that as the viscosity is increased and the other system parameters held constant, the plate efficiency decreases. Similarly as the relative volatility is increased, the plate efficiency decreases. This relationship shows that for multicomponent systems whose components have the same viscosity,the lighter phase has the lower plate efficiency and that all component plate efficiencies are different.

Investigations into the effect of surface tension on plate efficiency have been numerous. Defining a positive system as one where the less volatile component has the higher surface tension and a negative system has the lower surface tension, Zuiderweg (118) showed that the positive systems have higher plate efficiencies than those of neutral or negative systems. Sawistowski and Bainbridge (6) showed that this generalisation is untrue and that the reason it was arrived at was that Zuiderweg worked only in the foam regime.

The surface tension affects efficiency in two ways:-

1) the surface tension alters the interfacial area between the liquid and the vapour either by promoting foam formation in positive systems or by inhibiting foam formation in negative systems.

2) the surface tension/composition relationship may also affect surface renewal and liquid mixing and was investigated by Ellis and Biddulph (26,27).

Haselden (40) in his work, verified that one of the most important factors affecting plate efficiency was the magnitude of the surface tension gradients in the dispersion on the plate.

3.7.3 The Effect of Heat Loss

It is inevitable that all fractionating columns lose heat to the surroundings. If, therefore, the column is non-adiabatic, the heat loss must be considered in the enthalpy balance (40). It can be shown by using the method of Ponchon and Savarit (104), that adiabatic columns where all the heat is extracted in the condenser are more efficient than non-adiabatic columns

3.7.4 Thermal Distillation

Thermal distillation is the separation achieved when there is local superheating or direct condensation of vapour due to local subcooling. Much work has been carried out on thermal distillation (21,27,64,85).

Sawistowski and Smith (86) stated that the contribution of thermal distillation to the effective mass transfer coefficient was greatest when the temperature difference between bulk phases was greatest. Dankwerts (21) pointed out that the effect of thermal distillation was to increase the rate of attaining equilibrium. Both Dankwerts (21) and Liang (65) pointed out that although thermal distillation may be the cause of discrepancy between results on plate efficiency, the true reason can only be stated after investigating all other factors especially the physical properties.

3.7.5 Tray Design

The tray design variables that affect the plate efficiency are numerous and a review is given by Smith (93). The way in which they affect the efficiency is not given here but a list of the variables is given in Table 3.1.

3.7.6 The Effect of the Reflux Ratio

Ellis and Hardwick (28) showed that in the rectifying section plate efficiency decreased as reflux ratios fell below 0.4 and between 0.4 and 1 stayed constant. This was also verified by other investigations (14,41,43,84). Ellis and Hardwick (28) further showed that for the stripping section, as the reflux ratio increases, the plate efficiency falls off rapidly for an ordinary sieve plate column but remains constant for a constant interfacial pool column. From these investigations, it was concluded that for a reflux increase, the rectifying section decreased in efficiency due to the liquid film resistances increasing and that the decrease in the stripping section was due to decreasing interfacial area.

3.8 Discussion on Efficiency

The conclusion drawn by Holland (49) that the vaporisation efficiency was superior to that of the Murphree plate efficiency was disproved by Standart (96). The case taken was where one of the components of a ternary system was at a 'pinch point' and the vaporisation and Murphree plate efficiency were compared. Although the vaporisation efficiency has positive values, the Murphree plate efficiency can be estimated from realistic models of the detailed transfer mechanism on the tray.

The Murphree plate efficiency has been generally accepted in distillation design work and is easily used in simulating unsteady-state operations. If the liquid mixing characteristics are included in the simulation model, then the Murphree plate efficiency can be quite easily calculated while the vaporisation efficiency, which gives no more information and is less generally used, would require more data and would be less readily calculated.

Standart (95) showed that the Murphree plate efficiencies had limitations in that certain streams had to be saturated (i.e. also in thermal equilibrium). The Murphree vapour plate efficiency demands that the liquid leaving the nth plate with composition X_n be saturated in order that a vapour phase of composition Y_n^* can exist in equilibrium with it. The question of saturation has been ignored by Murphree for both liquid and vapour-phase efficiences. For a binary system if the two phases leaving the plate were in temperature equilibrium, they cannot be saturated or they would be in full equilibrium. However, there is no reason to suppose that the streams leaving a plate are saturated and in many practical cases they are not. From this point it would seem that if the generalised efficiency was used, it would be more accurate than that of the Murphree plate efficiency for systems whose streams leaving the plate were saturated, but less accurate if they were not saturated.

From this brief review on efficiency, it can be seen that the Murphree plate efficiency has the advantage of being easily calculated and has been generally accepted and widely used in column design. The generalised plate efficiency has been shown to be more accurate if the phases leaving the plate are saturated, but as most practical cases are not and as the generalised plate efficiency is less readily calculated, there seems no advantage in using the generalised plate efficiency instead of the Murphree plate efficiency for simulation purposes. Similarly, as the vaporisation efficiency has shown no advantage over the Murphree plate efficiency, it can be concluded that the Murphree plate efficiency is superior for simulation purposes as it is more readily estimated.

Table 3.1 Variables affecting efficiency

Sieve tray operating variables

Operating variables	1) Temperature (heat losses)
	2) Pressure
	3) Liquid flow rate; internal reflux; L/V
	4) Vapour flow rate
	5) Liquid diffusion coefficient
System variables	1) Density of liquid
	2) Viscosity of liquid
	3) Relative volatility of components
	4) Gas properties
	5) Surface tension
Plate design variables	1) Free hole area
	2) Hole diameter
	3) Hole pitch
	4) Hole arrangement
	5) Plate thickness
	6) Overflow weir height
	7) Inlet weir design
	8) Length of active tray and settling areas
	9) Splash baffles
	10) Flow patterns
Column design variables	1) Downcomer area
	2) Downcomer arrangement
	3) Plate spacing
	4) Flow arrangement (counter current or parallel flow)

4. LITERATURE SURVEY ON LIQUID MIXING ON DISTILLATION PLATES: MIXING EFFECTS.

4. Literature Survey on Liquid Mixing on Distillation Plates: Mixing Effects

4.1 The Importance of Liquid Mixing for Plate Efficiency

The existence of a concentration gradient in the liquid phase on a distillation tray was realised as early as 1936 by Lewis (64). He made the first successful attempt at relating the Murphree plate efficiency to the local point efficiency. If the liquid on the plate is perfectly mixed, then the Murphree plate efficiency is equal to the point efficiency. However, if the plate is large, then a concentration gradient exists due to partial liquid mixing and the plate efficiency is different from the point efficiency as only the liquid leaving the plate is considered when calculating the equilibrium vapour composition. From these observations, it can be seen that in order to simulate a distillation column, the liquid phase mixing must be considered and accurately modelled.

Since the first work of Lewis (64), many investigations have been carried out on liquid mixing on distillation plates and the following sections briefly review some of the general models derived.

4.2 Diffusion Model

The first known residence time distribution was done by Stewart (98) in 1894. He studied the passage of a salt tracer through blood vessels. In 1953 the concept was revised by Dankwerts (19) who emphasised its importance in chemical engineering. Zwietering (119), Spalding (94) and Dankwerts (20) presented generalised discussions on the residence times and distribution functions, whilst Bischoff and McCracken (10) showed their importance in describing flow patterns in chemical reactors.

Foss (29) and Gerster (36) used Dankwerts approach to investigate the liquid mixing characteristics. Foss (29) applied step changes to the inlet salt concentration and plotted salt concentration against time after initiating the step for various distances along the tray.

Making the assumptions:-

1) The only concentration gradient existing in the liquid is in the direction of flow and that the liquid is perfectly mixed in the vertical and horizontal directions perpendicular to the direction of flow.

2) The depth of liquid on the plate is constant along the plate.

3) That the outer regions of the plate, outside the rectangular section between inlet and outlet weirs are neglected.

it was found that the concentration responses could be accurately represented by use of the diffusion model which has been extensively reviewed by Levenspiel (61,62,63) and Crank (18).

$$De \frac{\partial^2 c}{\partial z^2} - u_1 \frac{\partial c}{\partial z} = \frac{\partial c}{\partial t} \qquad \text{.......... 4.1}$$

From a plot of experimental data, Foss derived an empirical equation for estimating the eddy diffusivity:-

$$\frac{2De}{u_1^3 h_c} = 0.0247 \left\{ \frac{h_c u_1}{h_f} \right\}^{-2.8} \qquad \text{.......... 4.2}$$

Similar expressions were derived by other workers (8, 37). Gilbert (37) varied the tracer input sinusoidally for both bubble-cap and sieve plates and the relationship derived for the estimation of the eddy diffusivity for both types of tray was:-

$$\frac{2De}{u_1^3 h_c} = 0.458 \left\{ \frac{h_c u_1}{h_f} \right\}^{-2.4} \qquad \text{.......... 4.3}$$

Barker and Self (8) worked on sieve trays with varying weir heights. The empirical equations derived for the eddy diffusivity were:-

1) for weir heights of 3, 2 and 4 inches:-

$$\frac{2De}{u_1^3 h_c} = 0.0098 \left\{ \frac{h_c u_1}{h_f} \right\}^{-2.91} \qquad \text{.......... 4.4}$$

2) for weir heights of 1 inch:-

$$\frac{2De}{u_1^3 h_c} = 0.0024 \left\{ \frac{h_c u_1}{h_f} \right\}^{-3.02} \qquad \text{......... 4.5}$$

Welch et al (110) used 27" diameter bubble-cap tray column for investigating the liquid mixing for various systems and flow rates. He found that the Peclet number defined as:-

$$Pe = \frac{u_l z}{De} \qquad \text{.........} \; 4.6$$

stayed constant for systems of different physical properties and for different flow rates and thus only varied with length of liquid flow path z. He derived a simple expression for the eddy diffusivity:-

$$De = 0.29u_l \qquad \text{.........} \; 4.7$$

The work generally used for estimating the eddy diffusivity is that presented in the A.I.Ch.E Manual (1). This work gives an equation for both bubble-cap and sieve trays:-

$$De = (0.0124 + 0.0171u_g + 0.0025L_g + 0.015W)^2 \quad ... \; 4.8$$

Thomas and Campbell (109) investigated liquid mixing on sieve trays with downcomers, using dye injections and measuring the colour intensity at the outlet weir. They stated that the diffusion model was a good representation of the liquid mixing on sieve trays.

Further discussions on the results and experimental procedure of these works are given in section (5.4) comparing them with the author's experimental results.

4.3 Pools-in -Series Model

Although the diffusion model has been generally accepted as an accurate representation of the liquid mixing model, it is awkward to use in numerical simulation and a simpler yet accurate model would be preferred.

The concept of modelling the liquid phase by a series of perfectly-mixed pools was first introduced by Kirschbaum (53,54). This type of model was later revised by Gautreaux and O'Connell (31) who stated that about every 1.25 ft of tray length could be represented by a perfectly mixed liquid pool. This type of model is illustrated in Fig 4.1a and the differential equation for the pool j on the nth plate for constant molal holdup is:-

$$\frac{dX_{n,j}}{dt} = L_n (X_{n,j-1} - X_{n,j}) + V_{n,j} (Y_{n-1,j} - Y_{n,j}) \quad .... \; 4.9$$

In these investigations, the vapour phase was assumed to be perfectly mixed between the plates.

Eduljee (25) tried to relate the number of pools to be used in series with the degree of mixing on the plate (i.e the Peclet number) and derived the empirical equation:-

$$Pe = (2.155 - 0.75 \log_{10} (\lambda E_{OG})) (M-1) \quad .... \; 4.10$$

where M is the number of mixed pools to be used. This equation is used to find the number of pools to be used in the equation derived by Gautreaux and O'Connell for a known

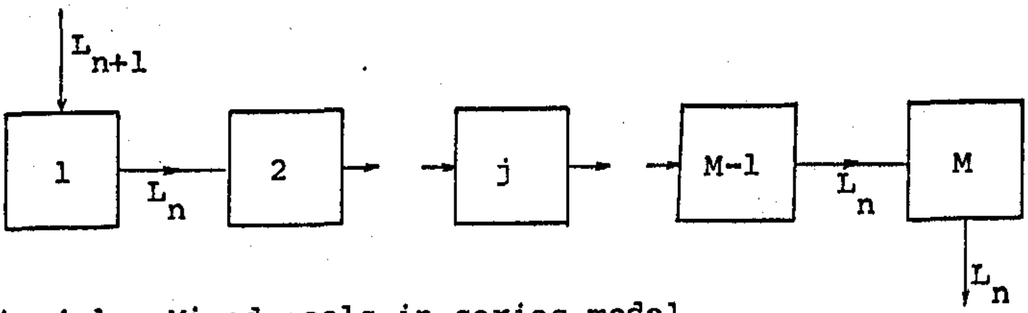


Fig 4.1a Mixed pools-in-series model.

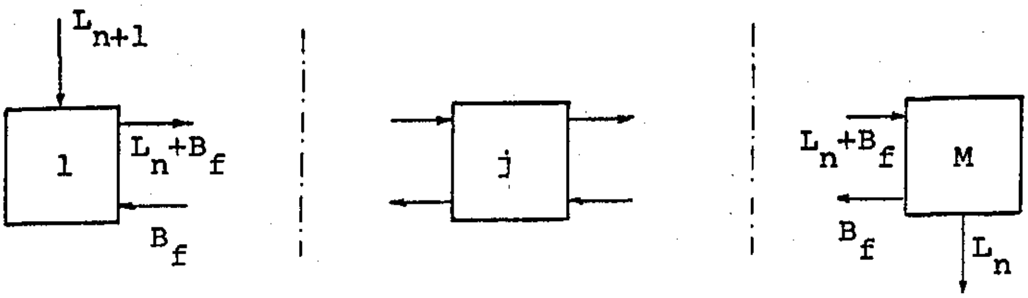


Fig 4.1b Mixed pools-in-series with backflow model.

Peclet number. However, there seems no real reason why the parameters λ and E_{OG} should be included in describing the mixing. This point also applies to the equation 4.14 given by Ashley and Haselden (5).

From the impulse response of a series of continuous stirred tank reactors (22), Ashley and Haselden (5) used gamma functions to derive a theoretical relationship for the dimensionless variance:-

$$\sigma^2 = \frac{1}{M} \qquad \text{........ 4.11}$$

The dimensionless variance for the impulse response to the diffusion model was derived by Foss (29), Crank (18) and Levenspiel (63) and when the theoretical relationship is reduced for large values of the Peclet number (above 6), the relationship becomes:-

$$\sigma^2 = \frac{2}{Pe} \qquad \text{...... 4.12}$$

Combining the equations 4.11 and 4.12 a relationship between the number of liquid pools and the Peclet number is derived:-

$$2M = Pe \qquad \text{...... 4.13}$$

Ashley and Haselden (5) pointed out that when the Peclet number is zero, the number of mixed pools must equal unity. From this they derived two analytical equations:-

$$\left.\begin{array}{ll} M = (Pe - 1)/2 & \text{for large } Pe \\ M = Pe/2 + 1 & \text{for } Pe > 2,\ \lambda E_{OG} < \tfrac{1}{2} \end{array}\right\} \quad \text{... 4.14}$$

These two equations converge for high values of Pe whilst the latter is intuitively more satisfactory for low values of Pe.

4.4 The Pools-in-Series with Backflow Model

The perfectly mixed pools-in-series model as described above does not allow for any backmixing between the pools. By introducing a recycle between these pools, the backmixing and splashing can be represented in the model (Fig 4.1b).

Steady-state operations in liquid-liquid extraction columns have been described using the recycle model by Miyauchi and Vermeulen (68) while Colburn (17) and Sleicher (92) used similar models with recycle to account for the effect of entrainment on plate efficiency in distillation and for extractor efficiency in mixer-settlers.

The backflow model was investigated by Retallick (79) who treated the model probabilistically in terms of the number of recycles a tracer particle undergoes.

Roemer and Durbin (80) give the transient response and a moments analysis of the backflow model, while Dickey and Durbin (23) analysed the backflow model of a distillation plate with linear interface mass transfer. The analysis was for a component disappearing at a rate proportional to its concentration and is analogous to first-order isothermal reactors as shown by Baldwin and Durbin (7).

Buffham (12) gives analytical expressions for the impulse response for an infinite series of stirred tanks with backflow when the impulse is introduced at the first tank (unilateral), and when there is an infinite number of stirred tanks before the injection point as well as after (bilateral). When a disturbance is introduced in the feed to a distillation column, operating with a low relative volatility, the early part of the response is accurately described by the useful approximations given.

4.5 The Comparison of Mixing Models and the Estimation of Efficiency

It was quoted in section 4.3 that by comparing variances between the diffusion model and the pools-in-series model a relationship between the Peclet number and the number of pools to be used could be derived. Using the same procedure as van der Laan (74,106), the variance for the diffusion model in its full form is:-

$$\sigma^2 = 2\left\{ \frac{1}{Pe} - \frac{1}{Pe^2}\left(1 - e^{-Pe}\right)\right\} \qquad \text{........ 4.15}$$

which for large values of Pe can be reduced to the equation given by Foss (equation 4.12).

Retallick (79) derived the residence time distribution for the backflow model probabilistically, but also incorporated an error by ignoring the different characteristics of the two end pools. Residence time distributions in a cascade of mixed vessels with backflow were derived by Klinkenberg (55). The relative variance of the backflow model was derived by van der Laan (74) for the same boundary conditions and is given as:-

$$\sigma^2 = \frac{1+2B}{M} - \frac{2B(1+B)}{M^2}\left\{ 1 - \left(\frac{B}{1+B}\right)^M \right\} \qquad \text{........ 4.16}$$

where $B = B_f / L$ 4.17

From the relationship given by equation 4.13, for a large number of pools equation 4.16 can be reduced to:-

$$\sigma^2 = \frac{1 + 2B}{M} \qquad \text{.........} \; 4.18$$

The boundary conditions for all the above equations are the same.

Miyauchi and Vermeulen (68) derived the relative variance of the backflow model from the material balance on each stage of a liquid extration column:-

$$\sigma^2 = \frac{1}{M} - 2\left[\frac{B}{M} + \left[\left(\frac{1}{2M} + \frac{B}{M}\right)^2 - \frac{1}{M^2}\right]\left[1 - \exp\left[\frac{-(\frac{1}{2}+B)\ln(1+1/B)}{1/(2M) + B/M}\right]\right]\right] \qquad \text{.........} \; 4.19$$

Comparing this variance with that of the reduced diffusion model (equation 4.12) gives:-

$$B = \frac{M - \frac{1}{2}}{Pe} - \frac{M(M - \frac{1}{2})}{(M - 1)(2M + 1)} \qquad \text{.........} \; 4.20$$

The more common relationship (23,68,79) can be obtained by substituting the variance of equation 4.18 in the reduced diffusion model variance equation 4.12 to give:-

$$B = \frac{M}{Pe} - \frac{1}{2} \qquad \text{........} \; 4.21$$

Kramers and Alberda (56) presented the analytical equation:-

$$B = \frac{M - 1}{Pe} - \frac{1}{2} \qquad \text{for small } M \qquad \text{.........} \; 4.22$$

Thus by comparing the variances of the models, relationships between the mixing parameters can be found. In the case of the backflow model two of the three parameters must be known. For work using bubble-cap trays, Holland (48) suggested that the number of pools in series to be used could for convenience be equal to the number of bubble-cap rows and the backflow worked out later.

The estimation of the plate efficiency for these various models can now be made for steady-state conditions. It has been mentioned before (see section 3.4) that if the liquid on the plate is perfectly mixed then the Murphree plate efficiency is equal to the point efficiency.

$$E_{MV} = E_{OG} \qquad \text{......... } 4.23$$

and if the liquid travels in plug flow across the plate:-

$$E_{MV} = \frac{1}{\lambda}\left(e^{\lambda E_{OG}} - 1\right) \qquad \text{......... } 4.24$$

The diffusion model as proposed by Gerster et al (1,29,36) gives an accurate representation for the case with complete vapour mixing and partial liquid mixing, but the relationship for the estimation of efficiency is complex:-

$$\frac{E_{MV}}{E_{OG}} = \frac{1 - e^{-(n+Pe)}}{(n+Pe)\left(1+\frac{n+Pe}{n}\right)} + \frac{e^{n} - 1}{n\left(1+\frac{n}{n+Pe}\right)} \qquad \text{...... } 4.25$$

where $n = \frac{Pe}{2}\left\{ 1 + \frac{4\lambda E_{OG}}{Pe} - 1 \right\}$

The mixed pools-in-series model as described by Gautreaux and O'Connell (31) gives a much simpler relationship for the Murphree plate efficiency:-

$$E_{MV} = \frac{\left(1 + \frac{\lambda E_{OG}}{M}\right)^{M} - 1}{\lambda} \qquad \text{........ } 4.26$$

Table 4.1 shows the comparison of the predicted Murphree plate efficiency using the equation derived from the diffusion model (4.25) and the pools-in-series model (4.26) for various fixed point efficiencies and absorption

Table 4.1 The comparison of estimated plate efficiencies

λ	E_{OG}	M	E_{MV} % from pools	E_{MV} % calculated from diffusion model			
				Pe=2M	Pe=2M-1	Pe=2M-2	Eduljee
0.5	0.5	2	53.12	54.04	53.50	52.76	53.24
		5	55.26	55.43	55.30	55.14	55.47
		9	55.93	55.99	55.94	55.89	56.09
		14	56.24	56.26	56.24	56.22	56.35
		20	56.41	56.42	56.41	56.40	56.49
		27	56.51	56.52	56.51	56.51	56.57
		35	56.58	56.58	56.58	56.57	56.63
0.5	0.8	2	88.00	90.54	89.12	87.15	88.13
		5	93.87	94.38	94.02	93.59	94.31
		9	95.80	95.95	95.84	95.70	96.15
		14	96.70	96.77	96.71	96.65	96.95
		20	97.19	97.23	97.20	97.17	97.38
		27	97.49	97.51	97.50	97.48	97.64
		35	97.69	97.70	97.69	97.68	97.81
1.5	0.5	2	59.37	62.99	61.14	58.63	59.33
		5	67.42	68.28	67.76	67.15	67.76
		9	70.35	70.65	70.46	70.25	70.65
		14	71.76	71.89	71.80	71.71	71.99
		20	72.54	72.89	72.57	72.52	72.72
		27	73.03	73.07	73.04	73.02	73.17
		35	73.35	73.38	73.36	73.35	73.46
1.5	0.8	2	104.00	115.46	110.04	102.96	103.73
		5	128.78	132.09	130.36	128.36	129.16
		9	138.98	140.26	139.58	138.83	139.41
		14	144.16	144.75	144.43	144.09	144.51
		20	147.14	147.45	147.28	147.11	147.42
		27	149.01	149.19	149.09	148.99	149.24
		35	150.27	150.37	150.32	150.25	150.45

factors. The number of pools is varied and the Peclet number used in the diffusion model equation is estimated using equations 4.10,4.13 and 4.14.

4.6 *The Effect of Liquid Mixing on Plate Efficiency*

Lewis (64) in his classical paper considered three limiting cases:-

1) Liquid unmixed on the plate with vapour perfectly mixed between the plates

2) Both the liquid and the vapour unmixed and the liquid flowing in the same direction on each plate (parallel flow)

3) Both the liquid and the vapour unmixed and counter-current liquid flow on adjacent plates.

He showed that a greater overall efficiency was achieved with parallel liquid flow on adjacent plates (case 2) than that of counter-current liquid flow (case 3). However, little use is made of this fact because a high point efficiency and a low liquid to vapour ratio is needed before the difference becomes significant and at these conditions the extent of the mixing in the main phases is impossible to calculate. Ashley and Haselden (5) also showed that parallel flow gave a higher efficiency than counter-current flow while Shore (89,90) obtained experimental values of the plate efficiency for both parallel, and counter-current flow of liquid using a three-plate column and accounted for the difference between parallel and counter-current flow in terms of the system properties.

By substituting constant values of λ and E_{OG} in equation 4.26 and varying the number of mixed pools, it can be shown that as the degree of mixing decreases, so the Murphree plate efficiency increases.

4.7 The Effect of Vapour Mixing on Plate Efficiency

In all the models and equations discussed so far, the assumption of perfect vapour mixing between plates has always been made. If the liquid on the tray is assumed to be perfectly mixed, it does not matter how the vapour flow is described.

A relationship between the point and Murphree plate efficiency for partial liquid mixing using the diffusion model and no vapour mixing was derived by Diener (24). Both counter-current and parallel flow (Lewis case 3 and 2) of the liquid on successive plates was considered. Assuming a point efficiency of 0.8, he calculated the Murphree plate efficiency for the case where the vapour phase was unmixed and for where it was perfectly mixed. Dividing the Murphree plate efficiency calculated for unmixed vapour by the efficiency calculated for perfectly mixed vapour gave a ratio. This ratio was shown to be greater than unity for the case where the liquid was in parallel flow and it increased with an increase in λ. When the liquid was in counter-current flow, this ratio decreased with an increase in λ. Diener concluded that most systems have point efficiencies less than 0.8 and that the effect of vapour mixing on plate efficiency was of little importance.

Due to the complexity of Diener's equations, it is doubtful whether his method can be exploited. A simpler approach was illustrated by Ashley and Haselden (5). They used a series of perfectly mixed pools to describe the liquid mixing and the vapour leaving these pools joined vapour leaving adjacent liquid pools to form a vapour cell, see Fig 4.2. The number of vapour cells to be used to describe the mixing was arbitrary as not enough data existed for vapour mixing characteristics.

They further showed that for $\lambda = 1$, a point efficiency of 0.8 and counter-current flow of liquid, that the plate efficiency increases with decrease in the number of vapour cells but this decrease was so small that the only significant difference can be seen between using one and two vapour cells. For parallel liquid flow the plate efficiency increased with increase in vapour cells and again, as shown in Fig 4.3, the only major increase was seen using one and two vapour cells. Fig 4.3 shows the effect of liquid and vapour mixing on the Murphree plate efficiency. For a point efficiency of 0.8 and $\lambda = 1$, when the number of liquid pools and vapour cells are both infinity, the plate efficiency for parallel liquid flow is 133% and that

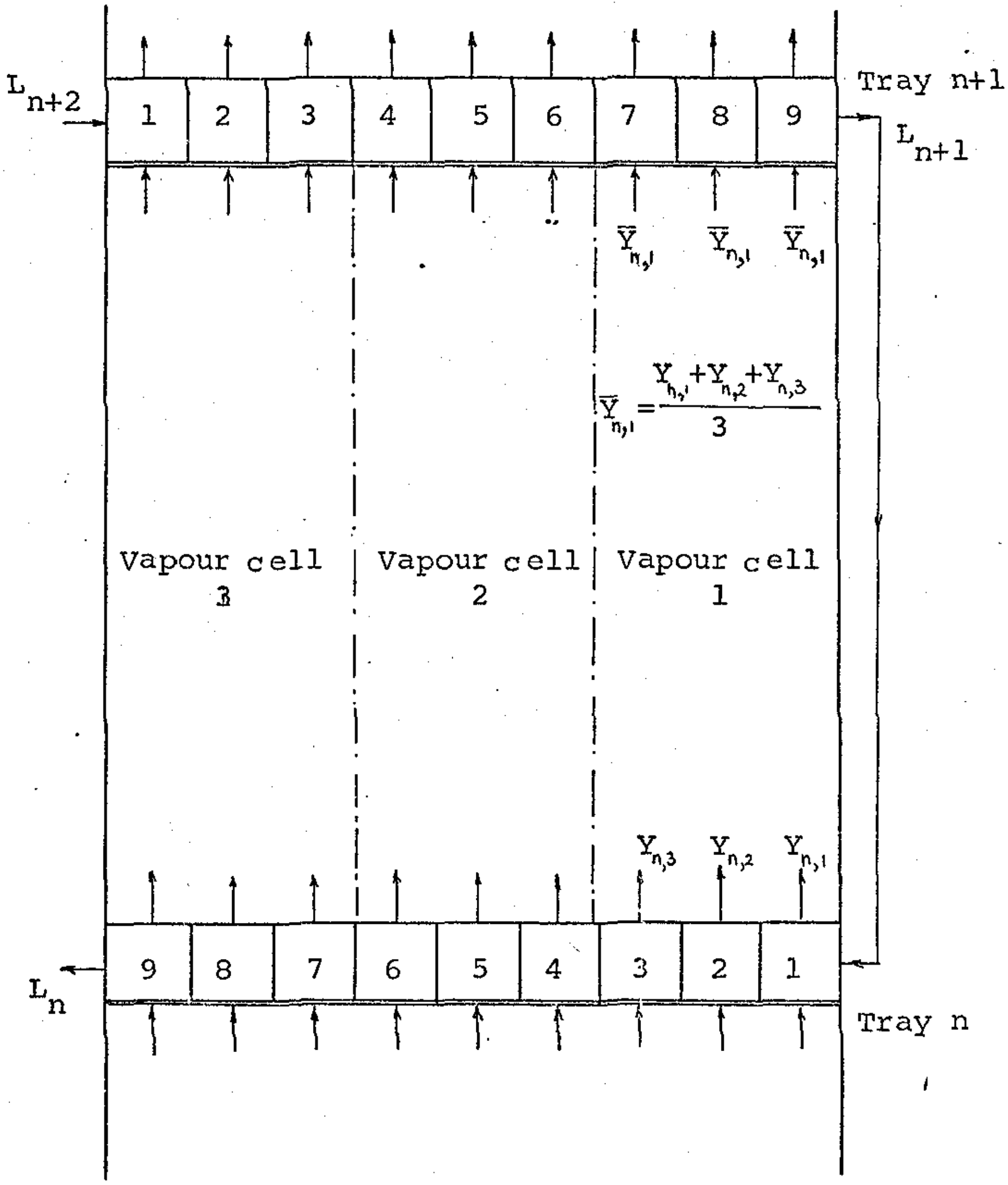


Fig 4.2 Vapour cell model using 3 liquid pools to 1 vapour cell.

for counter-current liquid flow is 116%. Although the difference in plate efficiency values for vapour phase unmixed and mixed, varies at the most by 5% for the liquid in counter-current flow, this is only true for steady-state conditions and the difference may be more important in unsteady-state conditions.

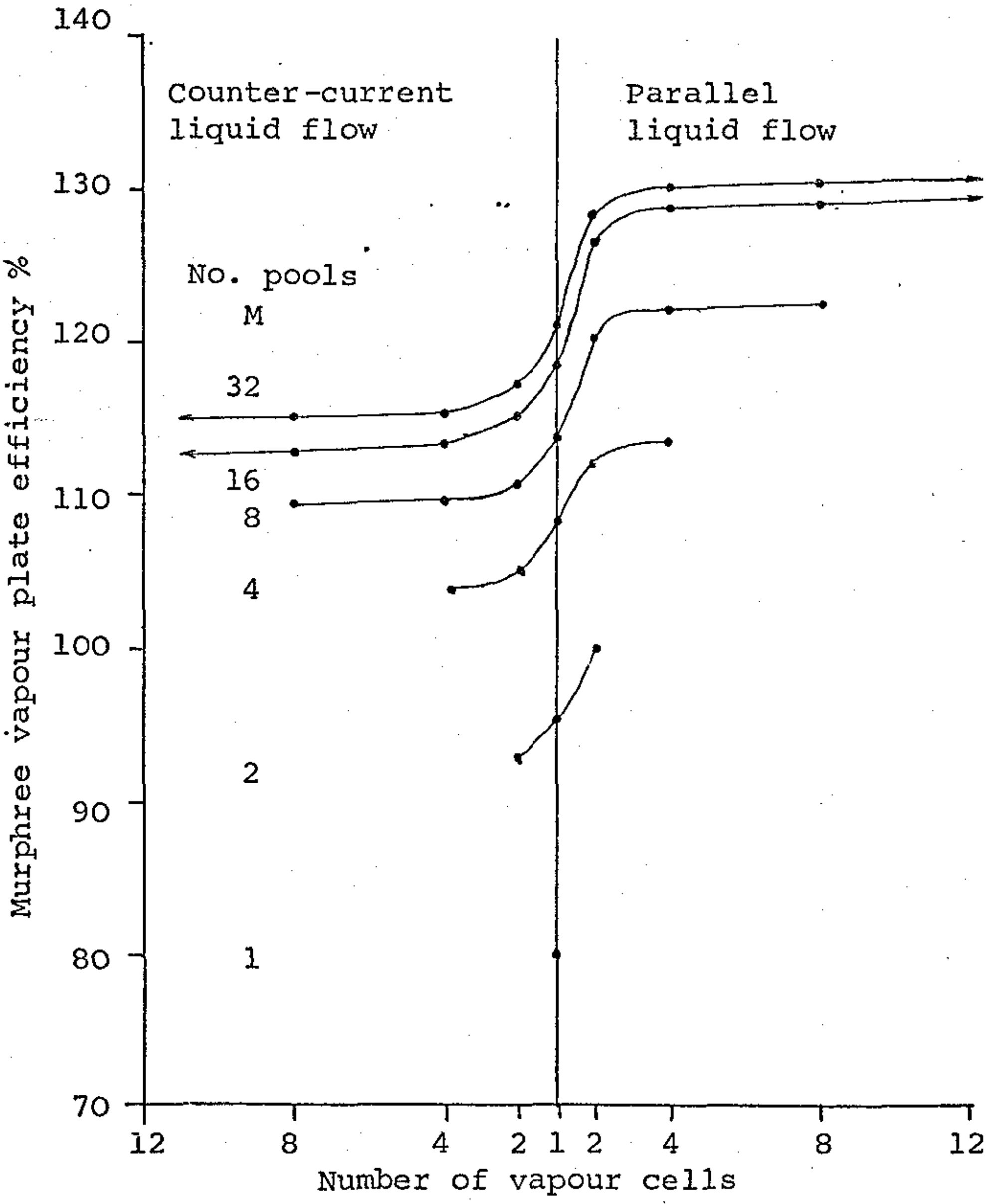


Fig 4.3 The effect of liquid and vapour mixing on the Murphree vapour plate efficiency.

5. EXPERIMENTAL INVESTIGATION OF MIXING ON SIEVE PLATES: LIQUID MIXING.

5. Experimental Investigation of Mixing on Sieve Plates: Liquid Mixing.

5.1 Experimental Apparatus

To investigate the liquid and vapour mixing taking place in a large plate column, a long rectangular column with three plates and segmental downcomers was constructed. The plates represented the centre section of a large circular plate. The top section of the column was made of half-inch thick clear perspex sheets, giving an open view of the top two plates. Marine-plywood was used for the bottom section, the plates and the inlet and outlet downcomers. All joints between the sections were sealed with strip rubber. The system air-water was used to represent the vapour and liquid streams. Fig. 5.1 shows the design of the column and the variable sections.

The sieve plates, weirs and downcomer sections were constructed such that they could easily be replaced with different sections or, in the case of the inlet downcomer on the observation tray, could be reduced in sectional area. The investigation was carried out using two different hole diameters and free areas for the plate and different weir heights.

The active tray length of 6 ft. was the distance between the inlet and outlet weirs and the actual active part depended on the number of holes and the pitch. At the outlet and inlet weirs, there was about two inches of unperforated space to act as a smoothing section.

The centre tray was the observation tray, the other two trays were present to ensure equal vapour and liquid distribution. A list of the column and tray dimensions used in the investigation is given in Table 5.1.

Table 5.1 Column dimensions

Column length	8 ft.	
Column width	1.5 ft.	
Tray length	7 ft.	
Active tray length	6 ft.	
Outlet downcomer length	1 ft.	
Inlet downcomer length	1" to 1 ft.	(variable)
Weir height	1.5" and 2.35"	(variable)
Tray thickness	.14"	
Tray spacing	1.5 ft.	

	Trays	
	1	2
Hole diameter	.375"	.1875"
No. holes	1172	4819
Free area	7.5%	7.7%
Triangular pitch	1"	.5"
Apron clearance	1"	1"

The downcomer length on the observation tray was variable and was the distance between the end blanking section and the downcomer outlet (see Fig. 5.1)

5.2 Experimental Procedure

5.2.1 Experimental Work

Many investigations into liquid mixing have been carried out (1,8,9,29,37,66,77,100,106), not only for calculating the eddy diffusivity, but more recently to investigate channelling at the outer regions of the column (i.e. near the perimeter wall). The main experimental work carried out in this investigation was concerned with the determination of the eddy diffusivity from impulse responses for the whole tray (active area and downcomer) and not just the active area as in previous investigations. Impulse response data at intermediate positions on the plate was compared with that predicted by the theoretical model to ensure that the model was suitable for describing the whole tray.

The impulse response experiments started by operating the column for at least thirty minutes to allow steady-state conditions to be reached. At the start of an actual experimental run, the conductivity of the water at the sampling positions was continuously read in to a computer at a sample interval of 0.2 seconds for a period of 15 seconds. These values were stored after being smoothed by a second order digital filter with small time constants of 0.08 to give a mean base-line for the dynamic response. The digital filter was confined to the calculation of the base line.

A small injection of potassium chloride was then introduced into the liquid stream between the outlet weir

Computer Specification and Use:-

The computer used for data logging was a Digital PDP11. The data points recorded were first read on to the disk and at the end of a batch of experiments, these results were then transfered to magnetic tape for storage. As the PDP11 was only a small capacity machine with no graph plotting facilities, the results were then transfered from the magnetic tape to paper tape for processing on a larger computer (an ICL 1900) with graph plotting facilities.

Conductivity Cell:-

The conductivity cell used was a 2 ml. capacity cylindrical cell. The conductivity of the sampled stream was continuously measured and had a small residence time in the cell, thus no significant lag occurred.

of the top plate and the holdup in the downcomer of the observation plate. The injection was evenly distributed across the tray to ensure that a constant concentration existed across the plate perpendicular to the liquid flow.

The conductivity at the outlet weir was continuously measured for at least four times the mean residence time of the liquid on the tray. The data-logging program was written so as to sample every second for the first two thirds of the run time, then to sample every three seconds for the tail end of the response. The data points and the sample time were stored and the same experiment was repeated for at least ten experimental runs. Intermediate points along the tray were also investigated for the same operating conditions and the conductivity and sample times were logged.

The dimensions of the weir height, downcomer length, hole diameter and the range of operating conditions over which the conductivity experiments were performed are given in Table 5.2.

From the steady-state theory given by Kropholler and Buffham (13),(Appendix B) further experimentation was carried out to illustrate the presence of backmixing and to compare the results between the impulse response and steady-state experimental methods of determining the eddy diffusion coefficient.

Table 5.2 Operating conditions for experimental runs on liquid phase.

Experiment	Hole diameter (inches)	Weir height (inches)	Downcomer length (inches)	Operating ranges		
				u_g ft/sec	Lg gal/min ft	Pe
1	.1875	2.25	6	3.7 to 4.4	4.2 to 17.5	5 to 13
2	.1875	1.5	6	3.9 to 4.3	3.5 to 11	4.5 to 14
3	.1875	1.5	3,6,9,12	3.9 to 4.1	4.2 to 9.2	4 to 13
4	.375	1.5	3,6,9,12	2.5 to 4	6 to 11.5	5.5 to 15

The experimental rig was operated and potassium chloride was continuously injected across the plate, (perpendicular to the liquid flow) two feet from the exit weir. After thirty minutes running steady-state was reached and samples of the liquid phase towards the inlet weir were taken and the conductivity measured. The concentration and sample distances from the injection point were stored and later plotted. Readings of the liquid phase, towards the exit weir were also taken to check for uniform concentration. The experimental results obtained showed this to be true. These steady-state experiments were carried out using the large hole diameter and a weir height of 1.5".

5.2.2 Tests to Validate Assumptions

A few tests were carried out to test the validity of the assumption of a constant concentration profile across the plate perpendicular to the liquid flow. This was done by taking a long probe and moving it quickly across the tray and continuously measuring the conductivity. It was found that the assumption was true if the injection was evenly distributed across the tray.

Further tests were also carried out to ensure that there was no channelling in the vertical direction (i.e. concentration constant in the vertical direction). By moving the sample probe up and down in the froth and measuring the conductivity continuously, the assumption of perfect vertical mixing was shown to hold.

Visual tests using nigrosine were also carried out to show the presence of backmixing in the liquid phase. Fig. 5.2 shows two photographs of the observation tray after an impulse dye injection has been introduced. The photograph 'A' was taken 10 seconds after the injection and photograph 'B' was taken 20 seconds after the injection. The weir height for these visual runs was 1.5". It can be seen from these photographs that the liquid is not perfectly mixed and also that there is backmixing. The visual tests also showed the validity of the previously made assumptions of perfect mixing in the vertical and horizontal directions.

5.3 Response Analysis

For each operating condition there was a set of ten experimental responses. Each one of these responses in the set was examined. The mean time and variances were calculated to see if the set could be combined to give one statistically significant, smooth response curve for that set. From the examination, it was found that the responses could be operated on to produce one response curve.

Each response in the set had been subjected to exactly the same disturbance and sampled at the same times and thus for each individual sample time in the set there were ten data points. A root mean squared value of these ten points was taken to give a modal point for that sample time. The response curve was then normalised to give an area of unity under the curve. This normalised curve was then compared with that of the diffusion model operating under the same flow conditions. The eddy diffusion coefficient used in the diffusion model was operated on until the coefficient giving the closest fitting response was found. The limit for finding the closest response was set by minimising the actual area between the two response curves. The time range over which this minimisation operated was 0 to 2 on the dimensionless time scale. The dimensionless time was the true time divided by the mean residence time of the liquid on the tray and the dimensionless concentration was the actual concentration divided by the average concentration had the injection been evenly distributed over the whole tray.

The conductivity experiments were carried out for varying weir heights, hole dimensions, hole pitch, and downcomer lengths. The comparison of experimental data with the diffusion model can be seen in Fig. 5.3 to 5.5.

The height of clear liquid, froth height and froth velocity were all calculated using the empirical equations given by Gerster (1) and Treybal (104) and are reproduced in Appendix A. From the calculated eddy diffusion coefficients and the computed heights and velocities, the parameter $\frac{2\ De}{u_1^3 h_c}$ was plotted against $\frac{h_c u_1}{h_f}$ on log-log paper and a least squares fit was used to find the best linear correlation through the points. From this a relationship for the eddy diffusion coefficient was derived and is given in the section comparing other work.

From the experiments carried out where the downcomer lengths were varied, and dye injections were introduced, it was seen that the downcomer segment when extended to its fullest length (i.e. 1 ft.), appeared to have some inactive regions which were reduced as the downcomer length was shortened. Further, there seemed little visual difference between the responses for the four different sized downcomers, but numerically the difference was quite significant. The Peclet number increased as the downcomer segment decreased. The downcomer length, Peclet number and operating conditions are given in Table 5.3.

From this table it can be seen that as the downcomer decreases in length, so the liquid mixing for the whole tray becomes less. It would seem from this, that a segmental downcomer tends to increase the liquid mixing characteristics. The response curves of Runs 4 and 6 for the four different sized downcomers are shown in Fig. 5.6a, b.

Table 5.3 Comparison of predicted Peclet numbers for different length downcomers.

Run	Weir height (inches)	Peclet numbers				Liquid flow (gal/min ft.)
		Downcomer lengths				
		3"	6"	9"	1'	
1	1.5	9.897	8.274	6.99	5.923	6.187
2	1.5	12.15	10.61	8.5	7.27	7.631
3	1.5	12.45	10.21	9.09	9.073	9.167
4	1.5	11.26	11.75	8.676	7.798	9.9
5	1.5	14.29	13.53	10.12	7.34	11.293
6	2.25	10.57	9.52	8.661	8.088	11.073

The gas flow range was u_g = 2.5 to 4 ft/sec, each run being different

From the steady-state experiments, a plot of concentration against distance from the injection axis was drawn and by using an optimisation technique (72) for minimising the least squares fit between the experimental data and the values computed from the equation (Appendix B):-

$$De = u_1 z \,/\, \ln \frac{c}{c_{out}} \qquad \ldots\ldots\ 5.1$$

the best single value of the eddy diffusivity can be calculated. Fig. 5.7 shows the results of two of the steady-state experiments and Table 5.4 compares some of the experimental steady-state values of the eddy diffusivity with those predicted by the empirical equation derived by the impulse experimental values. The values of the eddy diffusion coefficient from the steady-state experimentation were higher than those predicted by the dynamic experiments except for the very small values of Run 4 and 5 in Table 5.4.

From experience in fitting curves to the experimental points in the steady-state experimentations, it was found that only a slight deviation in the experimental points was needed to cause a significant shift in the curve and alter the value of the eddy diffusion coefficient predicted. This illustrates the sensitivity of the steady-state experimentation as the dynamic experiments using varying downcomer lengths showed that the absence of a downcomer would give less mixing and hence a smaller eddy diffusion coefficient. The absence of a downcomer makes little difference when the mixing tends to 'plug flow'. The Peclet numbers associated with Run 4 and 5 of Table 5.4 are:-

4) Pe = 11.32 5) Pe = 19.6 which are intermediate cases of liquid mixing.

5.4 Comparison with Other Work

The investigation into liquid mixing on sieve trays done by Barker and Self (8) was carried out using trays 13½" wide by 5 ft. 8½" long with 3/16" diameter holes based on ¾" triangular centres. Both dye and salt tracers were used. The salt injections were of the steady-state type of experiments where the injection was continuous and the amount of backmixing upstream was measured. Although the theory was the same as that described in Appendix B, the author's work indicates that this method of experimentation gives rise to more scattered results than does dynamic experimentation. Further the work done by Barker and Self did not include downcomers. On correlating their results, they derived two equations for the determination of the eddy diffusion coefficient.

$$\frac{2De}{u_1^3 h_c} = 0.0098 \left\{ \frac{h_c u_1}{h_f} \right\}^{-2.91} \qquad \text{2, 3 and 4 inch weirs} \qquad \ldots\ldots\ 5.2$$

$$\frac{2De}{u_1^3 h_c} = 0.0024 \left\{ \frac{h_c u_1}{h_f} \right\}^{-3.02} \qquad \text{1 inch weir} \qquad \ldots\ldots\ 5.3$$

Gilbert (37) investigated the liquid mixing on both bubble-cap and sieve trays using frequency response methods. The plates were 1 ft. wide with an active area length of 3 ft. perforated by 3/16" diameter holes in a ¼" equilateral pitch. The tracer, sodium thiosulphate, was varied in a sinusoidal manner. The equation derived for determining the eddy diffusivity from the bubble cap tray experiments was also a good estimate for that of the sieve trays.

$$\frac{2De}{u_1^3 h_c} = 0.458 \left\{ \frac{h_c u_1}{h_f} \right\}^{-2.4} \qquad \ldots\ldots\ 5.4$$

The main work done on eddy diffusion is that of the A.I.Ch.E. Manual (1) and further details of the work given in the manual are based on the work carried out at Delaware. The liquid flow Lg ranged from 6 to 30 gal/min ft. and the liquid flow length was 2.6 ft. The column used consisted of bubble-cap trays with split flow. It would seem that the flow rates used were rather large for the actual liquid flow path and thus only gave short residence times. The equation derived for the eddy diffusivity and the heights of clear liquid and froth are all given in detail.

$$De = (0.0124 + 0.0171\ u_g + 0.0025\ Lg + 0.0015W)^2 \quad ...5.5$$

Welch, Durbin and Holland (110) used a 27 inch diameter bubble-cap plated column with segmental downcomers to investigate the liquid mixing. A salt tracer was injected in the form of an impulse input and the conductivity was measured in the downcomers at the end of the first and second plates. Within the limits of the investigation ($4.4 \leqslant Pe \leqslant 27$), it was found that the ratio of Peclet number over liquid path length was independent of liquid and vapour flow as well as the physical properties (density, viscosity and surface tension). From this work they derived a simple equation for determining the eddy diffusion coefficient.

$$De = 0.29\ u_l \quad .... 5.6$$

Foss (29) used a rectangular column 9.5 inches wide by 3 ft. active length. There was a preaeration section which was partitioned off from the active area to stop backmixing. Various trays were used all with 3/16" diameter holes but with 7/8", 5/8", 17/32" triangular pitch. A unit-step salt injection was introduced at the inlet, thus directly giving the cumulative residence time distribution which was then differentiated with respect to time to give the response of a unit impulse. From a plot of his data, Foss derived the empirical equation:-

$$\frac{2De}{u_1^3 h_c} = 0.0247 \left\{\frac{h_c u_1}{h_f}\right\}^{-2.8} \qquad \text{.......} \quad 5.7$$

From a least squares fit of the data points obtained in the author's work, the following empirical relationships were derived:-

1) 1.5" weir, ⅜" holes giving 7.5% free area and a downcomer length of 6"

$$\frac{2De}{u_1^3 h_c} = 0.088 \left\{\frac{h_c u_1}{h_f}\right\}^{-3.06} \qquad \text{.......} \quad 5.8$$

2) 2.25" weir, 3/16" holes giving 7.7% free area with a downcomer length of 6"

$$\frac{2De}{u_1^3 h_c} = 0.1289 \left\{\frac{h_c u_1}{h_f}\right\}^{-2.56} \qquad \text{.......} \quad 5.9$$

3) 1.5" weir, 3/16" holes giving 7.7% free area with all downcomer lengths

$$\frac{2De}{u_1^3 h_c} = 0.137 \left\{\frac{h_c u_1}{h_f}\right\}^{-2.68} \qquad \text{.......} \quad 5.10$$

A plot of $2De/u_1^3 h_c$ against $h_c u_1/h_f$ for all the investigations mentioned above can be seen in Fig. 5.8a,b. The difference in the value of De predicted by the above relationships for the same values of u_1,h_c and h_f is in some instances quite large.

If the computed values of De are then used to calculate a Peclet number for the mixing parameter of a distillation plate, it can be seen from Table 5.5 that for reasonable working conditions, large Peclet numbers occur using some of the empirical equations. If large Peclet numbers, (above 40) are predicted, then the model can easily be represented by a plug flow model with no mixing.

With the exception of Barker and Self and the work done by the author, tray lengths of 3 ft. maximum were used. It was stated (1) that liquid paths greater than 3 ft. would tend to that of plug flow. However, this has been shown to be untrue. The difference in the works might be due to large liquid flow rates being used over small liquid paths which would give the appearance of plug flow. Further only the work of Welch and the author have included downcomers which tends to give better mixing characteristics and thus their work would give lower values of Peclet number.

From the wide range of results derived, it would seem that there are more factors that effect the liquid mixing on a plate than those incorporated in the parameters u_l, h_c, h_f. One of these factors appears to be the hole diameter and the triangular pitch used. Downcomers have been ignored by other investigators except that of Welch, and thus they only predicted the eddy diffusivity for the active area and not the whole tray. The empirical equations derived in this work are based on segmental downcomers of about 10% of the column cross sectional area which is the usual tray design.

For these reasons, the empirical equations derived by this work will be used in the mathematical model along with the associated column dimensions (i.e. weir height, hole diameter, tray length etc.) to ensure a true prediction of the eddy diffusion coefficient and hence the Peclet number.

5.5 Theoretical Simulation Using Mixed Pools in Series with Backflow.

Although the diffusion model has been shown to be a good representation of the liquid mixing on a distillation plate, it is awkward to use for continuous simulation work. For this reason, a more suitable yet accurate model is needed. It has been mentioned in the previous chapter, that the mixed pools-in-series with backflow model is the most appropriate.

The concept of a perfectly mixed pool is purely hypothetical and as to the number of pools to be used is uncertain. The first major work using pools-in-series was done by Gautreaux and O'Connell (31) who used a series of pools without backflow to describe the liquid mixing on a distillation plate. The parameters affecting the degree of liquid mixing and hence the number of pools to be used are:-

1) the length of liquid path
2) the vapour flow rate
3) the liquid flow rate
4) the actual tray design

Gautreaux and O'Connell (31) showed that the limiting number of stages to be used is a function of the length of the liquid path. They found that every 2.5 ft. of liquid path could be represented by 2 mixed pools.

Durbin (7,23,80) worked with the pools-in-series with backflow model and by introducing the 'phi number', which is a characteristic parameter of the pool model:-

$$\phi = \frac{2M}{1+2B} \qquad \ldots\ldots\ 5.11$$

it was shown that for a given number of pools M and with the backflow ratio B set for equal variances at the outlet, the deviations in moments at intermediate points were of the same value as those evaluated when B was determined by setting $\phi = Pe$. Thus for a set number of pools and a calculated Peclet number, the backflow ratio can be calculated by the theoretical equation:-

$$B = \frac{M}{Pe} - \frac{1}{2} \qquad \ldots\ldots\ 5.12$$

This is suitable if $M \geqslant Pe/2$. Again the actual number of pools was undefined.

An investigation into defining the number of pools and optimum B to be used was carried out by comparing the normalised variances and the root mean squared deviations of the computed analytical impulse response of the diffusion model (48) with those of the pools-in-series model with backflow for different pairs of M and Pe. Starting with an equation of the sort described by Durbin but with unknown coefficients $K_1 \rightarrow K_4$:-

$$B = K_1 \frac{M^{K_2}}{Pe^{K_3}} - K_4 \qquad \ldots\ldots\ldots \quad 5.13$$

an optimisation routine (72) was used to determine these coefficients. The objective function to be minimised was $0.5\sigma^2$ and 0.5 RMSD. It was found that the optimum value was in a shallow area and the best fitting coefficients gave the equation:-

$$B = 0.9698 \frac{M^{1.009}}{Pe^{0.9966}} - 0.54 \qquad \ldots\ldots\ldots \quad 5.14$$

which is almost the same as that defined by Durbin but giving better fits for large Pe. A plot of the backflow ratio against the number of pools for given Peclet numbers using equation 5.12 is shown in Fig. 5.9. The equation derived by this work (equation 5.14) can be approximated to that of equation 5.12 for ease in calculation.

To test the limiting factors described by Gautreaux and O'Connell (31) a series of computations were performed where the number of pools was increased for fixed Peclet number and tray dimensions. The tray length was 6 ft. and the number of pools was increased for different runs. It can be seen from Fig. 5.10a,b that the responses for the fixed positions are similar and as the number of pools is increased, the backflow ratio using equation 5.12 compensates to give roughly the same response. As can be seen from Fig. 5.10b as the number of pools exceeds that of 1 pool per foot of length, the responses at the same positions are virtually identical.

The Peclet number can be estimated from the eddy diffusion coefficient, evaluated from the empirical equations derived, the liquid flow path and the froth velocity. Thus from the above results it appears that if the backflow ratio is determined by equation 5.12, then as long as the number of pools used is greater than Pe/2, then the mixing is well represented. For the simulation work, the limits of Gautreaux were observed and one pool per foot of tray was used except where the liquid path was so large as to make the number of pools excessive for computation.

Table 5.4

Comparison of eddy diffusivity between steady-state and dynamic experiments. ft^2/sec.

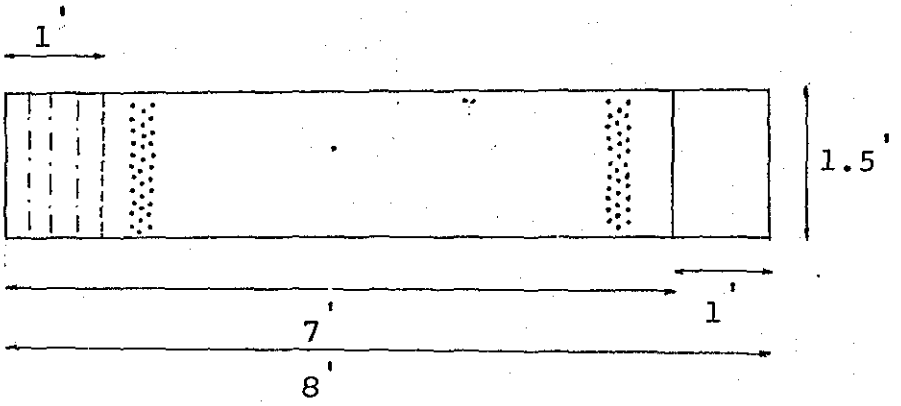
Run	Experimental values from steady-state	Predicted values from correlation derived from dynamic experiments
1	.230112	.17985
2	.205743	.169106
3	.218054	.140827
4	.052265	.052434
5	.045716	.046963
6	.185131	.153922
7	.164354	.153922

Table 5.5

Comparison of Peclet numbers for various works.

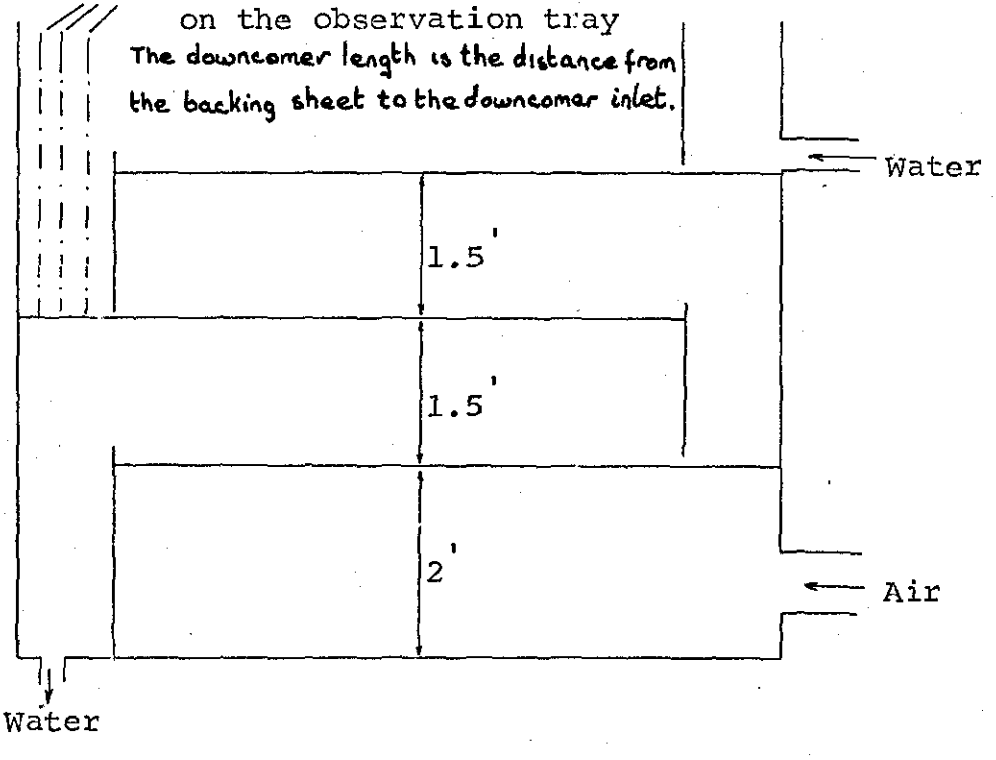
Weir height 2.25"			Vapour flow u_g 3 ft/sec.			h_f = .3744 ft.	
h_c ft	u_l ft/sec	A.I.Ch.E.	Barker and Self.	Foss	Welch	Gilbert	Sheppard
.1291	.023	13.57	13.81	9.33	20.69	3.49	5.86
.1325	.056	30.01	32.62	19.92	20.69	5.16	10.09
.1358	.0874	42.71	51.29	29.75	20.69	6.38	13.5
.1391	.1172	52.47	70.22	39.32	20.69	7.43	16.57
.1425	.1457	59.93	89.55	48.84	20.69	8.38	19.45
.1458	.1729	65.56	109.35	58.38	20.69	9.26	22.22
.1491	.1988	69.56	129.69	68.01	20.69	10.11	24.92
.1558	.2474	74.9	172.03	87.65	20.69	11.73	30.21
.1692	.333	77.3	263.81	128.9	20.69	14.83	40.66

Plan of the observation tray



Removable backing sheets to vary downcomer length on the observation tray

The downcomer length is the distance from the backing sheet to the downcomer inlet.



Front elevation of the experimental rig

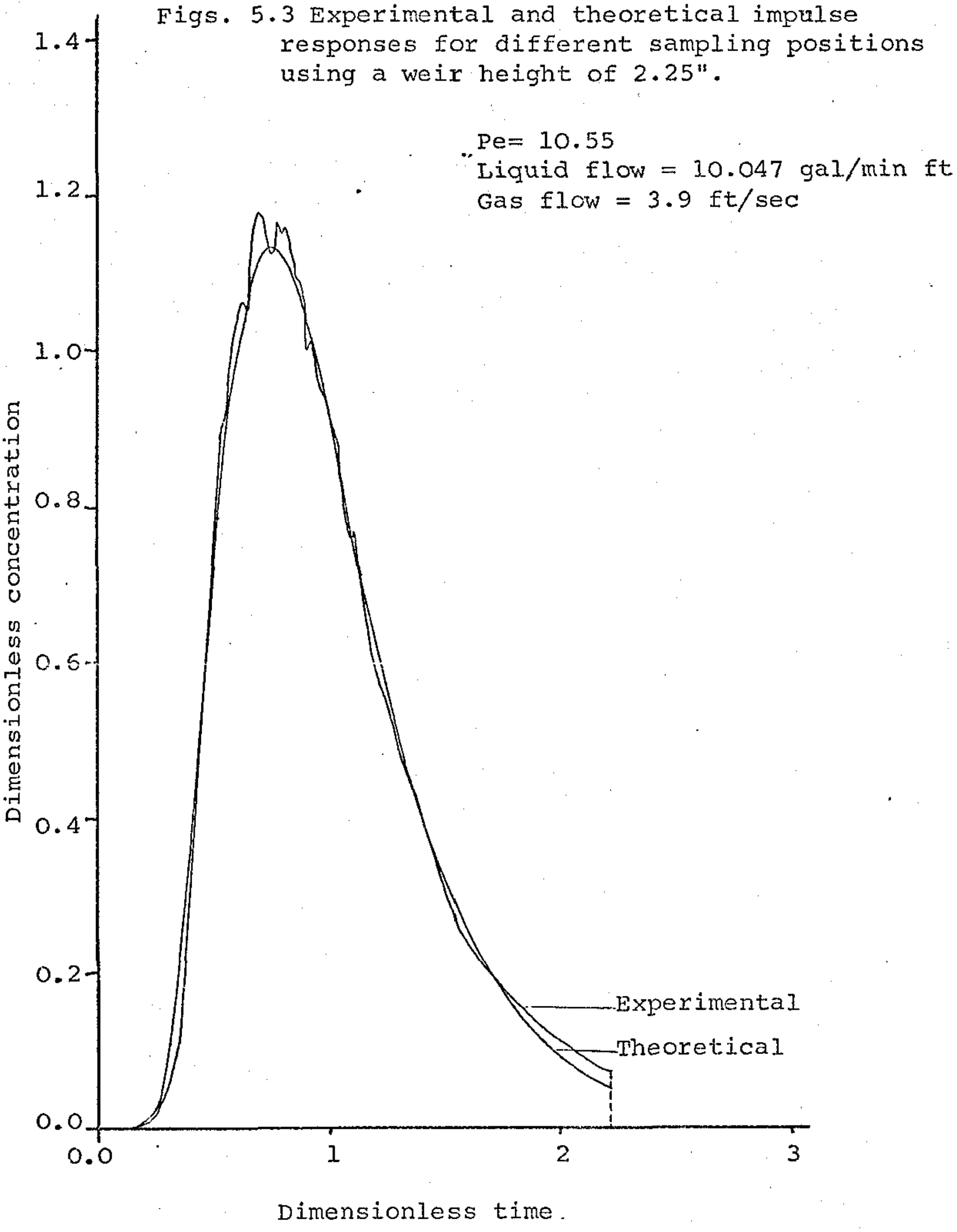
Fig 5.1 Plan and front elevation of the experimental rig. The removable backing sheets allow the downcomer length on the observation tray to be 3",6",9"and 1 ft. The inlet downcomer and tray clearance on all trays was 1" and the weir heights except for the observation tray were 1.5".

Fig 5.2 Photographs of observation tray after dye injection has been introduced.

Photograph A. Taken 10 seconds after the injection. Injection point.
The photograph shows the absence of perfect mixing and the presence of backmixing.

Photograph B. Taken 20 seconds after the injection. Injection point.
The photograph shows the dye being more evenly distributed, but the liquid is not perfectly mixed nor is it travelling in plug flow.

Figs. 5.3 Experimental and theoretical impulse responses for different sampling positions using a weir height of 2.25".



Pe= 10.55
Liquid flow = 10.047 gal/min ft
Gas flow = 3.9 ft/sec

Fig 5.3a Sampled at the outlet weir.

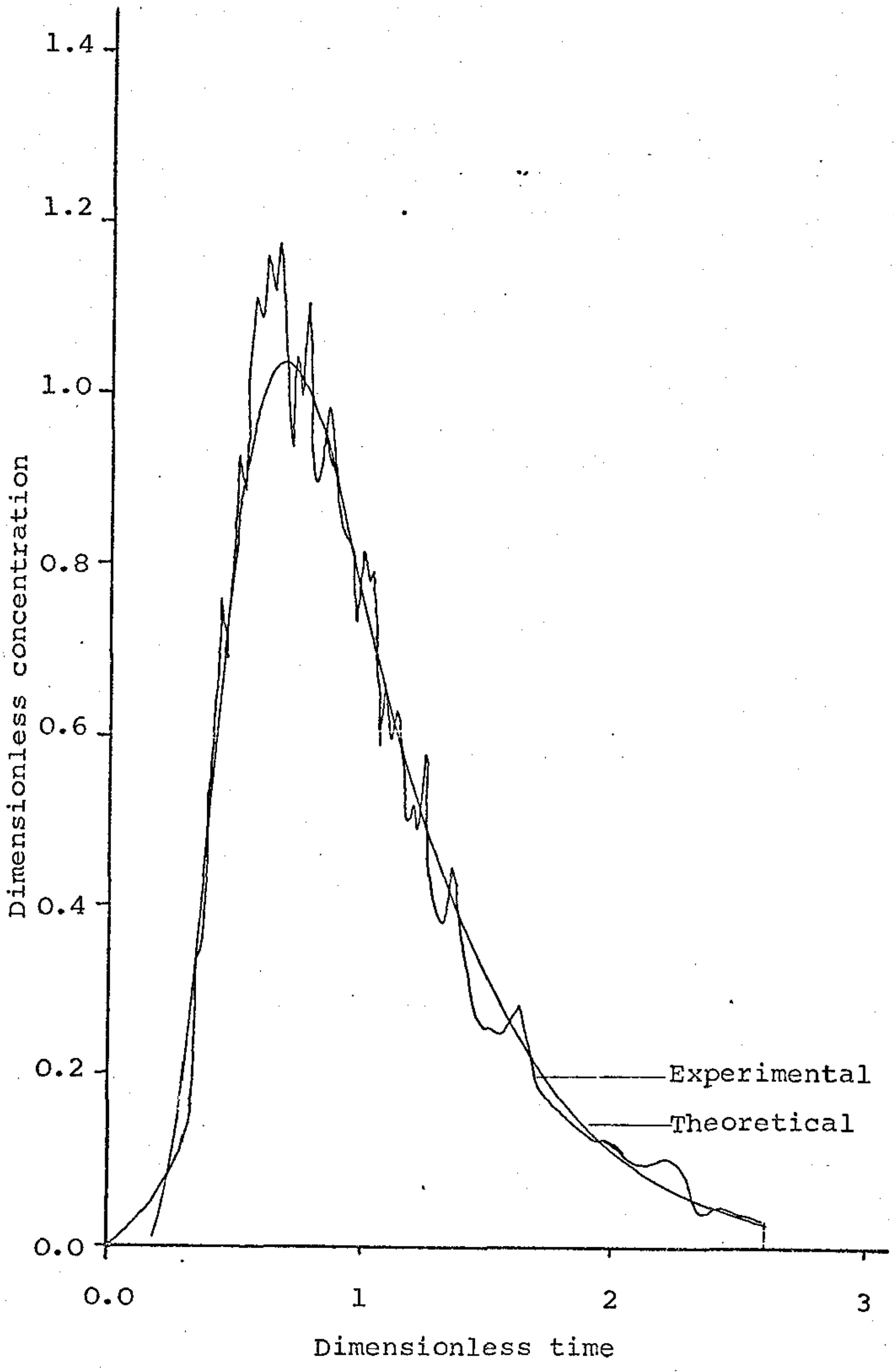


Fig 5.3b Sampled 2 ft before the outlet weir.

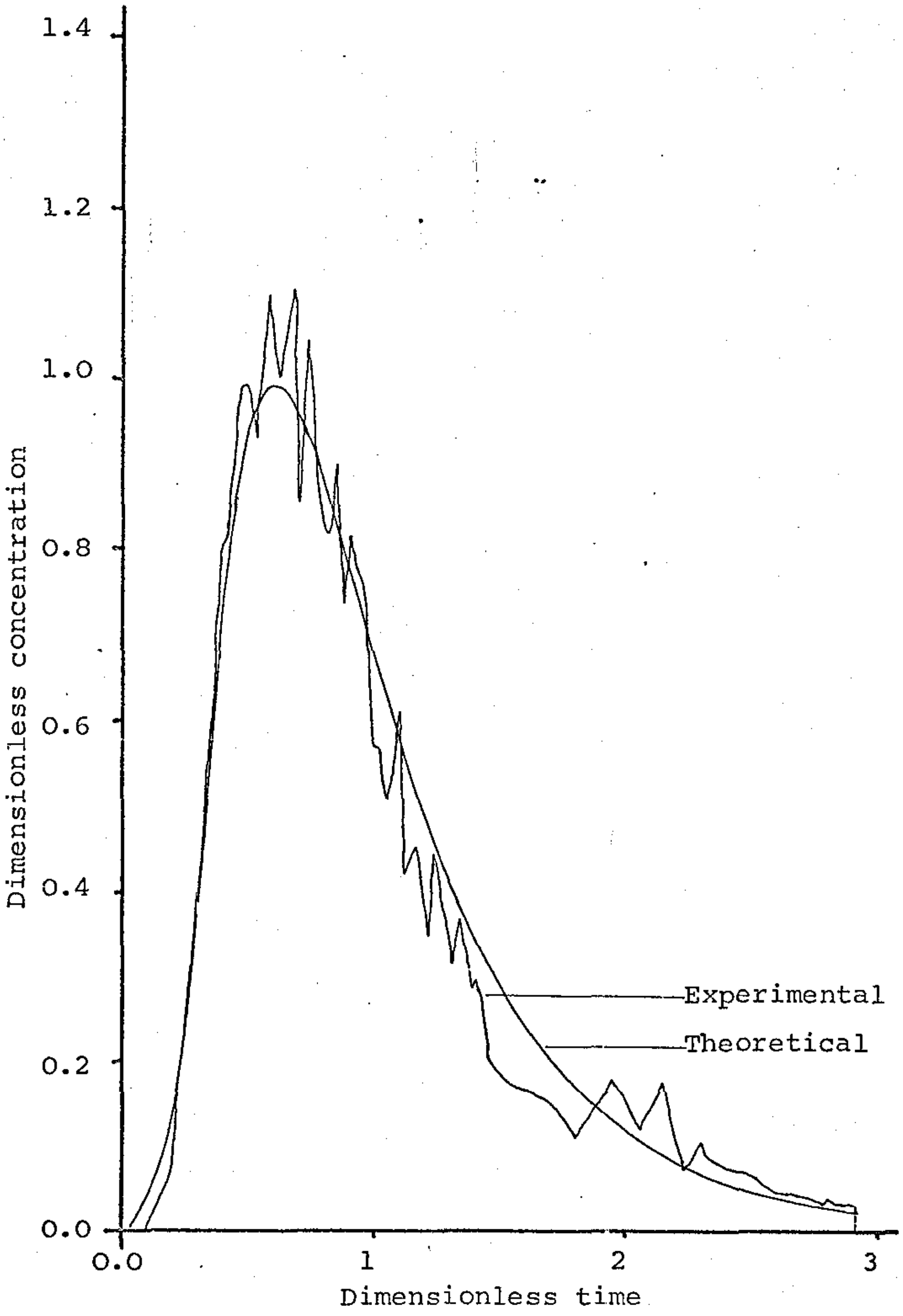


Fig 5.3c Sampled 3 ft before the outlet weir.

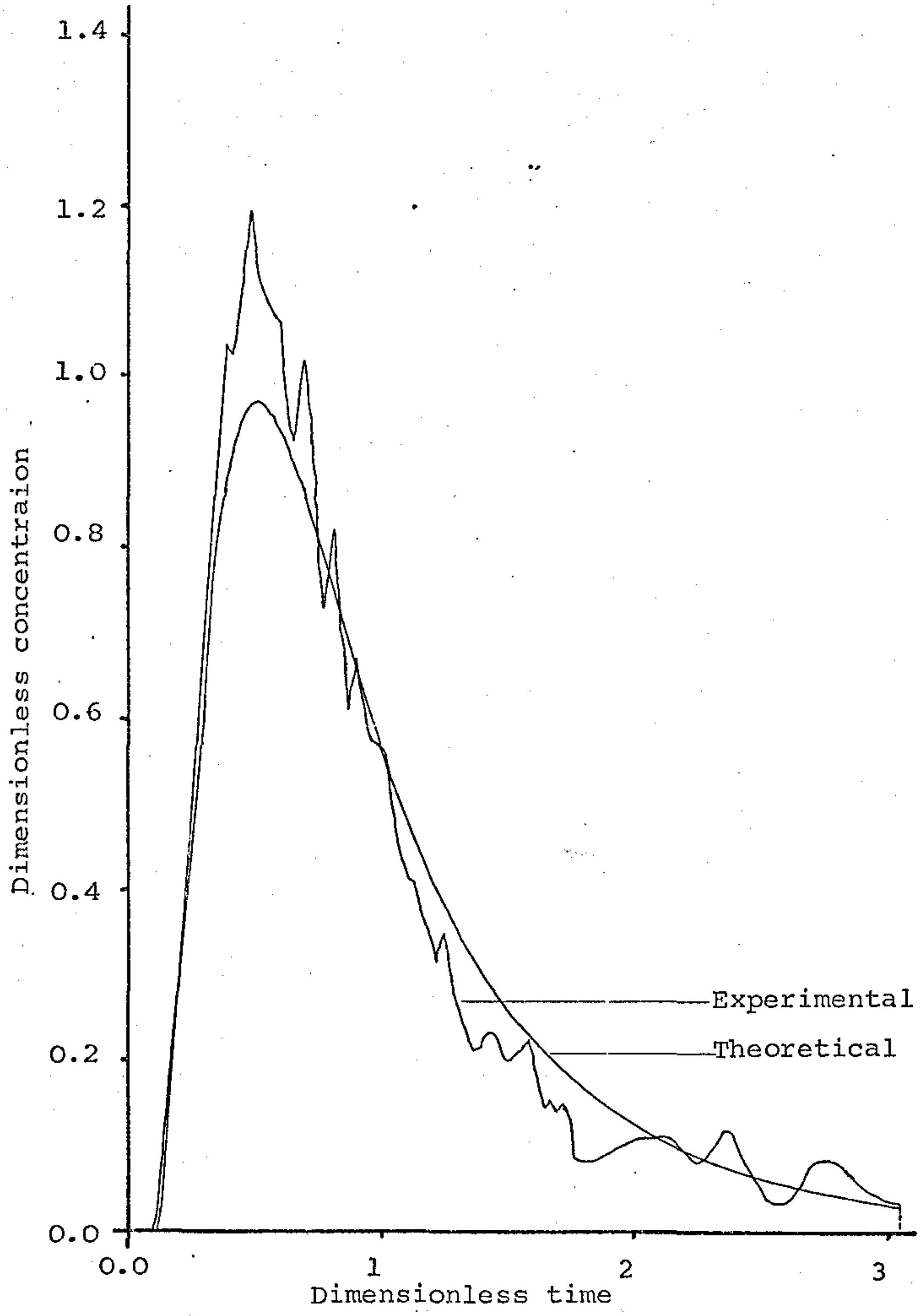


Fig 5.3d Sampled 4 ft before the outlet weir.

Figs. 5.4 Experimental and theoretical impulse responses for different operating conditions ,using a weir height of 1.5" and sampling at the outlet weir.

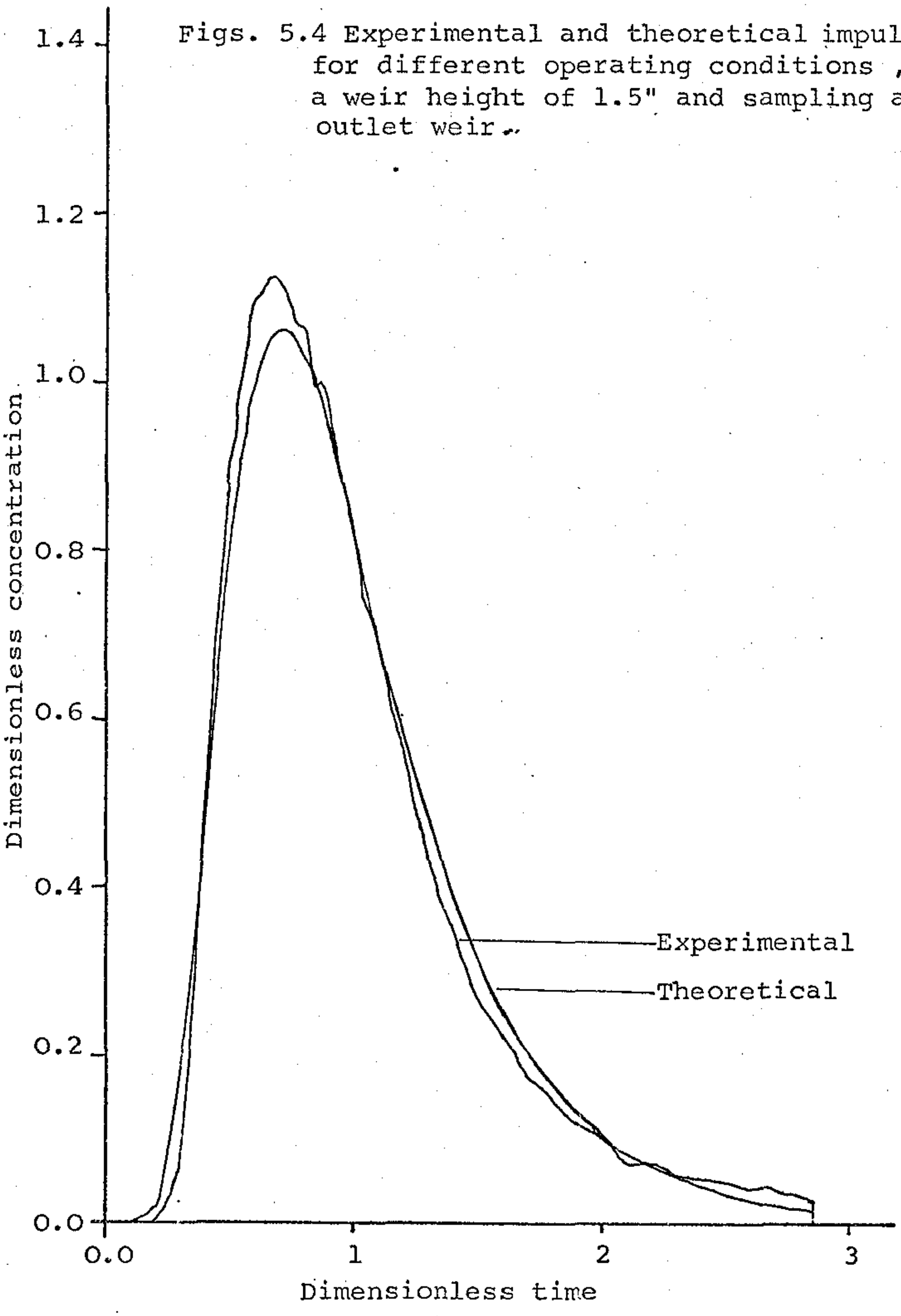


Fig 5.4a Response for Peclet No. = 8.274, Liquid flow = 6.187 gal/min. ft

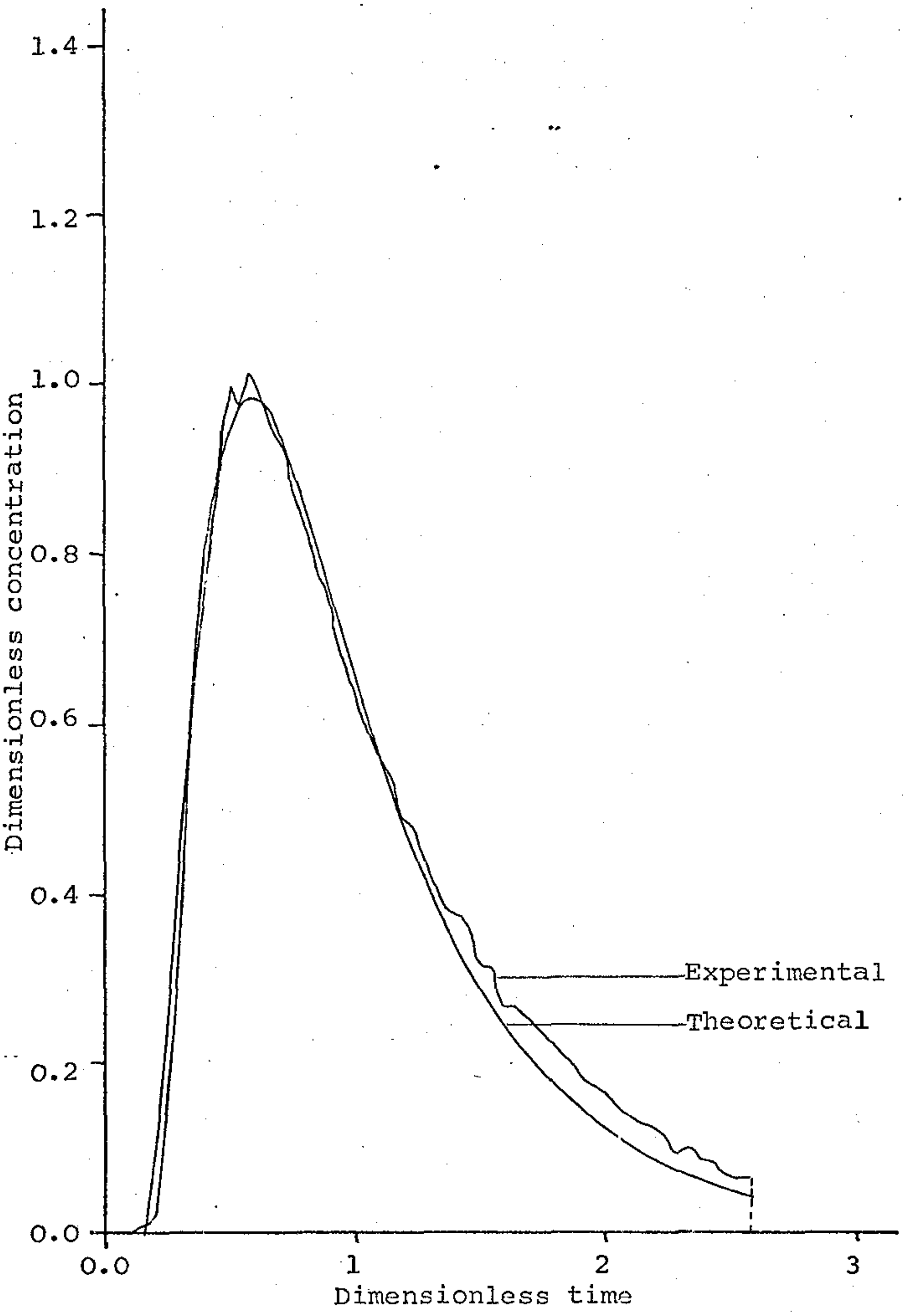


Fig 5.4b Response for Peclet No. = 5.327,
Liquid flow = 4.194 gal min ft

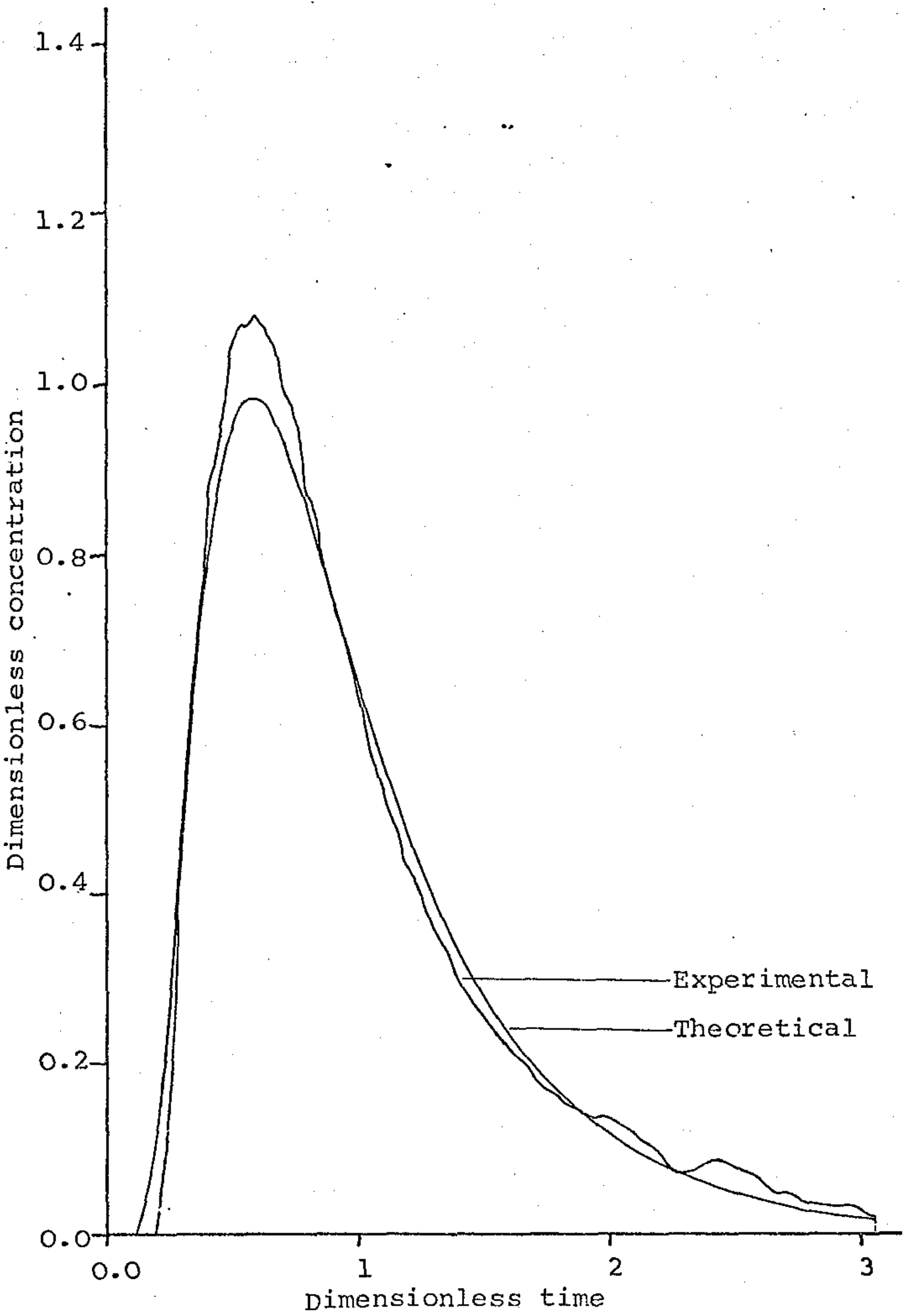


Fig 5.4c Response for Peclet No. = 5.087,
Liquid flow = 5.078 gal/min ft

Figs. 5.5 Experimental and theoretical impulse responses, sampled at the outlet weir for the tray with the large diameter holes.

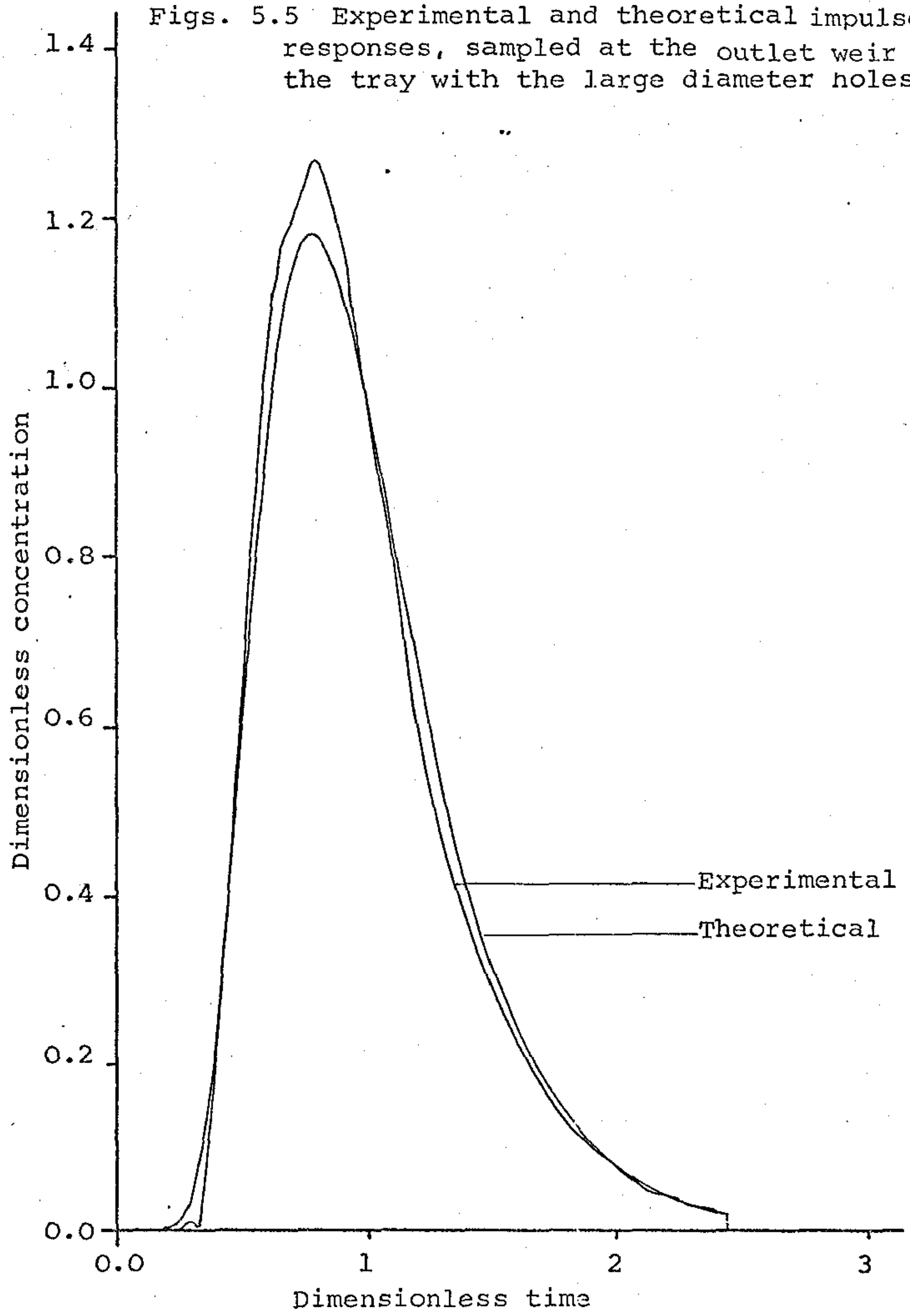


Fig 5.5a Response for Peclet No. = 12.22,
Liquid flow = 8.662 gal/min ft
Air velocity = 2.98 ft/sec

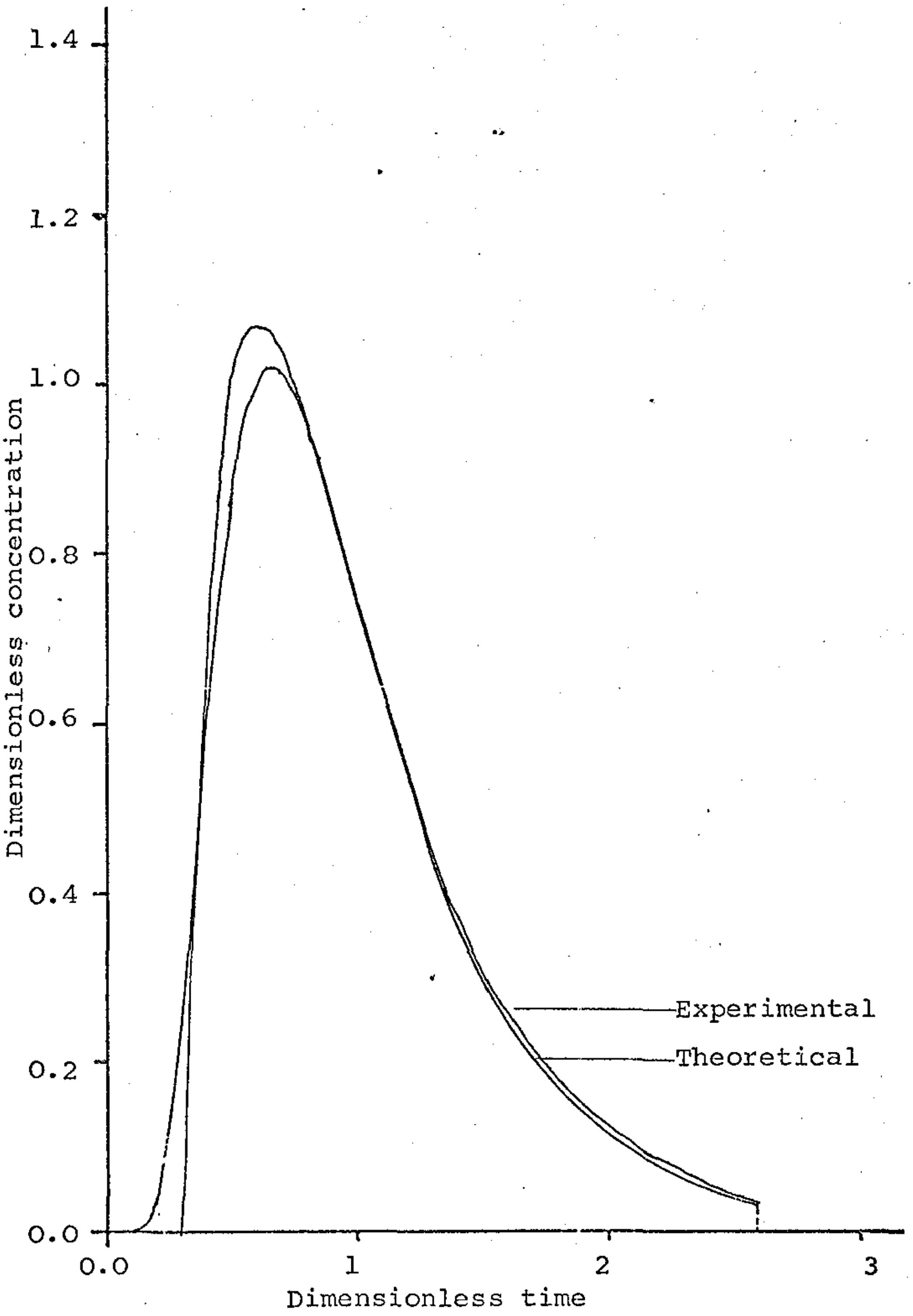


Fig 5.5b Response for Peclet No. =6.882,
Liquid flow = 4.537 gal./min ft
Air velocity = 3.06 ft/sec

Figs. 5.6 Experimental impulse responses sampled at the outlet weir for four different sized downcomer segments.

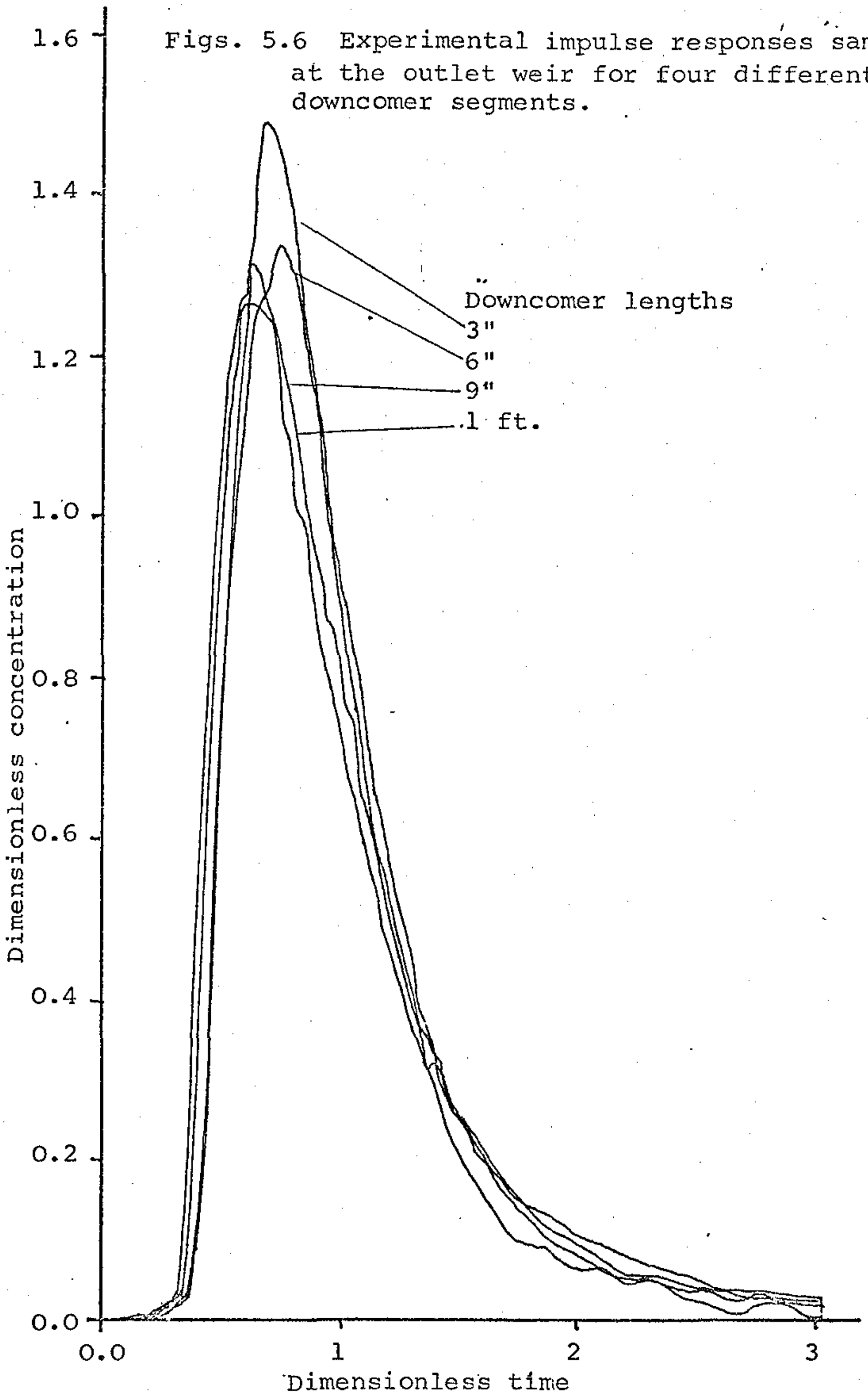


Fig 5.6a Responses of experiment No. 4 Table 5.3

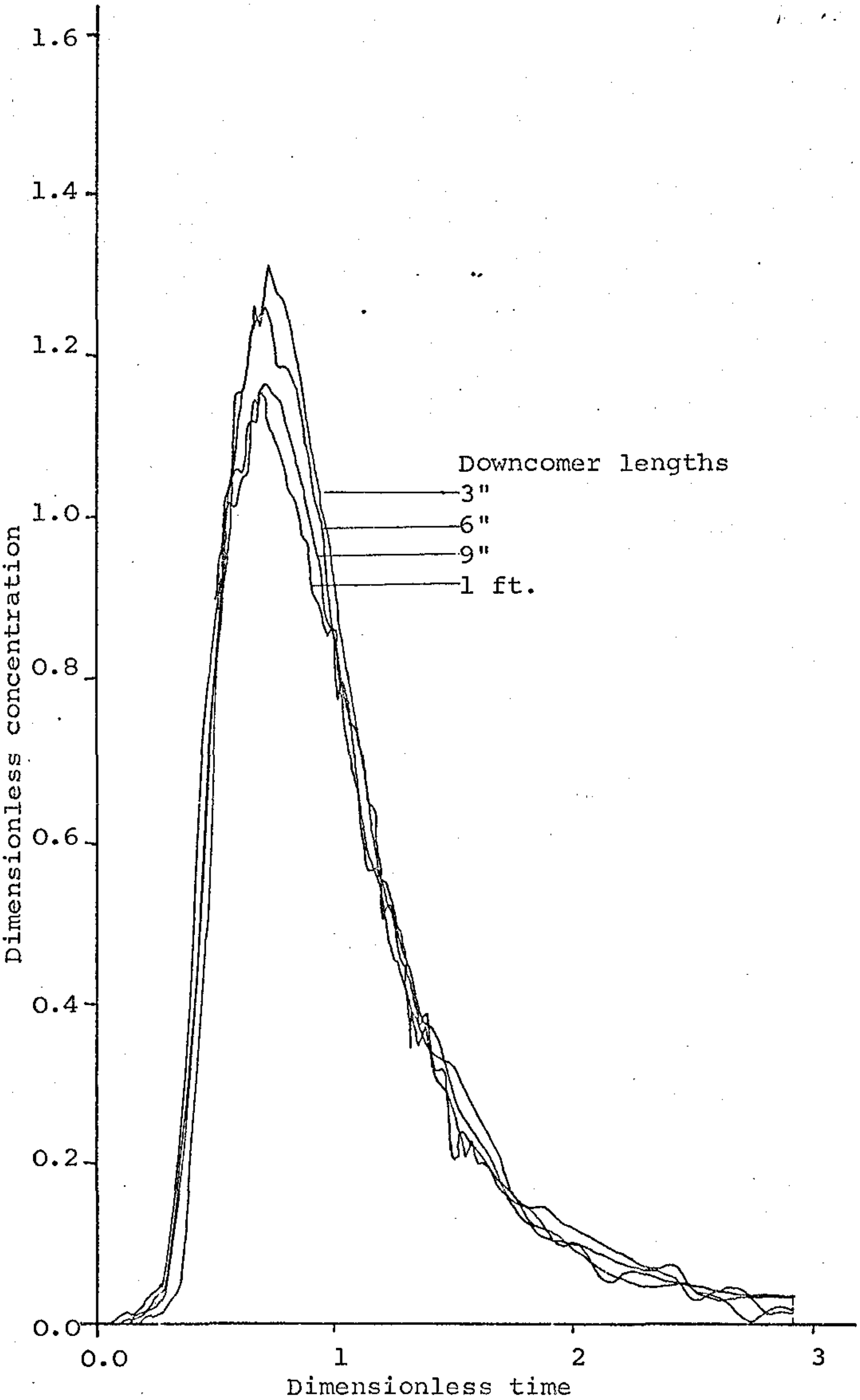


Fig 5.6b Responses of experiment No. 6 Table 5.3

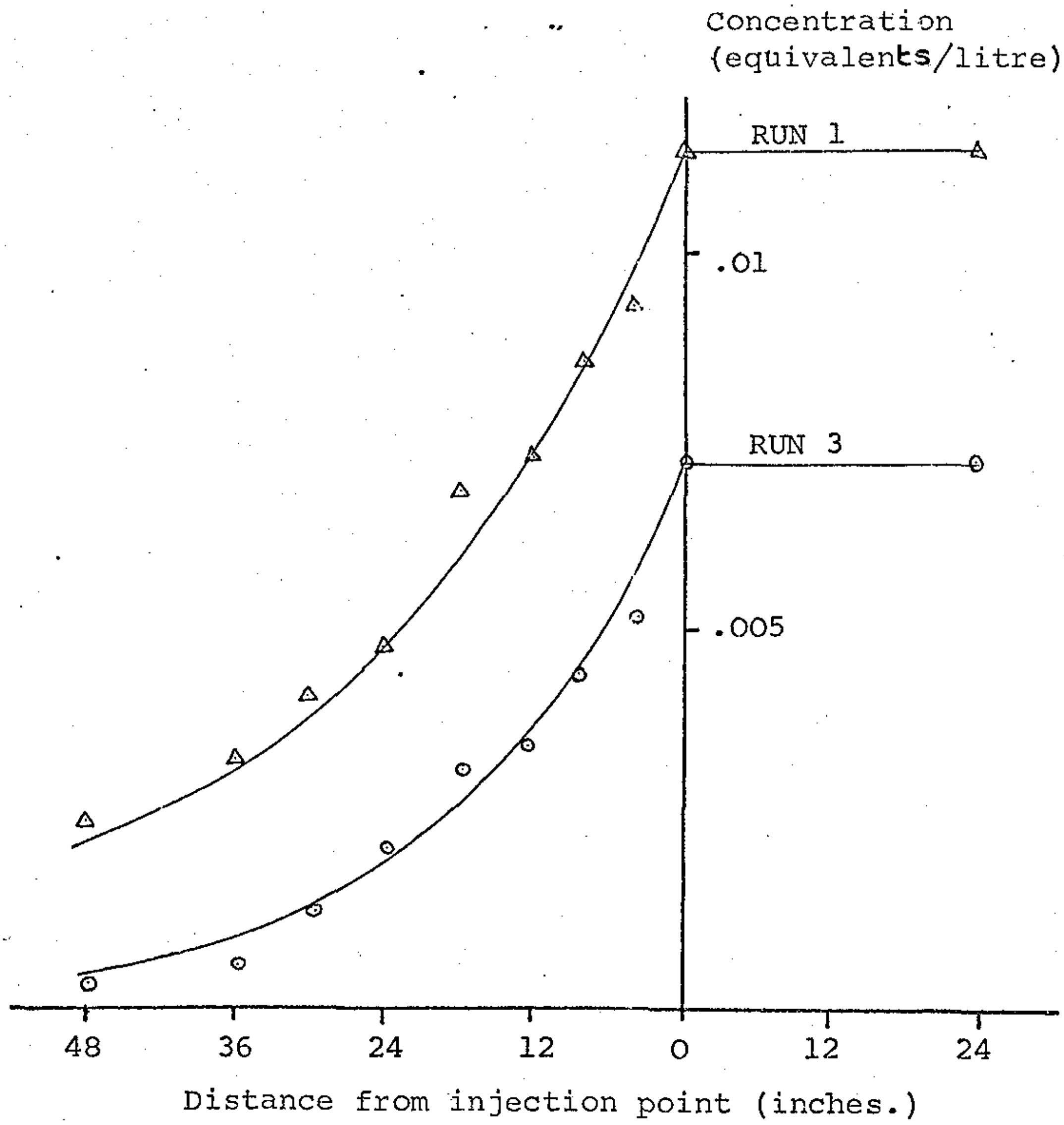


Fig 5.7 Experimental results of two of the steady-state experiments showing computed response and experimental points.

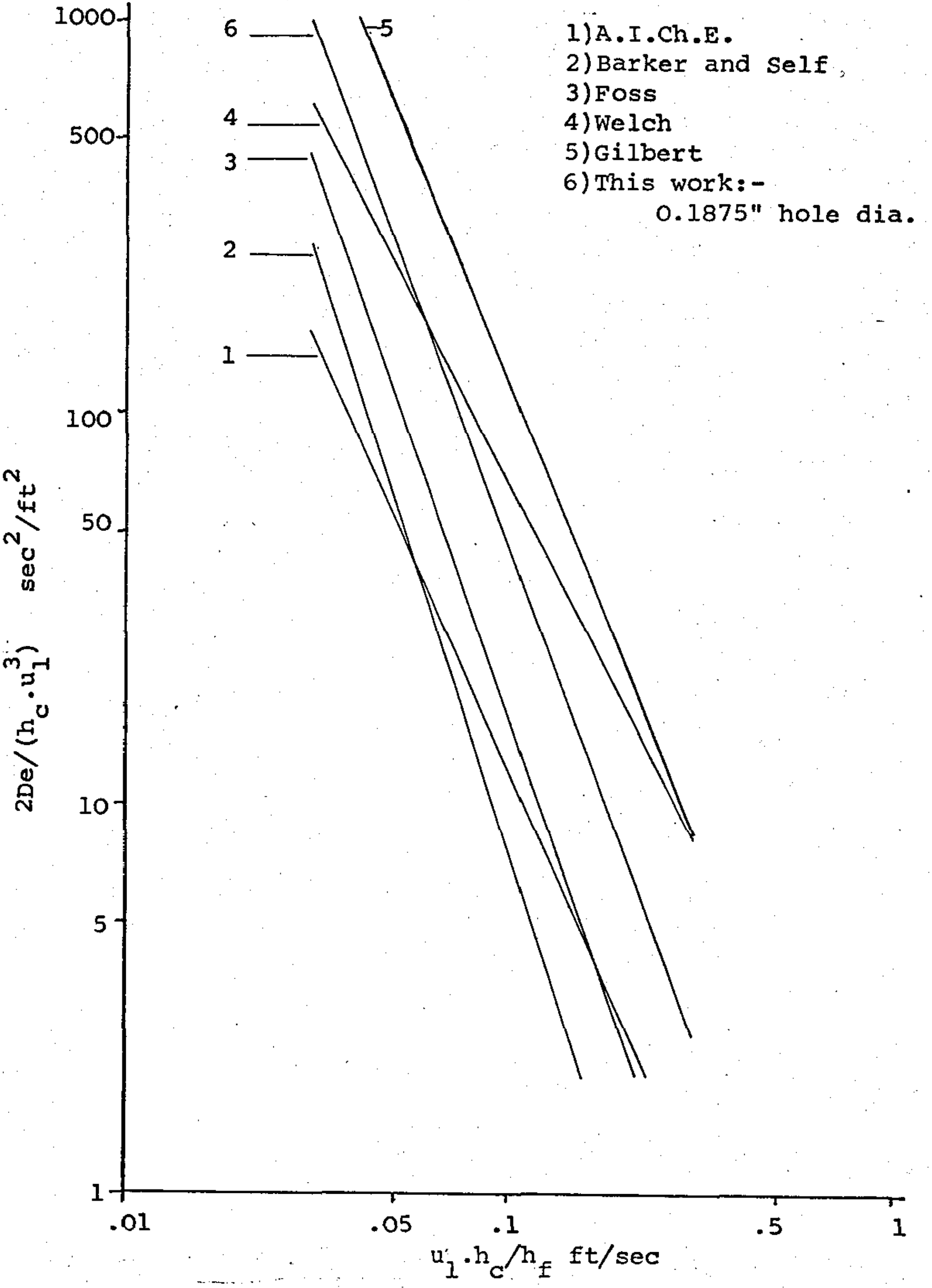


Fig 5.8a Comparison of empirical equations derived using weir heights of 2.25".

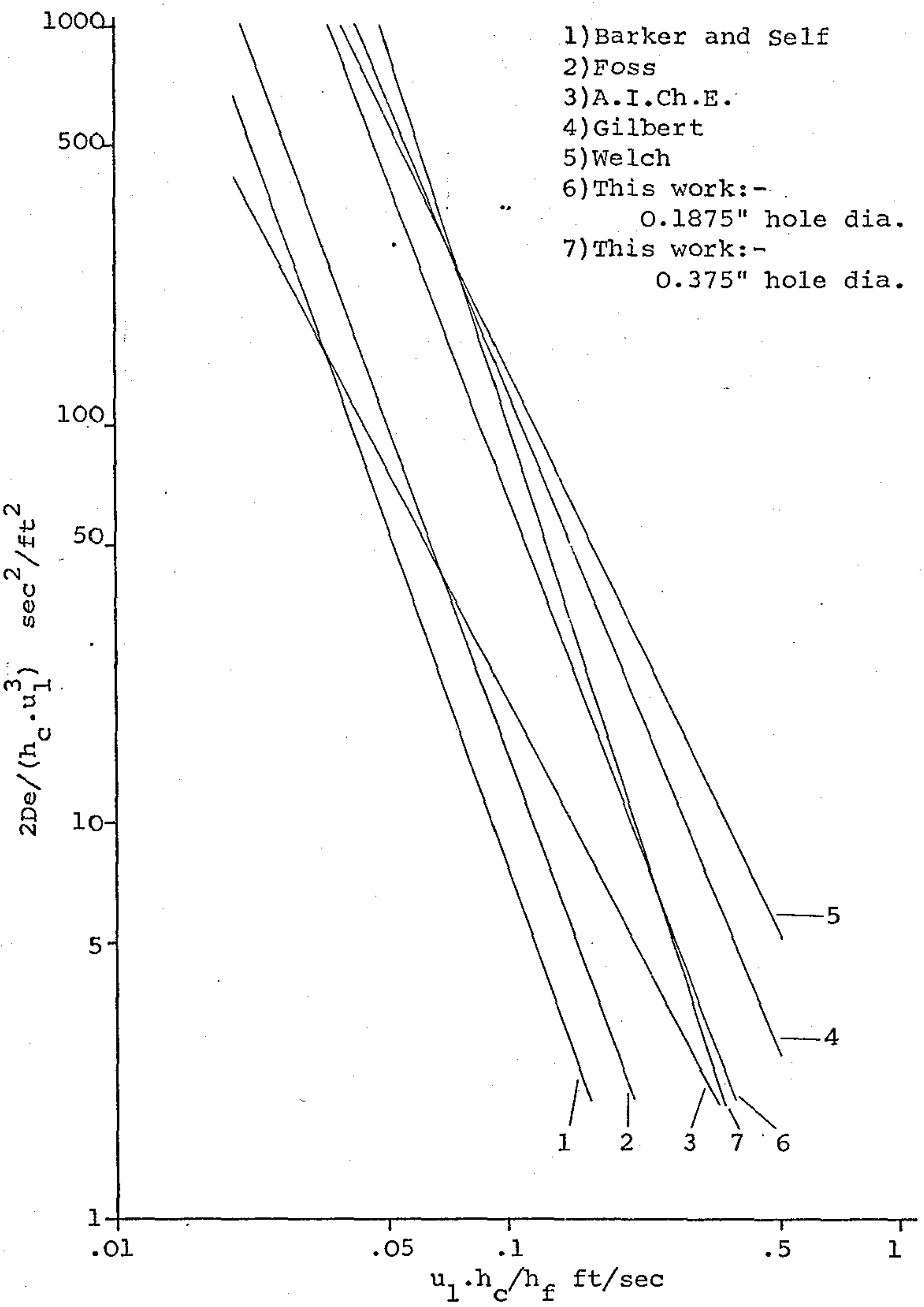


Fig 5.8b Comparison of empirical equations derived using weir heights of 1.5".

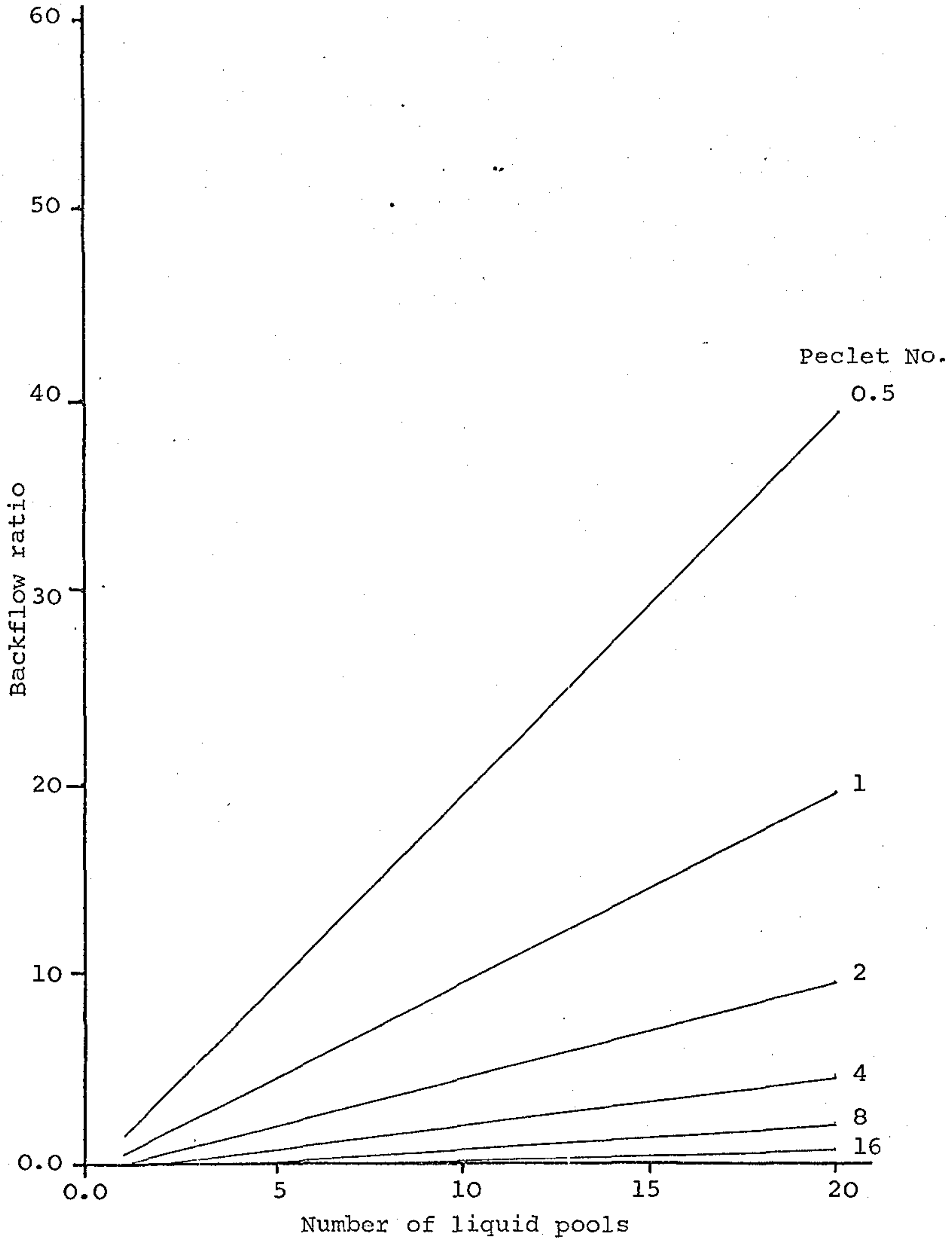


Fig 5.9 Theoretical comparison of backflow ratio with the number of mixed pools used.

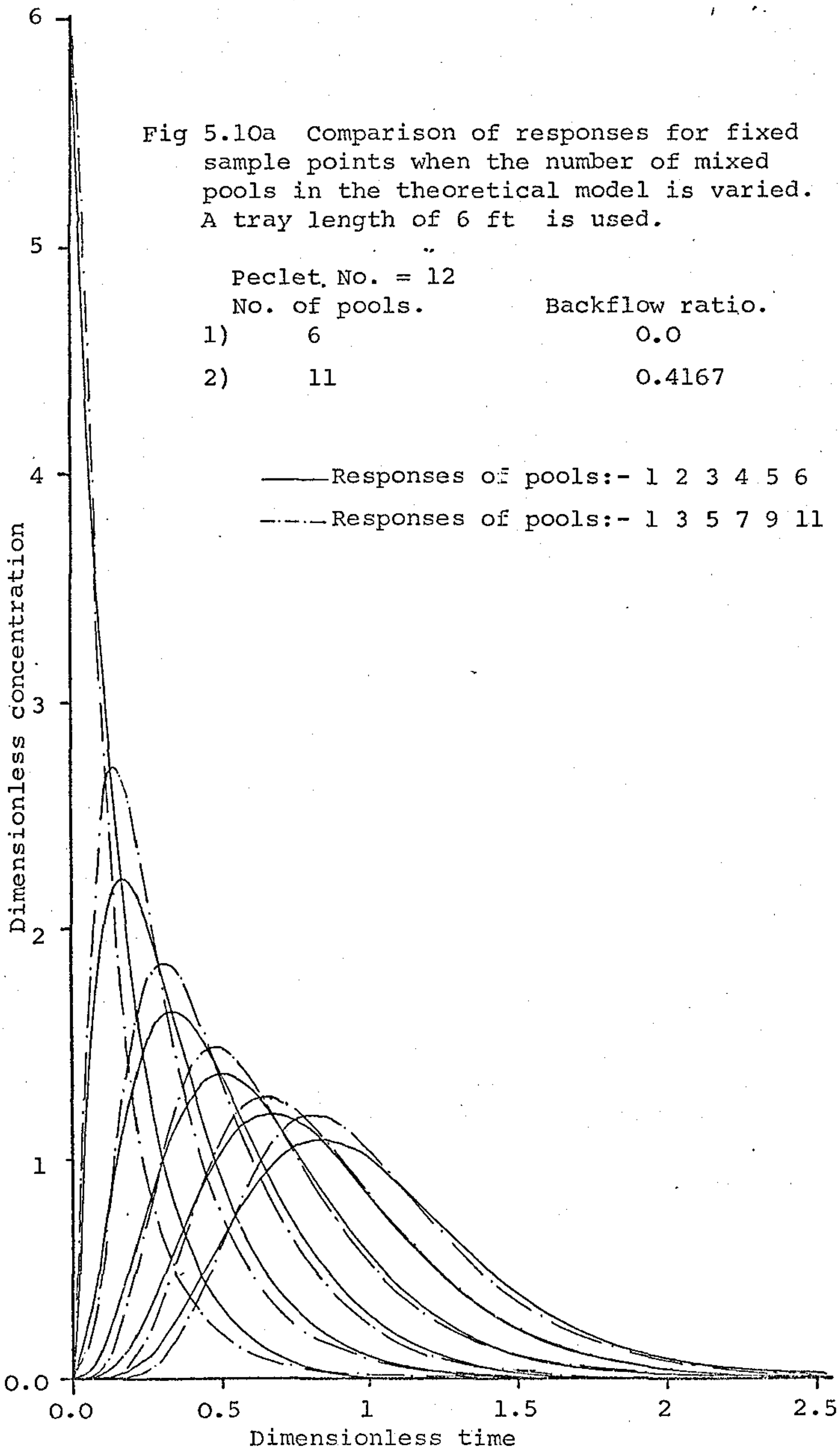


Fig 5.10a Comparison of responses for fixed sample points when the number of mixed pools in the theoretical model is varied. A tray length of 6 ft is used.

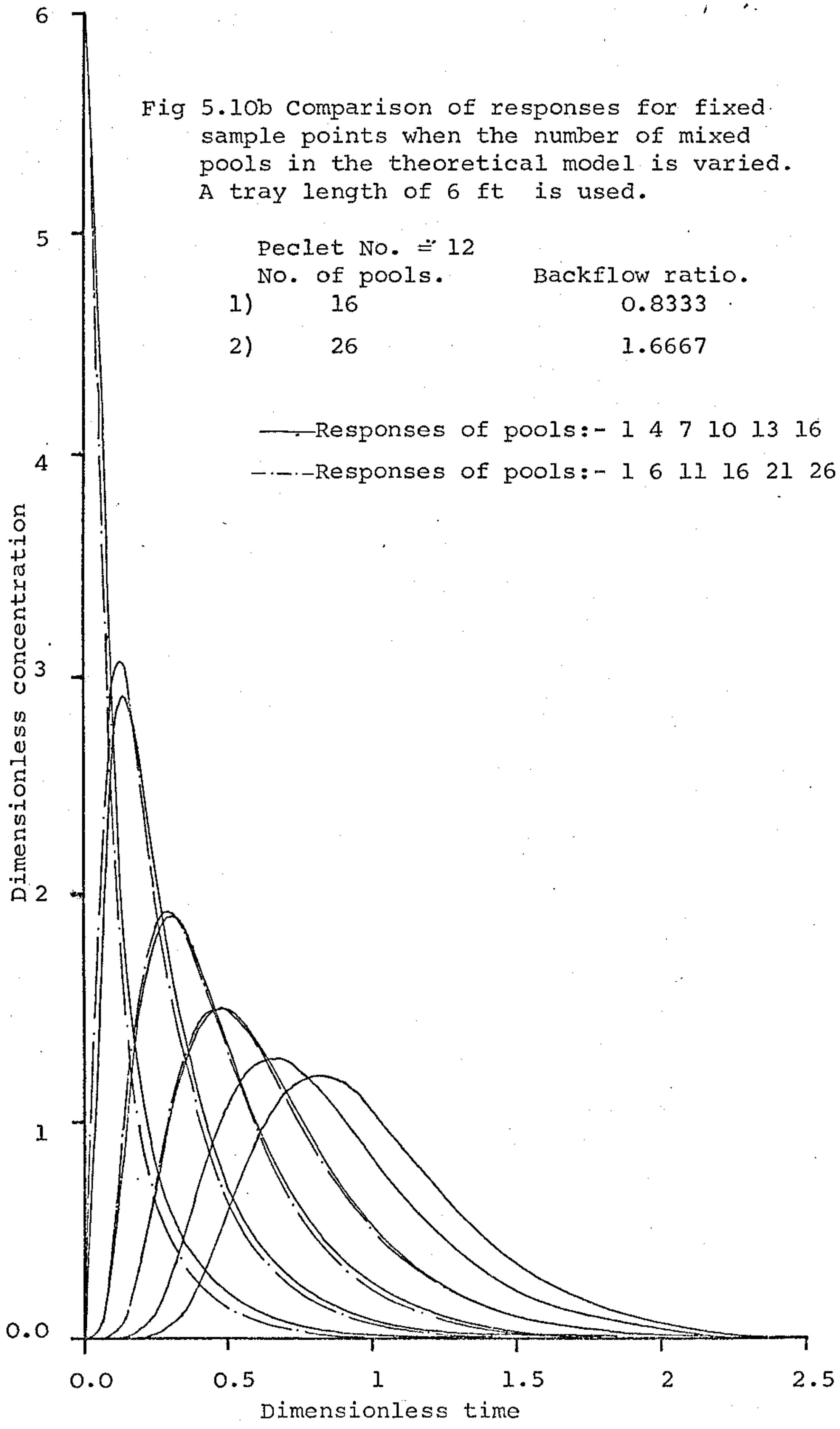


Fig 5.10b Comparison of responses for fixed sample points when the number of mixed pools in the theoretical model is varied. A tray length of 6 ft is used.

Peclet No. ≐ 12

	No. of pools.	Backflow ratio.
1)	16	0.8333
2)	26	1.6667

6. VAPOUR MIXING BETWEEN DISTILLATION PLATES

6. Vapour Mixing Between Distillation Plates

6.1 Vapour Mixing Experiments

Reports on the extent to which liquid or vapour mixes between closed boundaries were scarce before the 1950s due to the difficult techniques involved in studying the eddy diffusivity. Experiments in water have been carried out by Van Driest (105), who injected a mixture of benzene and carbon tetrachloride and recorded the distribution of the tracer in the main stream by photography. Investigations into the diffusion in gases have been carried out by Towle (102,103), who investigated the eddy diffusion of carbon dioxide and hydrogen in a turbulent air stream using different sized pipe work; and by Sherwood and Woertz (88) and Slattery (91).

The theory used by the author to determine the eddy diffusivity of the vapour phase between two distillation plates has been given in Appendix C and is the same as that described by Slattery (91) for diffusion of gases in pipes. The purpose of the experimentation is not to determine accurately the vapour mixing characteristics, but to determine whether the vapour is:-

1) Perfectly mixed

or 2) Partially mixed but not enough to justify the assumption of perfect mixing

or 3) Not mixed at all

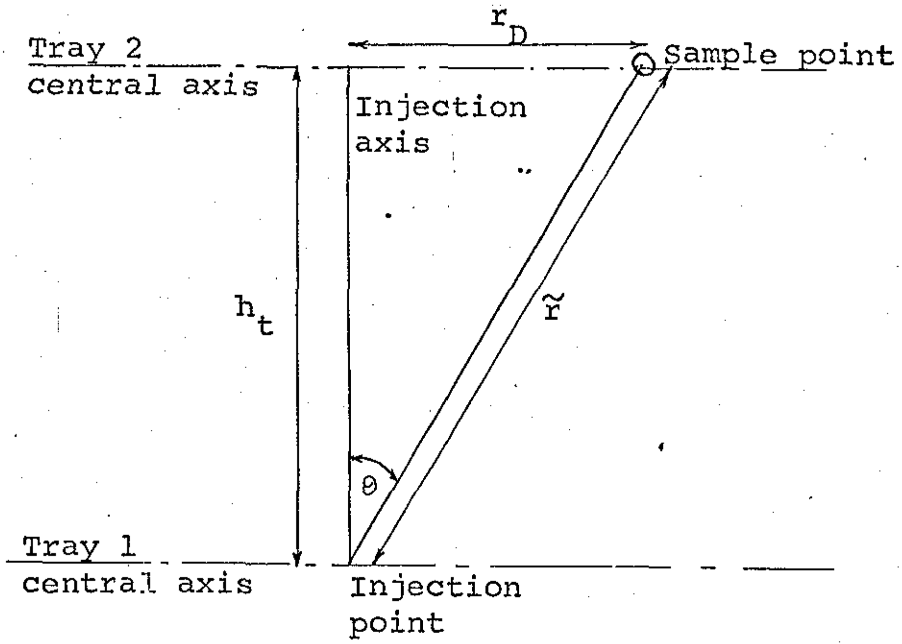


Fig 6.1 Schematic representation of vapour mixing apparatus .

6.2 Experimental Procedure of Steady-state Calculation

A micro-katharometer with a maximum travel time constant of a few milli-seconds was used to measure the thermal conductivity of the continuously sampled stream. The main flow stream was air, which was assumed to be far removed from any bounding walls or surfaces and to move in steady-state turbulent flow with a uniform and constant velocity u_g.

From the diagram of the apparatus Fig. 6.1, the tracer was continuously and uniformly injected into the main air stream at the origin of the central axis in the diagram. The tracer used was helium and its mass injection rate C was constant and small compared to the mass flow rate of air, thus giving a negligible increase in total flow. Diffusion of the helium occurs in both the axial and radial directions as it moves up between the two plates. The eddy diffusion coefficient can be found from the concentration distribution of the helium in the air stream.

Due to the sensitivity of the element in the micro-katharometer any contact with liquid would overload and damage it. For this reason the initial experimentation was carried out with no liquid present on the trays. At the start of an experiment, only the main air stream was passed and readings from the micro-katharometer were continuously recorded for a period of twenty seconds and stored in a computer to determine a mean base line for the experiment. Having calculated the mean base line, the helium tracer was introduced and steady-state was allowed to be reached. At steady-state, continuous readings of the thermal conductivity of the sampled stream were taken every 0.1 of a second over a period of ten seconds and stored in the computer. The value of the mean base line was subtracted from the experimental values and the area under the curve given by the experimental readings was calculated. The mean value was found by dividing the area by the sampling time and was the mean voltage given as output from the micro-katharometer. This voltage was then converted into the mean concentration of tracer passing. The experiments at each sample point were repeated several times to form a set of mean concentrations for each particular sample point.

The sample probes were small stainless steel tubes of 1 mm. internal diameter. Being of such a small diameter, the sample probes would not affect the vapour mixing and the volume of sample taken was very small compared to the main flow. The sample probes were placed at various radial distances around the injection axis. The helium injection probe was a point source resting on the centre of distillation tray while the sample probes were positioned directly below the tray above.

From the set of mean concentrations for each sample probe position, one value was calculated using a least squares fit to represent the average concentration for that particular sample position. From the theory given in Appendix C, the equation:-

$$c = \frac{Q_h}{4\pi De\,\tilde{r}} \exp\left\{\frac{-u_g}{2De} (\tilde{r} - h_t)\right\} \qquad \ldots\ldots\; 6.1$$

can be arranged to give:-

$$\ln c\tilde{r} - \ln \frac{Q_h}{4\pi De} = \frac{-u_g}{2De}(\tilde{r} - h_t) \qquad \ldots\ldots\ldots \quad 6.2$$

The eddy diffusion coefficient can be found by plotting $\ln c\tilde{r}$ against $(\tilde{r} - h_t)$ and by least squares fitting finding the best straight line joining the experimental points, the slope of the line being equal to $-u_g/2De$ from which De can be calculated.

Having performed the experiments with no liquid present on the trays, the same experiments were repeated with the water flowing. To safe-guard the micro-katharometer, a small volume water trap was inserted in the sample probe. The injection probe was no longer resting on the tray but was raised to the height of the froth as only the vapour mixing between the froth and the tray above is of interest. The Peclet number was calculated using the superficial gas velocity and the vertical distance between the injection probe and the tray above.

6.3 Experimental Results

6.3.1 Experiments with Liquid Absent

The eddy diffusion coefficient was determined for various vapour flow rates. Table 6.1 lists the run numbers, the main stream velocity, the determined eddy diffusion coefficient and the Peclet number. The mole percentage of helium in the sampled stream at various points can be determined. Fig. 6.2 shows the two extreme cases in the eight experimental runs. In Run 8 where the vertical velocity was largest, the helium tracer shows a greater dispersion than that of Run 1. The values for these two extreme cases calculated using equation 6.1 are given in Table 6.2.

From Fig. 6.3, showing the cumulative percentage of tracer passed for Run 1 it can be seen that within a radial distance of 15 cms about the injection axis, 83% of the tracer is passed and within a radial distance of 20 cms 96% of the tracer is passed.

From these results it would seem that the vapour is not perfectly mixed though at this stage the effect of liquid present on the tray is not known.

6.3.2 Experiments with Liquid Present

The same experimental procedure as previously explained was carried out except that the injection probe was raised to the height of the froth. The eddy diffusion coefficient was determined by plotting the data points and by least squares fit finding the best line joining the points, the slope of which is given by $-u_g/2De$. The eddy diffusion coefficient was determined for various vapour flow rates, and a list of air velocities, eddy diffusion coefficients and Peclet numbers are given in Table 6.3. The actual experimental values and the best fitting lines are shown in Fig. 6.4a,b.

The mole fraction of helium tracer in the sample stream for the two extreme experimental cases (no. 6 and 7) are shown in Fig. 6.5 and their percentage of helium tracer passed within radial distances from the injection axis are shown in Fig. 6.6. It can be seen from Fig. 6.6 that the dispersion of the helium tracer is approximately the same even though the air velocities are different.

Comparing the eddy diffusion coefficients determined under dry conditions with those determined with liquid present, the first set give higher values for the same air velocity. This is thought to be due to the air leaving the tray perforations in the form of jets which later create a greater disturbance when they mix giving better mixing characteristics and thus a higher value of the eddy diffusion coefficient. As the vapour velocity is lowered, the eddy diffusion coefficients of those determined with liquid present tend to those determined with liquid absent. This is due to the froth height or holdup on the tray decreasing and giving less resistance to the jet streams. When liquid is present, the vapour leaves the perforations as jets which are then squashed by the presence of the liquid thus giving a smoother, less active vapour stream leaving the froth.

Plotting the values of the eddy diffusion coefficient against the Reynolds number (Fig. 6.7) for both sets of experiments, it can be seen that the best linear relationship joining the experimental points where the liquid is absent, is steep and is most likely some function of u_g^3.* The line joining the experimental points with liquid present, if assumed to pass through the origin, cuts the line predicted by the experiments with liquid absent. When the vapour flow rate reaches a certain limit weeping occurs and the liquid holdup decreases rapidly until it becomes negligible. This can be seen by the eighth experiment where the liquid holdup was negligible and gave good agreement with the experiments with liquid absent.

The comparison of the eddy diffusion coefficients determined by this work and the eddy diffusion coefficients determined by other works for free turbulent flow in pipes, show that the mixing characteristics of vapour between the distillation plates is greater than that occurring in pipes. This must be due to the presence of the sieve trays which disturb and cause mixing of the vapour. However, the extent of the mixing is small.

6.4 Discussion on Vapour Mixing

From Fig. 6.7 it can be seen that the sieve trays cause the vapour to mix and that the presence of liquid

* If the line is assumed to pass through the origin.

on the tray causes the vapour to be more uniformly dispersed (less mixing). If the Peclet numbers for all but the last experiment listed in Table 6.3 are considered, then the range is from 31 to 39 and the degree of mixing is thus very small. If the same concept as that shown in the liquid mixing section is used, then this could be represented by 15 vapour cells consisting of thin elements perfectly mixed in the direction of vapour flow which is approximately the same model as that of no mixing at all.

Ashley and Haselden (5) tried to describe the vapour mixing between plates by a series of vapour cells which crossed the tray perpendicular to the vapour flow (Fig. 4.2). The vapour leaving a specified number of mixed liquid pools on the tray was averaged and introduced directly to the tray above. The actual number of vapour cells to be used was unknown due to lack of data. The same vapour mixing model as described by Ashley and Haselden (5) was adopted by the author for the mathematical model used in the distillation simulation.

From this work on vapour mixing and assuming the superficial gas velocity to be within the same range (0.8 to 2.0 m/s), from Fig. 6.6 it can be seen that 91% of the tracer passes within a radius of 15 cms. From this it can be concluded that for standard tray spacings and superficial gas velocities of about 0.8 to 1.5 m/s, it can be assumed that there is no vapour mixing between plates for large plate columns. This means that for the liquid mixing model, the vapour leaving any liquid pool passes directly to the liquid pool above it.

From the best fitting linear relationship between the eddy diffusivity and the Reynolds number, a relationship between the eddy diffusivity and the vapour velocity has been derived.

$$De = u_g/95 \qquad \ldots\ldots\ 6.3$$

This relationship is only valid when the vapour velocity is great enough to eliminate weeping.

Table 6.1 The eddy diffusivity and Peclet numbers for the vapour phase

Run	Superficial gas velocity m/hr	Eddy diffusivity m^2/hr	Peclet number
1	2064.6	10.75	88.6337
2	2272.1	20.658	50.7587
3	2486.9	42.077	27.2762
4	2676.0	39.105	31.5836
5	2724.2	36.066	34.8588
6	2991.9	58.44	23.6269
7	3470.4	88.061	18.1866
8	3724.8	94.539	18.1829

S.I. units are used in this section on the vapour phase as there are no empirical equations dependent on old units involved.

Table 6.2 Experimental tracer distribution

Radius (inches)	Helium in sample mole %	
	Run 1	Run 8
0	$.6417 \times 10^{-1}$	$.7517 \times 10^{-2}$
1	$.6 \times 10^{-1}$	$.7406 \times 10^{-2}$
2	$.4909 \times 10^{-1}$	$.7082 \times 10^{-2}$
3	$.3520 \times 10^{-1}$	$.6576 \times 10^{-2}$
4	$.2218 \times 10^{-1}$	$.5934 \times 10^{-2}$
5	$.1234 \times 10^{-1}$	$.5209 \times 10^{-2}$
6	$.6086 \times 10^{-2}$	$.4452 \times 10^{-2}$
7	$.2676 \times 10^{-2}$	$.3710 \times 10^{-2}$
8	$.1055 \times 10^{-2}$	$.3019 \times 10^{-2}$
9	$.3756 \times 10^{-3}$	$.2402 \times 10^{-2}$
10	$.1214 \times 10^{-3}$	$.1872 \times 10^{-2}$
11	$.3586 \times 10^{-4}$	$.1430 \times 10^{-2}$
12	$.0974 \times 10^{-4}$	$.1073 \times 10^{-2}$

Table 6.3 Experimental eddy diffusivity and Peclet numbers for vapour phase with liquid present on the tray. Liquid Flow :- 350 gal/hr.

Run	superficial air velocity m/hr	Eddy diffusivity m^2/hr	Peclet number
1	2696.36	32.8466	31.5728
2	3200.71	36.7838	33.4656
3	3415.1	39.2483	33.4664
4	3688.7	39.4186	35.9899
5	4143.69	42.7781	37.2541
6	4205.0	41.3388	37.2554
7	2559.08	25.1392	39.1508
8	2061.99	10.8264	87.8926

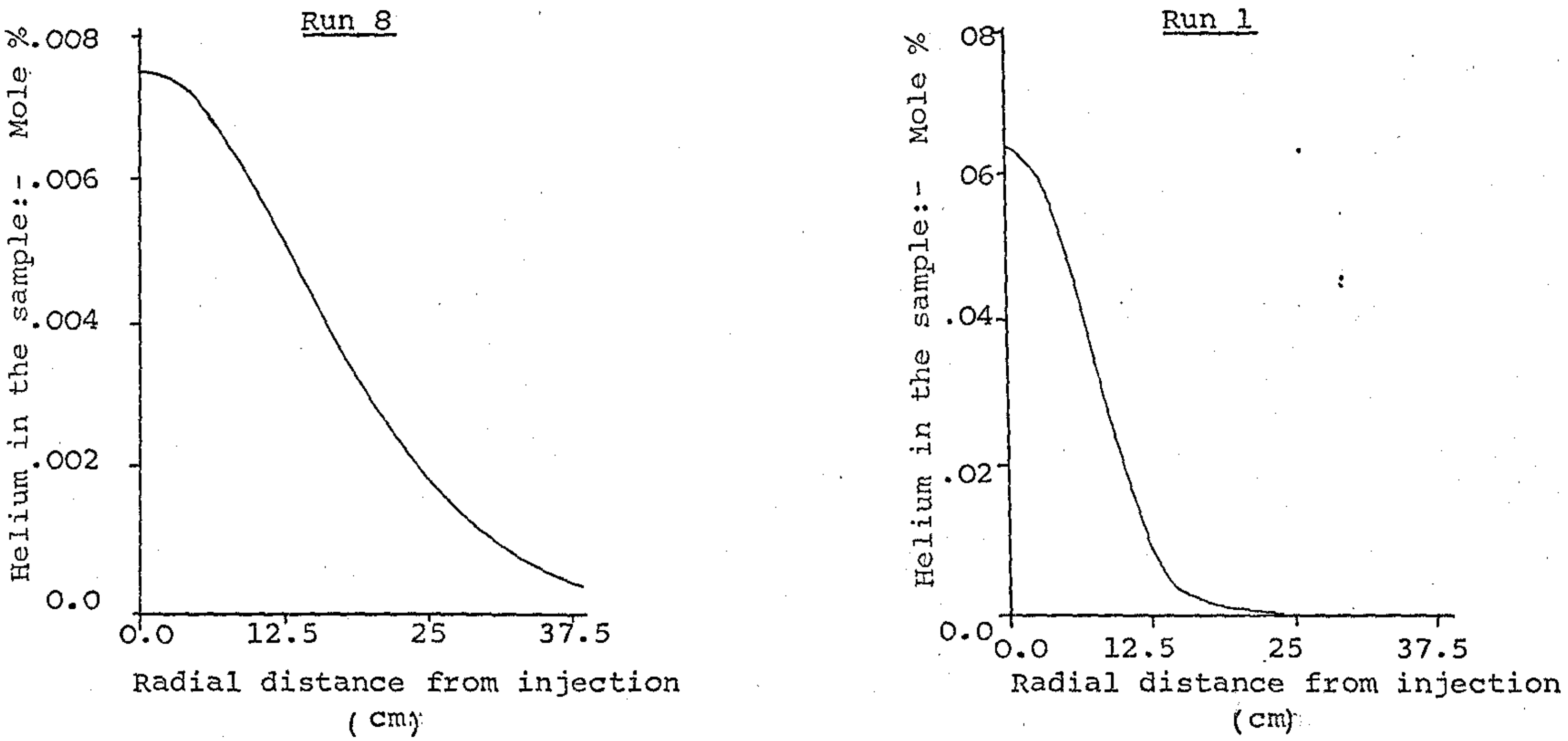


Fig 6.2 Helium in the sample stream for the two extreme experimental cases.

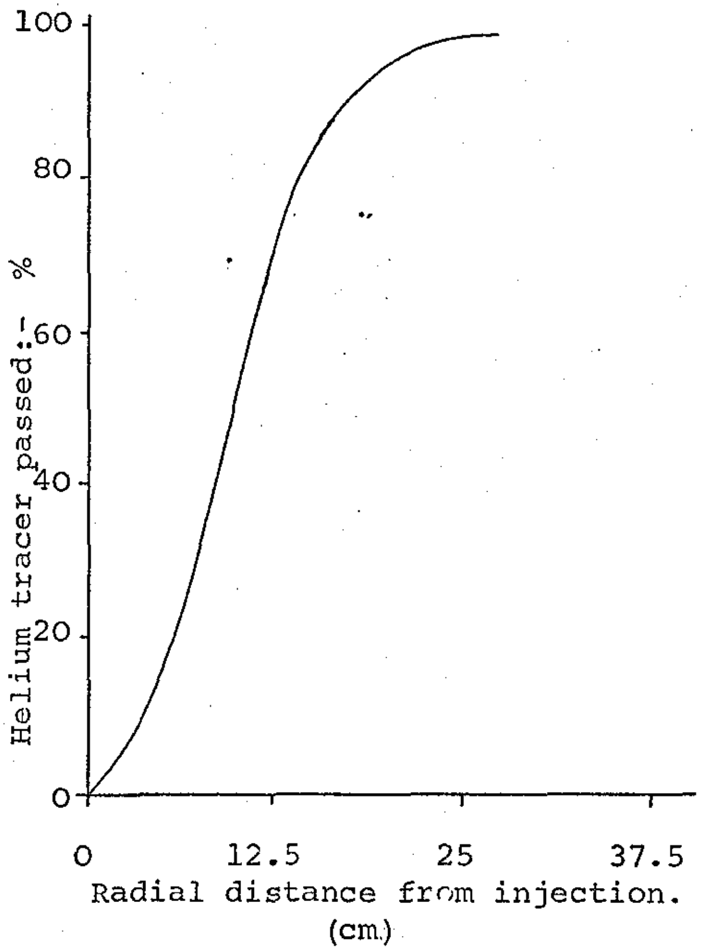


Fig 6.3 Cumulative percentage of tracer passed for Run 1 .

Fig 6.4a Experimental points for vapour mixing investigation with liquid present.

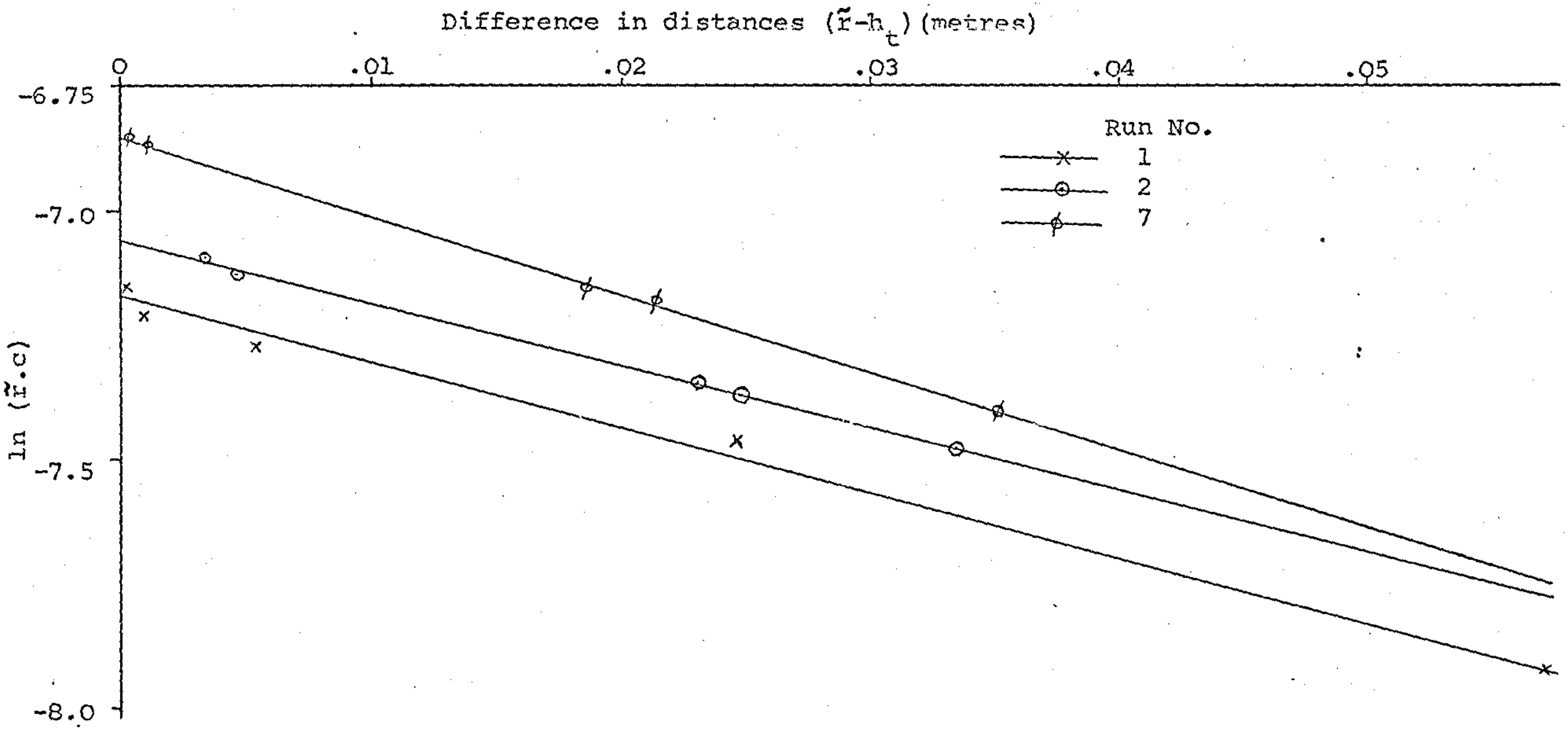
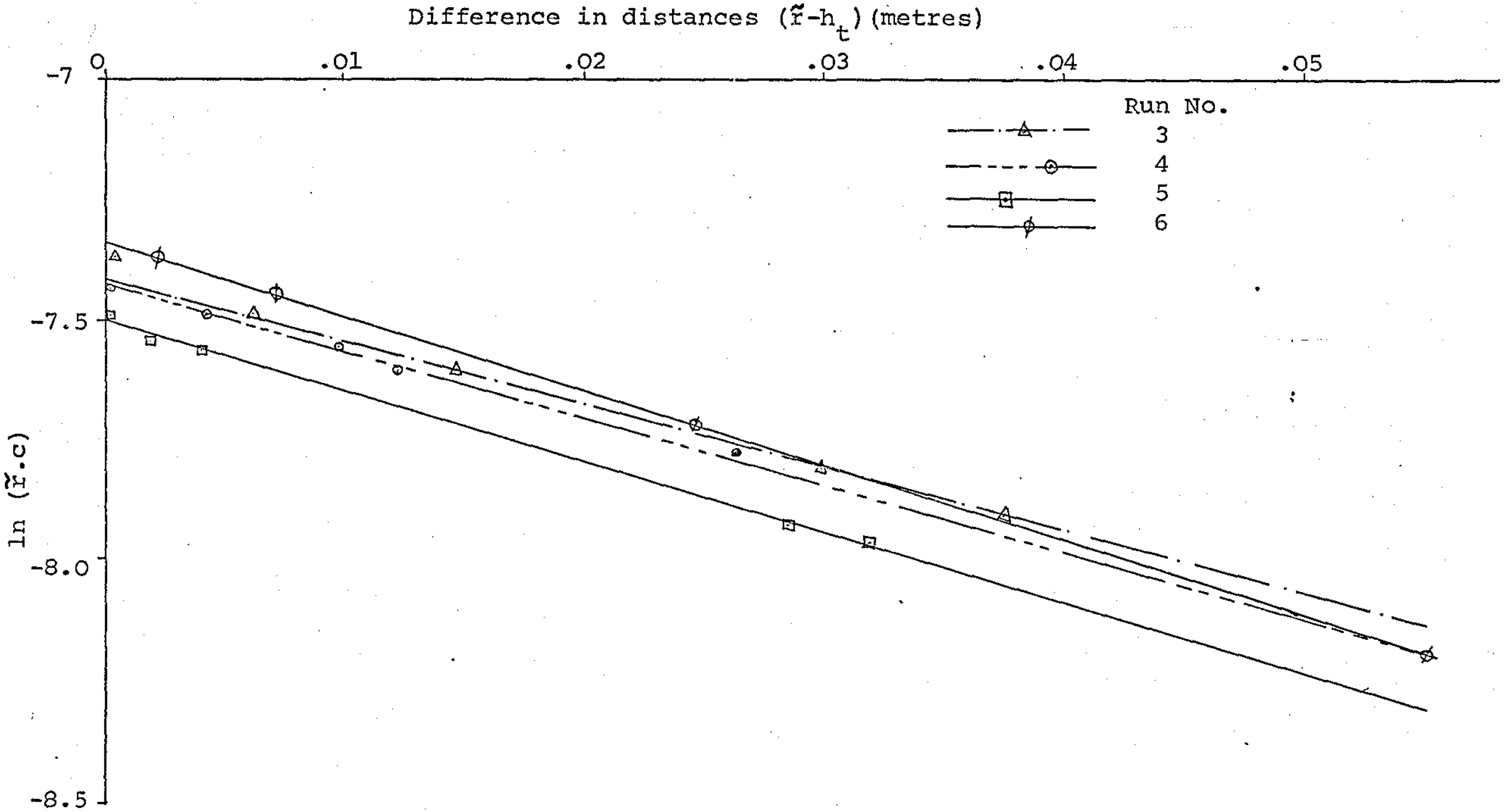


Fig 6.4b Experimental points for vapour mixing investigation with liquid present.



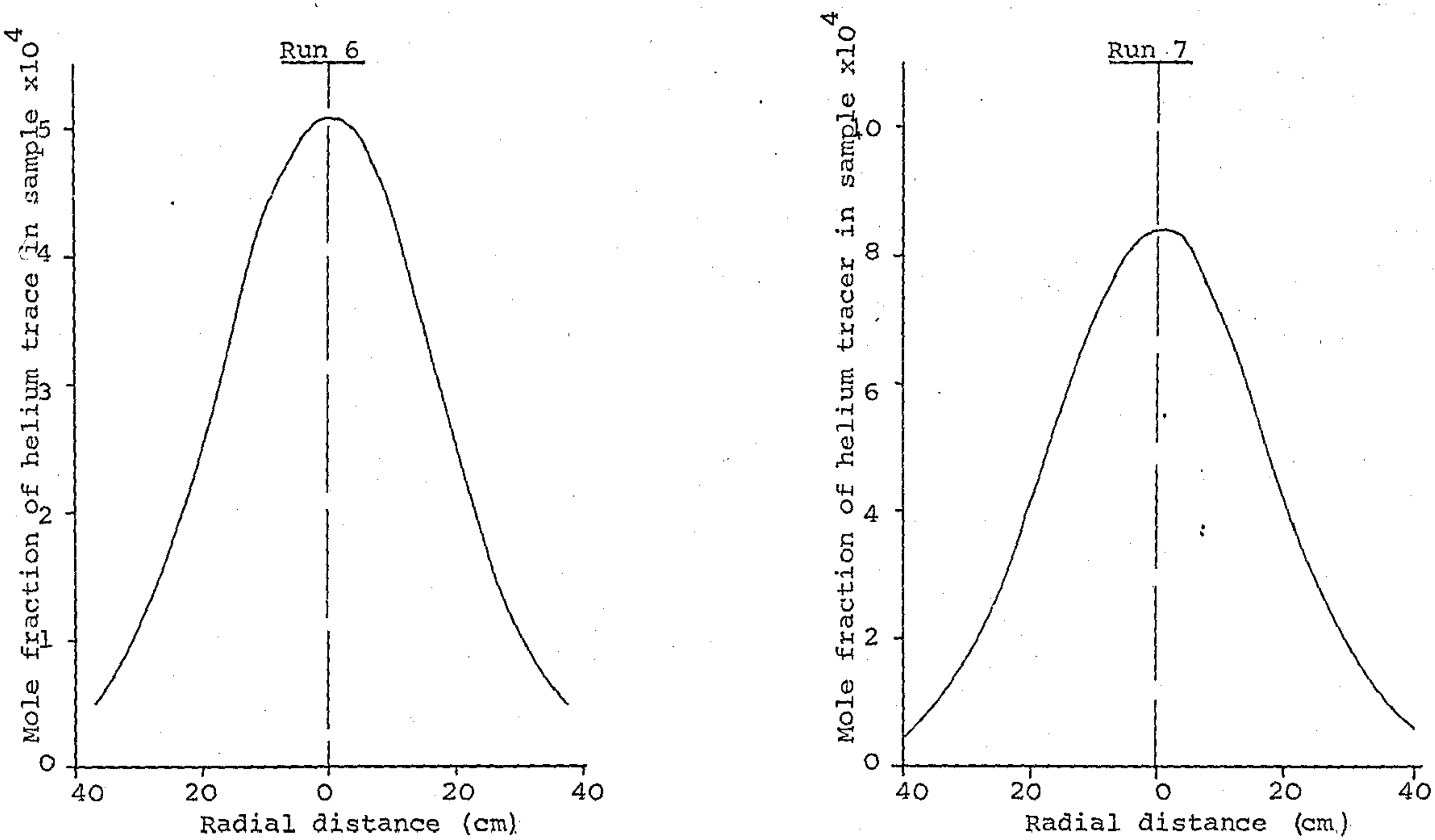


Fig 6.5 Mole fraction of helium tracer in the sample streams of the two extreme experimental cases, for when liquid was present on the tray.

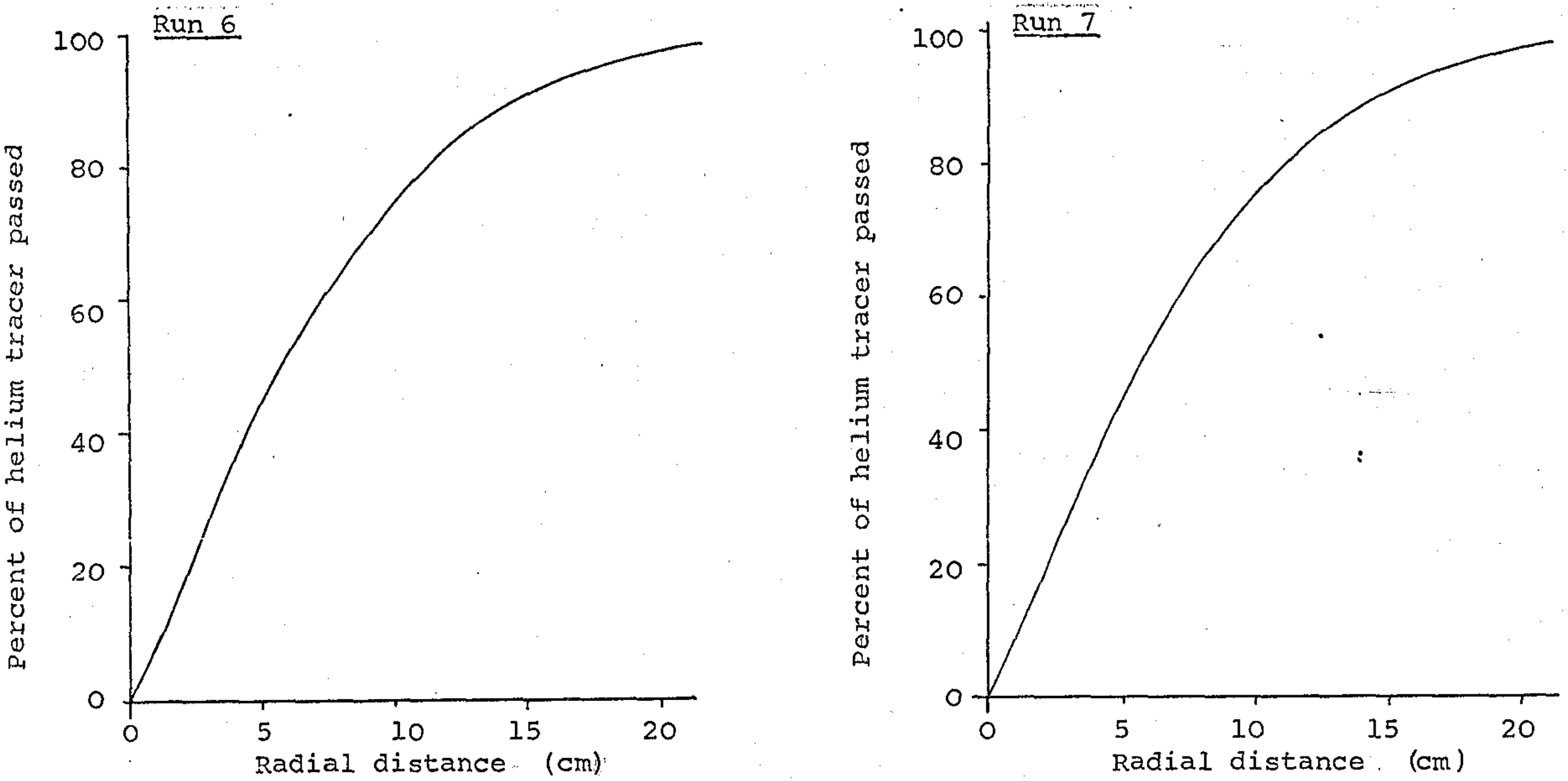


Fig 6.6 Helium tracer passed within certain radii about the injection axis.

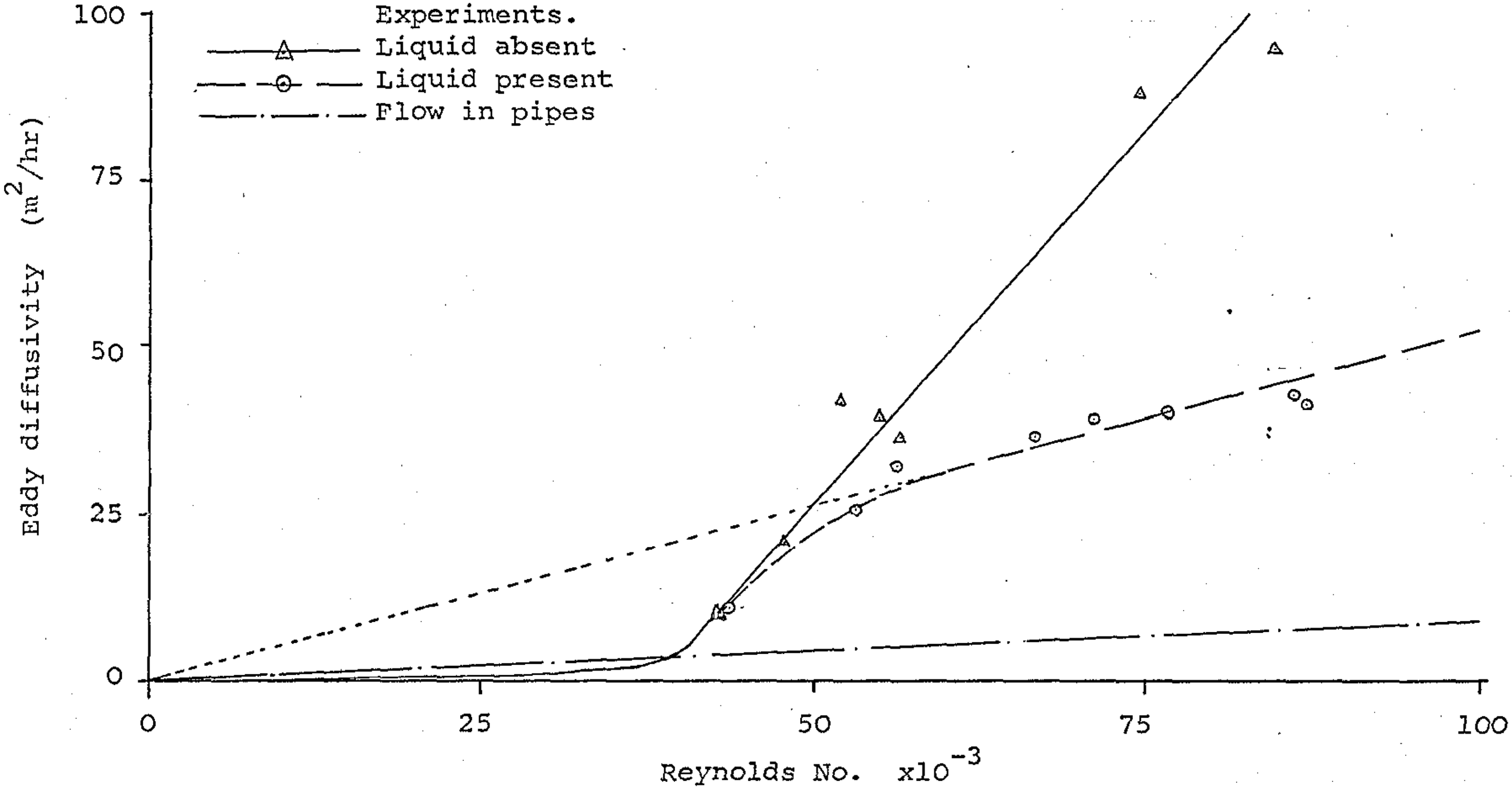


Fig 6.7 Comparison of eddy diffusivity with Reynolds No. for vapour flow in a distillation column.

7. UNSTEADY-STATE MATHEMATICAL MODELS OF PLATE COLUMNS

7. Unsteady-state Mathematical Models of Plate Columns

7.1 Literature Survey on Distillation Column Simulation

7.1.1 General Work

Critical reviews on the unsteady-state behaviour of plate distillation columns have been given by Holland (48); Rosenbrock (83), Archer and Rothfuss (2), Gould (38). Column dynamics were of main interest to chemical engineers but little work had been done before the introduction of computers when studies in batch distillation were presented by Lapidus and Amundson (58) and Pigford et al (75). Early publications concerning control of distillation columns were presented by Rosenbrock (83), Voetter (108) and Armstrong and Wilkinson (115).

The main approaches to solving the mathematical formulation of distillation processes has been the use of digital and analogue computers to obtain numerical solutions of the equations by the Laplace transformation of the equations followed by inversion to obtain the time response or by determination of the frequency response. Mah, Michaelson and Sargent (67), and Holland (48) present good reviews on numerical methods used to obtain solutions of the equations using a digital computer. Nonlinear equations are solved using integration routines such as Runge-Kutta and Adams Bashforth, while matrix integration methods are used for solving linear equations.

The disadvantage occurring with matrix methods is that the coefficients of the plant matrix are functions of time.

Methods of solution have been developed (67,85) where the coefficients of the plant matrix are assumed to be linear functions of time or treated as constants over a small range.

The frequency response and transient behaviour of a column with 16 theoretical trays was given by Lamb et al (57) who used a curved vapour-liquid equilibrium relationship and calculated by use of an analogue computer the oscillations in tray compositions due to input oscillations in reflux or feed composition. The response of top product composition to reboiler vapour composition was expressed as a transfer function which for low frequency gave a single transfer lag for which the time constant was approximately proportional to the square of the number of plates. Franks (30) also used an analogue computer when simulating the dynamics of a multicomponent, 60-plate column.

An implicit method involving an arbitrarily fixed constant θ was used by Rosenbrock (83) in his study on numerical methods. The definition of efficiency, equilibrium relationship and simple differential equations were set up for a binary system and he compared the theoretical step responses obtained with the experimental responses obtained by Wilkinson and Armstrong (3).

A different θ method for the solution of unsteady-state distillation equations was made by Waggoner and Holland (109), Groves (39) and Tetlow et al (100), while Holland (47) used the θ method in steady-state distillation column design. The θ method in these cases was effectively an implicit and corrector method.

Most of the work done using the analytical approach has been concerned with binary distillation where the equations are expressed in terms of transient deviations. Laplace transformation, derivation and inversion of the transfer function have been used by Wilkinson and Armstrong (115) to give the step response of the top product composition for feed composition changes in a column at finite reflux. A mathematical solution for the simplified equations is given showing good agreement between experimental and theoretical values for the early part of the response. The solution for the tail end of the response was given by Wood and Armstrong (117).

Izawa and Morinaga (51) presented the transfer functions based on material balance equations of a binary distillation column for changes in flow rates and feed composition. The existence of dead time was found as the liquid flow lag caused a large phase lag in the high frequency range. This adversely affects the controllability of the column.

The dynamic behaviour of a binary distillation column in the vicinity of an equilibrium state was investigated by Pohjola and Norden (76). The process was divided into two subprocesses, mass transfer and liquid flow and the total response for small perturbations in the feed composition was the sum of the two independently obtained responses. Further investigations in the attempt to derive simplified models and finding limiting cases have been carried out (4,56a,57).

Except for the work presented by Holland et al (39,48,100,109), the works discussed in this review on distillation column simulation all assume the liquid on the trays and the vapour between trays to be perfectly mixed. All assume constant plate efficiency and most make the specification of perfect liquid mixing as efficient numerical integration routines were not available for computing the responses of distillation columns where partial liquid mixing was present. Further the responses of large dimensional models assume perfect liquid mixing, giving the Murphree plate efficiency equal to the point efficiency.

Two of the main investigations on unsteady-state distillation simulation where liquid mixing occurs and the study of charges in plate efficiency are reviewed in the next section.

7.1.2 Column Models and Review of Main Work

Holland (48) studied distillation responses using a generalised model with perfect mixing on the plate and bypassing of the downcomer. A further extension of his work was to use the model presented by Tetlow (99), where any transfer lags, mixing and channelling are accounted for in the downcomer. This model is basically the same as his generalised model except the downcomer holdup is split into three different parts (Fig. 7.1):-

a) Perfect mixing b) Plug flow c) Bypassing

Molal holdups are calculated using the usual empirical equations, but the fraction of the liquid stream ($\lambda_M, \lambda_D, \lambda_C$) passing through the above three sections seems to be arbitrary, except they must sum to unity. From the responses of the top and reboiler products given in an example, it is shown that there is a great difference between assuming perfect mixing and partial mixing of the liquid phase. However, he considers that all the mixing can be described in the downcomer. This is more than likely untrue as only the downcomer holdup was considered while that of the plate and downcomer should have been considered.

When investigating the changes in plate efficiency

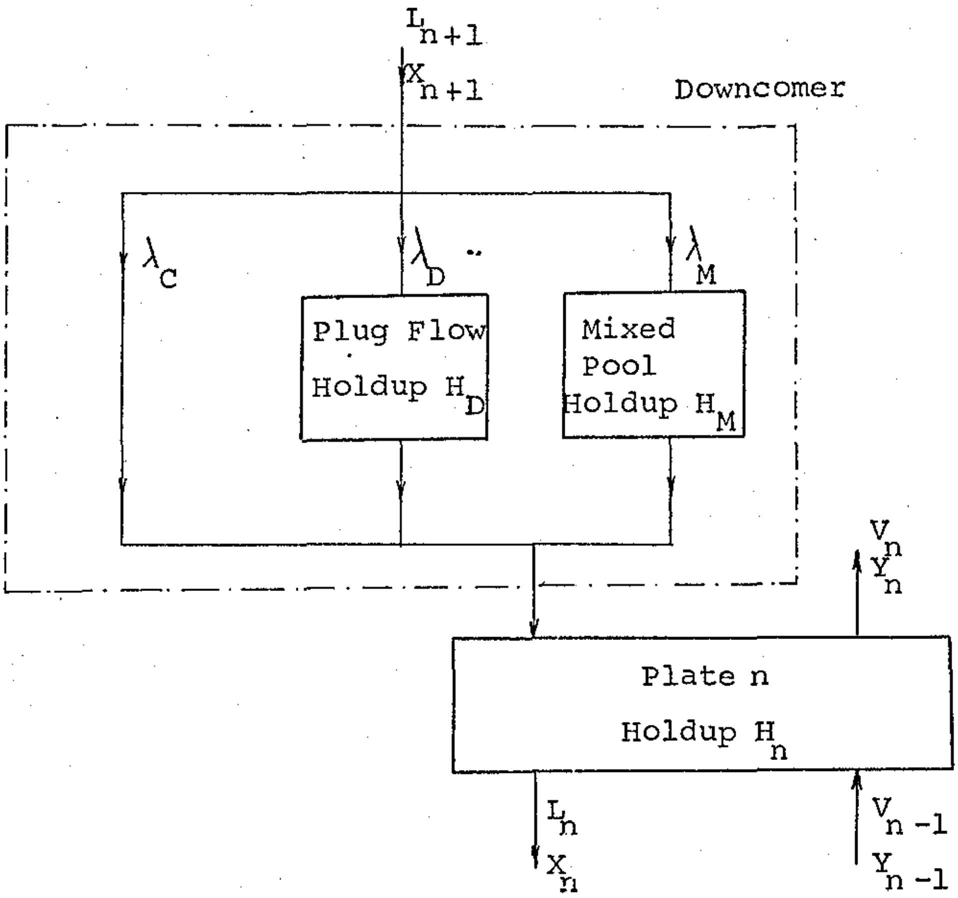


Fig 7.1 Theoretical model presented by Tetlow (99).
λ = the fraction entering each section of the downcomer,
$\lambda_C + \lambda_D + \lambda_M = 1$

for a multicomponent system, Holland (48) ignores the mixing effects by setting λ_C to unity thus giving a series of first order differential equations for the plates. Groves (39), who continued Hollands work on plate efficiencies, stated that the convergence of the (iteration) θ method for a new set of inputs was acceptable for the general model where λ_C is 1, but for the more realistic model convergence was slow. He further stated that more data must be known to be able to describe the model more fully and that the existing model is of limited use due to the great amount of computer storage needed.

A more explicit investigation into the effect of unsteady-state operating conditions on plate efficiency was presented by Huang and Yaws (50). A single bubble-cap tray was considered and the liquid mixing was described by the diffusion model. Impulse, sinusoidal and step inputs for the incoming liquid concentration were introduced to the plate. Keeping the mass transfer and point efficiency constant, the changes in efficiency ratio were plotted against dimensionless time for the disturbances introduced. From these plots it could be shown that the Murphree plate efficiency changes drastically during unsteady state but after about two residence times of the liquid crossing the plate, steady-state was approached. These observations were caused by the facts that

1) the model was dependent on constant vapour composition entering the plate.

2) the effect of the disturbance on adjacent trays was not considered.

3) from the definitions of the disturbances

introduced:- Xp, Xs the responses showed large differences in the efficiency ratio. This was due to the magnitude of the disturbances introduced.

Example:- in the responses given

$$\sigma_s = \frac{\Delta Xs}{X_{n+1}-X_e} = 1 \qquad \ldots\ldots\ldots \quad 7.1$$

where $X_e = (Y_{n-1}-b)/m$, $Y^* = mX + b$ 7.2

Taking m = .875 and b = .13 which is equivalent to α = 1.3 then for the feed plate at steady-state:-

$Xe = .427$, $X_{n+1} = .51218$, $\Delta Xs = .08518$

where X_{n+1} is the composition of the total incoming flow and is equal to the respective proportions of the actual feed composition of 0.5 at 300 moles/hr and the liquid from the plate above of 0.5174 at 700 moles/hr. Thus the new total input is 0.59736 which means that the actual feed composition must change from 0.5 to 0.7839, as the composition of the plate above is assumed constant, which is a 56% change in feed composition. For systems where the relative volatility is greater than 1.3 this change increases and for one real system of methanol and water which was studied, the feed composition had to change, in order to give a unit step, from 0.3 to 1.3 which is impossible.

In industry, disturbances of about 0 to 20% occur and it is therefore suggested that the above disturbances are practically unrealistic and that the plate efficiency will not vary as much.

7.2 Proposed Unsteady-State Models

7.2.1 Mixing Models and Assumptions

The model used to describe the characteristics of a large plate column included liquid mixing and vapour mixing models. The liquid mixing characteristics were represented by a pools-in-series with backflow model and the number of pools used was governed, with one exception described later, by one pool per foot of column diameter with the downcomer represented by one of these pools. The volumes of these pools were proportional to the molal holdups on the active tray and downcomer respectively. The backflow ratio was determined by the equation given in section 5.5:-

$$B = \frac{M}{Pe} - 0.5 \qquad \ldots\ldots \quad 7.3$$

The vapour model was represented by the vapour cell model, but as the vapour mixing has been shown to be negligible, the vapour was assumed to rise to the mixed pool directly above during simulation with the one exception when perfect vapour mixing was assumed.

The estimation of the eddy diffusivity was carried out using the empirical equations derived by the author.

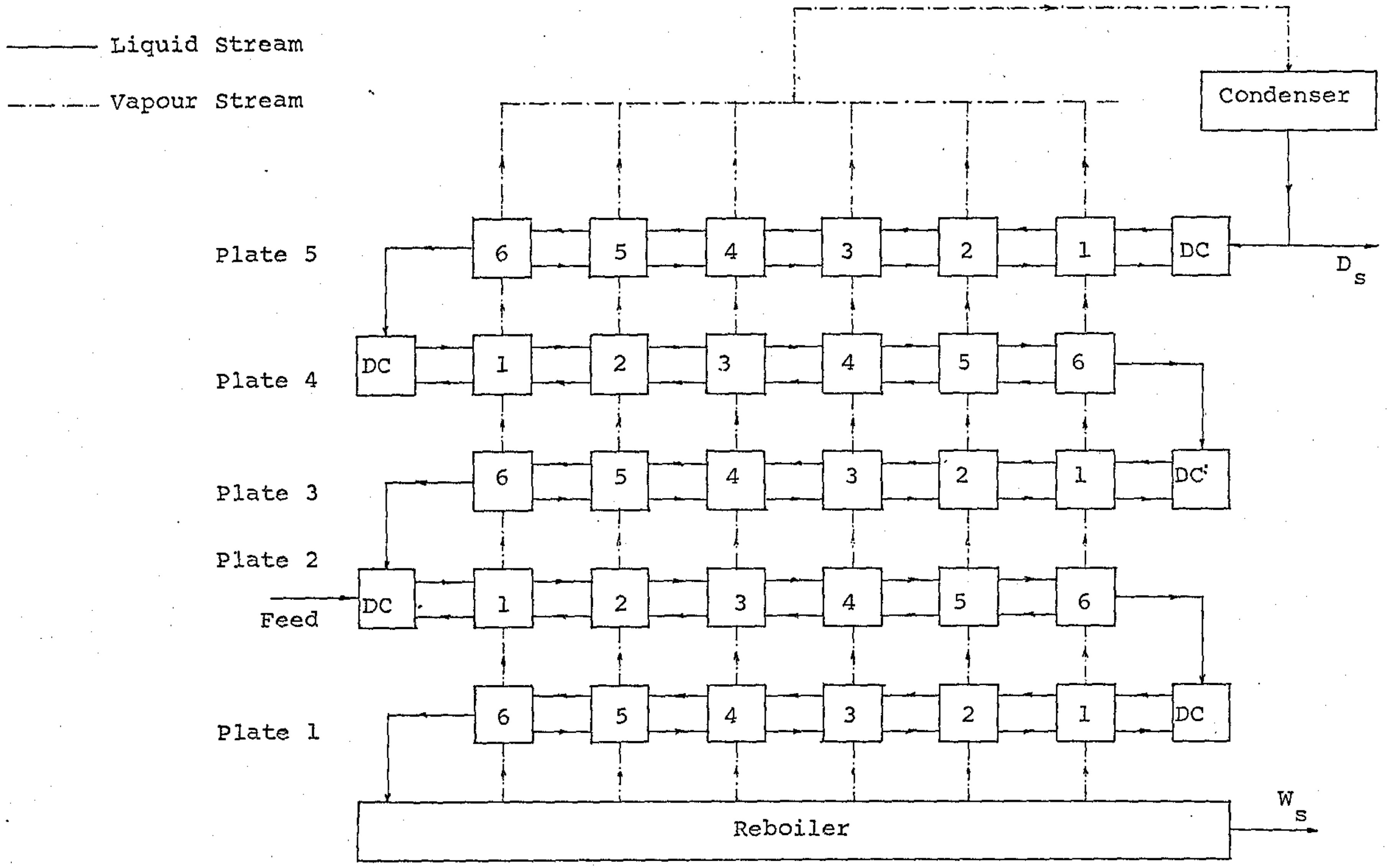


Fig 7.2 Theoretical column model showing liquid and vapour streams.

An illustration of the column model is given in Fig. 7.2 for a five-plate column of approximately 7 ft diameter. This model is referred to as the generalised model.

The main assumptions of the model are:-

1) the liquid crossing the plate is only partially mixed and the pools-in-series with backflow model is a good representation of the mixing.

2) the downcomer can be represented by one pool which no vapour leaves or enters.

3) the liquid flow leaving the tray and the backflow are constant across the tray.

4) there is no vapour holdup in the liquid phase and the latent heat of vaporisation is constant:-

$$V_{n-1} = V_n = V$$

5) as the vapour holdup above a plate is a function of pressure above the plate, the vapour holdup is assumed negligible compared to that of the liquid and can be ignored.

6) any changes in molar holdup on the trays are so rapid that they can be included in the arithmetic equations and do not warrant any differential equations.

7) the estimated point efficiency for each tray is the same for each liquid pool on that tray.

8) there is no heat lost by the column.

7.2.2 Material Balance Equations for the Generalised Model

The rate of change in molal holdup for a plate n, where F is zero except for the feed plate, is given by:-

$$\frac{dH_n}{dt} = L_{n+1} - L_n + QF \qquad \ldots\ldots \quad 7.4$$

For a single pool j on the plate,this equation is:-

$$\frac{dH_{n,j}}{dt} = \frac{H_{n,j}}{H_n} \left(L_{n+1} - L_n + QF\right) \qquad \ldots\ldots \quad 7.5$$

however, in the computation, these equations have been worked out arithmetically. The number of pools representing the active tray is M.

Plate n

Downcomer:- (Pool 0)

$$\frac{dX_{n,0}}{dt} = \left\{L_{n+1}X_{n+1,M} + B_{f,n}X_{n,1} - (L_n + B_{f,n})X_{n,0} - X_{n,0}(L_{n+1} - L_n)\frac{H_{n,0}}{H_n}\right\}\frac{1}{H_{n,0}} \qquad \ldots\ldots \quad 7.6$$

Dividing by L_n and defining T_0 as $\frac{H_{n,0}}{L_n}$ gives:-

$$\frac{dX_{n,0}}{dt} = \left\{\frac{L_{n+1}}{L_n}X_{n+1,M} + B_nX_{n,1} - (1+B_n)X_{n,0} - X_{n,0}\left(\frac{L_{n+1}}{L_n} - 1\right)\frac{H_{n,0}}{H_n}\right\}\frac{1}{T_0} \quad \ldots 7.7$$

Centre pools:- $1 < j \leqslant M$ Dividing by L_n and substituting for T_j, $V = V_s$ below feed plate and V_B above feed plate.

$$\frac{dX_{n,j}}{dt} = \Big\{ B_n X_{n,j+1} - (1+2B_n)X_{n,j} + (1+B_n)X_{n,j-1} + \frac{V}{ML_n}(Y_{n-1,j} - Y_{n,j})$$

$$- X_{n,j}\Big\{\frac{L_{n+1}}{L_n} - 1\Big\}\frac{H_{n,j}}{H_n}\Big\} \frac{1}{T_j} \qquad \ldots\ldots \quad 7.8$$

Exit pool:-

$$\frac{dX_{n,M}}{dt} = \Big\{ (1+B_n)X_{n,M-1} - (1+B)X_{n,M} + \frac{V}{ML_n}(Y_{n-1,1} - Y_{n,1})$$

$$- X_{n,1}\Big\{\frac{L_{n+1}}{L_n} - 1\Big\}\frac{H_{n,M}}{H_n}\Big\} \frac{1}{T_M} \qquad \ldots\ldots \quad 7.9$$

Feed plate

$$L_n = L_{n+1} + QF \quad ; \quad V_B = V_s + (1-Q)F \qquad \ldots\ldots \quad 7.10$$

Downcomer:-

$$\frac{dX_{n,0}}{dt} = \Big\{ \frac{(L_{n+1}X_{n+1,M} + QFX_f)}{L_n} + B_n X_{n,1} - (1+B_n)X_{n,0}$$

$$- X_{n,0}\Big\{\frac{(L_{n+1}+QF)}{L_n} - 1\Big\} \frac{H_{n,0}}{H_n}\Big\}\frac{1}{T_0} \qquad \ldots\ldots \quad 7.11$$

Centre pools:- $1 < j \leq M$

$$\frac{dX_{n,j}}{dt} = \Big\{ B_n X_{n,j+1} - (1+2B_n)X_{n,j} + (1+B_n)X_{n,j-1} + \frac{(V_s Y_{n-1,j} - V_B Y_{n,j})}{ML_n}$$

$$- X_{n,j}\Big\{\frac{(L_{n+1}+QF)}{L_n} - 1\Big\} \frac{H_{n,j}}{H_n}\Big\}\frac{1}{T_j} \qquad \ldots\ldots \quad 7.12$$

Exit pool:-

$$\frac{dX_{n,M}}{dt} = \Big\{ (1+B_n)X_{n,M-1} - (1+B_n)X_{n,M} + \frac{(V_s Y_{n-1,1} - V_B Y_{n,1})}{ML_n}$$

$$- X_{n,M}\Big\{\frac{(L_{n+1}+QF)}{L_n} - 1\Big\} \frac{H_{n,M}}{H_n}\Big\}\frac{1}{T_M} \qquad \ldots\ldots \quad 7.13$$

Q line conditions:-

Liquid below bubble point $Q < 1.0$
Saturated liquid $Q = 1$
Liquid and vapour $1.0 > Q > 0$ 7.14
Saturated vapour $Q = 0$
Superheated vapour. $Q < 0$

Reboiler:-

It was assumed that the molal holdup of the reboiler was constant and the bottoms flow rate 'W_s' was governed by:-

$$W_s = L_1 - V_s \qquad \text{.......} \quad 7.15$$

component balance (efficiency = 1):-

$$\frac{dX_R}{dt} = (L_1X_{1,M} - W_sX_R - V_sY_1^*)/H_R \qquad \text{.......} \quad 7.16$$

Condenser:-

It was assumed that the vapour was totally condensed and the liquid was returned at its boiling point. Further, any change in vapour flow was assumed only to affect the top product flow 'D_s'.

$$L_{N+1} + D_s = V_B \qquad \text{.......} \quad 7.17$$

$$R = \frac{L_{N+1}}{D_s} \qquad \text{.......} \quad 7.18$$

component balance:-

$$\frac{dX_{N+1}}{dt} = (\bar{Y}_N - X_{N+1})\,\frac{V_B}{H_C} \qquad \text{.......} \quad 7.19$$

The equilibrium relationship was estimated using constant relative volatilities or, when comparing with the linear model, linear equilibrium relationships.

7.2.3 Material Balance Equations for the Linearised Model.

The linearised model assumed constant Murphree plate efficiency and constant molal flow rates as only feed composition changes were made and compared with the results of the generalised model.

Defining

$X_n = \bar{X}_n + x_n$ where $\bar{X}_n$ = the steady-state value

and x_n = the deviation from steady-state

Equilibrium

$$y_n^* = m_n x_n \qquad \ldots\ldots\ldots \qquad 7.20$$

Efficiency

$$E_{MV,n} = \frac{y_n - y_{n-1}}{y_n^* - y_{n-1}} \qquad \ldots\ldots\ldots \qquad 7.21$$

The Q line was assumed to be unity. $V_s = V_R = V$

Reboiler:-

$$H_R \frac{dx_R}{dt} = L_1 x_1 - W_s x_R - V y_R^* \qquad \ldots\ldots\ldots \qquad 7.22$$

General plate n:-

$$H_n\frac{dx_n}{dt} = L_{n+1}x_{n+1} - L_n x_n + V(y_{n-1} - y_n) \qquad \ldots\ldots \quad 7.23$$

Feed plate f:-

$$H_f\frac{dx_f}{dt} = L_{f+1}x_{f+1} + Fx_{ff} - L_f x_f + V(y_{f-1} - y_f) \qquad \ldots\ldots \quad 7.24$$

Condenser:-

$$H_c\frac{dx_c}{dt} = -(L_{N+1} + D_s)x_c + Vy_N \qquad \ldots\ldots \quad 7.25$$

$$y_n^* = \frac{1}{E_{MV,n}}(y_n - y_{n-1}) + y_{n-1} \qquad \ldots\ldots \quad 7.26$$

$$x_n = \frac{1}{E_{MV,n}m_n}(y_n - y_{n-1}) + \frac{y_{n-1}}{m_n} \qquad \ldots\ldots \quad 7.27$$

$$\frac{dx_n}{dt} = \frac{1}{E_{MV,n}m_n}\left\{\frac{dy_n}{dt} - \frac{dy_{n-1}}{dt}\right\} \frac{+1}{m_n}\frac{dy_{n-1}}{dt} \qquad \ldots\ldots \quad 7.28$$

Defining $\lambda_n = \frac{Vm_n}{L_n}$ and $\tau_n = \frac{H_n}{L_n}$ 7.29

Condenser:-

Assumed perfect efficiency $y_{N+1} = y_{N+1}^* = m_{N+1}x_c$

$$\frac{dx_c}{dt} = \frac{\lambda_{N+1}}{\tau_{N+1}m_{N+1}}(y_N - x_c) \qquad \ldots\ldots \quad 7.30$$

General plate:- (where $\frac{L_{n+1}}{L_n}$ is omitted as it is assumed to be unity)

$$\tau_n \frac{1}{E_{MV,n}m_n}\left\{\frac{dy_n}{dt} - \frac{dy_{n-1}}{dt}\right\} + \frac{\tau_n}{m_n}\frac{dy_{n-1}}{dt} = -\frac{1}{E_{MV,n}m_n}(y_n - y_{n-1})$$

$$-\frac{y_{n-1}}{m_n} + \frac{1}{E_{MV,n+1}m_{n+1}}(y_{n+1} - y_n) + \frac{y_n}{m_{n+1}} \quad \frac{-V}{L_n}(y_n - y_{n-1}) \qquad \ldots \quad 7.31$$

Multiplying by $E_{MV,n}m_n$ and rearranging gives:-

$$\frac{dy_n}{dt}+(E_{MV,n}-1)\frac{dy_{n-1}}{dt} = \left\{\frac{E_{MV,n}m_n}{E_{MV,n+1}m_{n+1}}y_{n+1} - \left[\frac{E_{MV,n}m_n(1-E_{MV,n+1})}{E_{MV,n+1}m_{n+1}}\right.\right.$$

$$\left.\left.+(1+\lambda_n E_{MV,n})\right]+(1-(1-\lambda_n)E_{MV,n})y_{n-1}\right\}/\tau_n \qquad \text{.... 7.32}$$

Let $\mu_n=1-(1-\lambda_n)E_{MV,n}$; $\psi_n=\frac{E_{MV,n}m_n}{E_{MV,n+1}m_{n+1}}$;

$$\alpha_n=\psi_n(1-E_{MV,n+1})+(1+\lambda_n E_{MV,n}) \qquad \text{..... 7.33}$$

$$\frac{dy_n}{dt}-(1-E_{MV,n})\frac{dy_{n-1}}{dt}=\{\psi_n y_{n+1}-\alpha_n y_n+\mu_n y_{n-1}\}/\tau_n \qquad \text{... 7.34}$$

Feed plate:-

$$\frac{dy_f}{dt}-\{1-E_{MV,f}\}\frac{dy_{f-1}}{dt}=\left\{-y_f+y_{f-1}-E_{MV,f}y_{f-1}+\frac{L_{f+1}}{l_f}\psi_f(y_{f+1}-y_f)\right.$$

$$\left.+\frac{L_{f+1}}{L_f}\psi_f E_{MV,f+1}y_f-\lambda_f E_{MV,f}(y_f-y_{f-1})+\frac{E_{MV,f}m_f Fx_{ff}}{L_f}\right\}/\tau_f$$

$$\text{....... 7.35}$$

$$\text{If } \alpha_f=\frac{L_{f+1}}{L_f}\psi_f(1-E_{MV,f+1})+(1-\lambda_f E_{MV,f}) \qquad \text{........ 7.36}$$

$$\frac{dy_f}{dt}-(1-E_{MV}\psi_f)\frac{dy_{f-1}}{dt}=\frac{\left\{\frac{L_{f+1}}{L_f}\psi_f y_{f+1}-\alpha_f y_f+\mu_f y_{f-1}+\frac{E_{MV,f}m_f Fx_{ff}}{L_f}\right\}}{\tau_f}$$

$$\text{...... 7.37}$$

Reboiler:-

$$\tau_R=\frac{H_R}{L_1} \quad ; \quad y_R^*=y_R=m_R x_R \qquad \text{.... 7.38}$$

$$\frac{\tau_R}{m_R}\frac{dy_R}{dt} = \frac{W_s y_R}{L_1 m_R} + \frac{1}{E_{MV,1} m_1}(y_1 - y_R) + \frac{y_R}{m_1} - \frac{V\ y_R}{L_1} \quad \ldots\ldots \ 7.39$$

$$\frac{dy_R}{dt} = \left\{ \frac{-W_s y_R}{L_1} - \lambda_R y_R + \psi_R y_1 - \psi_R y_R + \psi_R E_{MV,1} y_R \right\} / \tau_R \ \ldots 7.40$$

$$\frac{dy_R}{dt} = \left\{ \psi_R y_1 - \left(\lambda_R + (1-E_{MV,1})\, \psi_R + \frac{W_s}{L_1} \right) y_R \right\} / \tau_R \quad \ldots\ldots \ 7.41$$

Four unsteady-state computation equations (7.41,7.34,7.37 and 7.30) were used and solved in that order. To solve these double differential equations, the reboiler equation is solved first then the value of the differential equation for the reboiler is used in the differential equation for the plate above and so on up the column.

7.2.4 Gas Absorption Model

The gas absorption model had the same form as that of the generalised model except that there was no feed plate, reboiler or condenser. Therefore, the actual column had the same equations as that of the general plate in the generalised model (equations 7.7,7.8,7.9).

The liquid composition entering was X_{N+1} and leaving was X_O, while the vapour composition entering was Y_O and leaving was Y_{N+1}. The molal holdups, point efficiencies and all other column parameters were calculated the same way as that described for the generalised model.

7.3 Computation

7.3.1 Integration Method Used

Numerical methods used for solving the unsteady-state equations have been mentioned in section 7.1. Until recently Runge-Kutta and variations of the Runge-Kutta method, many having adjustable step lengths, were the main integration programs used. A more powerful method of integration has been presented by Gear (32,33,34). He gives a multistep predictor-corrector method whose order is automatically chosen as the method proceeds. The integration routine contains two methods of integrating,

1) an Adams predictor-corrector (32), suitable only for ordinary, non-stiff differential equations.

2) a multistep method suitable for stiff equations.

The generalised model is a stiff system and requires the multistep method as use of the Adams method makes the computing time longer by a factor of 20 or more.

The starting procedure is automatic and information retained by the program about previous steps is stored in such a way as to make the interpolation to a non-mesh point straightforward.

7.3.2 Disturbances Introduced for Unsteady-state Operation

It was shown by Huang and Yaws (50) that if large changes in the feed composition are made then there are large changes in plate efficiency during the unsteady-state period. In industry the disturbances that occur in stagewise processes, are usually not greater than 20% of the original value of the parameter changed and in linear models used for simulation, changes are not greater than 10% of the steady-state value. For this reason, disturbances introduced for unsteady-state operation will not exceed 20% of the parameters original value.

The disturbances introduced for the generalised model of a large plate binary distillation column were:-

1) changes in feed composition

2) changes in feed flow with vapour and liquid reflux flow held constant

3) changes in liquid reflux with vapour flow held constant

4) changes in vapour flow with liquid reflux held constant

For vapour changes the simulation has been simplified for the case where liquid dumping occurs. For this case it is assumed that the liquid dumped from the plate above is instantaneously mixed with the liquid in the downcomer of the plate below and that after the new molal holdups have been calculated, constant molal flow exists.

The disturbances introduced for the linear model were changes in feed composition. The equilibrium relationship was linear and the results were compared with that of the generalised model with a linear equilibrium relationship.

The generalised model of the gas absorption column was used to investigate extreme values of the absorption factor and only liquid feed compositions changes were investigated.

7.4 Determination of Steady-State Variables

The steady-state values for both liquid and vapour compositions were estimated using the unsteady-state simulation program, but no disturbances were introduced and the program was allowed to continue until the estimated top product determined by:-

$$ETP = QFX_{ff} - W_s X_R \qquad \ldots\ldots\ 7.42$$

was within 0.0001 of the calculated top product defined by:-

$$CTP = D_s X_c \qquad \ldots\ldots\ 7.43$$

and the values of liquid and vapour composition throughout the column were constant.

Steady-state values were estimated using the generalised model for the actual systems of methanol-water; benzene-toluene and methylcyclohexane-toluene. The equilibrium relationships were represented by polynomials calculated by doing a least squares fit on the data presented in the literature (16). Working with real systems causes the problem that any results found cannot be directly related to any specific distillation parameter and therefore most work was carried out using constant relative volatilities or a linear equilibrium relationship.

The steady-state values for the linear model were taken from those estimated by the generalised model. The plate efficiencies were also those calculated for the generalised model.

The gas absorption column steady-state values were estimated using a linear equilibrium relationship whose slope was in the range of 10 to 500. These values are typical of the removal of carbon dioxide from water by air stripping. The values of the plate efficiency were in the range of 0.1% to 10% as the system was liquid film limiting while the binary distillation was gas film limiting and thus gave higher values of plate efficiency.

8 UNSTEADY-STATE SIMULATION OF LARGE PLATE COLUMNS

8 Unsteady-State Simulation of Large Plate Columns

8.1 The Simulation Program

The column dimensions, number of plates, liquid pools, feed plate number, weir height, flow rates, physical properties of the system and the steady-state compositions of all the liquid pools were required as input data for the program. The simulation time for which the program runs at steady-state was also read in. At the end of this time, a step change of not more than 20% of the original value was made in one of the system inputs. The total simulation time was 4 times that of the period operated at steady state.

Although the number of plates and liquid pools could be varied, most simulations were carried out using a 5-plate column with the feed input situated on the second tray and 6 liquid pools representing the active area. Each column consisted of a reboiler and condenser of specified size and the column diameter was usually 7 ft.

A stiff integration method was needed for solving the generalised model otherwise the computation time increased by a factor of 20.

Equilibrium data for real systems (16) were used for the initial simulations, but as no direct relationship between transient responses and system parameters could be made, constant relative volatilities were used instead.

For the linear model simulation, the plate efficiencies calculated at steady-state, were held constant during the unsteady-state period and used to calculate the transient behaviour of the liquid and vapour phases. A listing and an explanation of the use of the integration routine and the generalised model simulation program are given in Appendix D.

8.2 Unsteady-State Plate Efficiency Results

The liquid and vapour flows and other operating conditions for the sections in this chapter are given in Table 8.1. In all the simulation work carried out by the author, the liquid composition responses and the vapour leaving the liquid pools was calculated first then the plate efficiencies were calculated from these values. The initial steady-state values for most of the simulations can be found at the beginning of Appendix F.

8.2.1 The Effect of Feed Composition Changes

The feed was introduced on the second plate with a steady-state composition of 0.5 mole fraction and step changes of ±0.1 were made in it. Constant relative volatilities were used for the equilibrium relationship.

For the system with a constant relative volatility of 2, the change in the plate efficiency was found to be small. The largest change in the plate efficiency was that of the feed plate which for a step increase in feed composition went from 0.6295 to 0.6476 in 32 seconds then dropped again to its new steady-state value of 0.6203. The transient responses of the liquid compositions for the reboiler, condenser, exit pools and downcomers for the 5 plates are shown in Fig 8.1. The transient responses for the plate efficiencies are shown in Fig 8.2. The values of the point efficiencies rose only slightly (by 0.17 of a percent) and thus can be assumed constant. Systems with a constant relative volatility greater than 2 showed even less of a change in plate efficiency.

Using a lower relative volatility of 1.5 and introducing step changes in feed composition of ±0.1, it was seen that the change in the plate efficiency was rather greater. For the step decrease of 0.5 to 0.4 the plate efficiency for the feed plate dropped from 0.6321 to 0.5817 in 32 seconds then rose to 0.6380 and for the step increase of 0.5 to 0.6 it rose from 0.6321 to 0.6718 then dropped to 0.6276. The transient responses for the liquid phase are shown in Fig 8.3 for both these cases and the plate efficiency responses are shown in Fig 8.4. Again the point efficiency remained virtually constant.

The lowest value of the relative volatility used was 1.1. For a step increase in feed composition of 0.5 to 0.6 the plate efficiency for the feed plate rose from 0.6407 to 0.7466 in 32 seconds, an increase of 0.1 which is an increase of 16.5% on the original value. The plate efficiency then quickly fell and settled at a new steady-state value of 0.6418. Unlike the systems with higher values of relative volatility, the plate to show the greatest change in plate efficiency was the plate above the feed plate. This plate efficiency rose from 0.5713 to 0.8235 in 39 seconds then fell to a new steady-state value of 0.5761. It will be seen in later sections that when the vapour is perfectly mixed and when the equilibrium relationship is linear giving a slightly higher value of 1.1 for the relative volatility, the feed plate always shows the greatest changes in the plate efficiency for composition disturbances. The efficiency responses for the step increase in feed composition are shown in Fig 8.5 and the plot of efficiency ratios is shown in Fig 8.6. The exit pool liquid composition responses are shown for both an increase and decrease in feed composition in Fig 8.7. When the feed changes from 0.5 to 0.4 the plate efficiency drops rapidly and even goes negative. The values of the efficiency for the feed plate go from 0.6407 to -0.8177 in 36 seconds then rise to 0.6322. From these results it can be seen that a step decrease in feed composition affects the value of the plate efficiency more than a step increase.

The mean residence time for the liquid crossing a plate in the stripping section was 14 seconds and 10 seconds in the rectifiying section. From the plate efficiency response, it can be seen that the major changes only occur in the first minute after the disturbance has been introduced which is 4 times the mean residence time. For systems with relative volatilities greater than 2 the plate efficiency changes are so small that they can be ignored.

In all the above cases, the vapour was assumed to travel from the liquid pool it was leaving to the one directly above it.

8.2.2 Comparison with Vapour Phase Perfectly Mixed

For a steady-state feed composition of 0.5, step changes of $\pm$ 0.1 were introduced. Using a relative volatility of 2 showed the changes in plate efficiency to be as small as those when the vapour is unmixed but the changes were more rapid and died away before the end of one minute.

In the case when the relative volatility was 1.1, the plate efficiency responses showed the changes were large, but not as great as those where the vapour was unmixed.

When the step decrease was introduced the plate efficiency for the feed plate did not go negative but dropped from 0.6407 to 0.3727 in 18 seconds then rose again to 0.6396. The transient responses of the liquid compositions leaving the plates for the two step changes can be seen in Fig 8.8. The responses of the plate efficiencies for the decrease in feed composition are shown in Fig 8.9 and for the increase in Fig 8.10.

The difference in overall separation between the vapour being perfectly mixed and unmixed was very small and a comparison of the final steady-state liquid compositions in the reboiler and condenser for relative volatilities of 1.1 and 2 is given in Table 8.2.

Table 8.2 Comparison of simulation results for the vapour phases being unmixed and perfectly mixed.

	$\alpha = 1.1$		$\alpha = 2.0$	
Vapour	Reboiler	Condenser	Reboiler	Condenser
Unmixed	0.45967	0.54032	0.23821	0.76178
Mixed	0.45956	0.54046	0.23757	0.76229

It can be concluded from these results that whether the vapour is perfectly mixed or not, the changes in plate efficiencies are large for systems with small relative volatilities though the systems where the vapour is unmixed have slightly larger changes and take longer to respond.

8.2.3 The Effect of Feed Flow Changes

The feed flow rate at steady-state was 300 lb moles/hr and step changes of ± 50 lb moles/hr were introduced. The boilup was kept constant and any changes in the liquid flow to the reboiler were immediately followed by changes in the bottom product flow, thus keeping the reboiler holdup constant. The point efficiency changed by ± 0.01 for the bottom two trays, thus giving slightly different final steady-state values in the plate efficiency.

For systems with constant relative volatilities of 2 the changes in plate efficiency were no greater than ± 0.007 and thus the plate efficiency can be assumed constant.

Systems with relative volatilities of 1.1 showed changes of ± 0.055 in plate efficiency during the first 2 minutes of unsteady-state, but quickly reach the new steady-state values. Fig 8.12 and 8.13 show the plate efficiency responses for a decrease and increase disturbance in the feed flow.

Again it has been found that for systems with relative volatilities of 2 or over, changes in the flow rate do not affect the plate efficiencies to any appreciable extent. Liquid composition changes due to the disturbances in feed flow are small and the exit pool composition responses for the system with constant relative volatility of 1.1 are shown in Fig 8.11.

8.2.4 The Effect of Reflux Changes

The liquid flow being recycled to the column from the condenser at steady-state was 700 lb moles/hr. Step changes of ± 50 lb moles/hr were introduced and the top product flow was accordingly changed. The bottom product flow was varied to keep the reboiler holdup constant. The steady-state reflux ratio was 4.667 and the flow rate changes caused the reflux ratio to change to 3.25 or 7.5 depending whether the change was positive or negative. The liquid composition changes and the point efficiency changes were of the same sign as the disturbance introduced. The point efficiency changes were no greater than 0.02.

The plate efficiency responses for the system with a constant relative volatility α of 2 were no greater than 0.04. As the relative volatility was lowered, the unsteady-state changes in plate efficiency increased. For systems with α equal to 1.1 the plate efficiency of the top plate changed from 0.57 to 0.66 in 11 seconds for an increase in liquid reflux. The top plate showed the greatest changes and when a decrease in the liquid reflux was introduced, the plate efficiency changed from 0.57 to 0.46 in 11 seconds then rose to a new steady-state value of 0.555. The responses of the plate efficiencies for both positive and negative disturbances are shown in Fig 8.15 and 8.16. The exit pool liquid concentration transients are shown in Fig 8.14 and the point efficiency responses for both cases are shown in Fig 8.17

Again it can be concluded that liquid flow rate changes in the recycle from the condenser do not affect the plate efficiency for systems with relative volatilities greater than or equal to 2, but the changes increase with a decrease in α.

8.2.5 The Effect of Vapour Flow Rate Changes

In simulations concerning the changes in vapour flow rate, difficulties in describing the immediate changes in liquid flow rate due to dumping or accumulation of liquid on the plate arose. When the liquid molal holdup on the plate rose, as the change in molal holdup was small for the disturbances introduced,the liquid holdup was adjusted to its new steady-state value but the liquid flows and liquid compositions were not adjusted. For the case where liquid dumping from the plates occurred, it was assumed that the liquid dumped was the same composition as that of the exit pool on the plate and was dumped into the downcomer of the plate below. The amount of liquid dumped from one plate to the next was cumulative as it travelled down the column and the liquid composition in the downcomer was the average value after the liquid had been dumped.

These assumptions give inaccurate overall material balances and it was found that small amounts of liquid were either lost or gained depending on the disturbances introduced. However, it will be shown that these discrepancies can be ignored as the main effect on the system was due to the actual vapour flow rate change and not the dumping and accumulation of liquid.

For the disturbances where liquid dumping occurred, very small step changes in the downcomer liquid compositions were observed. These composition changes were so small that their effect on the system was negligible. In the case of liquid accumulation, no instantaneous changes in composition occurred.

The steady-state vapour flow rate was 850 lb moles/hr and disturbances of ± 50 lb moles/hr were introduced with the liquid recycle from the condenser remaining constant. For an increase in the vapour flow of 50 lb moles/hr, the reflux ratio changed from 4.667 to 3.5 and the point efficiency dropped by 0.01. The liquid compositions gradually dropped due to more of the volatile being taken off in the top product. The changes in composition were more noticeable for the systems with large relative volatilities and the transient responses of the exit pools are shown in Fig 8.18 for $\alpha = 2$ and for an increase in vapour flow.

The point efficiency changes were the same size regardless of the relative volatility used and the changes in plate efficiencies were also the same size and followed the same response pattern. The plate efficiencies increased for a decrease in vapour flow rate and the maximum change for a system with $\alpha = 2$ was that of plate 3 whose plate efficiency rose from 0.5577 to 0.5825 in 34 seconds then settled at a new steady- state value of 0.5644. Fig 8.19 shows the plate efficiency responses for both disturbances using a relative volatility of 1.1 and Fig 8.20 those using a relative volatility of 2.

From these responses it can be seen that the main effect was due to the changes in top product flow and point efficiency changes. It is doubtful whether the liquid dumping and accumulation, had they been accurately described in the model, would have had any effect on the overall responses.

8.2.6 The Effect of Larger Trays

The generalised model was used to investigate the difference in using larger trays of 15 ft diameter. The average Peclet number during the operation was 12 and the number of liquid pools for the active part of the tray was 9. This was the only case where 1 liquid pool for the active area was not equal to 1 ft of tray. The same physical properties of the system as those of the 7 ft diameter column were used. Only step changes in the feed composition were investigated for varying values of relative volatilities. The steady-state feed was introduced on the second plate with a composition of 0.5 and step changes of ± 0.1 were made.

For systems with large relative volatilities of 1.5 and over, very little change in the plate efficiency occurred and the changes decreased with increase in α. However, for systems with α of 1.1 or less, the changes become larger. The largest change in the plate efficiency was that of the feed plate which for a decrease in feed composition went from 1.011 to 0.5367 in 27 seconds then rose to a new steady -state value of 1.001. The changes of the other plate efficiencies were not as great, but took longer to reach their maximum changes and to settle back down to their new steady-states.

The exit pool liquid composition transients for both disturbances and relative volatility of 1.1 are shown in Fig 8.21. Their corresponding plate efficiency responses are shown in Fig 8.22 and 8.23.

8.3 Comparison of the Generalised Model with Simplified Models

8.3.1 The Linearised Model

For the comparison of the generalised model results with those of the linearised model, constant linear equilibrium relationships were used. The values of the gradient and constant were chosen so as to give relative volatilities that varied only slightly throughout the column. Using the equilibrium relationship of:-

$$Y^* = 0.975X + 0.05 \qquad \ldots\ldots\ldots \; 8.1$$

gave a relatively constant α value of 1.16. The feed entered the five plate column on the second plate with a composition of 0.5. The linearised model used constant plate efficiencies in the liquid composition calculations during the unsteady-state period.

Only step changes of ± 0.1 in the feed composition were investigated. The liquid composition transients given by the generalised model for a step decrease in composition are shown in Fig 8.24 and the plate efficiency responses are shown in Fig 8.25. The liquid composition responses given by the linear model were so close to those of the generalised model that graphically they were indistinguishable and the comparison of the results is listed in Table 8.3. From this table it can be seen that the difference between the two models was very small and the end steady-state values vary only by about one in the third decimal place.

Using different equilibrium relationships to give the effect of a larger relative volatility, showed the end steady-state values and the intermediate unsteady-state values of the end plate compositions to agree more closely between the two different models.

Table 8.3 Comparison of liquid phase responses for the generalised and linearised models.

Generalised model feed composition 0.5 to 0.4

Time (sec.)	Reboiler	1	2	3	4	5	Condenser
0.0	.43586	.4691	.4867	.5032	.5215	.5418	.56440
12.9	.43544	.4656	.4766	.5003	.5199	.5406	.56270
21.7	.43443	.4629	.4713	.4983	.5185	.5393	.56136
39.3	.43131	.4564	.4628	.4940	.5149	.5362	.55800
56.9	.42719	.4505	.4578	.4898	.5112	.5327	.55443
74.5	.42277	.4457	.4541	.4859	.5075	.5290	.55080
92.1	.41843	.4418	.4508	.4822	.5038	.5254	.54725
144.9	.40661	.4317	.4423	.4721	.4936	.5152	.53725
294.5	.38211	.4107	.4241	.4501	.4710	.4926	.51527
505.7	.36082	.3923	.4080	.4305	.4508	.4725	.49566
646.5	.35188	.3846	.4012	.4222	.4424	.4640	.48740
857.7	.34330	.3772	.3947	.4143	.4342	.4559	.47984

Linearised model feed composition 0.5 to 0.4

Time (sec.)	Reboiler	1	2	3	4	5	Condenser
0.0	.43586	.4691	.4867	.5032	.5215	.5418	.56440
12.9	.43533	.4648	.4727	.5013	.5205	.5412	.56346
21.7	.43402	.4599	.4667	.4987	.5189	.5401	.56225
39.3	.43027	.4532	.4614	.4939	.5153	.5371	.55923
56.9	.42593	.4488	.4573	.4898	.5116	.5336	.55580
74.5	.42138	.4445	.4536	.4858	.5077	.5298	.55202
92.1	.41700	.4407	.4503	.4819	.5039	.5260	.54830
144.9	.40500	.4304	.4415	.4714	.4931	.5153	.53773
294.5	.37941	.4080	.4225	.4490	.4695	.4918	.51497
505.7	.35917	.3911	.4070	.4298	.4495	.4714	.49470
646.5	.35070	.3835	.4003	.4211	.4413	.4631	.48653
857.7	.34226	.3763	.3940	.4133	.4332	.4550	.47856

It must be remembered that in the linearised model, the plate efficiencies were constant whereas in the generalised model the plate efficiencies were allowed to vary and that the two steady-state values may vary a little. This fact is the reason for the slight differences between the end steady-state composition values of the two models as it can be seen that the transients of the two models follow each other closely over the period when the plate efficiency of the generalised model was changing. If the final steady-state plate efficiencies differ appreciably from those of the initial values, then to eliminate errors in the end steady-state values, the final plate efficiencies must be used in the linear model for the unsteady-state operation.

8.3.2 The Simplified Model: Perfect Liquid Mixing

The simplified model assumes a plate with the same molal holdup, but with only one liquid pool. For normal operations, this model gives the value of the Murphree plate efficiency equal to that of the point efficiency. The simulation was carried out using the equilibrium relationship given in equation 8.1. Initially the Murphree plate efficiencies estimated by the generalised model for partial liquid mixing was held constant and used in the unsteady-state estimation. Transient responses the same as those for the linearised model were obtained for feed composition changes of ± 0.1. This shows that both the linearised and simplified model can be used if the Murphree plate efficiency estimated for partial liquid mixing is available and can be held constant without incorporating any errors.

The simplified model was also used with the Murphree plate efficiency being set equal to that of the point efficiency which was calculated according to the operating conditions present. This value of the plate efficiency was lower than that estimated by the generalised model. The overall column efficiency for this case was less than that of the previous case. The liquid compositions on each plate were lower than those predicted by the generalised model and the liquid transient responses for both this case and the previous case, where the plate efficiency was not equal to that of the point efficiency, are shown in Fig 8.26 for a step increase in feed composition.

It can be seen from this investigation that simplified models assuming perfect liquid mixing on the plate can only be used if the plate efficiency used in the unsteady-state estimation has taken the actual partial liquid mixing into account.

8.4 The Effect of Feed Composition Changes in Gas Absorption

The generalised model was used to describe the liquid and vapour mixing characteristics of a large sieve tray gas absorption column. The simulation was similar to the decarbonation of water by air and the physical properties of this system were used in the computation. The equilibrium relationship was linear and the gradients used were high giving extreme values of the absorption coefficient. The steady-state liquid feed entered the downcomer of the top plate with a composition of 0.005 and step changes of ± 0.0025 were made.

The concentration gradient on the lower plates was very small due to the low composition of the liquid. The top plate showed a very large concentration gradient thus verifying the need for the liquid mixing model.

For the system with an equilibrium slope of 31.667 giving an absorption coefficienct λ of 25.9, the plate efficiencies were about 0.09. For a step decrease in the feed composition the point efficiencies of the top plates changed (Fig 8.27) and the plate efficiencies changed giving oscillatory responses about the new steady-state plate efficiencies (Fig 8.28). It can be seen from Fig 8.28 that the final and initial steady-state plate efficiencies do not change appreciably, but the top tray plate efficiency goes to a new value corresponding to the change in the point efficiency. The liquid composition responses of the exit pools and the top two downcomers are shown in Fig 8.29. Step increases in the feed composition gave similar plate efficiency responses but in the opposite direction to those with a decrease in composition. The plate efficiency responses showed that during the unsteady-state period, the values of the efficiencies vary appreciably and the time period of the changes was greater than that for distillation.

A more extreme value of 155 for the absorption coefficient was obtained using an equilibrium slope of 190. The plate efficiencies for this system were about 0.0185. The responses of the exit pool compositions for the five plates and the downcomer compositions of the top two plates for a step decrease in feed composition are given in Fig 8.30. The corresponding responses of the point and plate efficiencies are shown in Fig 8.31 and 8.32 respectively. It can be seen that only the top tray plate efficiency strays from its original steady-state value.

Larger values of the equilibrium slope were investigated only showing similar results to those already given. Although the step changes introduced were 50% of the original composition value, it can be seen that while large changes in the plate efficiencies do occur during the unsteady-state period, they do not change appreciably between the two steady-state values.

If only the initial and final steady-state values were of interest, a linearised model using constant plate efficiencies in the liquid composition calculations could be used and the end results would be accurate solutions.

8.5 Discussion on Unsteady-State Plate Efficiencies for Binary Systems

From the investigations of unsteady-state plate efficiencies using the generalised model, it can be concluded that for systems with relative volatilities greater than 2, feed composition and liquid flow rate changes do not affect the plate efficiency. As the relative volatility decreases, so the changes in plate efficiency increase during the unsteady-state period, but they soon settle down to the new steady-state value which is almost the same as the original value.

In the case of vapour flow rate changes, the effect of liquid dumping can be ignored as it is the change in reflux ratio which changes the liquid compositions. However, the changes in the point efficiencies give rise to changes in the two steady-state plate efficiencies and must be accounted for if a linear model is to be used.

The comparison with the linear model showed that as long as the initial and final steady-state plate efficiencies were the same, a constant plate efficiency would give the same transient responses of the liquid phase and the end steady-state values would be the same. Thus although the unsteady-state plate efficiencies showed large changes in the generalised model for systems with a low relative volatility, little difference is seen if these efficiencies are held constant.

From the investigation of the gas absorption column, the plate efficiencies were shown to be more oscillatory during unsteady-state, but the original and final plate efficiencies did not differ and use of a linearised model using constant plate efficiencies would give no inaccuracies in the final steady-state values of the liquid compositions.

The same results are obtained for columns with larger diameters of 15 ft, but the responses during unsteady-state take longer to reach their maximum deviations due to the larger liquid holdup on the plates. Again constant plate efficiencies and a linearised model would give accurate steady-state values.

Table 8.1 Liquid and vapour flows and other operating conditions for the unsteady-state simulation of binary systems.

	lb moles/hr					Pe	X_f	Reflux ratio
	L_s	L_R	V	W_s	D_s			
Steady-state	1000	700	850	150	150	7.7	0.5	4.667
Unsteady-state changes in:-								
Feed composition	1000	700	850	150	150	7.7	0.4 0.6	4.667
Feed flow 1)	1050	700	850	200	150	8.2	0.5	4.667
2)	950	700	850	100	150	7.3	0.5	4.667
Reflux 1)	1050	750	850	200	100	8.2	0.5	7.5
2)	950	650	850	100	200	7.5	0.5	3.25
Vapour 1)	1000	700	800	200	100	10.5	0.5	7.0
2)	1000	700	900	100	200	6.0	0.5	3.5
Large trays								
Steady-state	6500	4500	5500	1000	1000	12.0	0.5	4.5
Unsteady-state	6500	4500	5500	1000	1000	12.0	0.4 0.6	4.5

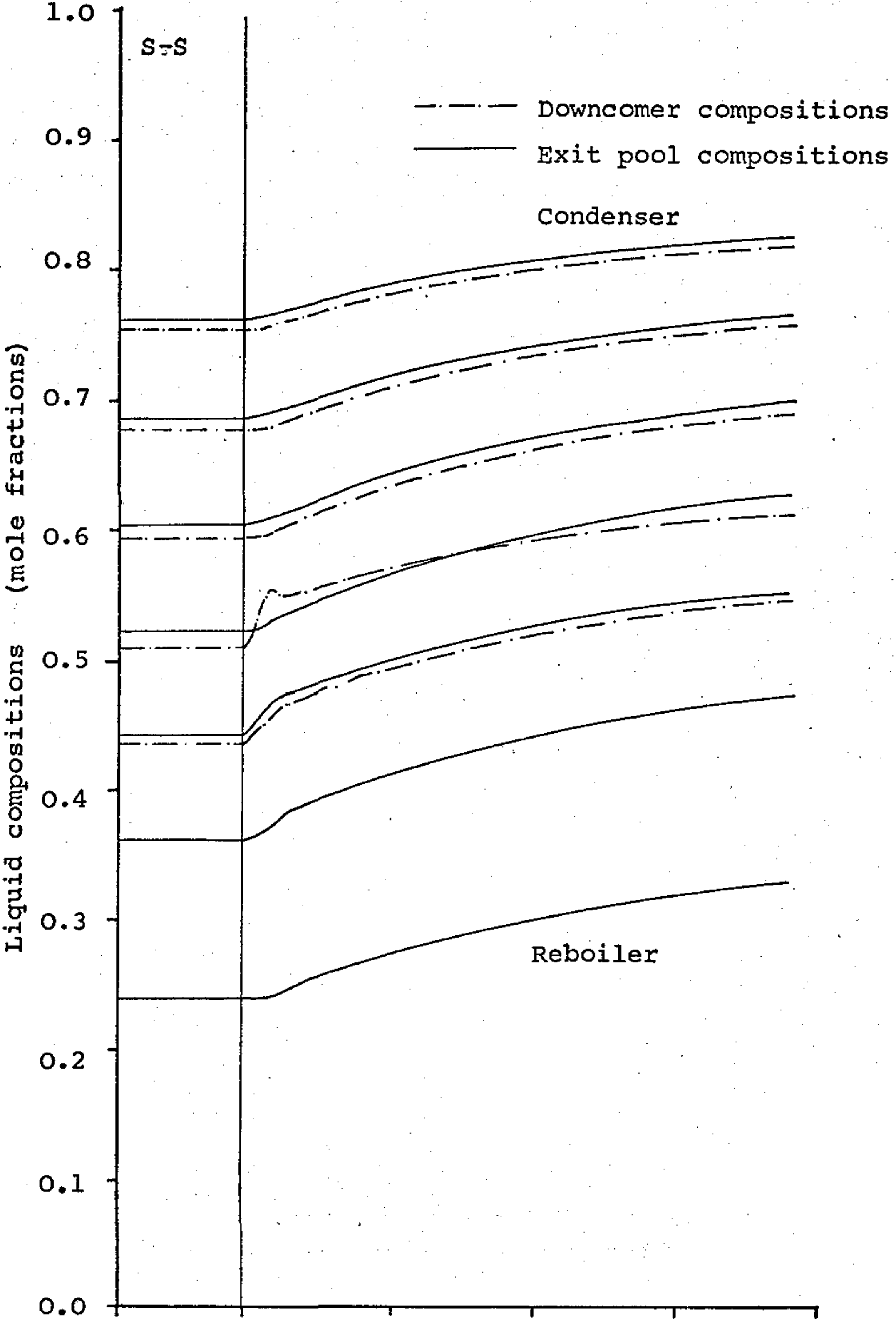


Fig 8.1 Exit liquid and downcomer composition responses for a step increase of 0.1 in the feed composition. Relative volatility of 2.

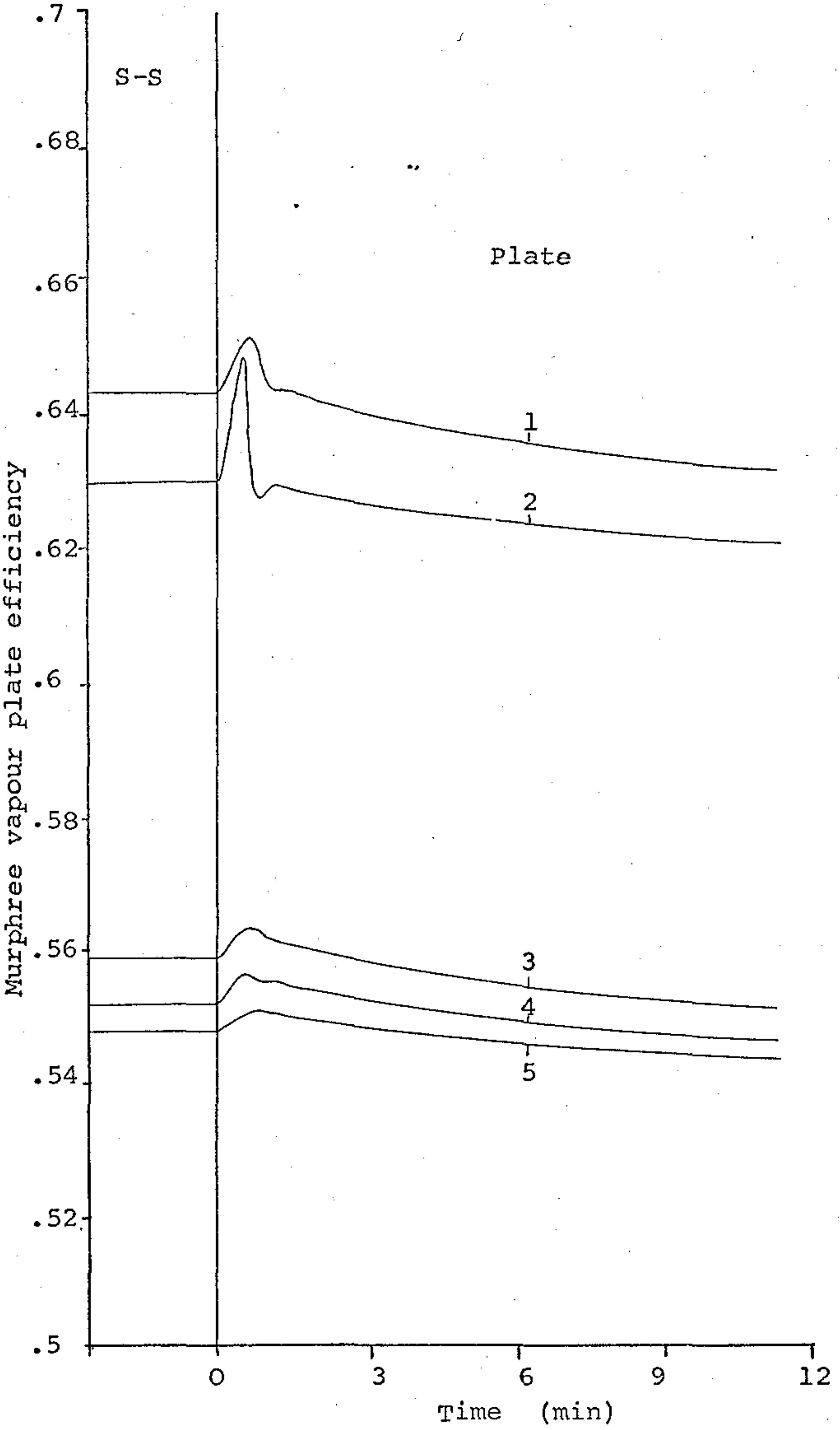


Fig 8.2 Murphree vapour plate efficiency responses for a step increase of 0.1 in the feed composition. Relative volatility of 2.

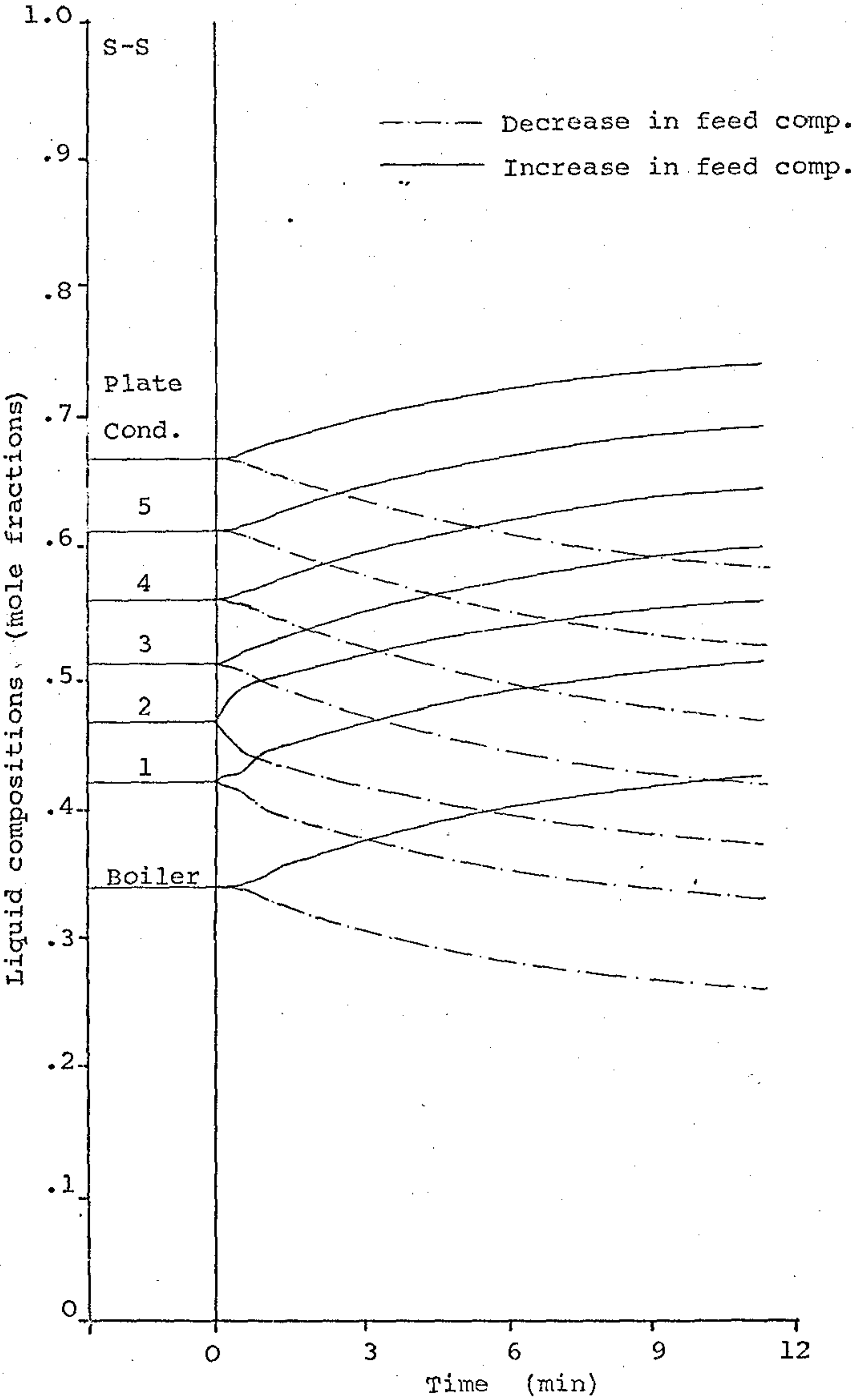


Fig 8.3 Condenser, reboiler and exit pool liquid composition responses for step changes in the feed composition. Relative volatility of 1.5.

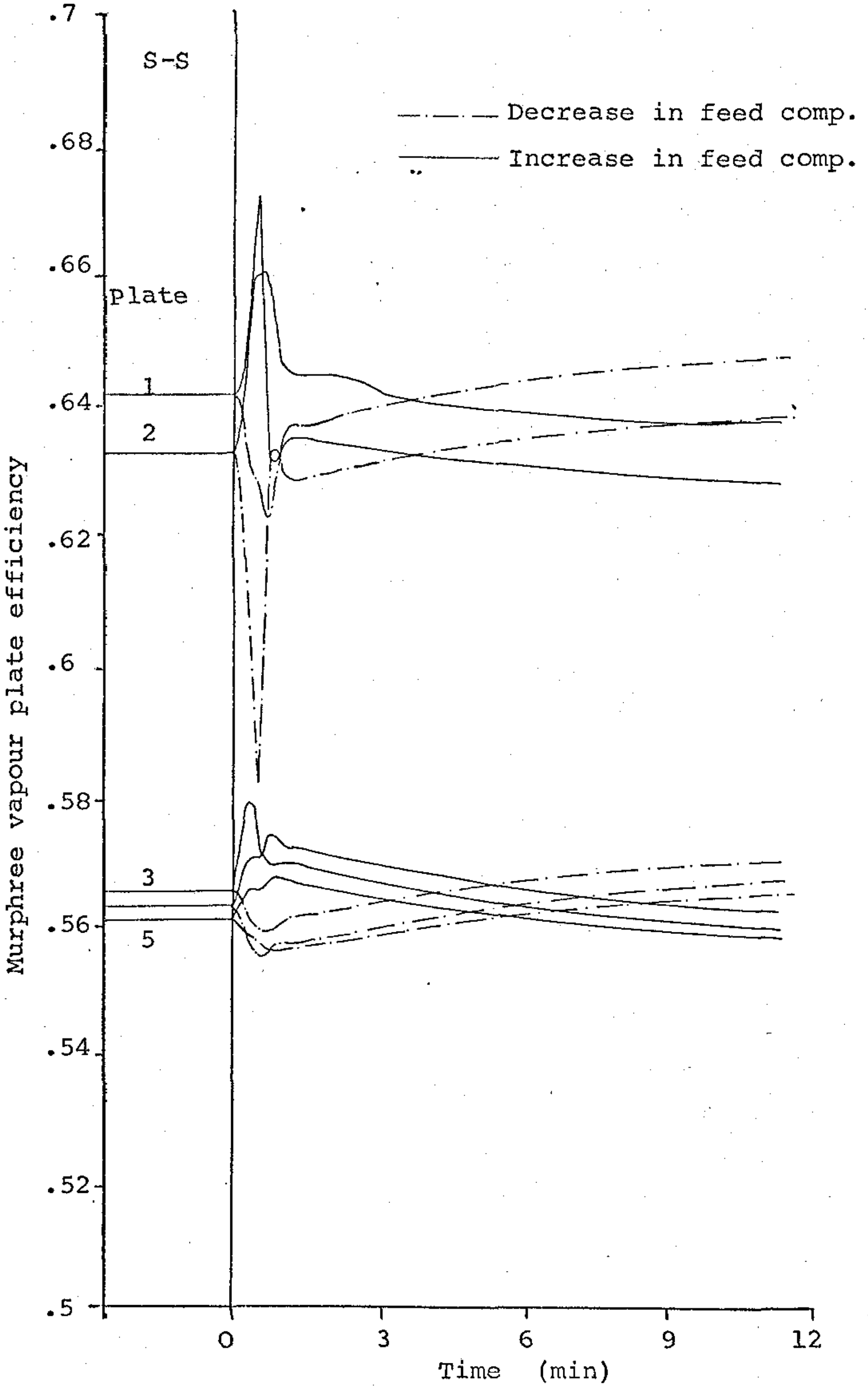


Fig 8.4 Murphree vapour plate efficiency responses for step changes in feed composition. Relative volatility of 1.5 .

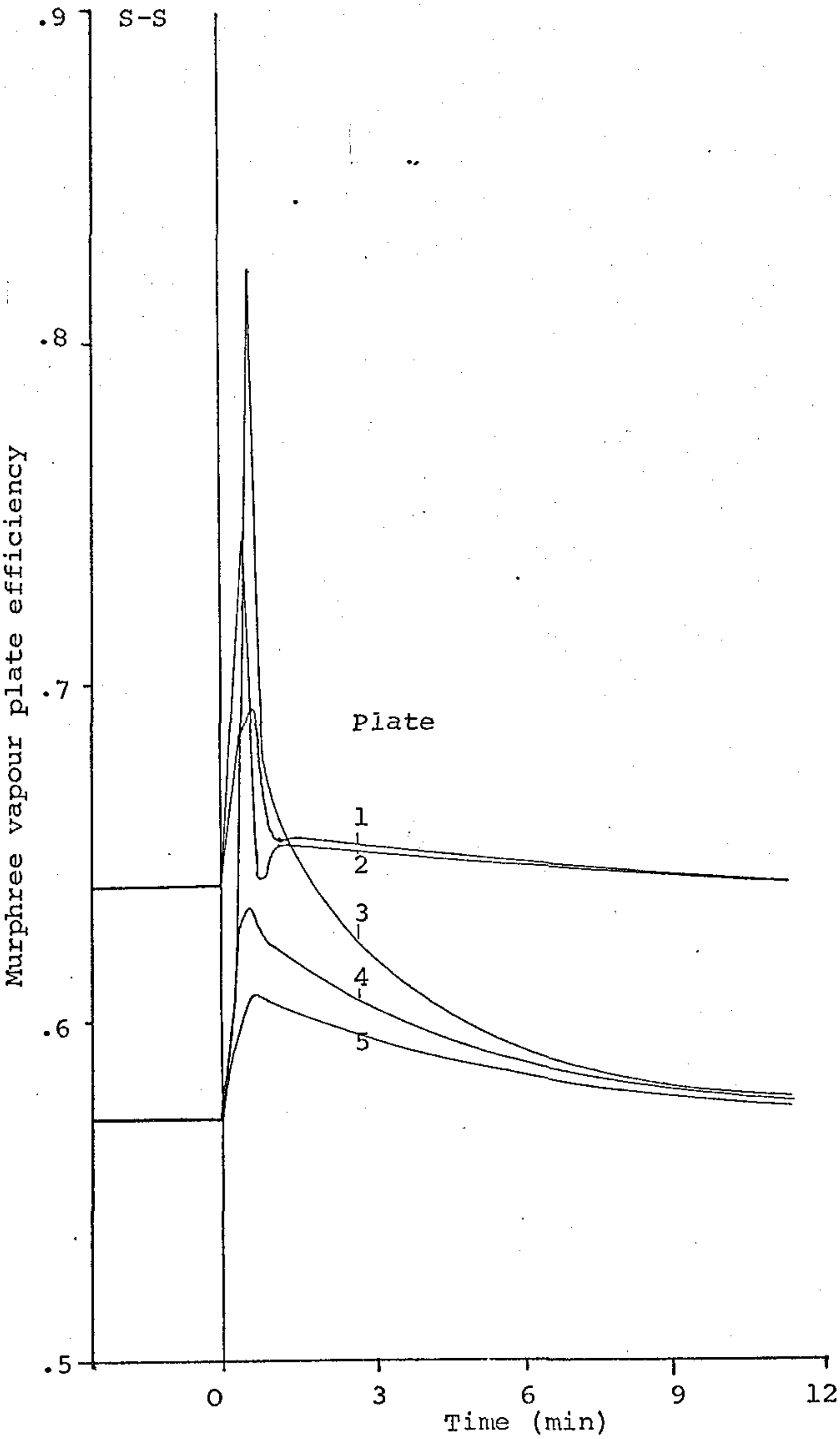


Fig 8.5 Murphree vapour plate efficiency responses for a step increase in feed composition. Relative volatility of 1.1 .

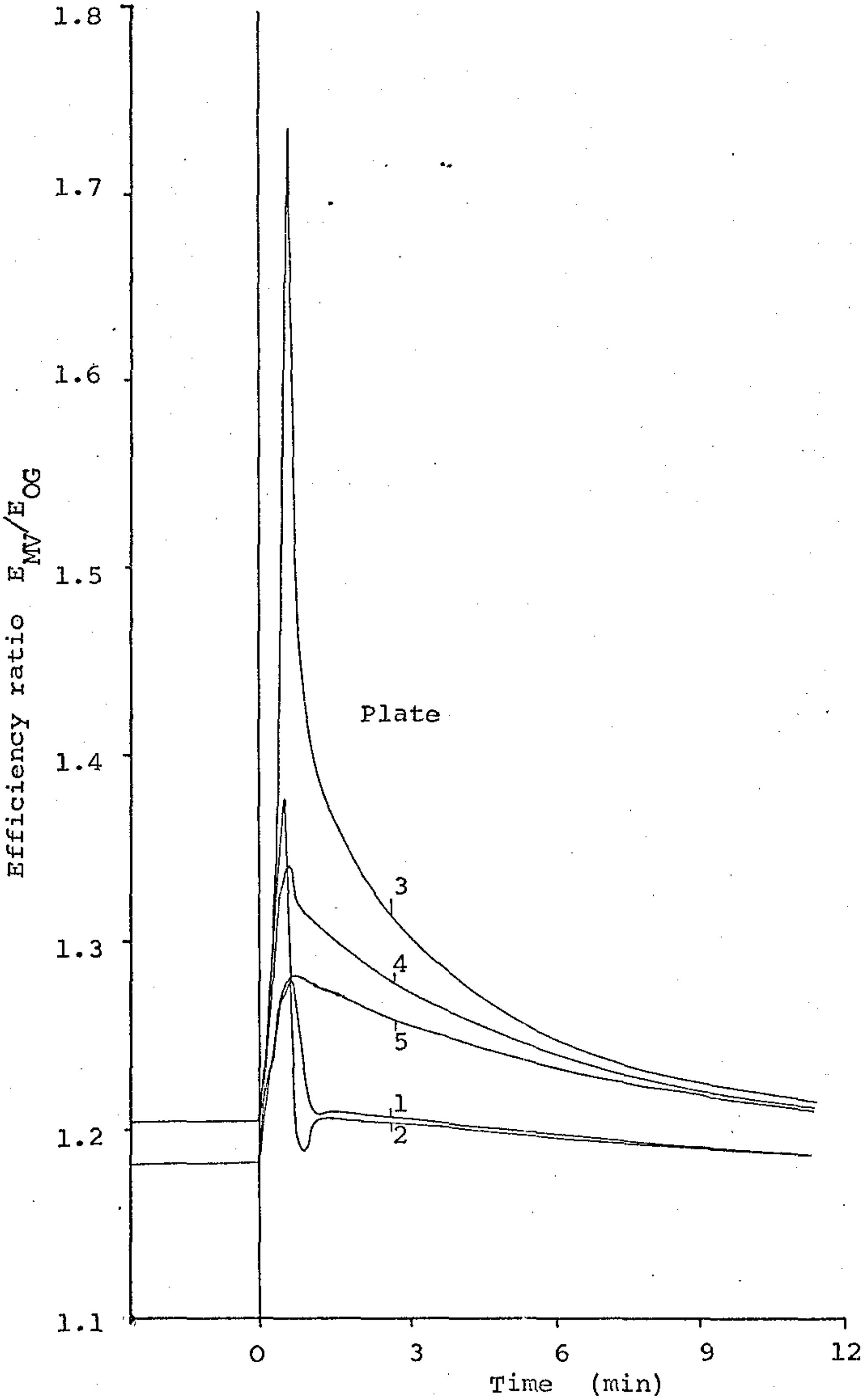


Fig 8.6 Efficiency ratio responses for a step increase in feed composition. Relative volatility of 1.1

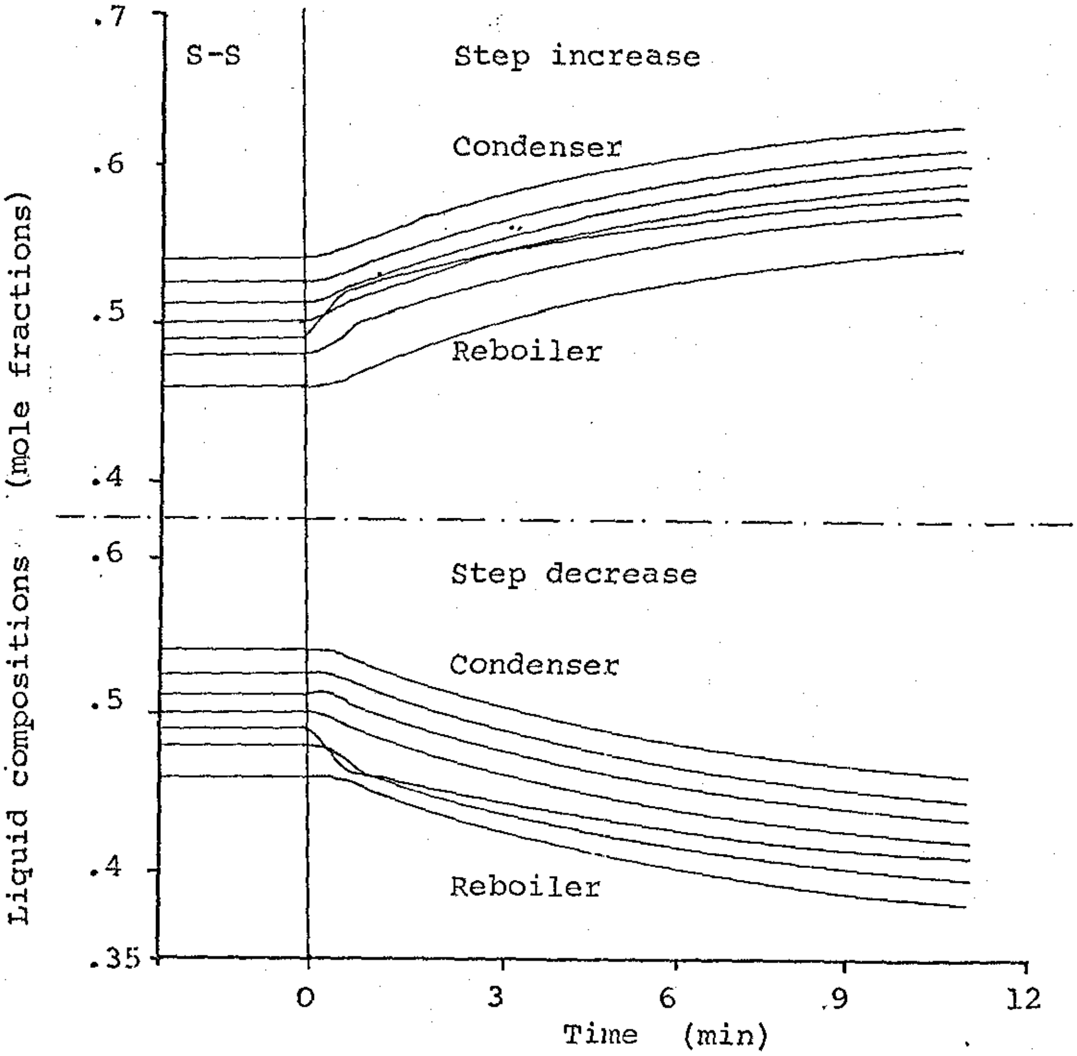


Fig 8.7 Condenser, reboiler and exit pool liquid composition responses for step changes in the feed composition. Relative volatility of 1.1 .

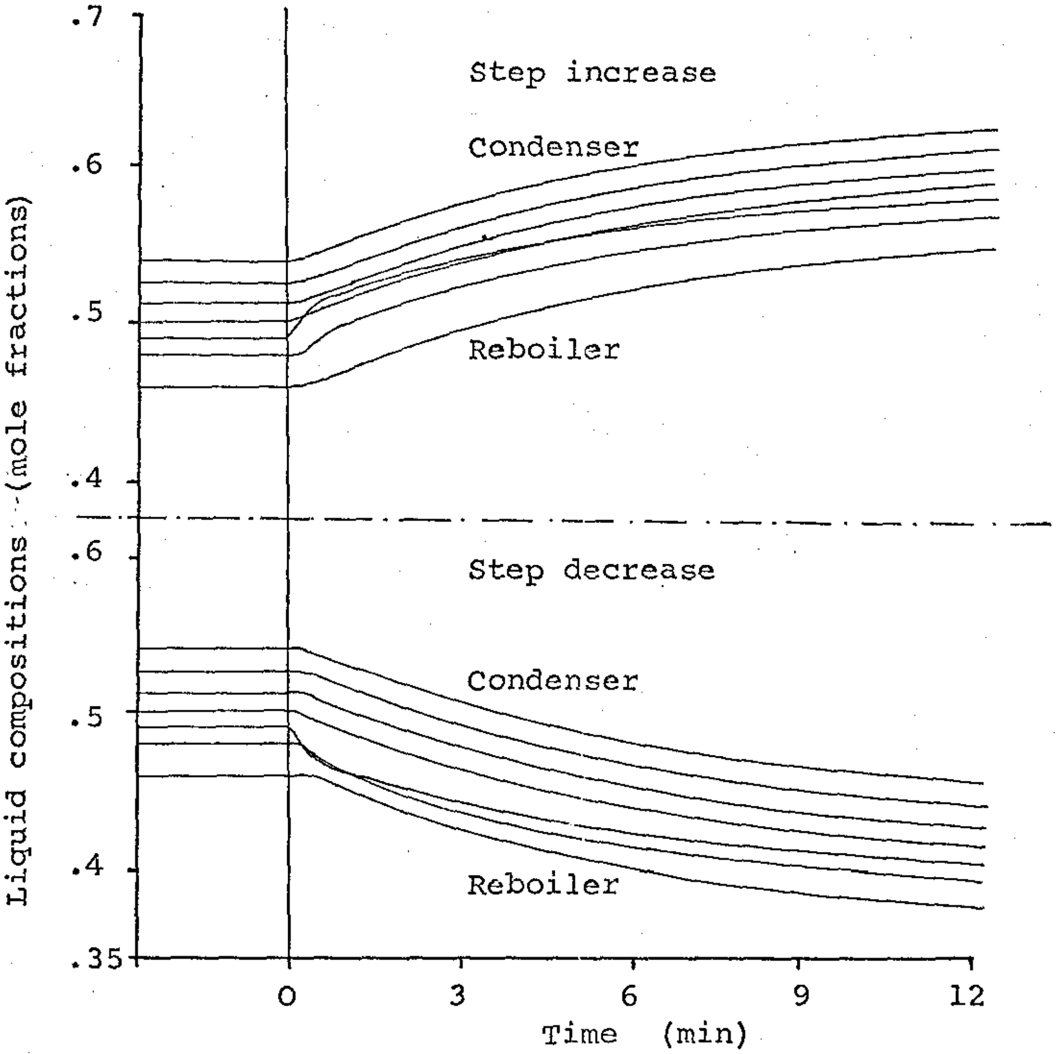


Fig 8.8 Condenser, reboiler and exit pool liquid composition responses for step changes in the feed composition. Relative volatility of 1.1 and vapour perfectly mixed.

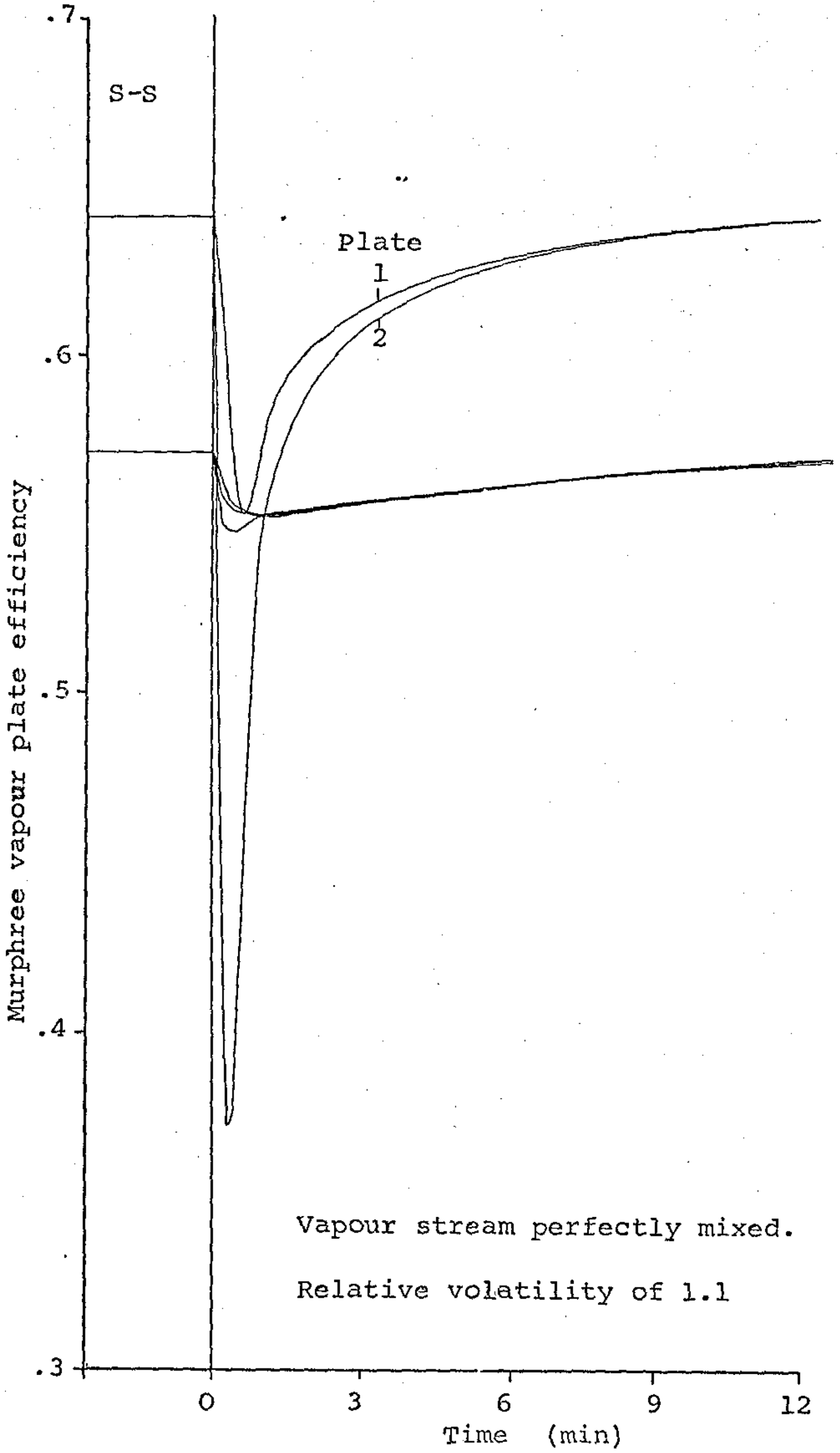


Fig 8.9 Murphree vapour plate efficiency responses for a step decrease in feed composition.

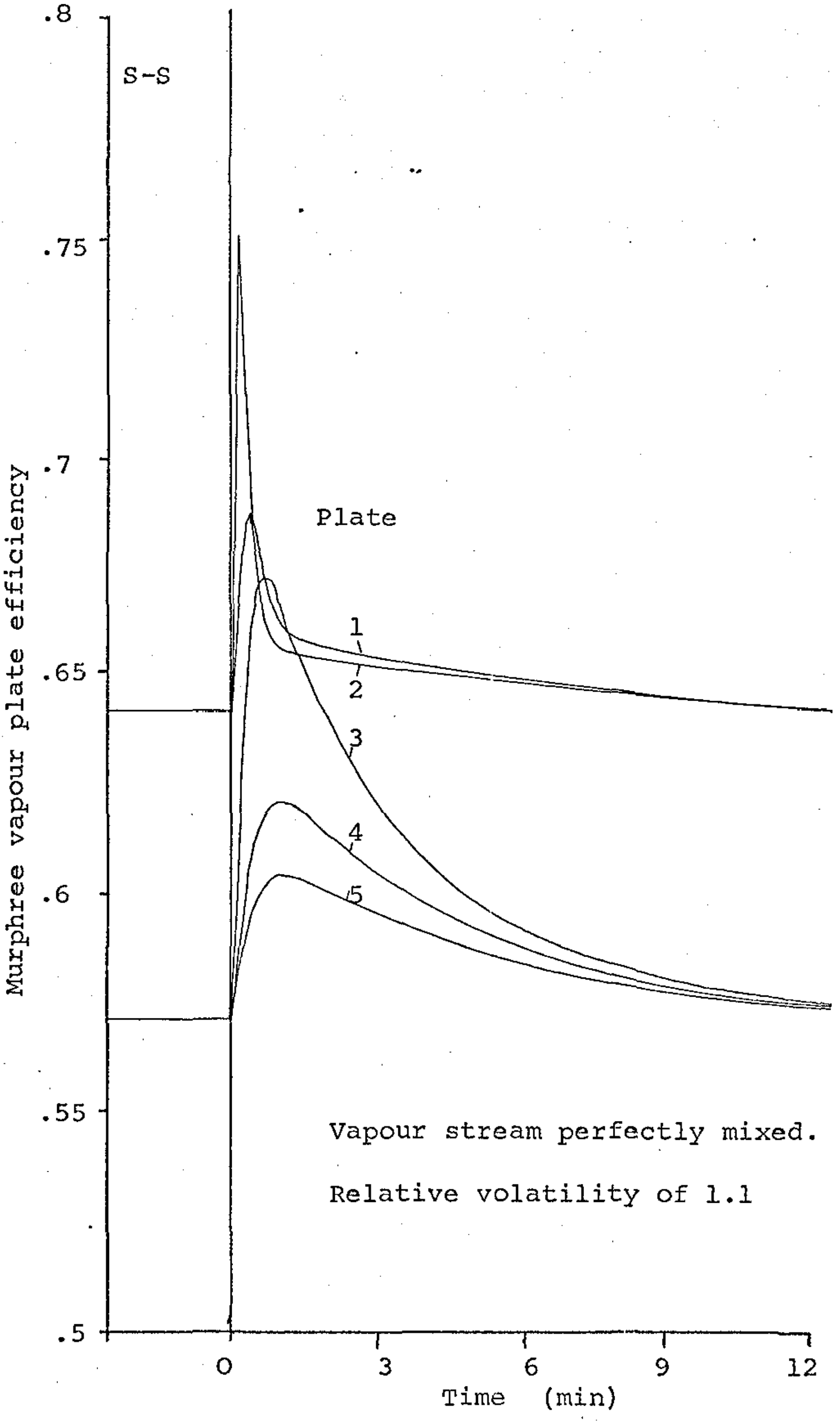


Fig 8.10 Murphree vapour plate efficiency responses for a step increase in feed composition.

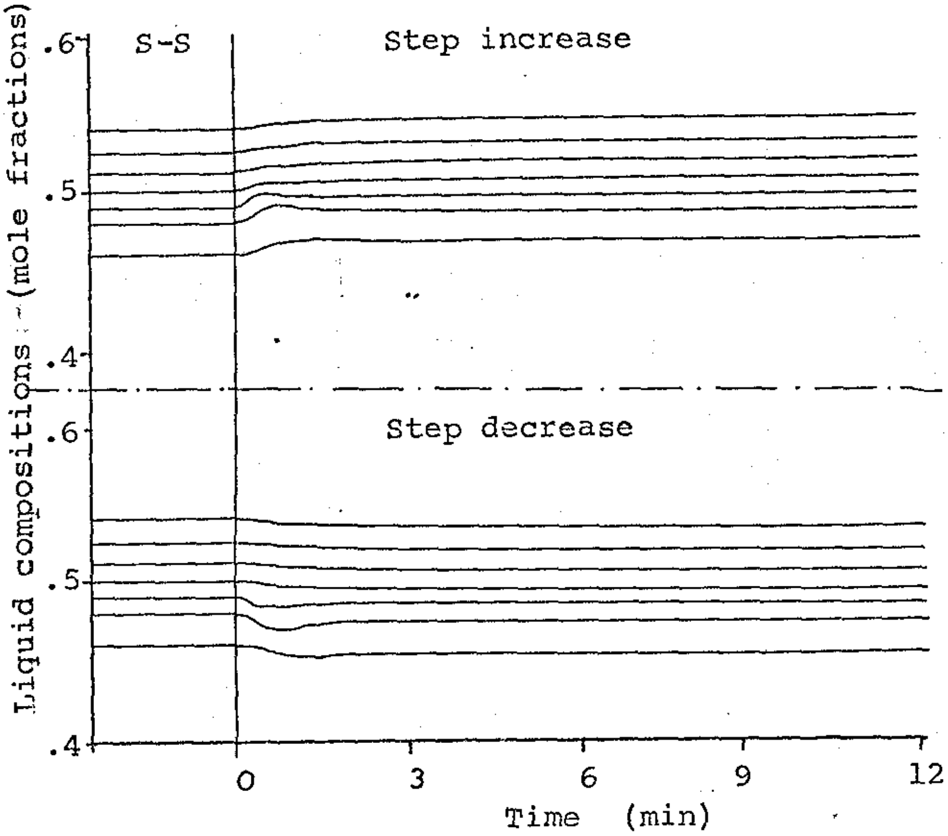


Fig 8.11 Condenser, reboiler and exit pool liquid composition responses for step changes in the feed flow rate. Relative volatility 1.1

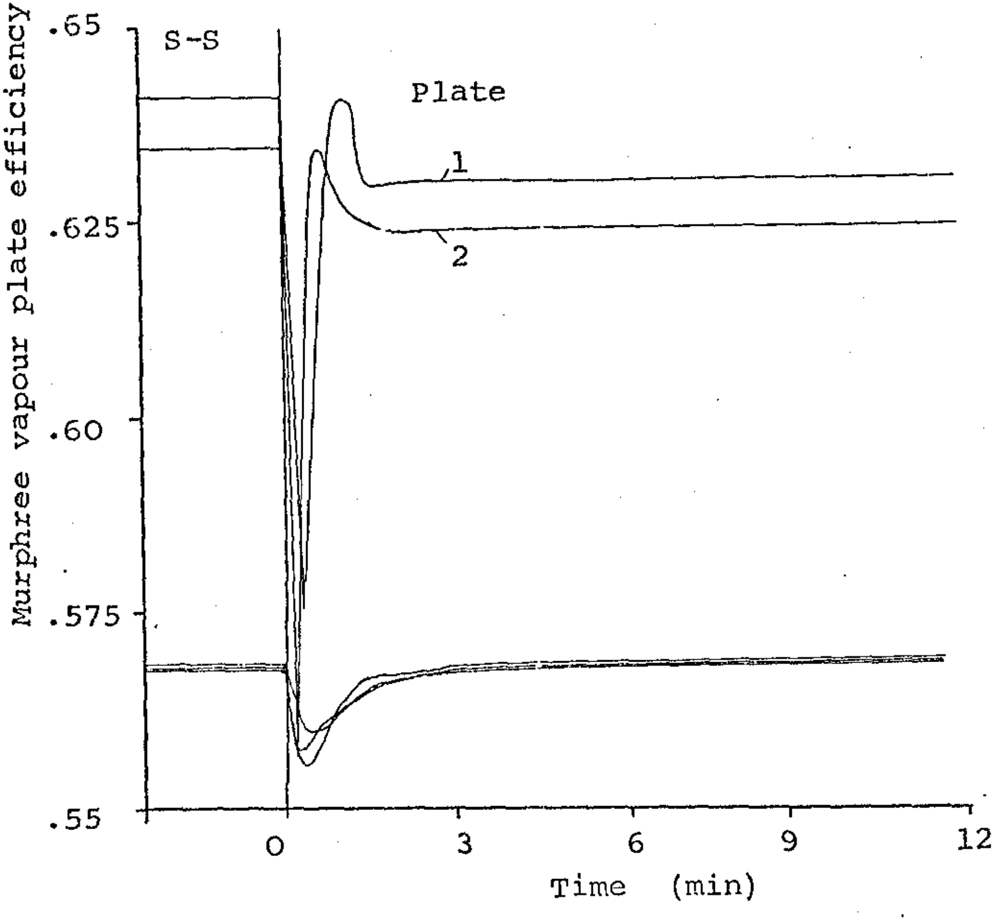


Fig 8.12 Plate efficiency responses for a step decrease in feed flow rate. Relative volatility of 1.1

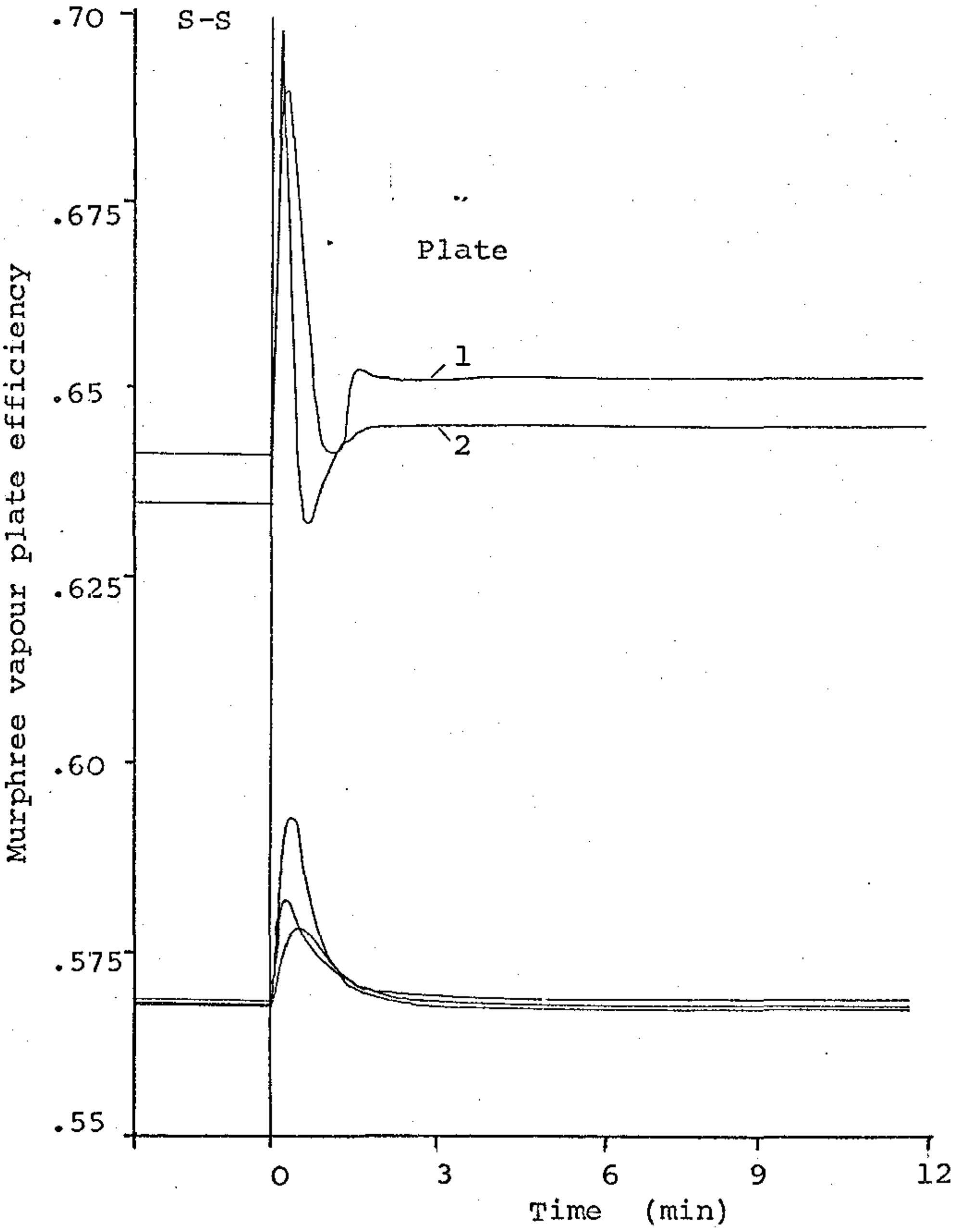


Fig 8.13 Plate efficiency responses for a step increase in feed flow rate. Relative volatility of 1.1

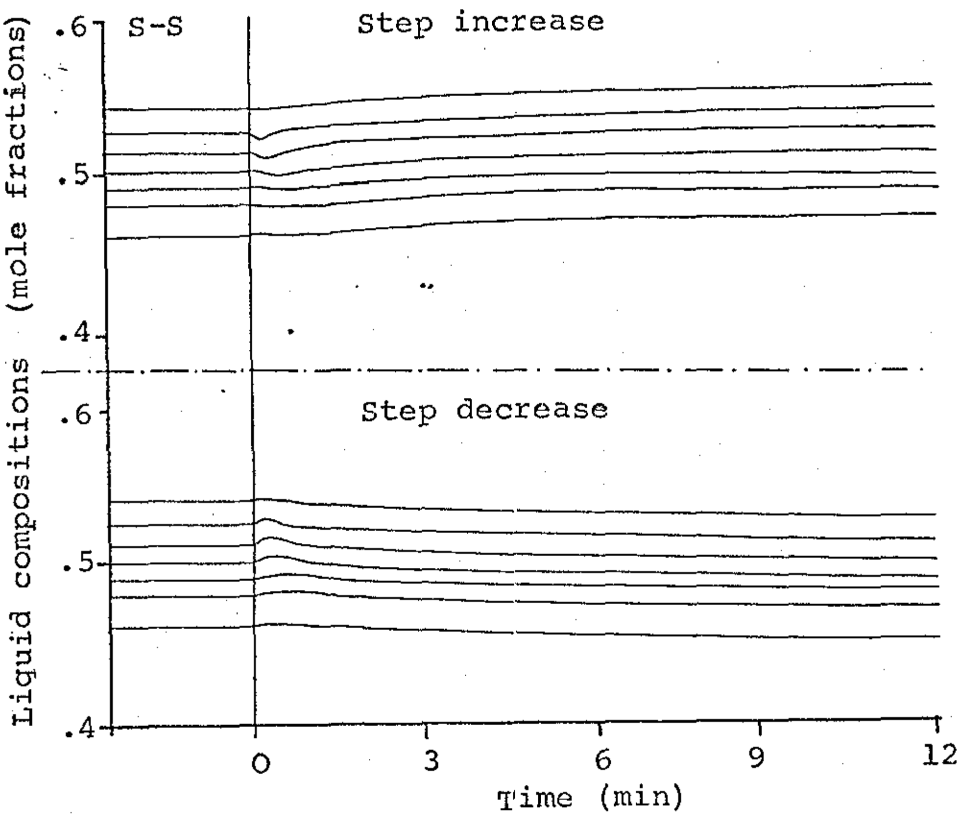


Fig 8.14 Condenser, reboiler and exit pool liquid composition responses for step changes in liquid reflux.

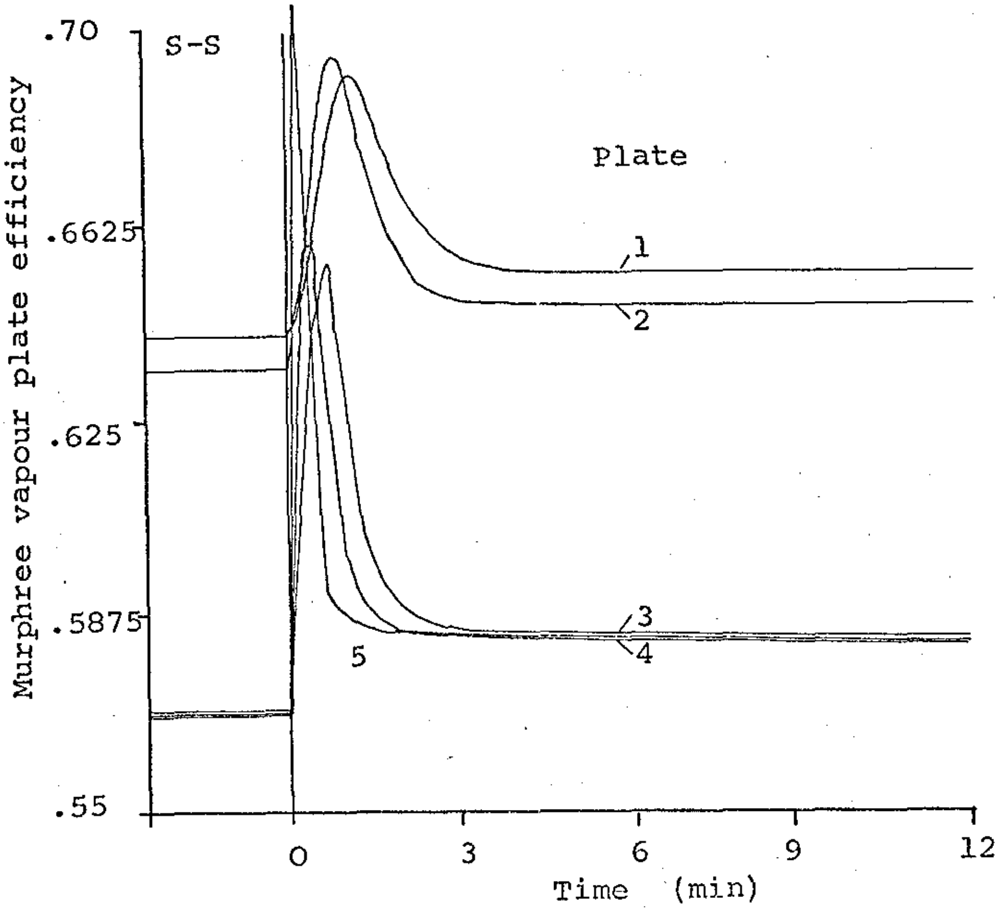


Fig 8.15 Plate efficiency responses for a step increase in liquid reflux. Relative volatility of 1.1

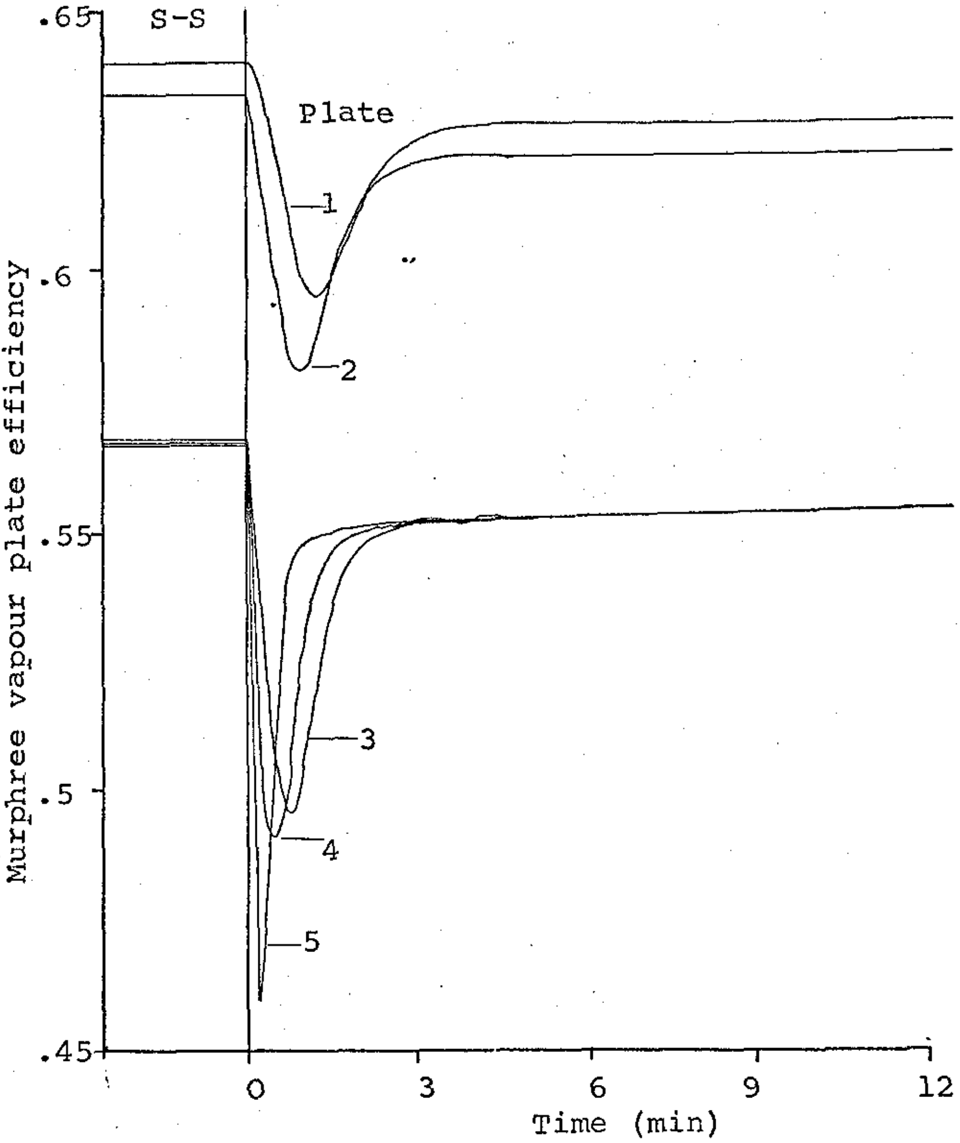


Fig 8.16 Plate efficiency responses for a step decrease in liquid reflux. Relative volatility of 1.1

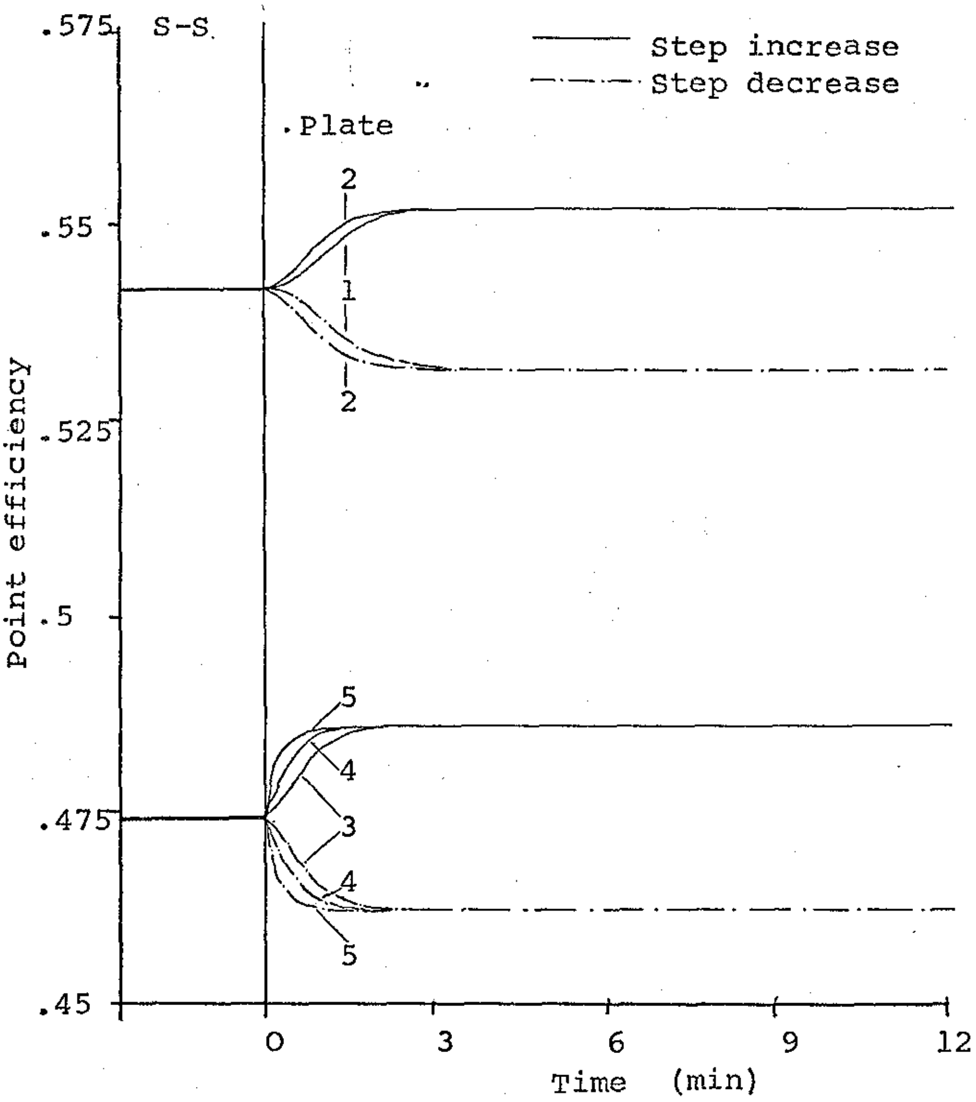


Fig 8.17 Point efficiency responses for a step change in liquid reflux. Relative volatility of 1.1

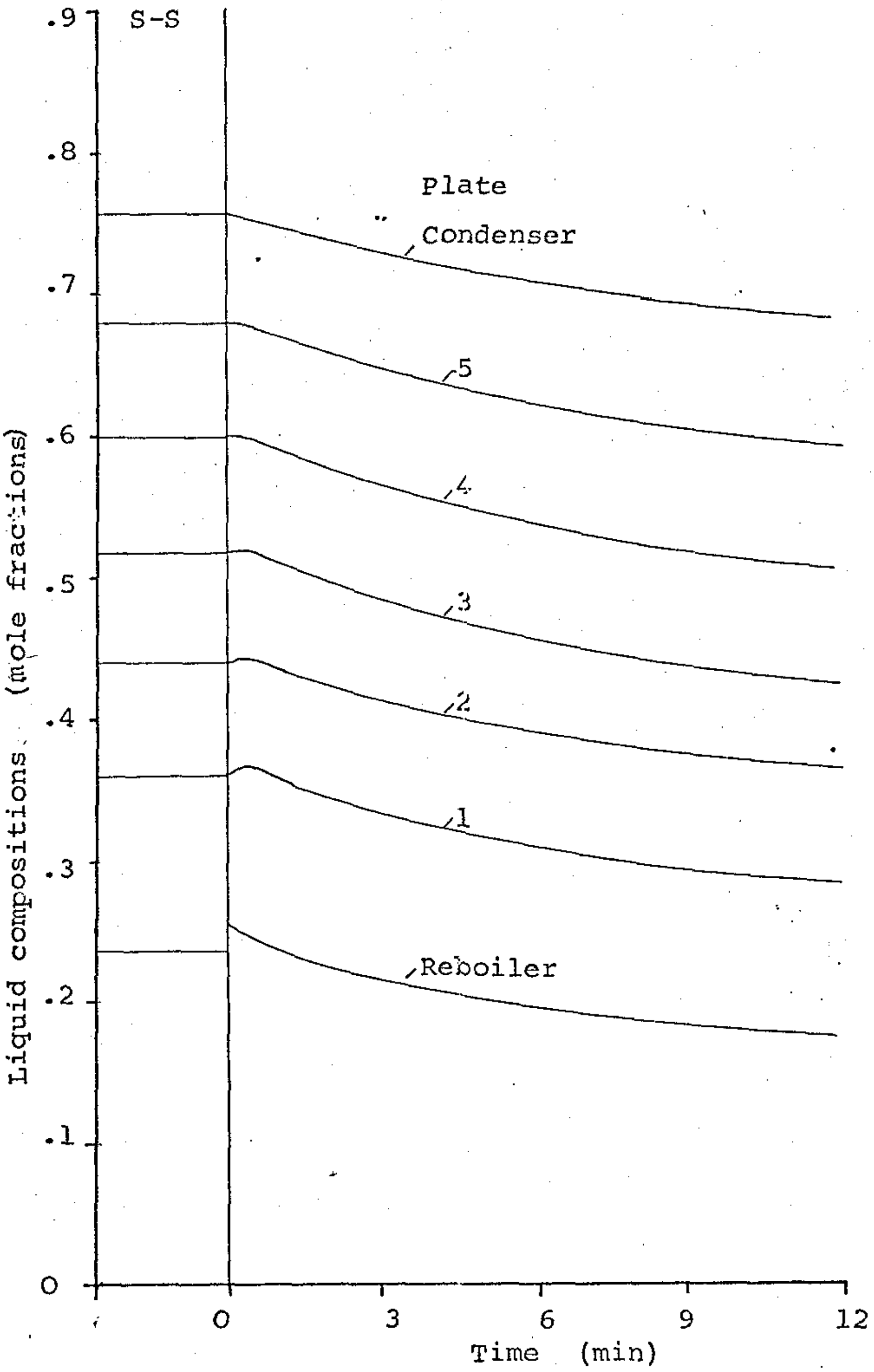


Fig 8.18 Condenser, reboiler and exit pool liquid composition responses for a step increase in vapour flow. Relative volatility of 2

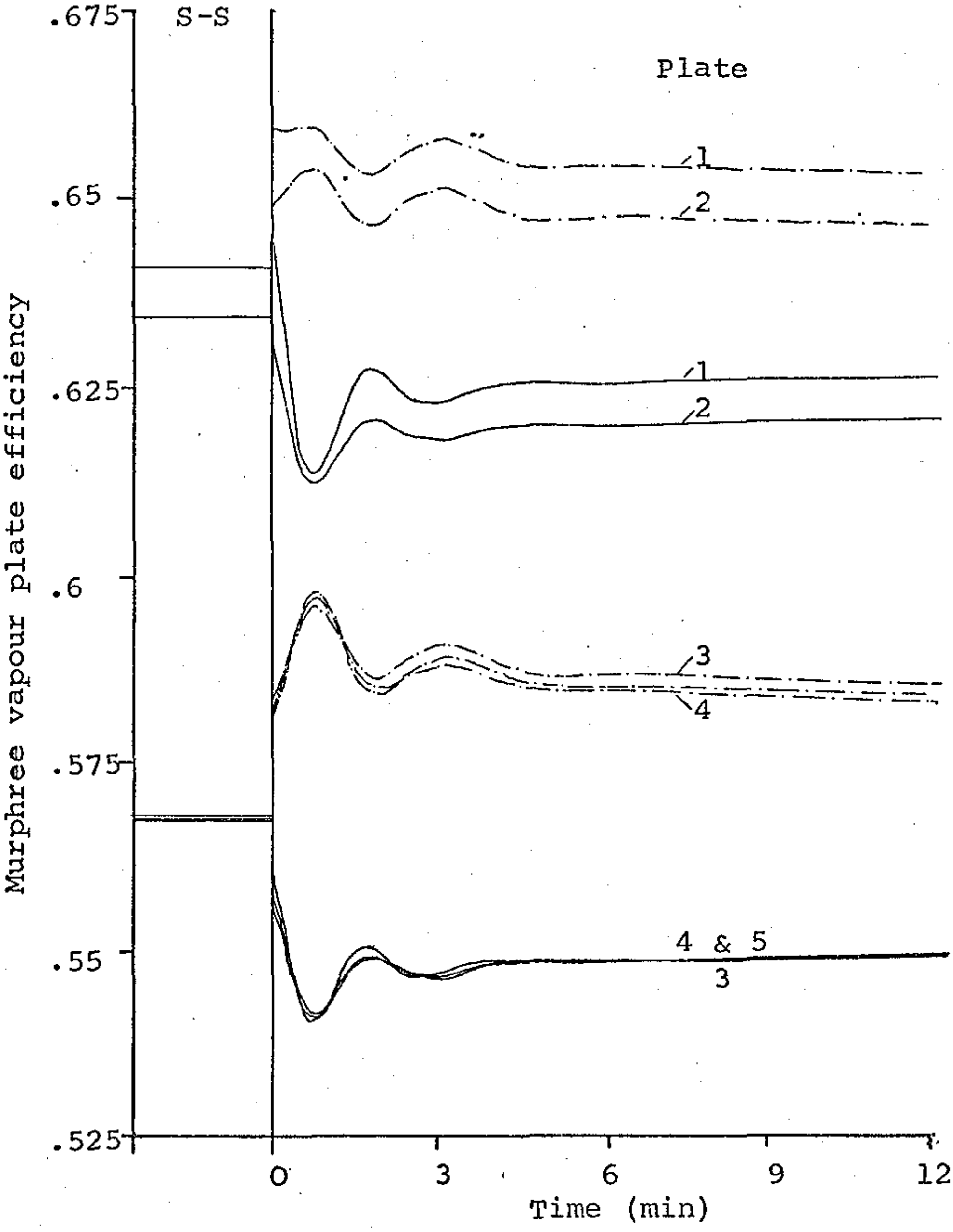


Fig 8.19 Plate efficiency responses for step changes in vapour flow. Relative volatility of 1.1

——— Step increase in vapour flow

—·—·— Step decrease in vapour flow

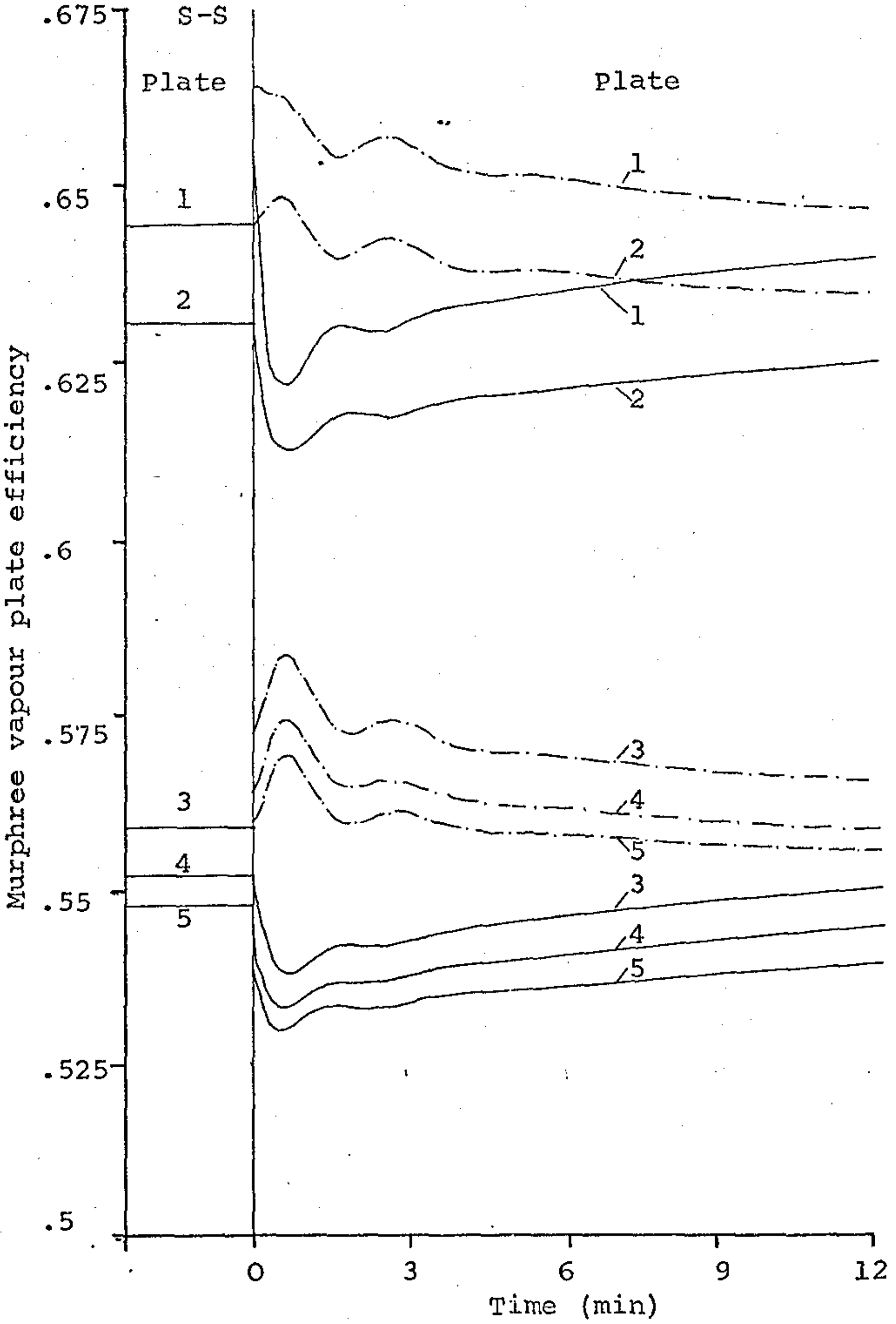


Fig 8.20 Plate efficiency responses for step changes in vapour flow. Relative volatility of 2

——— Step increase in vapour flow

—·—·— Step decrease in vapour flow

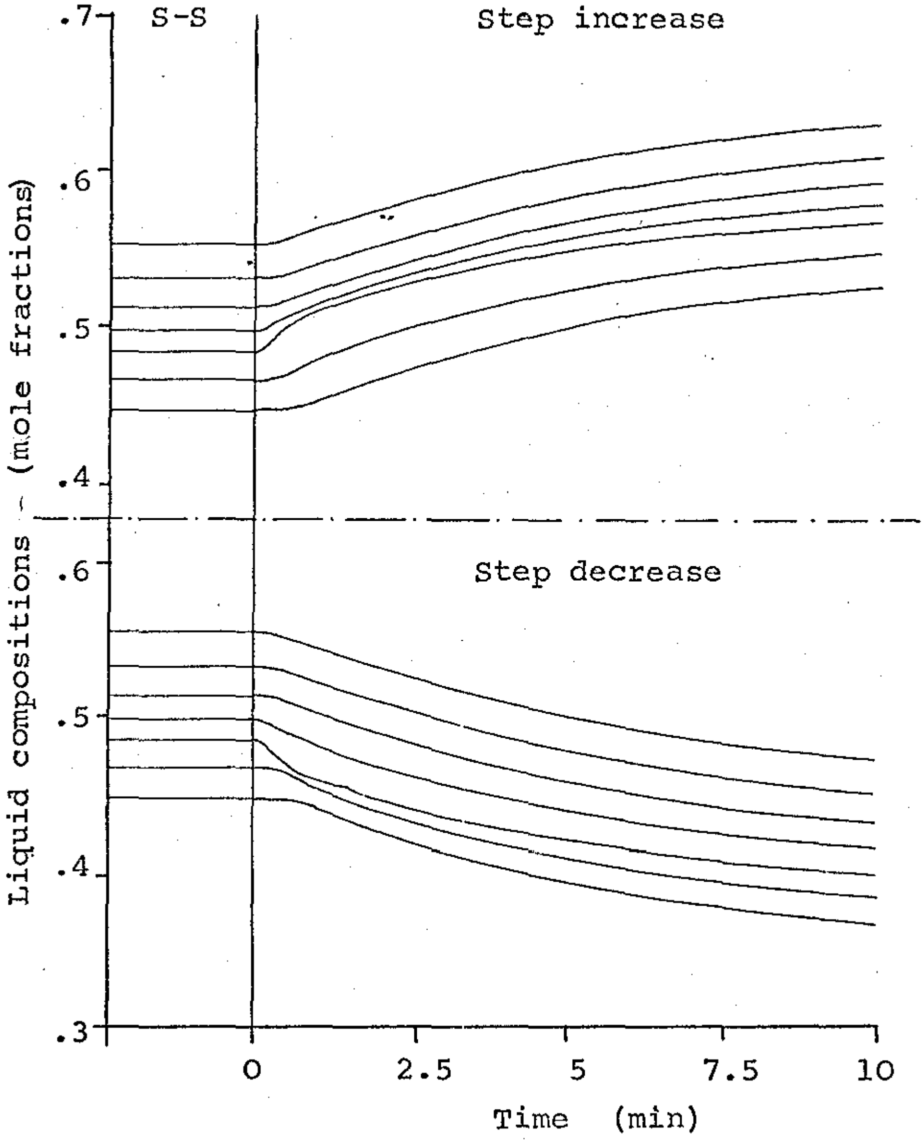


Fig 8.21 Condenser, reboiler and exit pool liquid composition responses for step changes in feed composition using 15 ft dia. column. Relative volatility of 1.1

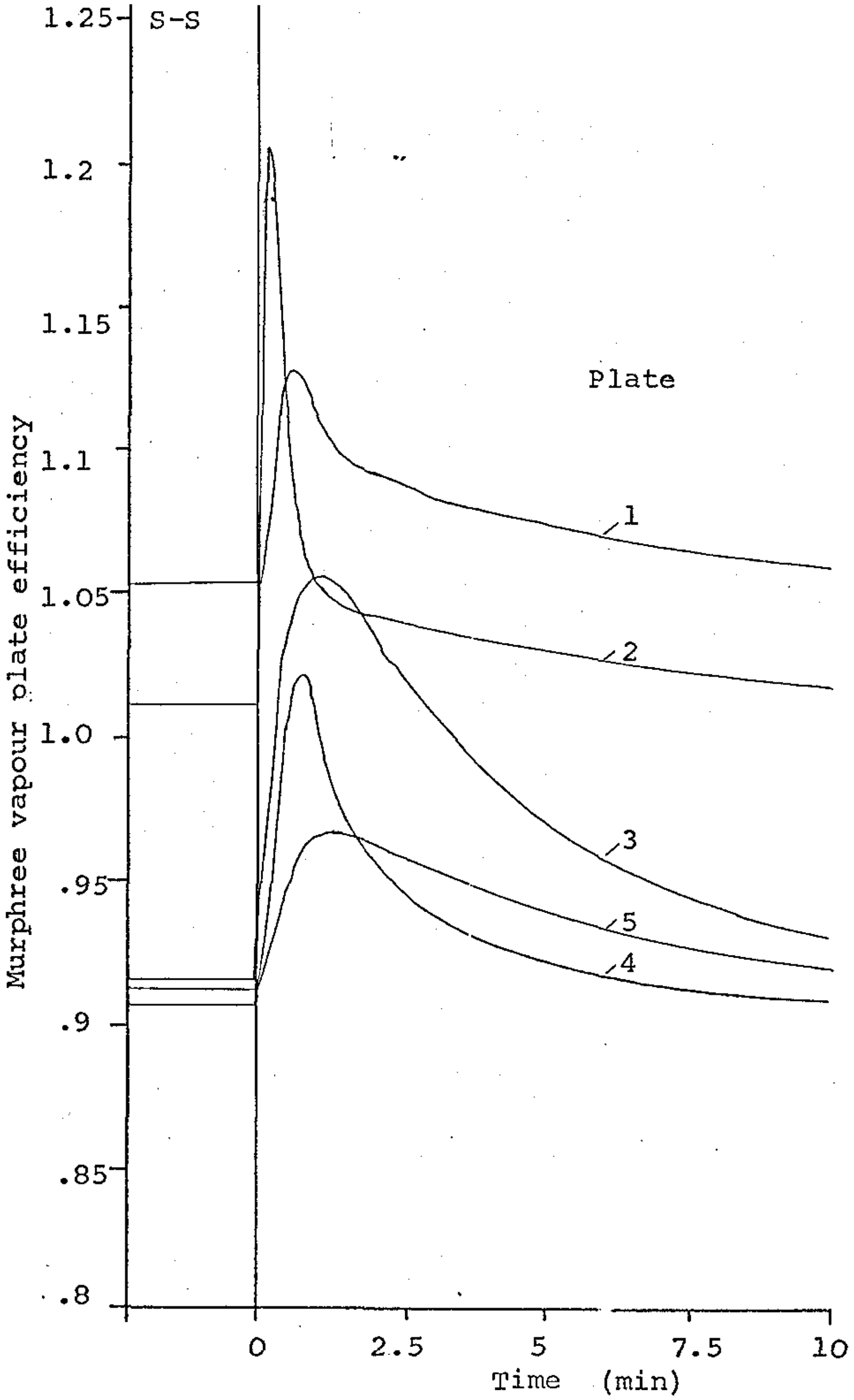


Fig 8.22 Plate efficiency responses for a step increase in feed composition using 15 ft dia. column. Relative volatility of 1.1

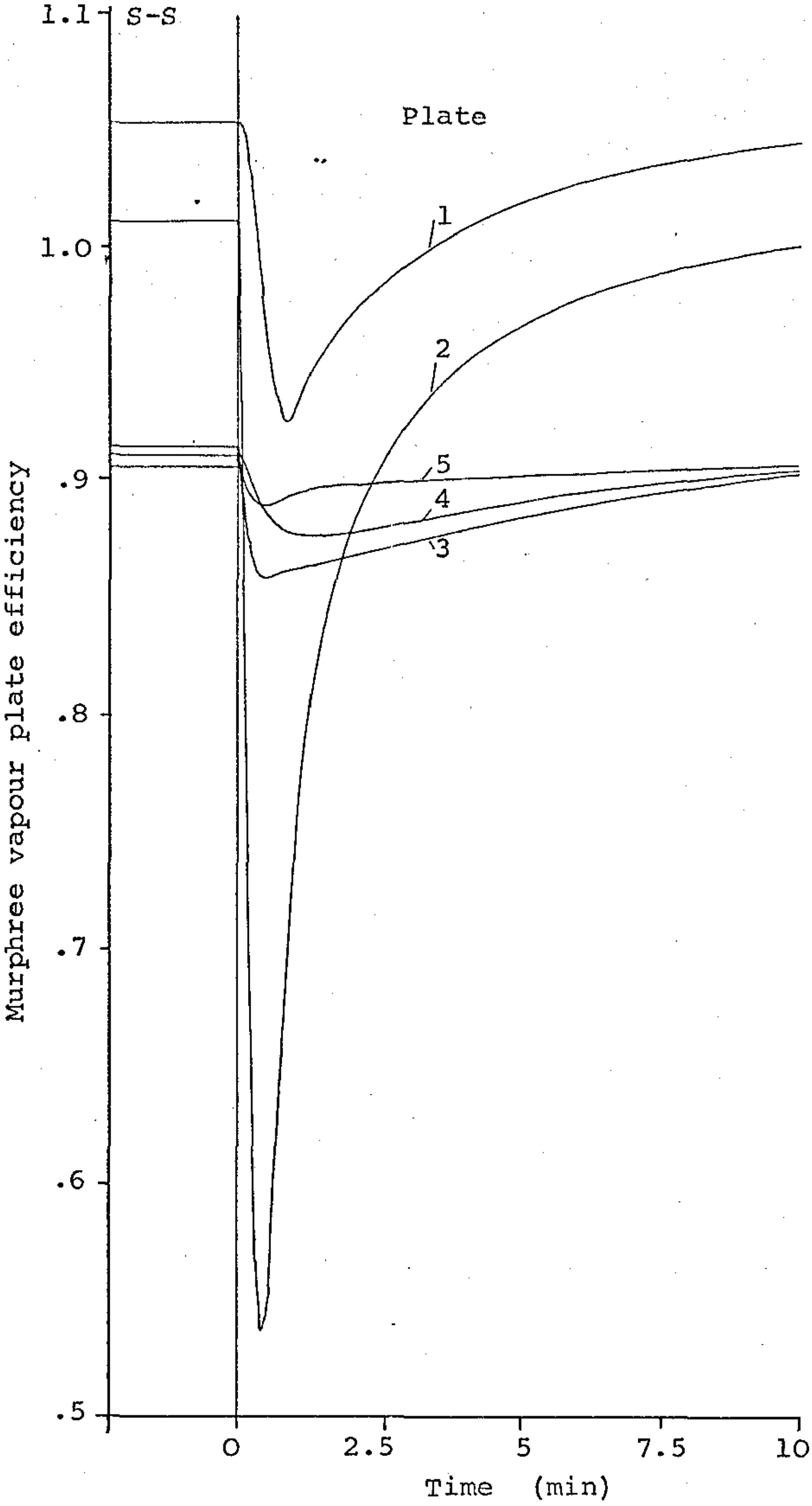


Fig 8.23 Plate efficiency responses for a step decrease in feed composition using 15 ft dia. column. Relative volatility of 1.1

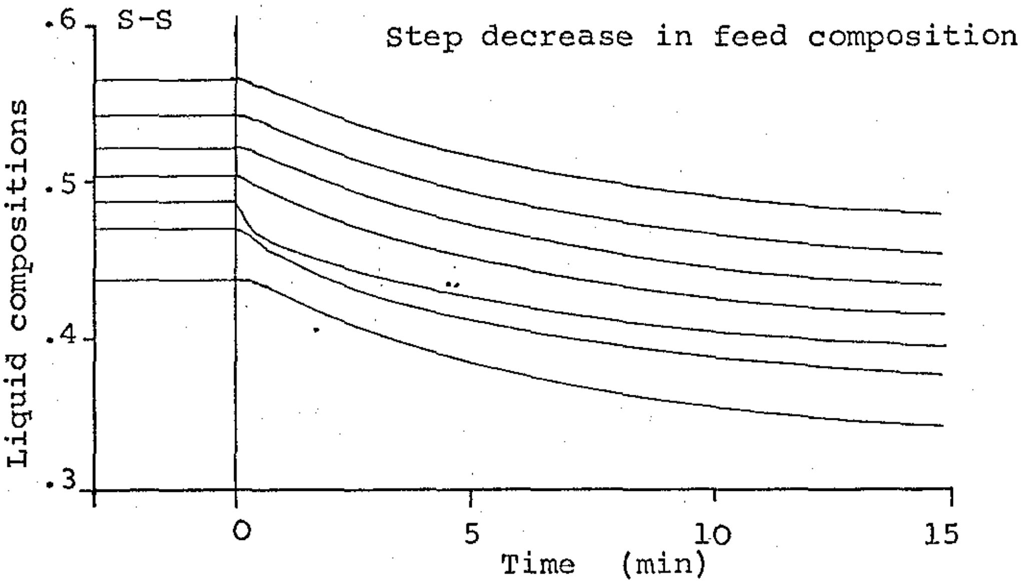


Fig 8.24 Condenser, reboiler and exit pool liquid composition responses for a step change in feed composition using a linear equilibrium relationship.

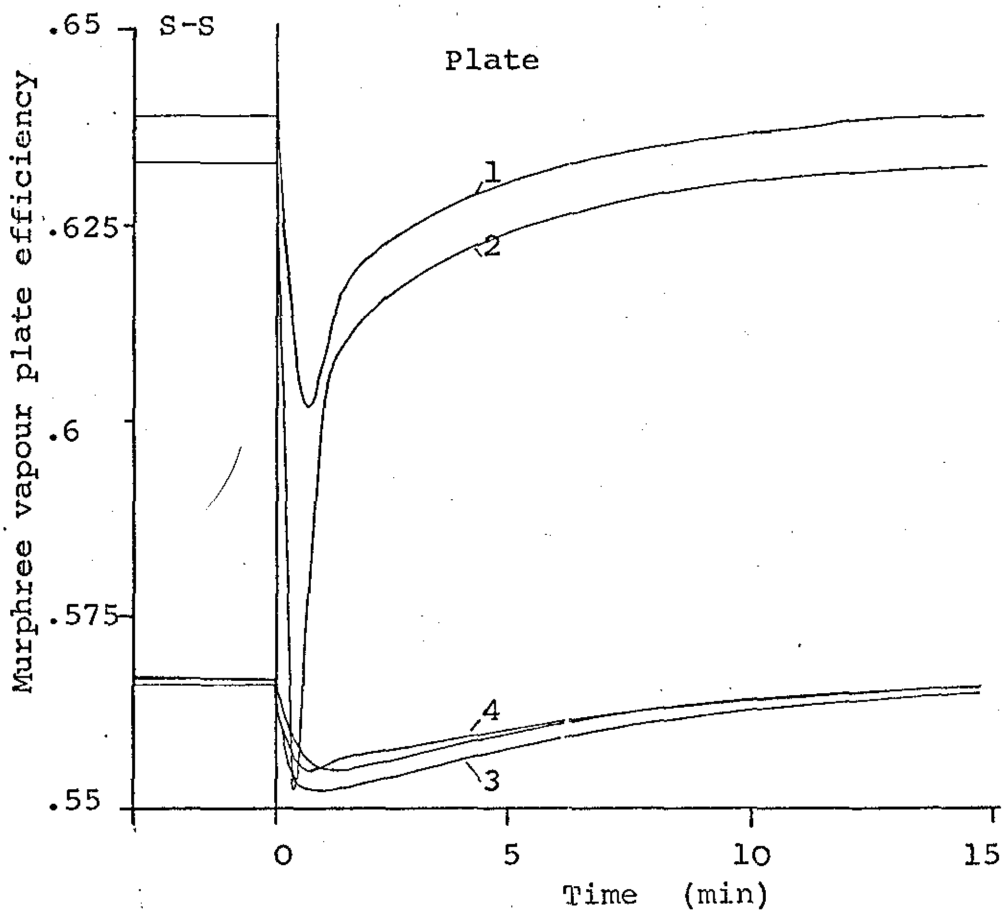


Fig 8.25 Plate efficiency responses for a step decrease in feed composition using a linear equilibrium relationship.

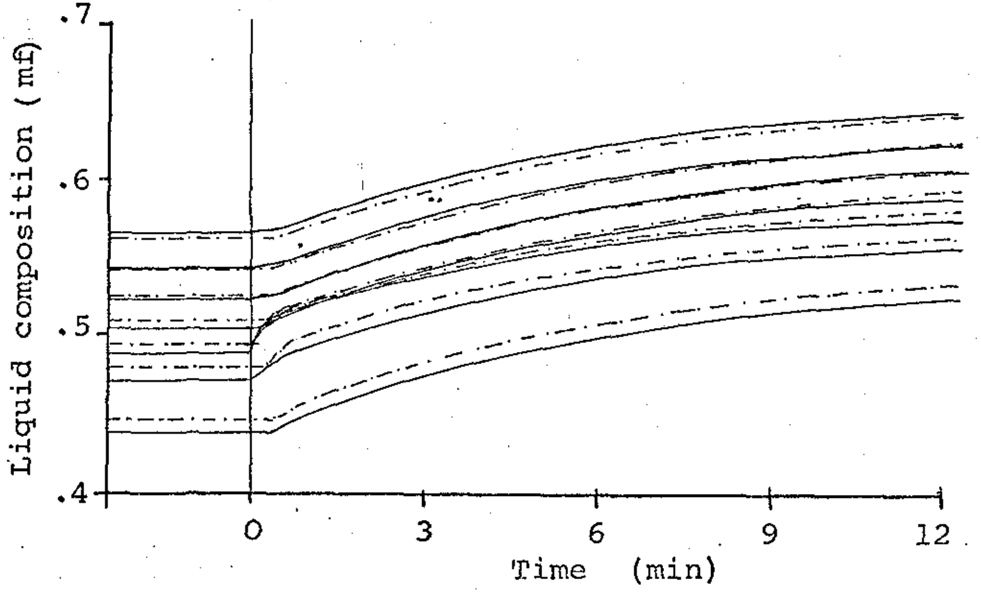


Fig 8.26 Liquid composition responses for a step increase in feed composition using simplified models.

——— Constant Murphree vapour plate efficiency used.

—·—·—Constant point efficiency used.

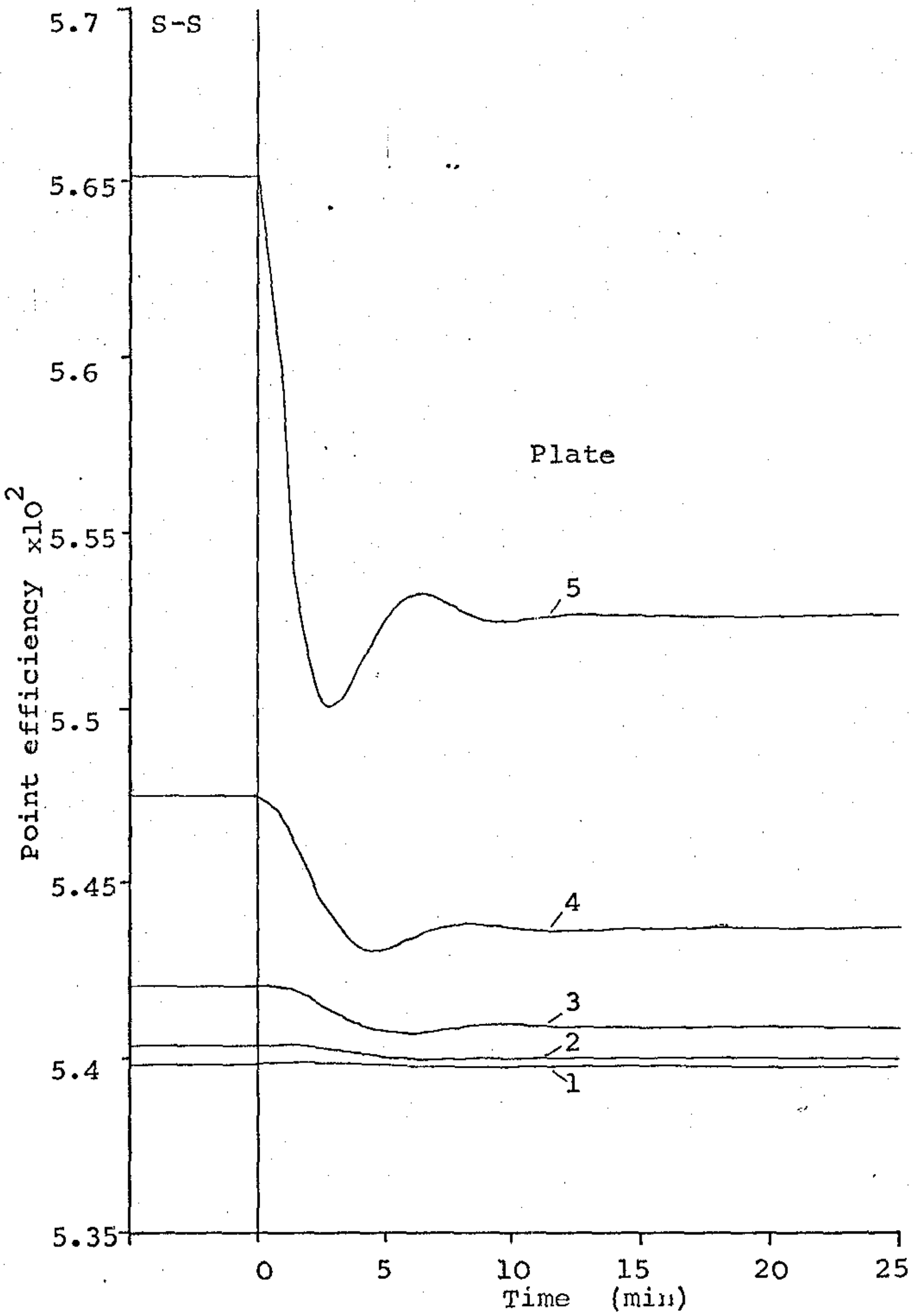


Fig 8.27 Point efficiency responses for a step decrease in liquid feed composition in gas absorption. Equilibrium slope of 31.667

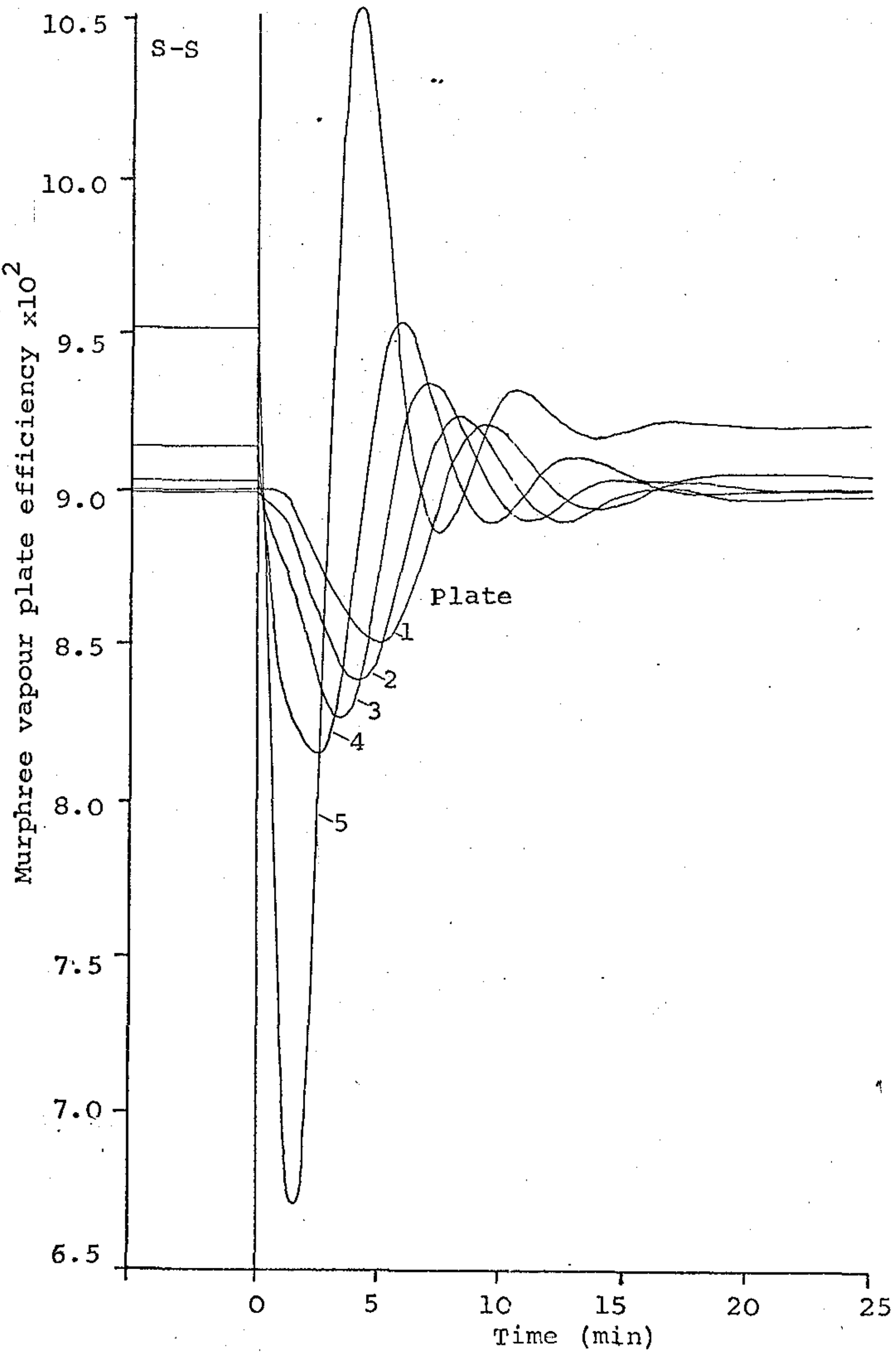


Fig 8.28 Plate efficiency responses for a step decrease in feed composition using an equilibrium slope 31.66

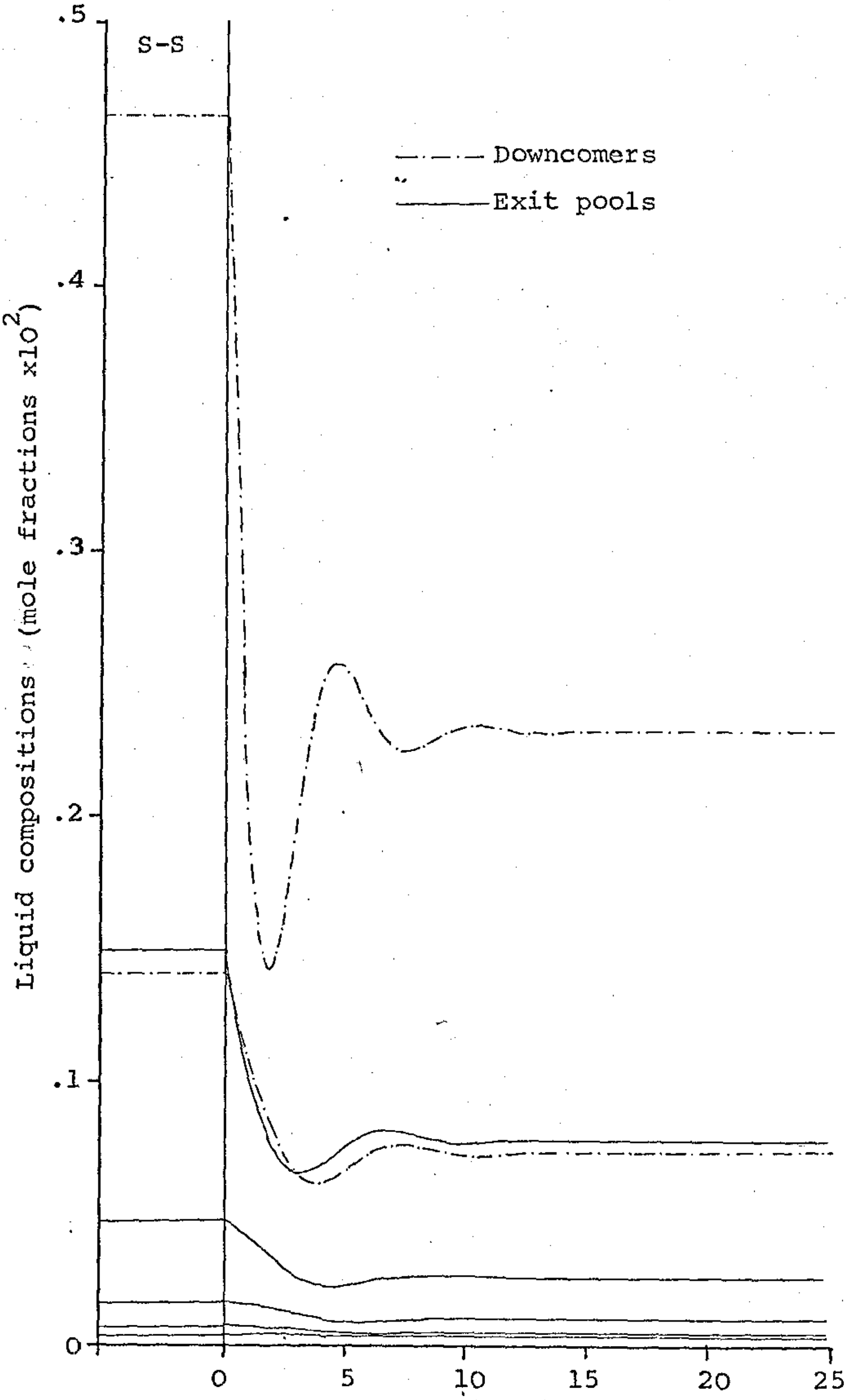


Fig 8.29 Exit pool and top two downcomer liquid composition reponses for a step decrease in feed composition. Equilibrium slope of 31.667

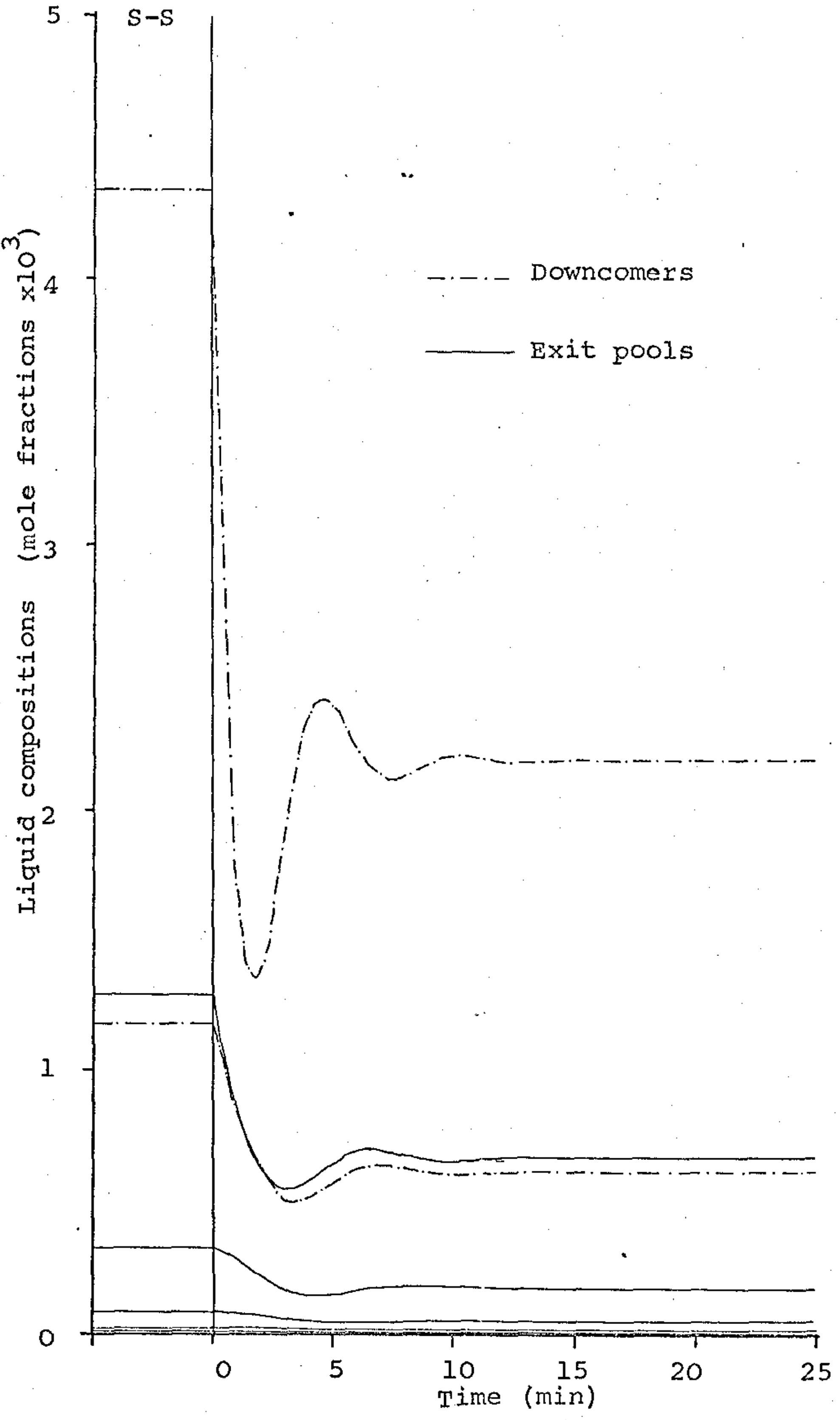


Fig 8.30 Exit pool and top two downcomer liquid composition responses for a step decrease in feed composition. Equilibrium slope of 190

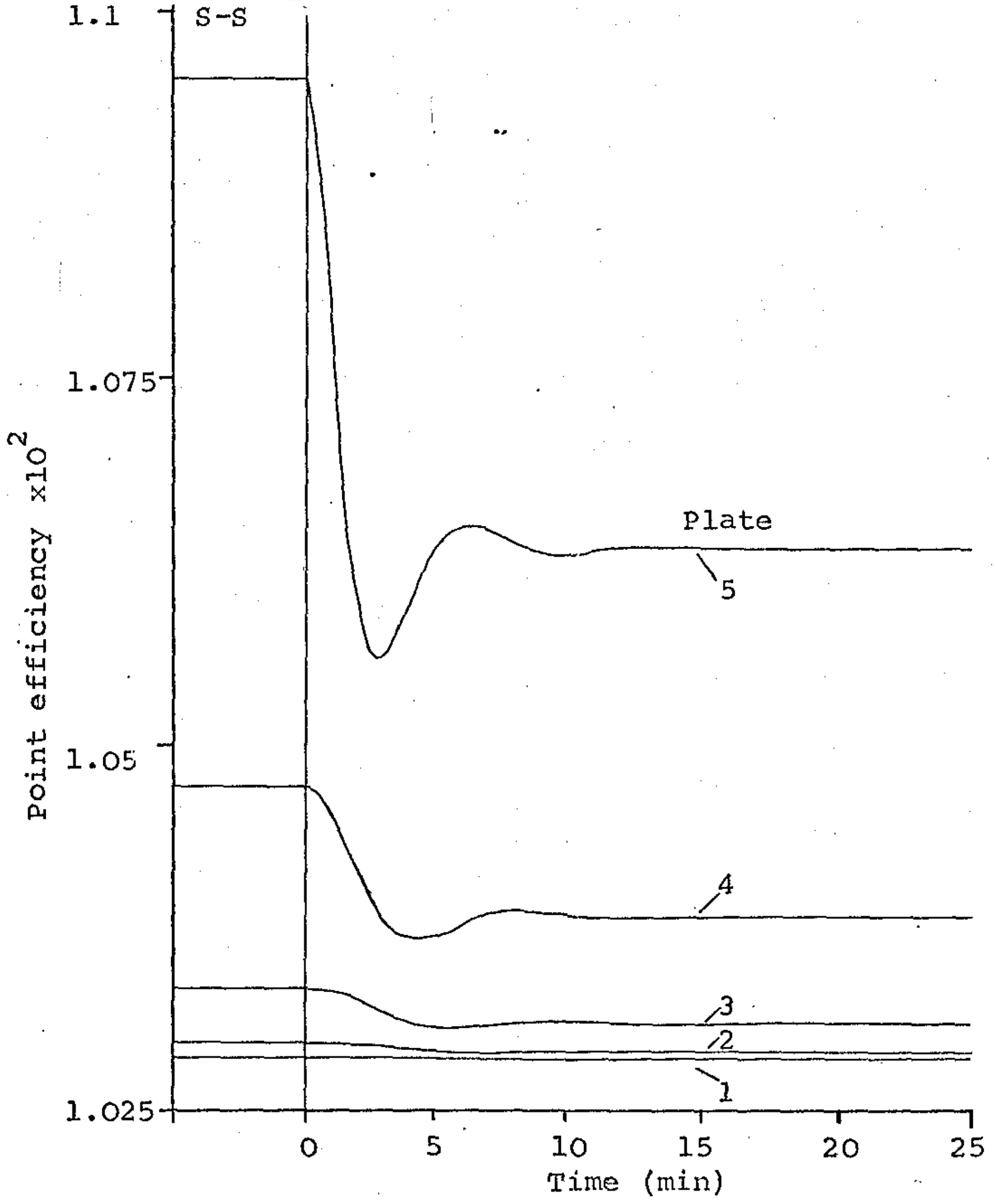


Fig 8.31 Point efficiency responses for a step decrease in liquid feed composition using an equilibrium slope of 190

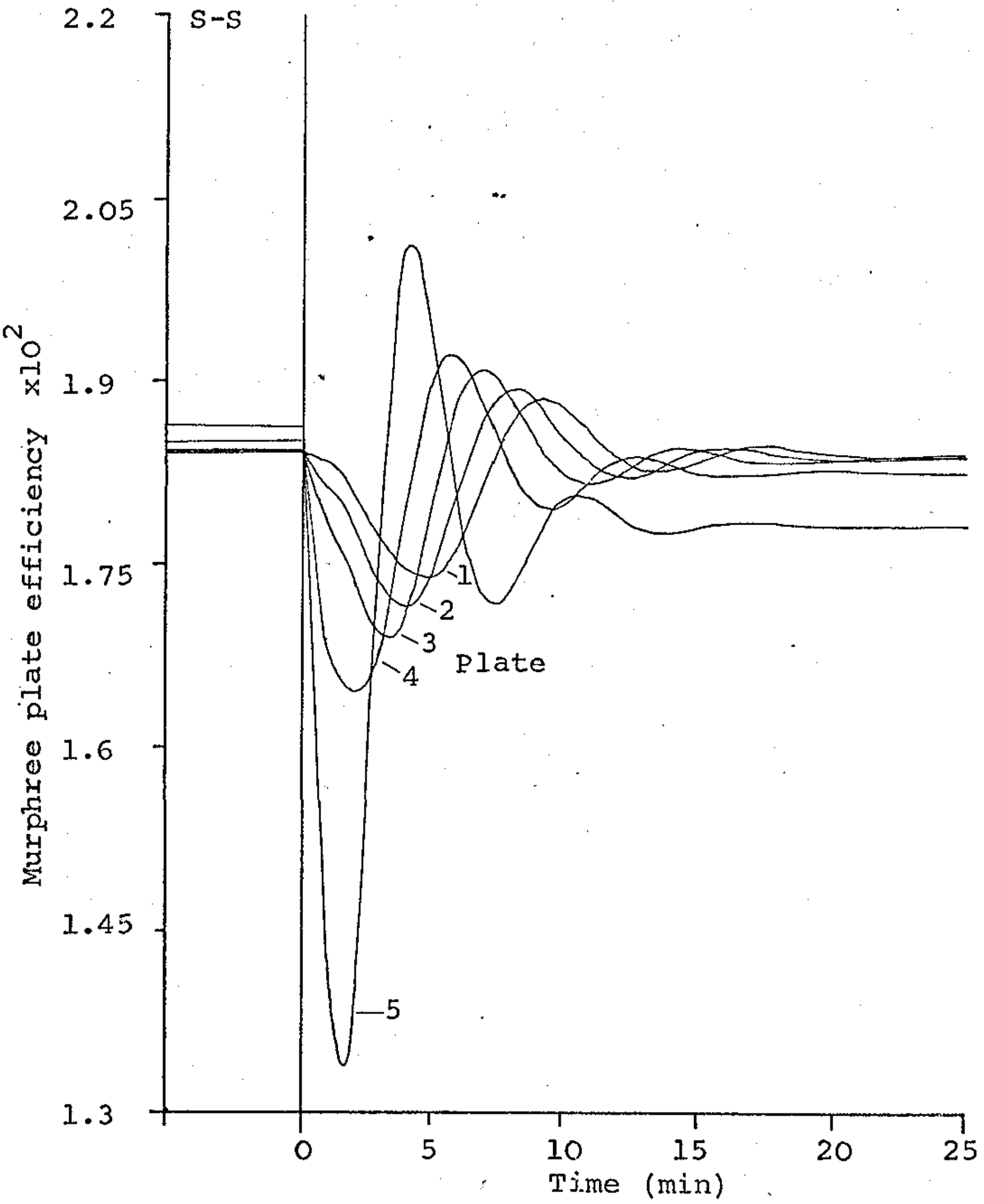


Fig 8.32 Plate efficiency responses for a step decrease in liquid feed using an equilibrium slope of 190

9. MULTICOMPONENT DISTILLATION

9 Multicomponent Distillation

9.1 Simulation Model

The generalised model of large plate columns presented for previous simulations of binary systems was used to describe the liquid and vapour mixing characteristics. The differential equations of the multicomponent system were the same as those for the binary system except that there were k - 1 of each equation, where k is the number of components. The value of the kth component was obtained from the sum of the others:-

$$X_k = 1 - \sum_{i=1}^{k-1} X_i \qquad \text{......} \; 9.1$$

Due to most distillation systems being gas film limiting, the following assumption was made to increase the ease in calculating the individual component point efficiencies for each pool:-

$$\frac{1}{N_{OG}} = \frac{1}{N_G} \qquad \text{......} \; 9.2$$

This assumption was verified by computing the vapour and liquid-phase resistances. The liquid-phase resistance was so small compared with that of the vapour-phase resistance that it could be ignored.

The modified equations presented by Wilke (111, 112) for calculating the average viscosity of the vapour and the diffusion coefficient of each component, were used in estimating the number of gas-phase transfer units. The point efficiency of each component was also calculated for each liquid pool and not just for the total tray.

Constant relative volatilities were used and the equation presented by Wood (116) for calculating the equilibrium vapour composition was also used:-

$$Y_i^* = \frac{\alpha_i X_i}{\sum_{r=1}^{k} \alpha_r X_r} \qquad \text{...... 9.3}$$

Further the slope of the equilibrium data was calculated from the partial derivatives:-

$$m_{n,i} = \frac{dY^*_{n,i}}{dX_{n,i}} = \sum_{r=1}^{k} \frac{\partial f_i}{\partial X_{n,r}} \qquad \text{...... 9.4}$$

where

$$\frac{\partial f_i}{\partial X_i} = \frac{\alpha_i \sum_{r=i}^{k} \alpha_r X_{n,r}}{\left\{ \sum_{r=1}^{k} \alpha_r X_{n,r} \right\}^2} \qquad \text{...... 9.5}$$

$$p=1 \text{ to } k \neq i \qquad \frac{\partial f_i}{\partial X_p} = \frac{\alpha_p \alpha_i X_{n,i}}{\left\{ \sum_{r=1}^{k} \alpha_r X_{n,r} \right\}^2} \qquad \text{...... 9.6}$$

as the slopes depend on all the changes in the component compositions, the equilibrium line is no longer of a fixed gradient but is a line in vector space.

The actual simulation program was set up only to deal with changes in the feed composition. The same column dimensions as those for the binary systems were used and the feed was introduced on plate 2 of the 5-plate column. Two different sets of systems were investigated;

1) relative volatilities of 1, 1.333, 1.667

2) relative volatilities of 1, 2, 3

and Table 9.1 gives the operating conditions for all the simulations.

Table 9.1 Liquid and vapour flows and operating conditions for the unsteady-state simulation of ternary systems.

Steady-state values for feed composition changes:-

Liquid and vapour flows (lb moles/hr)

	Systems:- relative volatilities	
	1, 1.333, 1.667	1, 2, 3
L_s	1000	1000
L_R	700	700
V	850	850
W_s	150	150
D_s	150	150
Pe	7	5
Reflux ratio	4.667	4.667

The increase in the number of components increases the number of computational operations carried out. This increase is equal to the square of the number of components being operated on multiplied by the number of operations carried out in the binary simulation. The actual computation time was roughly equal to the cube of the number of components operated on multiplied by the simulation time for the binary systems. Thus as the number of components was increased, the actual computation time increased rapidly. Although the simulation program was set up to handle up to 5 components, only ternary mixtures using the generalised model were investigated due to the computation time involved. The multicomponent program and instructions in use are given in Appendix E.

The generalised model operates on only two of the components in the material balances, but the third component is taken into account in the calculation of the point efficiencies. For this reason, it does not matter which two components are operated on as the results are the same.

9.2 The Effect of Feed Composition Changes on Ternary Systems

The first system investigated was that with constant relative volatilities of 1, 1.333, 1.667. The feed entered on the second plate with a feed composition of 0.30, 0.35, and 0.35 going from light to heavy components. The middle component feed composition was kept constant while step changes of ±0.05 were introduced into the light and heavy components.

For an increase in the light component feed composition and a corresponding decrease in the heavy, the liquid compositions on the trays showed the usual smooth step increase and decrease responses. The middle component compositions showed that this component was not separated off and that the top and bottom product streams of the middle component were of the same value. The highest middle component composition was found at the exit pool of the third plate. The composition responses of the middle component for the feed change showed a slight separation occurring; the reboiler value increased and the condenser decreased. The highest composition of this component was now found on the second plate.

The plate efficiency responses for both the light and heavy components rose by 0.02 in the value for the feed plate, then dropped quickly to the new steady-state value. Just like the binary cases, the feed plate showed the greatest change and the new steady-state values did not differ from the initial values by more than 0.01. The plate efficiency responses for the middle component showed large changes, the greatest being that of the plate above the feed. For this plate the Murphree plate efficiency rose from 1.28 to 1.52 in 19 seconds then fell gradually to a new steady-state value of 0.7176. The liquid composition responses of the light and heavy components are shown in Fig 9.1 and the plate efficiency responses of the middle component are shown in Fig 9.2.

The reason for the large differences between the two steady-state values of the middle component plate efficiencies was due to a substantial drop in the absorption factor λ . The initial and final steady-state values of λ for the feed plate were 2.786 and 1.732 respectively and the values for the plate above were 2.9 and 1.56 respectively. If these values of λ , Peclet number and point efficiency for this simulation are used in the equation for the prediction of the steady-state plate efficiencies (1), (equations 3.10 and 4.25), it can be shown that the simulation results are verified.

Increasing the heavy component feed composition and decreasing the light component by 0.05 for the same system and initial feed conditions, the light and heavy component composition responses were as expected and the plate efficiencies for the feed plate both fell by 0.02 then rose again to the new steady-state values which were approximately the same as the initial values. The middle component plate efficiencies showed very large changes, both positive and negative. The plate efficiency for the feed plate rose slowly at first then rapidly reach a huge value of 75 then fell to a negative value of -7 then rose to a new steady-state value of 0.61. From these responses, it can be seen that ridiculous and meaningless values of the plate efficiency are found which cannot be used in unsteady-state calculations.

The system with constant relative volatilities of 1, 2, and 3 and feed compositions of 0.30, 0.35 and 0.35 showed larger differences in the top and bottom products of both the light and heavy components. Further this system showed that if the middle component had a relative volatility equal to the mean value of the others, and the feed was nearly central, then the top and bottom products of this component would be equal and the highest composition value of this component would be found on the central tray.

An increase in the light and a decrease in the heavy component feed composition of 0.05, showed the usual smooth step increase and decrease in the light and heavy component liquid compositions respectively. The plate efficiency responses of the light and heavy components showed slight increases of about 0.01 then fell to new steady-state values which were about 0.01 less than the initial values. The middle component, however, showed its greatest change to occur on the plate above the feed. Both the feed plate and the plate above showed sharp rises in the plate efficiency of about 0.03 then fell to new steady-state values of 0.72 and 0.64 respectively, which were about 0.2 less than their initial values.

When the increase in the light component feed composition was 0.1 and corresponding decreases in both the middle and heavy components of 0.05 were made, similar, but more pronounced responses to those mentioned above, were obtained. The computed step increase and step decrease responses of the light and heavy component liquid compositions, are shown in Fig 9.3 and the middle component liquid composition responses, showing the high liquid composition in the middle of the column, are shown in Fig 9.5. The plate efficiency responses of the light and heavy components are only slight and can be assumed constant (see Fig 9.6 and 9.7). The middle component plate efficiency responses show large changes and are shown in Fig 9.4. The liquid and vapour compositions for every liquid pool on each plate, the condenser and the reboiler values, plate efficiency and all computed data for this simulation are given in Appendix G.

If large changes in the light and heavy component feed compositions are made, the changes in the middle component plate efficiencies become large and meaningless in both the +ve and -ve directions. Going from light to heavy, the feed composition changes were from 0.35, 0.35 and 0.3 to 0.1, 0.4 and 0.5. The middle component liquid responses are shown in Fig 9.8. The plate efficiency responses for the heavy component showed little difference for the rectifying section but the response for plate 1 showed a decrease from 1.31 to 1.23 in 40 seconds then gradually settled out to a new value of 1.15. The feed plate response for the heavy component showed a rapid drop of 0.557 to 0.523 in 6 seconds then a sharp rise to 0.546 in the next 13 seconds then gradually rose to a new steady-state value of 0.59 in the following 13 minutes.

The light component plate efficiency responses shown in Fig 9.10, show the plate efficiency of the feed plate to drop suddenly and rise again then gradually rise to a new steady-state value about 0.1 higher than the original. Further the plate efficiencies of the rectifying section all rise gradually to new steady-state values 0.1 higher than the initial values.

When the initial steady-state feed compositions were 0.45, 0.1 and 0.45, the middle component plate efficiency for the feed plate had a high value of 2.43. The initial steady-state values showing the middle component liquid composition distribution, plate efficiencies and the wide composition range through the column of both the light and heavy components are shown in Table 9.2. Changes in the feed composition for this system gave similar results to those where the feed compositions were 0.35, 0.35 and 0.3.

9.3 Multicomponent Distillation Investigations

From an extension of his earlier work on batch distillation (81), Rose et al (82) presented one of the first methods of solving multicomponent distillation problems by finite difference formulation. Since then improvements in numerical integration techniques have been made and alternative methods of solution have been given, the most important being that of matrix methods which have been presented by Mah et al (67) and Sargent (85).

Following the works of Armstrong and Wood (4) and Lamb, Pigford and Rippin (57), Wood (116) presented the frequency responses of a multicomponent system. Constant relative volatilities were used and the equations were linearised. Matrix methods were used for the solution and equations for estimating the equilibrium data and the equilibrium slope of each component was given.

The changes in plate efficiencies for multicomponent systems has been investigated by Holland (48). The model used by Holland in the investigation into binary systems was that presented by Tetlow (99, 100) which described all the mixing characteristics in the downcomer. However, for the multicomponent systems he ignores the mixing effects and has complete bypassing of the downcomer and just has a perfectly mixed plate. The plate efficiency used was that of the vaporisation efficiency which is defined as:-

$$E^{o}_{n,i} = \frac{Y_{n,i}}{Y'_{n,i}} \qquad \text{...... 9.7}$$

The vaporisation efficiency was further defined by two parameters introduced by Holland (47):-

$$E^{o}_{n,i} = \bar{E}_i \bar{B}_n \qquad \text{......} \ 9.8$$

where $\bar{B}_n$ was the plate factor for plate n and $\bar{E}_i$ was the component efficiency for component i. Equations for estimating these factors for the system and the disturbance investigated were given as a function of the time since the changes were made. For the 5-component system investigated, the major part of the unsteady-state was over after 3 minutes even for a 10 plate column and the initial steady-state values of the vaporisation efficiency of each component was the same for every plate but did not converge to a new constant value for each component at the new steady-state.

The advantage of using the vaporisation efficiency is that negative values are not obtainable. The advantage of using the Murphree vapour plate efficiency is that for a liquid composition change, the Murphree plate efficiency may remain constant while the vaporisation efficiency changes.

The system investigated by Holland was for feed composition changes of 0.1, 0.25, 0.3, 0.25, 0.1 to 0.35, 0.25, 0.2, 0.15, 0.05 going from light to heavy. The vaporisation efficiency responses for the light, middle and heavy components obtained by Holland (48), are shown in Fig 9.11 to 9.13 and numerical values of the vaporisation efficiency responses are given in Appendix H.

Groves (39) who did the latest work on Holland's technique of estimating plate efficiencies, stated that the convergence of the θ method for a new set of inputs was acceptable for the simple model where the downcomer was bypassed, but for the more general model convergence was slow. He further stated that more data had to be known to be able to describe the model more fully and that the model was of limited use due to the great amount of computer storage needed.

9.4 Discussion on Ternary Distillation Results

From the work carried out in this investigation, it has been shown that the Murphree vapour plate efficiency of the middle component changes drastically, and settles at new values which are far removed from the initial steady-state values for small changes in the feed composition. Responses of the light and heavy component plate efficiencies show very little change during the unsteady-state period and between the initial and final steady-state values.

In the work done by Holland (48), large changes in the feed compositions were made, but there were no dramatic changes in any of the component vaporisation efficiencies. This was due to the vaporisation efficiency not being dependent on the composition of the incoming vapour stream to the plate. Further, from the responses of the vaporisation efficiencies, it can be seen that the responses for each component are similar and that their response paths are dominated by the plate factor. It would seem unlikely that the vaporisation efficiency of each component was the same on every plate at the initial steady-state, but if this was so , then they should also converge to a new constant value at the final steady-state which they do not. Due to the inaccessibility of the complex functions for describing the plate factor and component efficiency for random systems and the great dependence of the vaporisation efficiency on the plate factor responses, it would seem that simplifications using the vaporisation efficiency cannot be made.

From the work carried out by the author on the comparison of liquid composition responses using a generalised model, the linearised model and the simplified model with only one liquid pool, it has been shown that if constant Murphree plate efficiencies are used which have been estimated for partial liquid mixing, then the simplified model and the linearised model show good agreement in responses with those obtained using the generalised model. It can be concluded, therefore, that if the light and heavy components are operated on and their Murphree plate efficiencies are kept constant during unsteady-state, then the simplified model or the linearised model may be used and good agreement in transients will be achieved.

It must be noted that the middle component must not be operated on in the simplified models or large errors will be produced due to the great changes occurring in plate efficiency values.

Table 9.2 Ternary steady-state values.

TIME	FEED	BOTTOMS	TOPS	R.RATIO	RB.HOLDUP	C.HOLDUP
0.00	700.0	150.0	150.0	4.667	13.9425	2.8390

COMPONENT	REL.VOL	FEED COMP.	BOILER	COND.	ETP.	CTP.
1	3.00	0.45	0.1463	0.7542	113.055	113.131
2	2.00	0.10	0.0903	0.1098	16.449	16.476
3	1.00	0.45	0.7634	0.1360	20.495	20.393

PLATE	HOLDUP	TEMP	PECLET NO.	BACKFLOW	LIQUID FLOW	VAPOUR FLOW
1	3.3113	104.67	4.72	0.9819	1000.00	850.0
2	3.5423	100.69	5.61	0.7473	1000.00	850.0
3	3.3952	96.38	4.86	0.9411	700.00	850.0
4	3.6524	93.70	5.62	0.7446	700.00	850.0
5	3.9076	89.51	6.39	0.5951	700.00	850.0

COMPONENT= 1		LIQUID COMPOSITION						VAPOUR COMPOSITION						MEAN	EMV
PLATE	DC	1	2	3	4	5	6	1	2	3	4	5	6		
1	.3744	.3562	.3393	.3239	.3102	.2990	.2917	.4556	.4467	.4382	.4306	.4241	.4197	.4358	0.8505
2	.4631	.4511	.4355	.4215	.4091	.3990	.3923	.5468	.5421	.5390	.5371	.5363	.5376	.5398	0.5810
3	.5737	.5563	.5400	.5250	.5118	.5009	.4939	.6291	.6233	.6189	.6156	.6137	.6139	.6191	0.5050
4	.6656	.6494	.6342	.6201	.6076	.5971	.5901	.6976	.6932	.6903	.6886	.6880	.6892	.6912	0.4949
5	.7458	.7317	.7183	.7058	.6943	.6845	.6777	.7608	.7567	.7537	.7517	.7508	.7514	.7542	0.4896

COMPONENT= 2		LIQUID COMPOSITION						VAPOUR COMPOSITION						MEAN	EMV
PLATE	DC	1	2	3	4	5	6	1	2	3	4	5	6		
1	.1291	.1284	.1276	.1268	.1259	.1251	.1246	.1348	.1358	.1366	.1373	.1378	.1382	.1367	0.9372
2	.1279	.1286	.1292	.1295	.1297	.1298	.1298	.1319	.1330	.1339	.1345	.1349	.1349	.1339	2.4383
3	.1359	.1369	.1377	.1383	.1387	.1389	.1390	.1287	.1298	.1306	.1311	.1313	.1312	.1305	1.0802
4	.1261	.1283	.1302	.1318	.1332	.1342	.1349	.1195	.1209	.1220	.1227	.1231	.1231	.1219	0.6154
5	.1116	.1146	.1173	.1196	.1217	.1234	.1245	.1076	.1090	.1100	.1106	.1110	.1108	.1098	0.5611

COMPONENT= 3		LIQUID COMPOSITION						VAPOUR COMPOSITION						MEAN	EMV
PLATE	DC	1	2	3	4	5	6	1	2	3	4	5	6		
1	.4965	.5154	.5331	.5493	.5639	.5759	.5837	.4096	.4176	.4252	.4322	.4381	.4421	.4275	1.2510
2	.4041	.4203	.4353	.4490	.4612	.4712	.4779	.3213	.3249	.3271	.3284	.3286	.3274	.3263	0.5687
3	.2904	.3068	.3223	.3366	.3495	.3601	.3670	.2422	.2468	.2505	.2533	.2550	.2549	.2505	0.4932
4	.2083	.2223	.2356	.2480	.2592	.2686	.2750	.1829	.1858	.1877	.1887	.1888	.1877	.1869	0.4822
5	.1426	.1537	.1644	.1745	.1840	.1921	.1978	.1316	.1344	.1363	.1376	.1382	.1377	.1360	0.4755

COMPONENT	L	M	H
FEED COMP.			
X_{ss}	·35	·35	·3
X_{ss}	·4	·35	·26

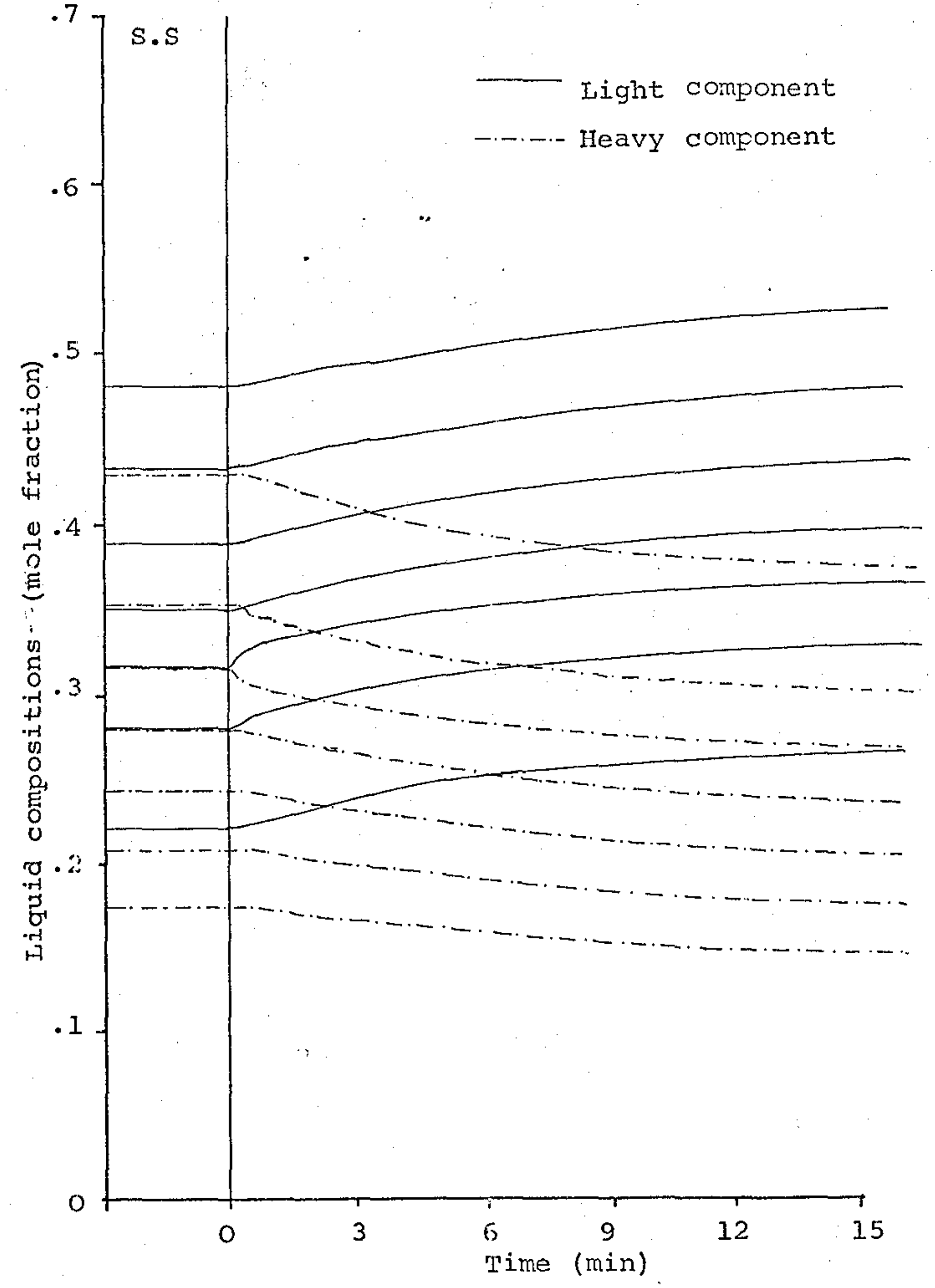


Fig 9.1 Liquid composition responses for step changes in feed composition. Light component goes from condenser to reboiler down the page, while the heavy component goes from reboiler to condenser. Relative volatilities of 1, 1.33 and 1.67 used.

COMPONENT	L	M	H
FEED COMP.			
X_{ss}	.35	.35	.3
X_{os}	.4	.35	.25

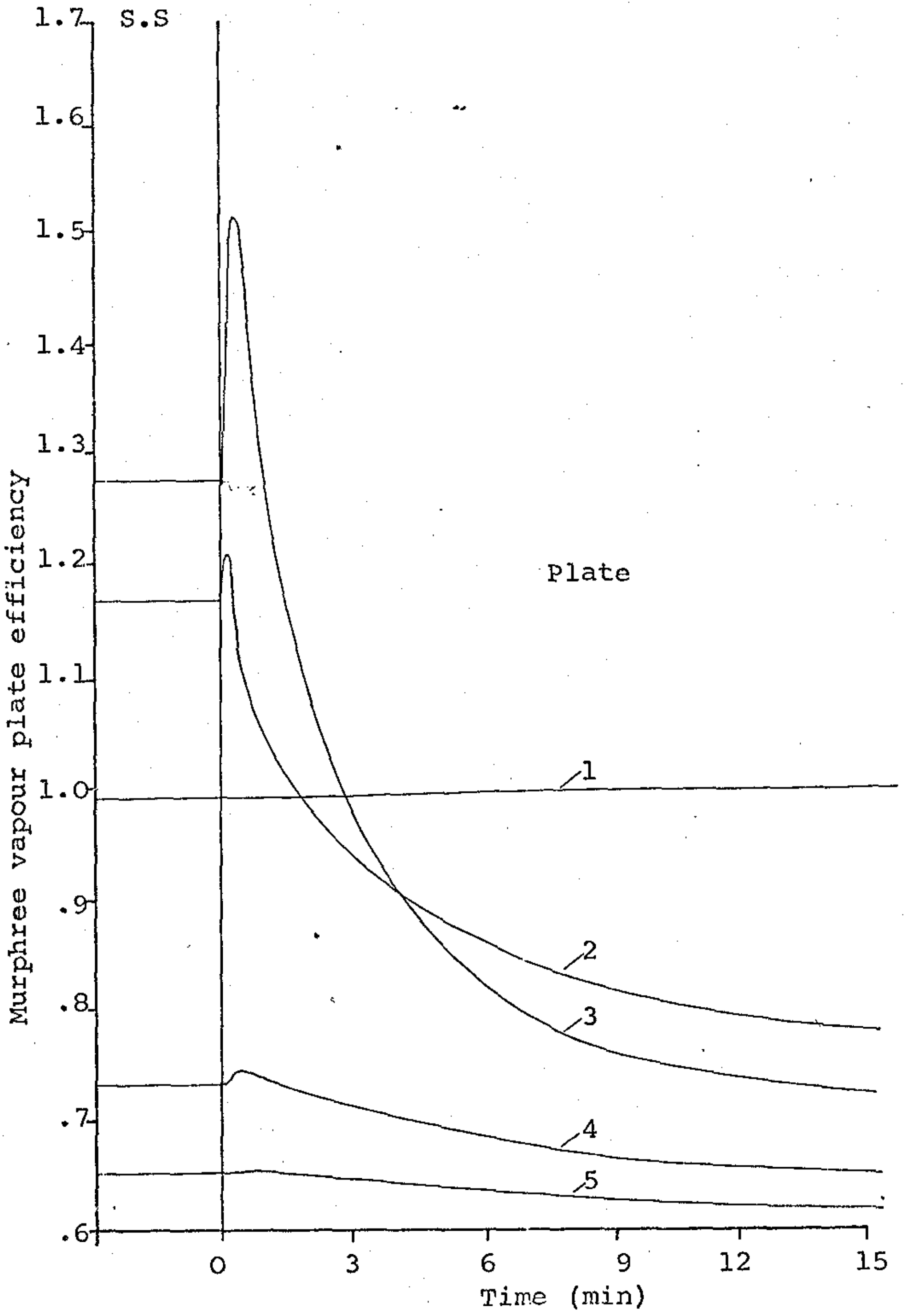


Fig 9.2 Middle component plate efficiency responses for step changes in feed composition.
Relative volatilities of 1, 1.33 and 1.67 used.

COMPONENT	L	M	H
FEED COMP.			
X_{SS}	·35	·35	·3
X_{OS}	·45	·3	·25

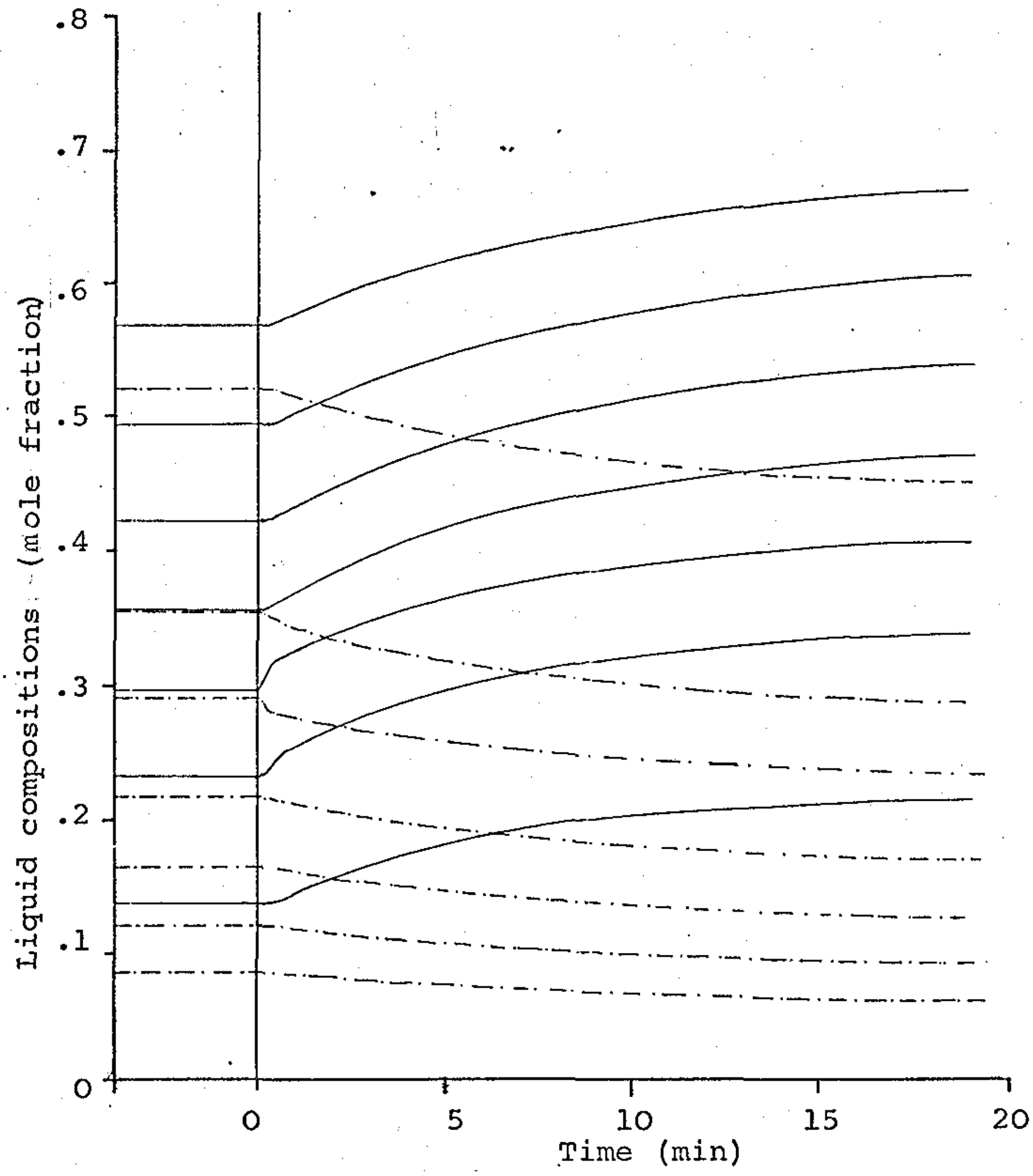


Fig 9.3 Liquid composition responses for step changes in feed composition.

——— Light component, going from condenser to reboiler down the page.

—·—·— Heavy component, going from reboiler to condenser down the page.

Relative volatilities of 3, 2 and 1 used.

COMPONENT FEED COMP.	L	M	H
X_{ss}	·35	·35	·3
X_{us}	·45	·3	·25

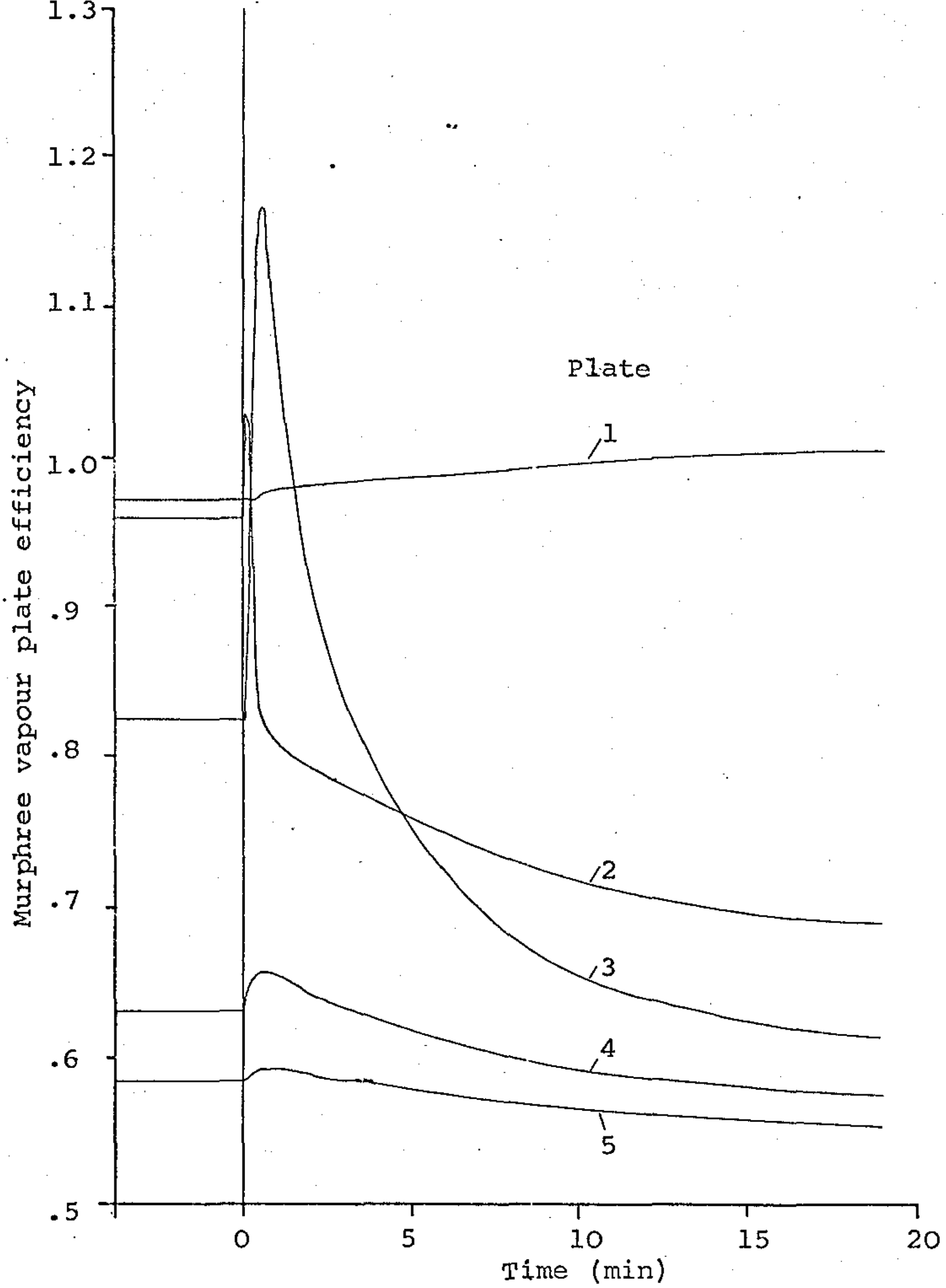


Fig 9.4 Middle component plate efficiency responses for step changes in feed composition.
Relative volatilities of 3, 2 and 1 used.

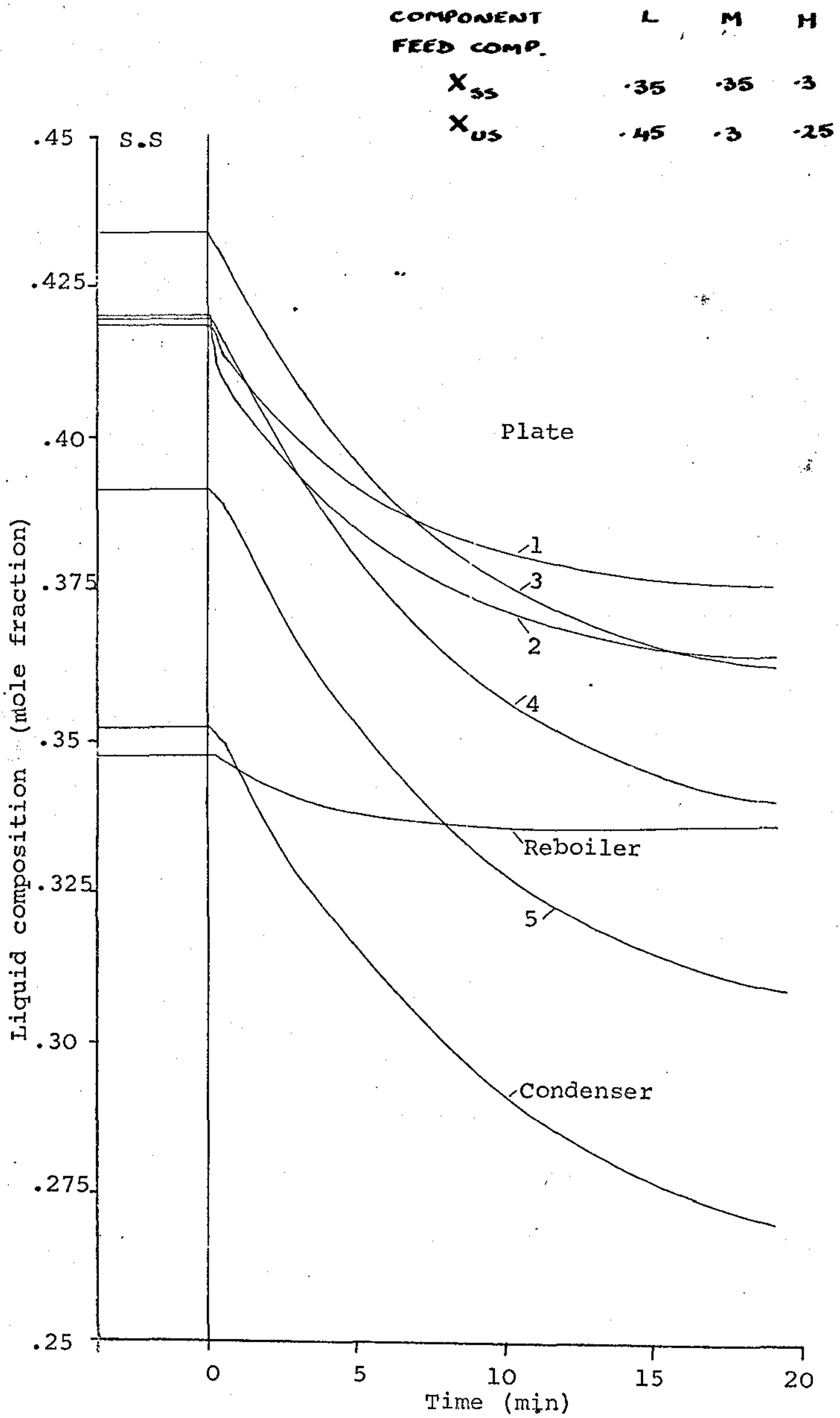


Fig 9.5 Middle component liquid composition responses for step changes in feed composition. Relative volatilities of 3, 2 and 1 used.

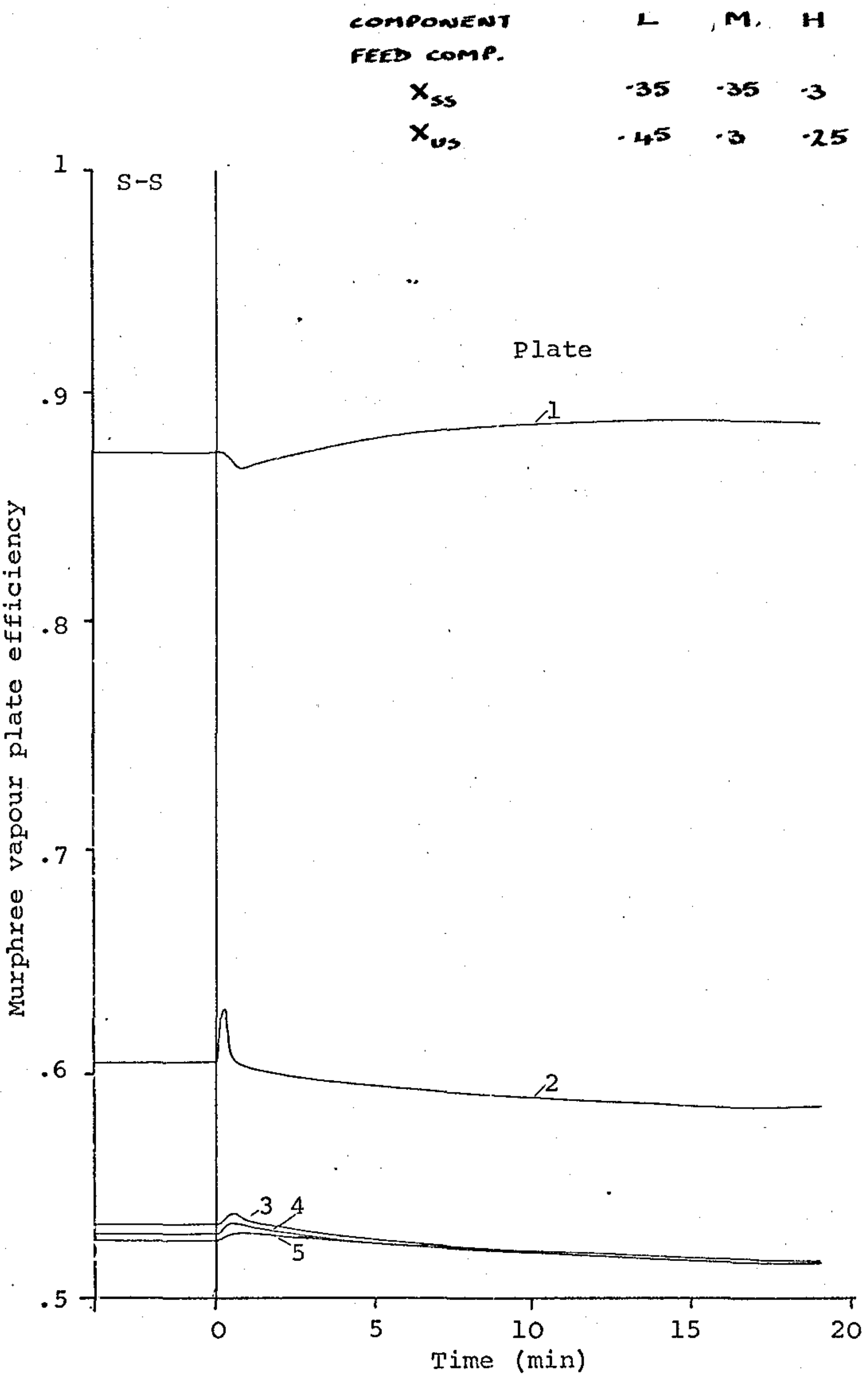


Fig 9.6 Light component plate efficiency responses for step changes in feed composition.
Relative volatilities of 3, 2 and 1 used.

COMPONENT	L	M	H
FEED COMP.			
X_{ss}	·35	·35	·3
X_{os}	·45	·3	·25

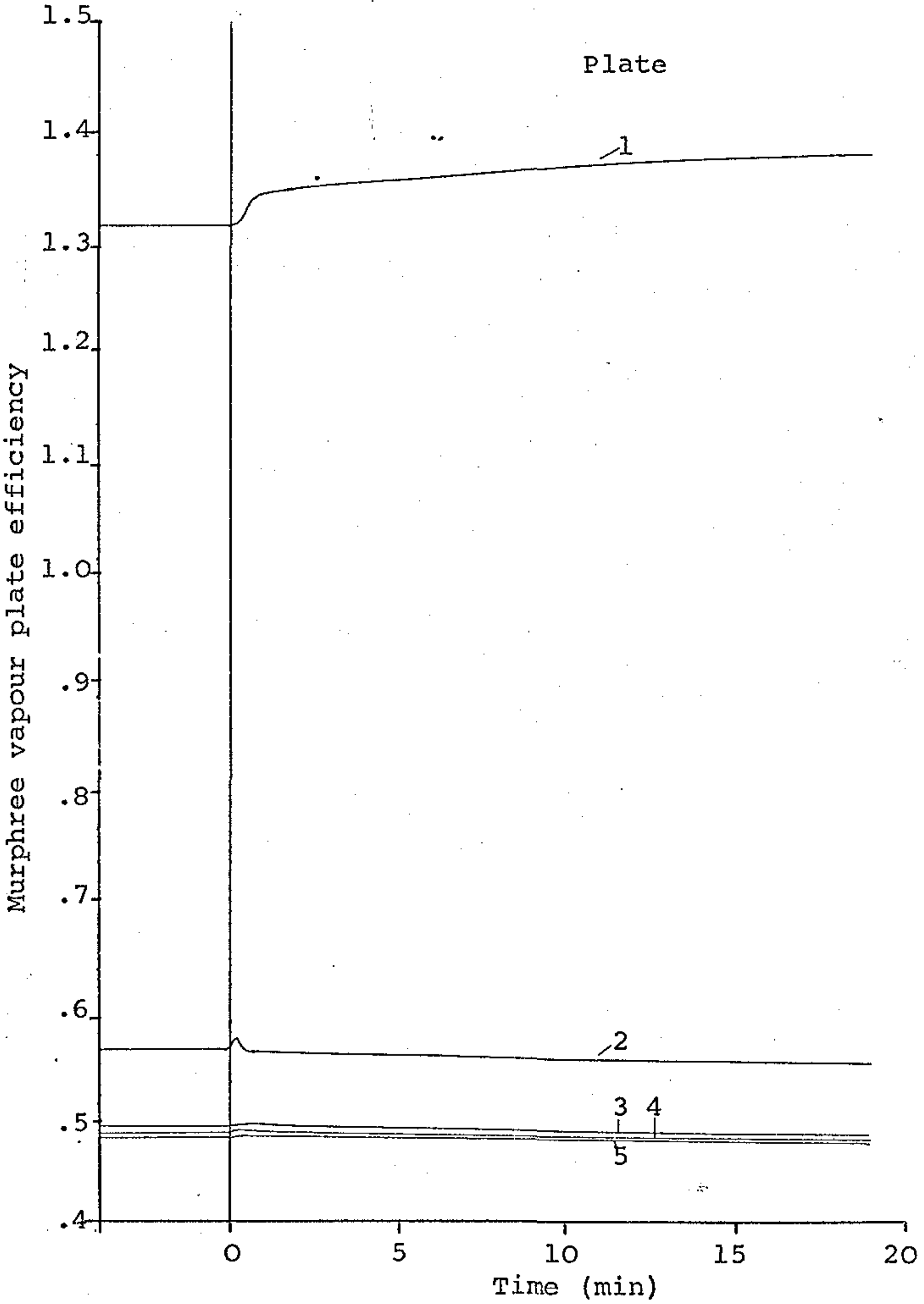


Fig 9.7 Heavy component plate efficiency responses for step changes in feed composition. Relative volatilities of 3, 2 and 1 used.

COMPONENT	L	M	H
FEED COMP.			
X_{ss}	·35	·35	·3
X_{us}	·1	·4	·5

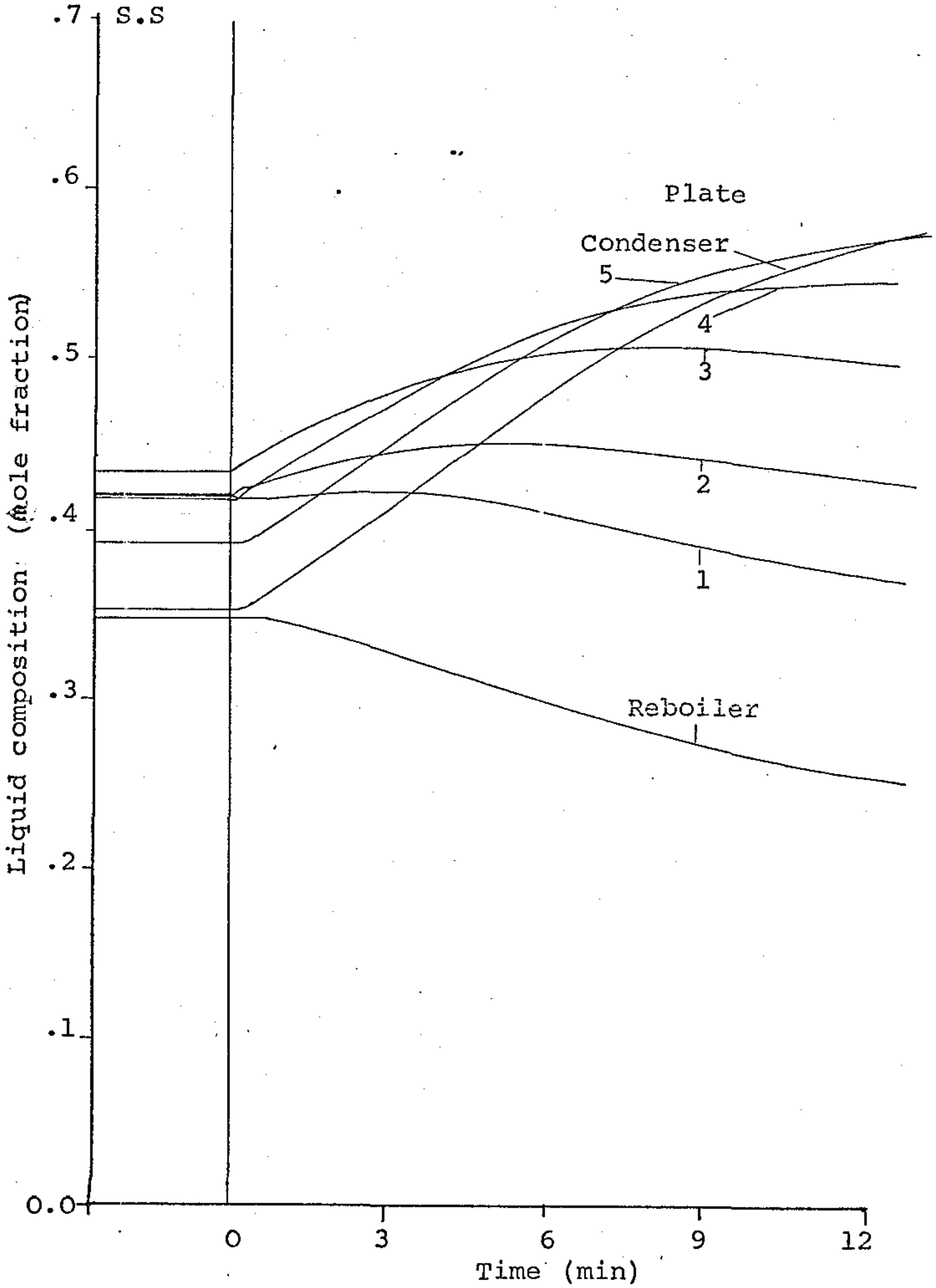


Fig 9.8 Middle component liquid composition responses for step changes in feed composition.
Relative volatilities of 3, 2 and 1 used.

COMPONENT	L	M	H
FEED COMP.			
X_{SS}	·35	·35	·3
X_{US}	·1	·4	·5

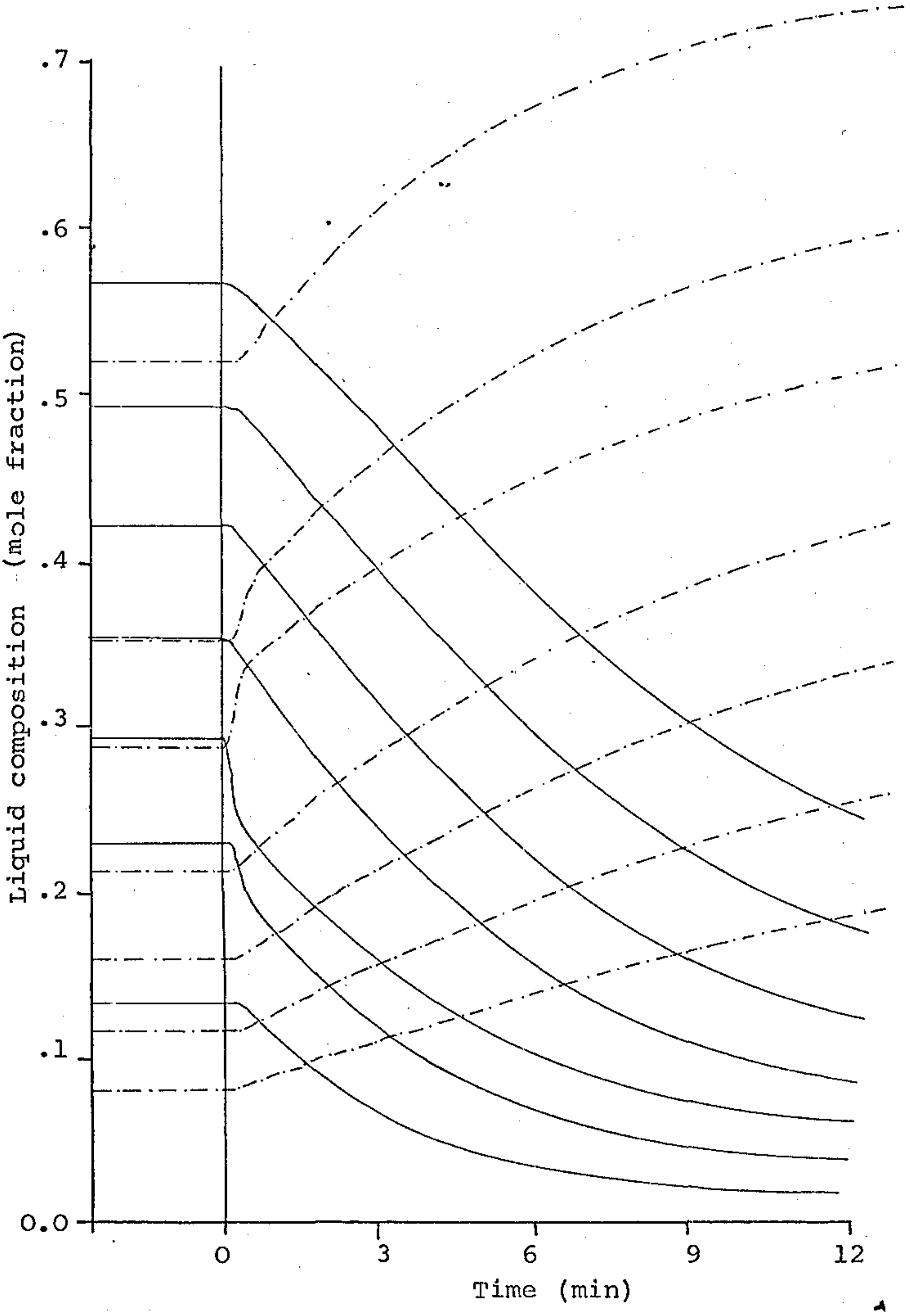


Fig 9.9 Liquid composition responses for step changes in feed composition.

——— Light component, going from condenser to reboiler down the page.

—·—·— Heavy component, going from reboiler to condenser down the page.

Relative volatilities of 3, 2 and 1 used.

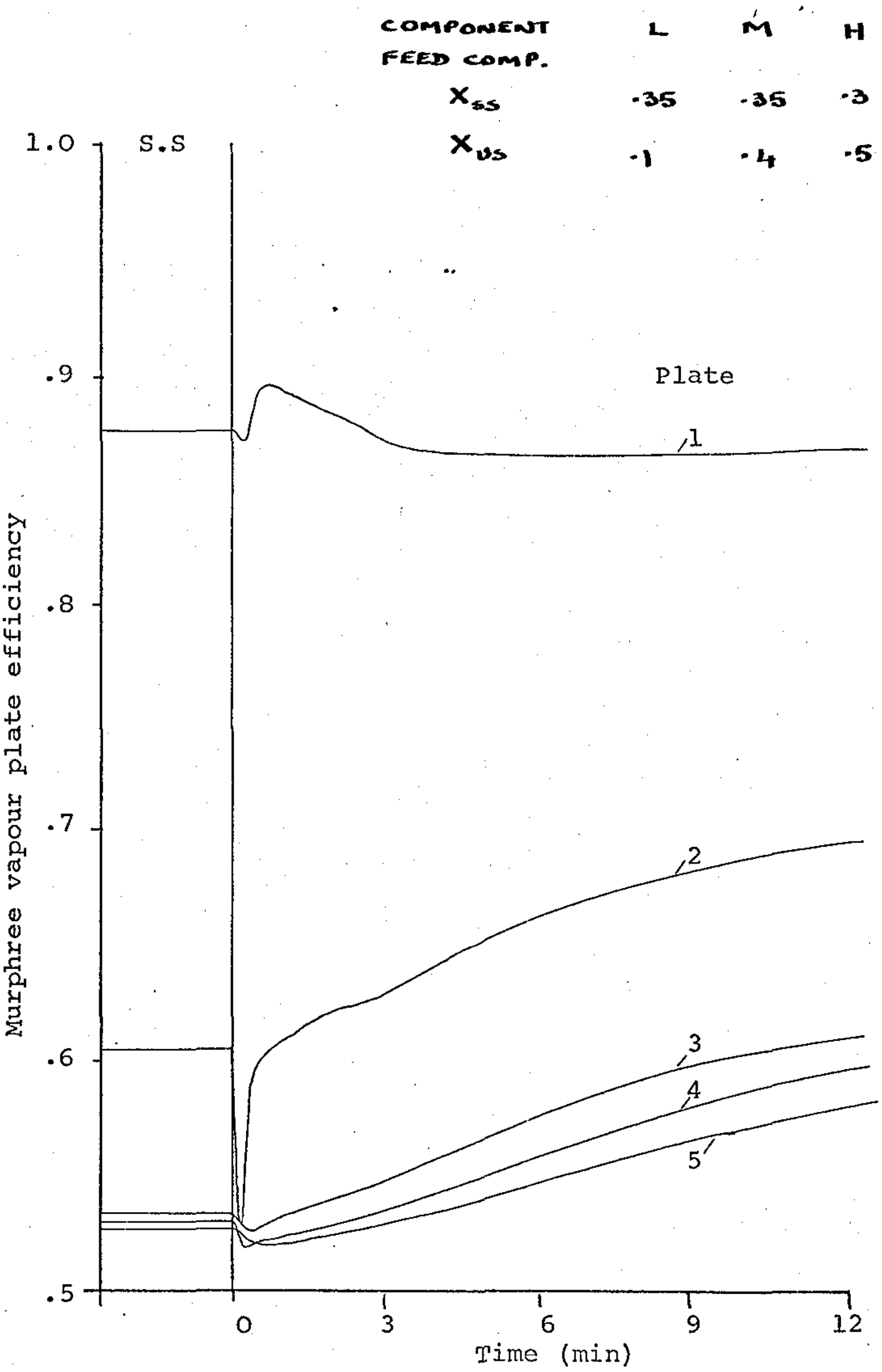


Fig 9.10 Light component plate efficiency responses for step changes in feed composition. Relative volatilities of 3, 2 and 1 used.

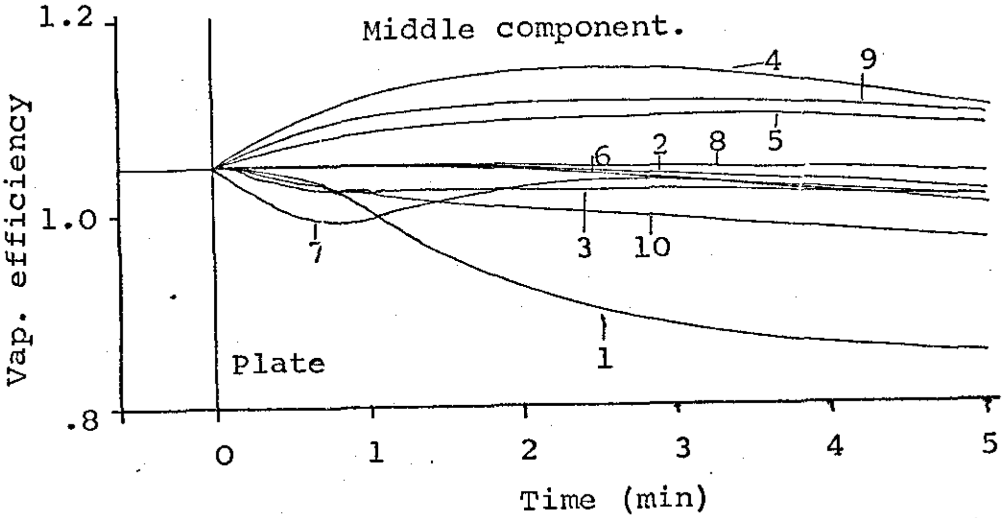


Fig 9.11 Middle component vaporisation efficiency responses for step changes in feed composition.

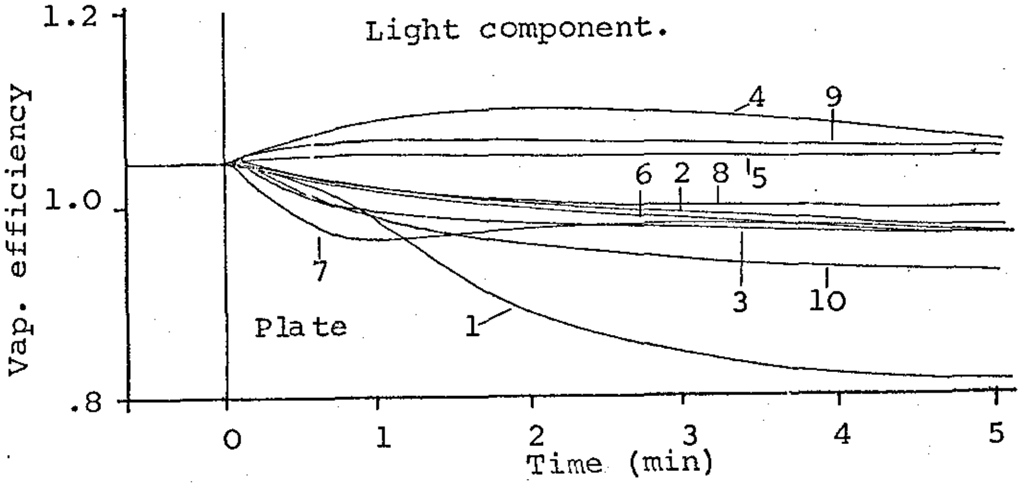
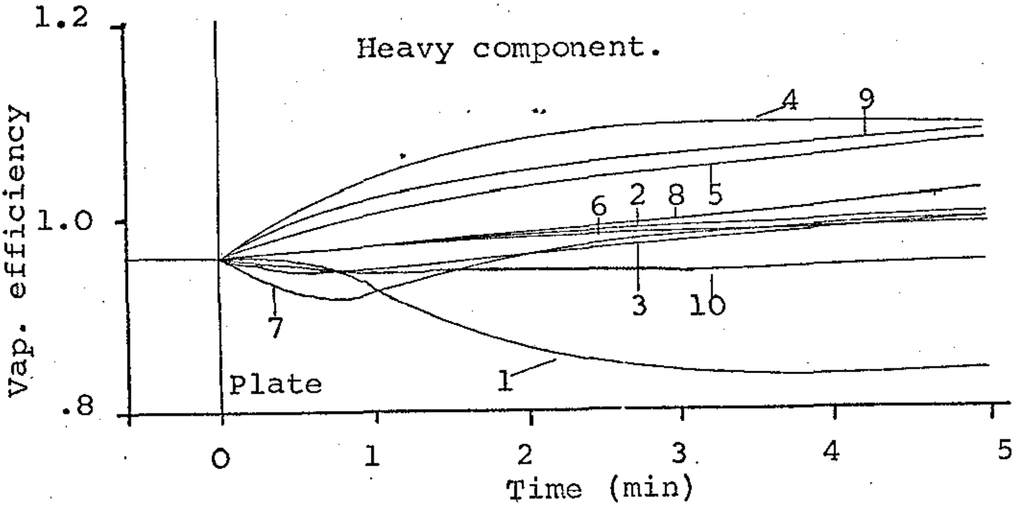


Fig 9.12 Light component vaporisation efficiency responses for step changes in feed composition.

Figures 9.11 to 9.13 are the vaporisation efficiency responses for the feed composition changes predicted by Holland (48).



Heavy component vaporisation efficiency responses for step changes in feed composition.

10. DISCUSSION

10 Discussion

10.1 Review of Work

From the computed unsteady-state liquid composition values the plate efficiencies were calculated. It can be seen from these responses that the magnitude of the unsteady-state plate efficiency responses have little effect on the liquid composition responses.

For a binary system whose relative volatility is greater than or equal to 2, step changes of less than 20% of the original value of the feed composition, liquid reflux from the condenser and liquid feed flow rate have no effect on the Murphree vapour plate efficiency. Systems with relative volatilities less than 2 and approaching unity show increases in plate efficiency changes as the relative volatility decreases. The plate showing the greatest change in its plate efficiency is the one on which the disturbance is introduced. Changes in the feed composition affect the plate efficiency responses more than the other disturbances.

Changes in boilup show that the size of the relative volatility is unimportant and that the responses for any relative volatility used follow the same pattern. Differences in the initial and final steady-state values of the plate efficiency are about 0.01 with maximum deviations of about 0.03 at unsteady-state. This difference is due to step changes in the point efficiency.

The assumption of perfect vapour mixing gives a more rapid response but also diminishes the magnitude of the maximum deviations. However, the dependence on the size of the relative volatility still holds. Similar responses are produced for larger diameter columns, only the response times are enlarged.

In gas absorption, the main changes in the plate efficiency are due to corresponding changes in the point efficiency. The time for the responses to settle seem to be longer, but the initial and final values of the plate efficiencies are similar.

Simplified models using perfect liquid mixing and linearised models can be used to produce the same liquid composition responses, if the plate efficiency used in the unsteady-state simulation has been calculated with the liquid mixing being accounted for. Use of the point efficiency in simplified models ignores the liquid mixing characteristics of the true model and gives different responses. From this it can be seen that the liquid mixing must be included either in the model or in the plate efficiencies used. The main point to be made is that if the initial and final steady-state values of the plate efficiencies differ, then if the final steady-state liquid compositions are required, it is best to use the final values of the plate efficiencies in the computation.

For the case of ternary systems, the plate efficiency responses of the light and heavy components showed little change during unsteady-state, while that of the middle component was erratic. Ternary systems can be evaluated using simplified models if the light and heavy components are the two operated on while the middle component is estimated by subtracting their sum from unity.

10.2 Suggestions for Further Work

From the work using the generalised model, model simplifications for the binary and ternary systems have been given. For these systems it has been shown that the changes in the plate efficiencies at unsteady-state have no effect on the liquid composition responses if the initial and final values of the plate efficiencies are the same.

The generalised model can be used to investigate systems that have more than three components. This work must be done as the work carried out by the author on ternary systems showed that the middle component was sensitive to disturbances and that the simplified models should operate on the light and heavy components in simulation work. If there are more than three components it may be that simplified models using constant plate efficiencies cannot be used.

NOTATION

Notation

B	backflow ratio B_f/L (eqn. 4.17)	-
$\bar{B}$	plate factor for vaporisation efficiency	-
B_f	recycle flow	lb moles/hr
c	concentration (specified units)	-
c_{out}	concentration at exit weir	-
CTP	calculated top product (eqn. 7.43)	lb moles/hr
De	eddy diffusion coefficient	ft^2/sec
D_s	top product flow	lb moles/hr
$\bar{E}$	component efficiency for vaporisation efficiency (eqn. 9.8)	-
E_a	plate efficiency if entrainment considered	-
E_G	generalised plate efficiency	-
E_H	thermal efficiency	-
E_{HG}	Hausen plate efficiency	-
E_L	overall liquid efficiency	-
E^M	modified Murphree plate efficiency	-
E_M	multicomponent Murphree plate efficiency	-
E_{ML}	Murphree liquid plate efficiency	-
E_{MV}	Murphree vapour plate efficiency	-
E_O	column efficiency (eqn. 3.2)	-
E^O	vaporisation efficiency	-
E_{OG}	vapour point efficiency	-
ETP	estimated top product (eqn. 7.42)	lb moles/hr
E_V	overall vapour efficiency	-

F	feed flow rate	lb moles/hr
H_c	condenser molal holdup	lb moles
h_c	height of clear liquid	ft
h_f	height of froth	ft
H_L	enthalpy of liquid stream	-
H_n	total molal holdup on tray n	lb moles
$H_{n,j}$	molal holdup of pool j on tray n	lb moles
$H_{n,O}$	molal holdup of downcomer on tray n	lb moles
H_R	reboiler molal holdup	lb moles
h_t	tray spacing	ft
H_V	enthalpy of vapour stream	-
L	liquid flow rate	lb moles/hr
Lg	liquid flow	gals/min ft
M	number of liquid pools representing the active area of the tray	-
m	slope of equilibrium line	-
N	number of plate in the column	-
N_G	number of stages for gas phase	-
N_L	number of stages for liquid phase	-
N_{OG}	overall number of stages for gas phase	-
Pe	Peclet number	-
Q	feed condition	-
$\tilde{r}$	vapour flow path (Fig 6.1)	ft
r_D	radial distance from injection axis	ft
T	residence time (eqn. 7.7)	-

u_g	superficial gas velocity	ft/sec
u_l	froth velocity	ft/sec
V	vapour flow rate	lb moles/hr
W	weir height	inches
W_s	bottom product flow rate	lb moles/hr
X	liquid phase composition	mole fraction
$\bar{X}$	steady-state liquid phase composition	mole fraction
x	liquid composition, deviation from steady-state	mole fraction
Y	vapour phase composition	mole fraction
y	vapour composition, deviation from steady-state	mole fraction
z	length of liquid path	ft
α	relative volatility	-
λ	absorption factor mV/L	-
μ_n, ψ_n, α_n	as defined by equation 7.33	-
σ	standard deviation	-
σ^2	normalised variance; area under the curve is unity	-
τ	space time , as defined by equation 7.29 and 7.38	unit time
ϕ	phi number; characteristic mixing parameter (eqn. 5.11)	-

Subscripts

B	rectifying section
c	condenser
ff	feed input
i	component number
j	pool number
n	tray number
	(these subscripts are always used in the order n,j,i ; the component subscript is always last)
R	reboiler
s	stripping section
ss	steady-state
us	unsteady-state

Superscripts

'	generalised equilibrium value
*	equilibrium value
-	average value in vapour cell model
-	steady-state value in linearised model .

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APPENDICES

Appendix A

Perforated Tray Hydraulics (1, 104, 107)

Liquid depth

The liquid depth on a tray should not be less than 2", and for a large tray 4" or greater is desirable. The liquid depth is the sum of the weir height h_w and the crest over the weir h_{ow}. The weir crest is calculated from the Francis formula:-

$$h_{ow} = 5.38 \left\{ \frac{L_c}{l_{eff}} \right\} \qquad \text{....... A.1}$$

Pressure drops

The pressure drop for the vapour phase is the sum of the effects for the vapour flow through the dry plate and those caused by the presence of the liquid:-

$$h_v = h_D + h_c + h_R \qquad \text{....... A.2}$$

The pressure drops are expressed as equivalent heads in inches of clear liquid of density ρ_L lb/ft^3 on the tray.

h_D, the dry pressure drop on entrance to the perforations, the friction within the passage through the plate, and the loss on exit from the plate is given by:-

$$h_D = \frac{12 C_D V_h^2 \rho_V}{2g \rho_L} \left\{ 0.4\left(1.25 - \frac{A_h}{A_n}\right) + \frac{4 l_t f}{d_h} + \left(1 - \frac{A_h}{A_n}\right)^2 \right\} \qquad \text{.... A.3}$$

The friction factor f is taken from a standard chart (9,114). The orifice coefficient is dependent on the plate thickness to hole diameter ratio. When the ratio is in the range 0.2 to 2 the coefficient is given by:-

$$C_D = 1.09\left\{\frac{d_h}{l_t}\right\}^{\frac{1}{4}} \qquad \text{.... A.4}$$

h_c, the hydraulic head, is the pressure drop due to the depth of liquid on the tray. In the active area of the tray, the liquid is actually froth. The equivalent depth of clear liquid h_c, is that which would be obtained if the froth collapsed. The height decreases with increase in vapour flow and is usually less than the height of the outlet weir. It is difficult to accurately determine this height, but it can be best calculated using:-

$$h_c = 0.24 + 0.725h_w - 0.29h_w u_g \rho_V^{0.5} + 4.48\,\frac{L_c}{l_w} \qquad \text{... A.5}$$

h_R, the residual vapour pressure drop is thought to be due to the necessity of overcoming surface tension as the vapour issues from a hole. The balance of the internal pressure in a static bubble required to overcome surface tension is:-

$$\frac{\pi d_p^2}{4}\,\Delta P_B = d_p \sigma_s \qquad \text{...... A.6}$$

$$\Delta P_B = \frac{4\sigma_s}{d_p} \qquad \text{...... A.7}$$

where ΔP_B is the excess pressure in the bubble owing to surface tension. The bubble of vapour grows over a finite time when the vapour flows and by averaging over time, it developes that the appropriate value is ΔP_R.

$$\Delta P_R = \frac{6\sigma_s}{d_p} \qquad \text{...... A.8}$$

As d_p cannot be readily calculated, the substitution of $d_h/12$ for d_p can be made to give an approximate diameter.

$$h_R = 12\,\frac{\Delta P_R}{\rho_L}.\frac{g_c}{g} = \frac{0.06\sigma}{\rho_L d_h} \qquad \text{....... A.9}$$

It has been stated (104) that the comparison of observed data with values of h_v calculated by these equations, show a standard deviation of 14.7%.

h_2, the pressure loss due to the liquid entering under the downcomer apron can be estimated by:-

$$h_2 = 0.558\left\{\frac{L_c}{A_{da}}\right\}^2 \qquad \text{....... A.10}$$

where A_{da} is the smaller of the two areas; the downcomer sectional area or the free area the downcomer apron and the tray.

h_3, the height of the backup in the downcomer, is the difference in liquid level inside the downcomer to that outside. This is equal to the sum of the pressure losses due to liquid and gas flow.

$$h_3 = h_v + h_2 \qquad \text{....... A.11}$$

It is inevitable that some of the liquid in the downcomer will be in the form of froth and thus a safe design requires that:-

$$h_w + h_{ow} + h_3 \leq \frac{h_t}{2} \qquad \text{...... A.12}$$

The visual froth height on the active tray can be estimated using:-

$$h_f = 2.53F^2 + 1.89h_w - 1.6 \qquad \text{...... A.13}$$

where the priming factor F is given by:-

$$F = u_g \rho_V^{\frac{1}{2}} \qquad \text{...... A.14}$$

The froth velocity is calculated by considering the tray length, the clear liquid height, and the average weir length w_1:-

$$w_1 = (T_d + T_w)/2 \qquad \text{...... A.15}$$

$$u_1 = h_c w_1 / L_c \qquad \text{...... A.16}$$

From these computed values of the clear liquid height, downcomer backup, froth height, froth velocity and the known column dimensions, vapour and liquid flow rates, the eddy diffusion coefficient can be determined from the empirical relationships derived in Chapter 5. From the eddy diffusion coefficient, the Peclet number and the backflow ratio can be calculated. Further, from calculated compositions and densities, the molal holdups for the theoretical model can be calculated.

Notation for Appendix A

A_{da}	smaller of the areas, downcomer section or downcomer and tray clearance.	ft^2
A_h	hole or free area	ft^2
A_n	net column cross-sectional area for vapour flow	ft^2
C_D	orific coefficient	-
d_h	hole diameter	inches
d_p	bubble or partical diameter	ft
F	priming factor	-
f	Fanning friction factor	-
h_2	head loss due to liquid flow under downcomer	inches liquid
h_3	backup in downcomer	"
h_c	hydraulic head	"
h_D	dry pressure drop , loss on entrance	"
h_f	froth height	inches
h_{ow}	liquid crest over weir	inches
h_t	tray spacing	inches
h_R	residual vapour pressure drop	inches liquid
h_v	sum of vapour pressure head	"
h_w	height of weir	inches
L_c	liquid flow rate	ft^3/sec
l_{eff}	effective length of weir	ft
l_t	tray thickness	inches

u_g	superficial gas velocity	ft/sec
v_h	velocity of vapour through holes	ft/sec
ρ_L	liquid density	lb/ft^3
ρ_V	vapour density	lb/ft^3
σ_s	surface tension	dynes/cm

Appendix B

Steady-State Experimental Theory: Liquid Mixing (13)

An axially-dispersed plug flow stream which is tracer free enters via a closed boundary at $z = -z_1$ with a velocity of u_{11}. A second stream containing tracer joins the first stream via another closed boundary at $z = 0$ and is instantaneously mixed. The cross section remains the same and the flow now has a velocity u_{12}, $(z > 0)$. The fluid stream then leaves the system at a third closed boundary $z = z_2$ with a concentration c_{out}. For steady-state there is no flux between $(-z_1, 0)$ and a constant net flux in $(0, z_2)$:-

$$u_{11}c - De_1\frac{dc}{dz} = 0 \qquad -z_1 < z < 0 \qquad \ldots\ldots\ B.1$$

$$u_{12}c - De_2\frac{dc}{dz} = u_{12}c_{out} \qquad 0 < z < z_2 \qquad \ldots\ldots\ B.2$$

The solution of equation B.1 is:-

$$c = c_{out}\exp(\ u_{12}z/De_1\ ) \qquad \ldots\ldots\ B.3$$

For experimental purposes, the tracer input was negligibly small compared with the bulk flow:-

$$u_{11} = u_{12} = u_1 \quad ; \quad De_1 = De_2 = De \quad \ldots\ B.4$$

From equation B.3, the eddy diffusivity can now be determined:-

$$De = \frac{u_1 z}{\ln(c/c_{out})} \qquad \ldots\ldots\ B.5$$

Appendix C

Theory of Eddy Diffusion in the Gas Phase (91)

It is assumed that the main flow is in free turbulence and that the mass flux may be expressed in terms of a constant eddy diffusivity. Therefore, it is justifiable to assume that there is only one nonzero component of the velocity in the cylindrical coordinate system Fig C.1.

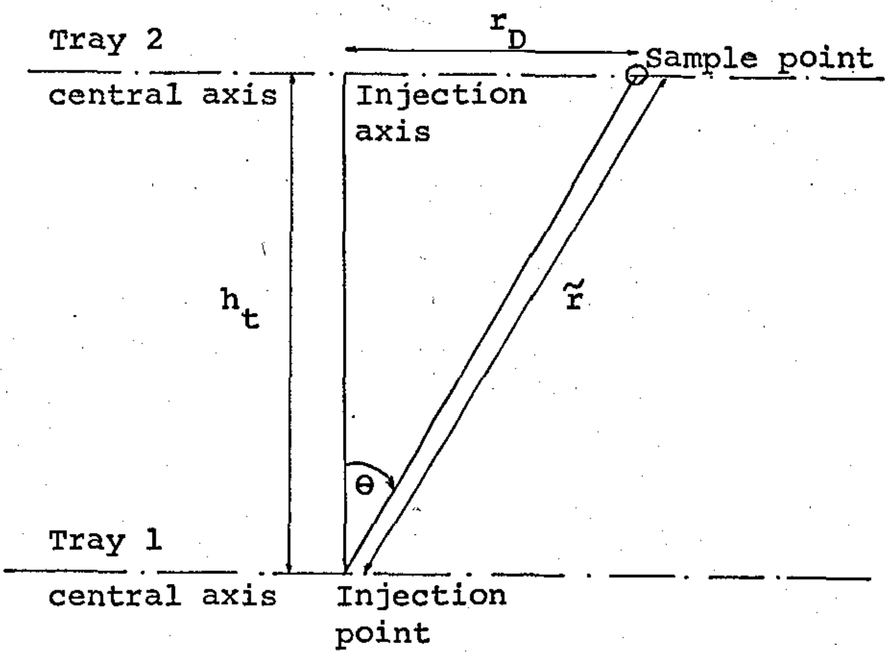

Fig C.1 Schematic representation of the vapour mixing system.

$$V_r = V_\theta = 0 \qquad \ldots\ldots\text{ C.1}$$

Assuming incompressible flow the equation of continuity becomes zero and the steady-state rate of change in concentration for any constant distance $\tilde{r}$ is zero. Thus the reduced form of the diffusion equation for the tracer is:-

$$V_{h_t}\left\{ \frac{dc}{dh_t} \right\}_{\tilde{r}} = De\left\{ \frac{d^2c}{dh_t^2} \right\}_{\tilde{r}} \qquad \text{..... C.2}$$

As the air is pure downstream from the point of injection, one of the boundary conditions is:-

$$\text{as } \tilde{r} \to \infty \qquad c \to 0 \qquad \text{..... C.3}$$

for a constant value of $\tilde{r}$, the mass flow rate of the tracer at the injection point is given by:-

$$C = \int_0^{2\pi}\int_0^{\pi} N_{\tilde{r}}\tilde{r}^2 \sin\theta d\theta d\psi = 2\pi\int_0^{\pi} (\bar{\rho}V_{\tilde{r}} - \rho De\frac{dc}{d\tilde{r}})\tilde{r}^2 \sin\theta d\theta \qquad \text{...... C.4}$$

If it is assumed that:-

$$\text{as } \tilde{r} \to 0 \qquad \bar{\rho}\tilde{r} \to k_1 = \text{a constant} \qquad \text{...... C.5}$$

then it follows that:-

$$\text{as } \tilde{r} \to 0\text{:} \qquad \int_0^{\pi} \bar{\rho}V_{\tilde{r}}\tilde{r}^2 \sin\theta d\theta = \int_0^{\pi} \bar{\rho}V_{h_t}\tilde{r}^2 \sin\theta\cos\theta d\theta = 0 \qquad \text{....... C.6}$$

$$\text{and as } \tilde{r} \to 0\text{:} \qquad C = -2\pi\int_0^{\pi} \rho De\frac{dc}{d\tilde{r}}\tilde{r}^2 \sin\theta d\theta$$

$$= -2\pi\rho De\frac{dc}{d\tilde{r}}\tilde{r}^2 \int_0^{\pi} \sin\theta d\theta$$

$$= -4\pi\rho De\frac{dc}{d\tilde{r}}\tilde{r}^2 \qquad \text{..... C.7}$$

Equation C.7 has been derived using equation C.5 in that

$$\text{as } \tilde{r} \to 0 \qquad \tilde{r}^2\frac{dc}{d\tilde{r}} = -\frac{k_1}{\rho} \qquad \text{..... C.8}$$

A solution to equation C.2 consistent with the boundary conditions C.3 and C.7 was suggested by Wilson (115) to be of the form:-

$$c = e^{-\alpha h_t} \phi(\tilde{r}) \qquad \ldots\ldots \text{C.9}$$

From this assumed solution the following equation can be calculated:-

$$\left\{\frac{dc}{dh_t}\right\}_r = -\alpha e^{-\alpha h_t}\phi + e^{-\alpha h_t}\left\{\frac{d\phi}{dh_t}\right\}_r \qquad \ldots\ldots \text{C.10}$$

From which equation C.2 becomes:-

$$V_{h_t}\left\{ -\alpha e^{-\alpha h_t}\phi + e^{-\alpha h_t}\left\{\frac{d\phi}{dh_t}\right\}_r \right\} = De\left\{ \alpha^2 e^{-\alpha h_t}\phi - \right.$$

$$\left. \alpha e^{-\alpha h_t}\left\{\frac{d\phi}{dh_t}\right\}_r - \alpha e^{-\alpha h_t}\left\{\frac{d\phi}{dh_t}\right\}_r + e^{-\alpha h_t}\left\{\frac{d^2\phi}{dh_t^2}\right\}_r \right\} \quad \ldots \text{C.11}$$

rearranging and dividing by $e^{-\alpha h_t}$ gives:-

$$\phi\left\{\frac{-V_{h_t}\alpha}{De} - \alpha^2\right\} + \left\{\frac{d\phi}{dh_t}\right\}_r \left\{\frac{V_{h_t}}{De} + 2\alpha\right\} = \left\{\frac{d^2\phi}{dh_t^2}\right\}_r \qquad \ldots\ldots\ldots \text{C.12}$$

If α is defined by:-

$$\alpha \equiv \frac{V_{h_t}}{2De} \qquad \ldots\ldots\ldots \text{C.13}$$

then equation C.12 reduces to:-

$$\phi\alpha^2 = \left\{\frac{d^2\phi}{dh_t^2}\right\}_r \qquad \ldots\ldots\ldots \text{C.14}$$

which in cylindrical coordinates gives:-

$$\phi\alpha^2 = \frac{d^2\phi}{d\tilde{r}^2} + \frac{2}{\tilde{r}}\cdot\frac{d\phi}{d\tilde{r}} \qquad \ldots\ldots\ldots \text{C.15}$$

Taking a change of variable:-

$$Y \equiv \tilde{r}\phi \qquad \text{....... C.16}$$

gives equation C.15 as:-

$$\alpha^2 Y = \frac{d^2 Y}{d\tilde{r}^2} \qquad \text{....... C.17}$$

for which the solution is:-

$$Y \equiv \tilde{r}\phi = Ae^{\alpha r} + Be^{-\alpha r} \qquad \text{....... C.18}$$

For boundary condition C.3 to be satisfied $r \rightarrow \infty$; $c \rightarrow 0$ constant $B = 0$ as α is always negative.

From this result and from equations C.9, C.16, C.18 it is found that:-

$$c = \frac{A}{\tilde{r}}.\exp(\alpha(\tilde{r} - h_t)) \qquad \text{....... C.19}$$

Boundary condition C.7 states that:-

$$\text{as } r \rightarrow 0 \; ; \; C \rightarrow 4\pi\rho DeA \qquad \text{....... C.20}$$

$$\text{or} \quad A \equiv \frac{C}{4\pi\rho De} \qquad \text{....... C.21}$$

Equations C.13, C.19, C.22 represent the mass fraction distribution for the tracer in free-turbulent flow described.

The useful result for experimental purposes is given by the equation for mass fraction:-

$$c = \frac{C}{4\pi De\rho\tilde{r}}\exp\left\{\frac{-V_h}{2De}t.(\tilde{r} - h_t)\right\} \qquad \text{....... C.22}$$

If the volumetric flow of the tracer is Q_h, then:-

$$c = \frac{Q_h}{4\pi De\tilde{r}}.\exp\left\{\frac{-V_h}{2De}t.(\tilde{r} - h_t)\right\} \qquad \text{....... C.22}$$

Notation for Appendix C

A, B	constants	-
C	mass flow rate of tracer	lb/sec
c	mole fraction of tracer	-
De	eddy diffusion coefficient	ft^2/sec
h_t	tray height	ft
k_1	constant	-
Q_h	volumetric flow rate of tracer	ft^3/sec
r	radial distance	ft
$\tilde{r}$	distance from injection to sample	ft
V_{h_t}	velocity of main stream	ft/sec
V_r, V_θ	velocity in cylindrical coordinate system	ft/sec
Y	change of variable (eqn. C.16)	-
α	(see eqn. C.13)	-
ρ	density of tracer	lb/ft^3
$\bar{\rho}$	molal density of tracer	lb $moles/ft^3$
θ	angle in coordinate system	radians
ϕ	(see eqn. C.9)	-

Appendix D

Generalised Model, Unsteady-State Simulation Program

D.1 Generalised Model Subroutines

Master unsteady

This subroutine gives a listing of the nomenclature used for all the subroutines. Any parameters that are not listed in this section but appear in the program are of no importance but have been introduced to ease the computation. At the end of the subroutine the initial working subroutine is called.

Subroutine Sec 1

This subroutine reads in the data needed to set the initial steady-state data and calculate column dimensions and system parameters.

Subroutine System

This subroutine is the centre of the program. The initial data read in includes the run time and printout step lengths for the initial steady-state calculation and the flag markers which indicate how the flow changes, if any, are made. The liquid and vapour flow rates are calculated and the integration parameters are initialised. The subroutine calls the integration routine and the calculated data is returned and written onto a disc in order to save on core storage. The maximum and minimum values of the

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Subroutine Diffun

This routine contains all the differential equations to be operated on. At the end of this routine the top product comparison is carried out to see how near to steady-state the calculation is. This subroutine calls two other subroutines which are needed to calculate the average densities of the liquid and vapour phases, the molal holdups and the point efficiency for each plate.

Subroutine Flow 1

This subroutine calculates the average densities of both the liquid and vapour phases and the volumetric flow rates. If liquid flow rate changes are introduced, the subroutine Flow 2 is called.

Subroutine Flow 2

The subroutine is only called if flow changes occur. The new flow rates are estimated and then the routine returns to Flow 1.

Subroutine Equida

The vapour composition in equilibrium with the liquid and the temperature of the liquid are calculated according to the equilibrium relationship used.

Subroutine AICHE

All the parameters involved with calculating the point and plate efficiencies are brought over into this routine. The slope of the equilibrium line and the point efficiency for each plate is calculated according to the relationships given in the A.I.Ch.E Bubble-Tray Design Manual (1).

Subroutine IPNEU

This subroutine contains a polynomial equation for calculating the collision integral for diffusion.

Subroutine IDNEU

This subroutine contains a polynomial equation for calculating the collision integral for diffusion.

Subroutine Writer

The calculated data stored on disc is read and written in a presentable form. The storage used for the integration routine is now used for storing this data. The liquid composition data, the plate and point efficiency data and the efficiency ratio data are stored in an array for carrying over to the plotting routine. This array is finally written on to disc to be stored for plotting.

Master Drawer

This is the master routine of another program that is run immediately after Master Unsteady. The reason for having to run another program is that the plotting routine available had a limitation on the core storage that could be used during the program and this was exceeded in the unsteady-state simulation. The values to be plotted are read from the disc and transfered to the plotting routine.

Subroutine Plot 1

This routine plots the liquid composition responses, the point efficiency, plate efficiency and efficiency ratio responses for the 5 plate column.

The data needed for running the program is always read in in free format. The parameters read in are in the order:-

N, NF, QFS, W, TL
AMA, BMA, DMA, DMB, VZ1, VZ2
AKI, FS, XFS, WS, LP, LQ, AREL
XS(1)
XS(2) to XS(total)
YV(1) to YV(N)
SLOPE, CONST, IREL, TB1, TB2
TCAN, STEP1, STEP2, HMAX1, HMAX3
IDIST
IREL
IMARK, NIMP, AIN, VVS, XFS

Nomenclature used:-

N	the number of plates including the reboiler as one plate.
NF	the feed plate number counting upwards from the reboiler.

QFS	the q-line feed condition.
W	the weir height (inches).
TL	the column diameter (ft).
AMA, BMA	the molecular weights of the volatile and nonvolatile components.
DMA, DMB	the liquid densities at the boiling temperature of the volatile and nonvolatile components. (gm/cc)
VZ1,VZ2	the liquid viscosity at the boiling points of the volatile and nonvolatile components. (centipoise)
AKI	the liquid flow rate leaving the bottom plate for the reboiler. (lb moles/hr)
FS	feed flow rate. (lb moles/hr)
XFS	the volatile component feed composition, (mf)
WS	the bottom product flow rate. (lb moles/hr)
LP	the number of liquid pools going to form 1 vapour cell. For no vapour mixing LP = 1 and for perfect mixing LP = LQ
Lq	the number of liquid pools for the active area.
AREL	the relative volatility.
XS	the liquid composition of the more volatile component, (mf). XS(1) is the reboiler value and XS(2) is the composition of the last pool on the bottom plate.
YV	the average vapour composition leaving the plate. YV(1) is the reboiler value, (mf).
SLOPE	slope of the equilibrium line if linear relationship

	is used. If not used then any values as they are not considered during computation.
CONST	the constant in the linear equilibrium relationship.
IREL	equilibrium relationship flag: set to 1 if constant relative volatility used: set to 2 if linear equilibrium relationship used.
TB1, TB2	the boiling points of the volatile and nonvolatile components respectively. (^{o}c)
TCAN	the time for the steady-state simulation before the disturbance is introduced. (sec)
STEP1, STEP2	the time between write statements being operated. (sec)
HMAX1, HMAX3	the maximum integration step lengths that will be allowed for the steady-state and unsteady-state simulations. (sec)
IDIST	an integration flag that if set to 2 uses the step length taken by the integration program, but if set to 1 doubles the allowable step length to decrease the computation time needed.
IREL	a flag set to 1 for ordinary use but set to 2 for vapour flow changes.
IMARK	a constant molal flow flag: set to 1 for variable molal flow and set to 2 for constant molal flow in the unsteady-state simulation.

NIMP	the plate number where the disturbance is introduced.
AIN	the change in the flow. (lb moles/hr)
VVS	the change in vapour flow. (lb moles/hr)
XFS	the new unsteady-state liquid feed composition. (mf)

```
      MASTER UNSTEADY
C
C
C     NF=FEED PLATE     N=NUMBER OF PLATES WHERE REBOILER IS PLATE ONE
C     QFS=Q LINE VALUE,ASSUMED = OR > THAN 1
C     COLUMN DIMENSIONS:- TW=WIDTH, TAA=ACTIVE LENGTH, TL=TOTAL LENGTH,
C     PHYSICAL PROPERTIES:-      VOLATILE         NONVOLATILE
C              DENSITIES           DMA                  DMB     GM/CC.
C           MOLE WEIGHTS           AMA                  BMA
C            VISCOSITIES          VZ1                  VZ2
C          BOILING TEMP.           TB1                  TB2
C     WEIR HT. INS.= W     ACTIVE AREA=TA       POINT EFFICIENCY=E : EP
C     NUMBER OF POOLS FOR ACTIVE AREA=LQ
C     NUMBER OF VAPOUR CELLS=KR, NUMBER OF LIQUID POOLS/VAPOUR CELL=LP
C     REFLUX RATIO=RR         BACKFLOW RATIO=BETA
C     LIQUID FLOW =ALS          VAPOUR FLOW =VV,VS,VR
C     MOLAR HOLDUPS= VD,VT,VR,VC   FOR DOWNCOMER,TRAY,REBOILER,CONDENSER
C     FEED RATE=FS,    FEED CONCENTRATION=XFS
C     BOTTOM RATE=WS    TOP RATE=DS
C     ESTIMATED TOP PRODUCT=ETP,   CALCULATED TOP PRODUCT=CTP
C     CONCENTRATIONS:-   LIQUID=XS ; Y      VAPOUR=YS     VAPOUR CELL=YC
C     PECLET NO.=PE,   EDDY DIFFUSION COEFFICIENT=DE
C     HEIGHTS :-  CLEAR LIQUID=HC        FROTH=HF
C     VELOCITIES:-  FROTH=UL     SUPERFICIAL GAS=UG
C     DENSITIES LB/CU.FT      LIQUID=PL      VAPOUR=PV
C     SPACE TIMES SECS.   REBOILER=TAUR    TEMPERATURES   TBOIL
C                        DOWNCOMER=TAUD                   TEXIT
C                            TRAYS=TAUT                   TEXIT
C                        CONDENSER=TAUC                   TCOND
C
C     EQUILIBRIUM VALUE OF VAPOUR=YE
C         AVERAGE VALUE OF VAPOUR=YV
C     MURPHREE VAPOUR PLATE EFFICIENCY=EMVP
C
C
C
```

```
C
C      PARAMETERS TO BE READ IN AND THE ORDER
C      ALL READS ARE IN FREE FORMATS.                  NUMBER OF DATA/CARD.
C      N,NF,QFS,W,TL                                                5
C      AMA,BMA,DMA,DMB,VZ1,VZ2                                      6
C      AK1,FS,XFS,WS,LP,LQ,AREL                                     7
C      XS(1)                                                        1
C      XS(2) TO XS(TOTAL)                                           7
C      YV(1) TO YV(N)                                               6
C      SLOPE,CONST,IREL,TB1,TB2                                     5
C      TCAN,STEP1,STEP2,HMAX1,HMAX3                                 5
C      IDIST                                                        1
C      MF                                                           1
C      IREL                                                         1
C      IMARK,NIMP,AIN,XFS                                           4
C
C      TCAN =ONE QUARTER OF THE TOTAL TIME TO BE RUN.
C      STEP1=THE PRINTING STEP AND  HMAX1=INTEGRATION STEP.
C      IDIST=1 IF THE INITIAL INTEGRATION STEP AFTER THE DISTURBANCE
C      HAS BEEN INTRODUCED CAN BE DOUBLED.
C      MF=0,1,2 DEPENDING ON THE NUMERICAL INTEGRATION USED.
C      IREL=1,2,3 DEPENDING ON THE BOIL UP RATE AND REFLUX.
C      IMARK=2 FOR CONSTANT MOLAR HOLDUP, OTHERWISE IMARK=1
C      NIMP=PLATE NUMBER FOR DISTURBANCE , AIN=THE SIZE OF FLOW CHANGE
C      XFS=THE NEW VALUE OF THE FEED COMPOSITION
C
C
C
C
C      -------------------------------------------------------------------
       CALL SEC1
       STOP
       END
```

```
      SUBROUTINE SEC1
      DIMENSION XS(40),ALS(10)
      DIMENSION EMVP(10),YV(10),TEXIT(10),YE(10)
      COMMON/FEED/XFS,E,FS,WS,DS,EMVP,YV,TEXIT,YF
      COMMON/ALPH/AREL,IREL
      COMMON/DATE/AMA,BMA,DMA,DMB,W,TAA,TL,TA,TR,AD,VZ1,VZ2,QFS,N,NF,TW,
     1LQ,XS,ALS,KR,LP,LZ,HDIA,TTH,HOA,ANT,AUD,COV
      COMMON/EQU/SLOPE,CONST,TB1,TB2
C
C     THIS SUBROUTINE CALCULATES THE COLUMN DIMENSIONS AND THE
C     RELAVANT AREAS FOR USE IN EMPIRICAL EQUATIONS.
C
C
C     SET COLUMN AND SYSTEM PARAMETERS
C
      READ(1,101)N,NF,QFS,W,TL
  101 FORMAT(2I0,3F0.0)
      W=W/12
      READ(1,103)AMA,BMA,DMA,DMB,VZ1,VZ2
  103 FORMAT(6F0.0)
      READ(1,104)AKI,FS,XFS,WS,LP,LQ,AREL
  104 FORMAT(4F0.0,2I0,F0.0)
      TW=.885*TL
      TA1=SQRT((TL/2)**2-(.385*TL)**2)
      TD1=TL/2-TA1
      TW1=.77*TL
      AD2=TW1*TD1
      THETA=ASIN(TW1/TL)
      TTA=3.1415926*TL**2/4
      TTA=3.1415926*(TL**2)/4
      AS=TTA*THETA/3.1415926
      AD=AS-0.5*TA1*TW1
      TD2=TD1*AD/AD2
      TR=TD2+2*TA1
      TAA=2*TA1
      TA=TTA-2*AD
```

```
      HDIA=.1875
      TTH=0.1
      HOA=.1*TTA
      ANT=TTA-AD
      AUD=TW1*1.0/12.0
      COV=1.09*(HDIA/TTH)**0.25
      PIT=0.5
      APR=1.0
      W2=W*12.0
      WRITE(2,200)
  200 FORMAT(1H1,/////////28X,31HDISTILLATION COLUMN DIMENSIONS.)
      WRITE(2,202)
  202 FORMAT(/30X,21H  WEIR AND DOWNCOMER.)
      WRITE(2,201)
  201 FORMAT(/15X,6HDIA.FT,4X,38HHT.INS     LENG.FT     SEC.AREA     APRON)
      WRITE(2,203)TL.W2,TW1,AD,APR
  203 FORMAT(17X,F3.1,6X,F4.2,6X,F6.4,5X,F6.3,8X,F3.1)
      WRITE(2,204)
  204 FORMAT(//21X,35HPERFORATIONS                   TRAY AREAS.)
      WRITE(2,205)
  205 FORMAT(15X,48HDIA.INS    PITCH    AREA     NET     ACTIVE    GROSS)
      WRITE(2,206)HDIA,PIT,HOA,ANT,TA,TTA
  206 FORMAT(15X,F6.4,5X,F3.1,3X,F7.4,2X,F6.3,2X,F6.3,2X,F6.3)
      WRITE(2.207)
  207 FORMAT(//30X,26HDATA USED FOR COMPUTATION.)
      WRITE(2.208)
  208 FORMAT(15X,56HM.WT.A    M.WT.B    ROE A    ROE B    VIS A    VIS B    RE
     1L.V.)
      WRITE(2.209)AMA,BMA,DMA,DMB,VZ1,VZ2,AREL
  209 FORMAT(15X,2(F6.2,3X),2(F5.3,3X),2(F5.2,3X),F6.2)
C
C     READ IN INITIAL ESTIMATES OF TRAY CONCENTRATIONS
C     SET FEED RATE AND COMPOSITION
C
      NLIM=(N-1)*(LQ+1)+2
      READ(1,102)XS(1)
```

```
      READ(1,102)(XS(I),I=2,NLIM)
  102 FORMAT(7F0.0)
      READ(1,100)(YV(I),I=1,N)
  100 FORMAT(6F0.0)
      READ(1,105)SLOPE,CONST,IREL,TB1,TB2
  105 FORMAT(2F0.0,I0,2F0.0)
      ALS(2)=AKI
      KR=LQ/LP
      LZ=LQ-LP
      CALL SYSTEM
      RETURN
      END
```

```
      SUBROUTINE SYSTEM
      DIMENSION EP(10),DL(10),DG(10),ANO(10),ANG(10),ANL(10),AB(10),ES(1
     10)
      DIMENSION ALE(10),ERA(10),TIM(100)
      DIMENSION TEXIT(10)
      DIMENSION XR(40),XS(40),ALS(10),YS(40),YC(50),YE(10),YV(10),EMVP(1
     10)
      DIMENSION Y(8,40),YMAX(40),SAVE(40,12),PSAVE(1600),ERROR(40),        1
     1A(8),PERTST(7,2,3)
      DIMENSION DENS(10),VDEN(10),PEN(10),B(10),SL(10),WD(10),WT(10)
      COMMON/SSL/ALLS(10)
      COMMON/METH/MF
      COMMON/NUM/IY
      COMMON/GRAF/EOMAX,EOMIN,ELMAX,ELMIN,AMAXE,AMINE
      COMMON/FEED/XFS,E,FS,WS,DS,EMVP,YV,TEXIT,YE
      COMMON/ISAR/I
      COMMON/PFS1/Y,SAVE,PSAVE,YMAX,ERROR,A,PERTST
      COMMON/EPRS/ALE,ERA,EP,DL,DG,ANO,ANG,ANL,AB,ES,XR,WD,WT,VR,VS,DENS
     1,VDEN,B,PEN,CTP,ETP,TCOND,VD,VT,VC,V8,YS,YC,TBOIL,IC,IZ,HT
      COMMON/DIST/IMARK,IDIST
      COMMON/DATE/AMA,BMA,DMA,DMB,W,TAA,TL,TA,TR,AD,VZ1,VZ2,QFS,N,NF,TW,
     1LQ,XS,ALS,KR,LP,LZ,HDIA,TTH,HOA,ANT,AUD,COV
      COMMON/ALPH/AREL,IREL
      COMMON/REFL/IREF
      COMMON/RELV/RV(10)
C
C     THIS SUBROUTINE IS THE CENTRE OF THE PROGRAM.
C     INITIALISES THE INTEGRATION PARAMETES. CALLS THE INTEGRATION
C     ROUTINE THEN WRITES ON TO DISC. IT THEN CALLS THE WRITING ROUTINE
C     WHICH READS THE STORED DATA ON DISC AND WRITES IT OUT.
C
      READ(1,100)TCAN,STEP1,STEP2,HMAX1,HMAX3
  100 FORMAT(5F0.0)
      READ(1,101)IDIST
      READ(1,101)MF
      READ(1,101)IREF
```

```
  101 FORMAT(I0)
 C
C      CALCULATE FLOW RATES OF STREAMS
C      SET VALUES FOR THE BOTTOM RATE
C
      HT=0.0
      IZ=0
      IC=0
      DS=FS-WS
      DO 2 I=2,NF-1
    2 ALS(I+1)=ALS(I)
      DO 8 I=(NF+1),(N+1)
    8 ALS(I)=ALS(NF)-FS*QFS
      DO 14 I=2,N+1
   14 ALLS(I)=ALS(I)
      VS=(ALS(2)-WS)/LQ
      VR=(QFS*FS+ALS(NF+1)-WS)/LO
      XWS=XS(1)
C
C      SET INTEGRATION PARAMETERS.
C
      EPS=0.1
      DO 3  I=1,40
      YMAX(I)=1.0
      ERROR(I)=0.
      DO 3 J=1,40
3     PSAVE(J+(I-1)*40)=0.
      DO 4 I=1,8
      DO 4 J=1,40
4     Y(I,J)=0.
      DO 5 I=1,12
      DO 5 J=1,40
5     SAVE(J,I)=0.
      DO 6 I=1,8
6      A(I)=0.
      DO 7 I=1,7
```

```
      DO 7 J=1,2
      DO 7 K=1,3
7     PERTST(I,J,K)=0.
       T=0.
C
C     SET THE MAXIMUM TIME TO RUN:- BIGT
C     SET THE MINIMUM, MAXIMUM AND INITIAL INTEGRATION STEP LENGTHS
C     SET THE INITIAL LIMITS FOR PLOTTING PURPOSES.
C
      H=1.0E-7
      HMIN=1.E-20
      BIGT=4*TCAN
      STEP=STEP1
      HMAX=HMAX1
      JSTART=0
      MAXDER=7
      NLIM=(N-1)*(LQ+1)+2
      DO 24 I=1,NLIM
      Y(1,I)=XS(I)
   24 CONTINUE
      NN=(N-1)*(LQ+1)+2
      N4=NN*NN
      TMAX=0.0
      I=0
      AMINE=2.0
      AMAXE=0.5
      EOMAX=0.0
      EOMIN=1.0
      ELMAX=0.0
      ELMIN=2.0
      RR=ALS(N+1)/DS
   20 CONTINUE
      IY=0
      IF(KFLAG.LT.0)IY=1
      CALL DIFSUB(NN,T,Y,SAVE,H,HMIN,HMAX,EPS,MF,YMAX,ERROR,KFLAG.JSTART
     1,MAXDER,PSAVE,N4)
```

```
      IF(KFLAG)11,11,12
   11 WRITE(2,13)KFLAG
      EPS=EPS*10
   13 FORMAT(3X,6HKFLAG=,I5)
      GO TO 20
   12 CONTINUE
      DO 1 JX=1,NLIM
    1 XS(JX)=Y(1,JX)
C
C     REFLUX SETTINGS.
C     IREF=1 FOR CONSTANT REFLUX WITH BOTTOMS VARYING.
C     IREF=2 CHANGE IN VAPOUR RATE
C     IREF=3 FOR NEW CONSTANT REFLUX WITH CHANGE IN BOTTOMS WITH
C     LIQUID FLOWS HELD CONSTANT AT NEW VALUE.
C
      WS=ALS(2)-VS*LQ
      RR=ALS(N+1)/DS
      DS=VR*LQ-ALS(N+1)
      IF(T.LT.TMAX)GO TO 20
      HT=T
      TMAX=TMAX+STEP
      I=I+1
      KC=0
      IF(IC.LT.1)TST=T
      TACT=T-TST
      TIM(I)=T/60
C
C     WRITE CALCULATED DATA ON TO DISC.
C
      WRITE(3)TACT
      WRITE(3)FS,XFS,WS,DS,RR
      WRITE(3)VB,VC
      WRITE(3)Y(1,1),Y(1,NLIM)
      WRITE(3)YS(1),CTP
      WRITE(3)TBOIL,TCOND,ETP
      WRITE(3)(Y(1,J*(LQ+1)+1),J=1,N-1)
```

```
      DO 26 JJ=1,LQ
      J=LQ+2-JJ
      WRITE(3)((Y(1,J+(KK-1)*(LQ+1)),YS(J+(KK-1)*(LQ+1))),KK=1,N-1)
   26 CONTINUE
      WRITE(3)(ALS(J),J=2,N)
      V8=VS*LQ
      VR1=VR*LQ
      WRITE(3)V8,VR1
      WRITE(3)(YV(J),J=1,N-1)
      WRITE(3)(YV(J),J=2,N)
      WRITE(3)(YE(J),J=2,N)
      WRITE(3)(TEXIT(J),J=2,N)
      WRITE(3)(PEN(J),J=2,N)
      WRITE(3)(B(J),J=2,N)
      WRITE(3)(ES(J),J=2,N)
      WRITE(3)(AB(J),J=2,N)
      WRITE(3)(EP(J),J=2,N)
      WRITE(3)(EMVP(J),J=2,N)
      WRITE(3)(ERA(J),J=2,N)
      WRITE(3)(RV(J),J=2,N)
      WRITE(3)(XR(J),J=2,N)
C
C     SET NEW LIMITS FOR PLOTTING.
C
      DO 10 JJ=2,N
      IF(EMVP(JJ).GE.AMAXE)AMAXE=EMVP(JJ)
      IF(EMVP(JJ).LE.AMINE)AMINE=EMVP(JJ)
      IF(EP(JJ).LE.EOMIN)EOMIN=EP(JJ)
      IF(EP(JJ).GE.EOMAX)EOMAX=EP(JJ)
      IF(ERA(JJ).LE.ELMIN)ELMIN=ERA(JJ)
      IF(ERA(JJ).GE.ELMAX)ELMAX=ERA(JJ)
   10 CONTINUE
C
C     CHECK FOR TIME THE DISTURBANCE IS TO BE INTRODUCED.
C     SET NEW LIMITS FOR INTEGRATION STEP LENGTHS.
C
```

```
      IF(IC.EQ.1)IC=2
      IF(IC.EQ.2)GO TO 22
      IF(IC.EQ.0.AND.T.GE.TCAN)IC=1
      IF(IC.EQ.0)GO TO 22
      IF(IC.EQ.1)TMAX=T
      IF(IC.EQ.1)H=0.02
      STEP=STEP2
      EPS=0.001
   22 CONTINUE
      IF(I.GE.90)GO TO 31
      IF(T.LT.BIGT)GO TO 20
   31 CONTINUE
C
C     SET FINAL LIMITS FOR PLOTTING
C
      EOMIN=(INT(EOMIN*10))/10.0
      EOMAX=(INT(EOMAX*10+1))/10.0
      ELMIN=(INT(ELMIN*10))/10.0
      ELMAX=(INT(ELMAX*10+1))/10.0
      AMINE=(INT(AMINE*10))/10.0
      AMAXE=(INT(AMAXE*10+1))/10.0
C
C     CALCULATED TRUE TIME IN MINUTES:- TIM(J)
C
      DO 9 J=1,I
    9 TIM(J)=TIM(J)-TST/60.0
      CALL WRITER(I,TIM,LQ,N)
      RETURN
      END
```

```
      SUBROUTINE DIFFUN(T,Y,SAVE,N2,NN)
      DIMENSION PWD(10),PWT(10)
      DIMENSION Y8(40),AB(10),ES(10),EPP(2),WTM(2),DSD(2),COD(2),VIS(2)
      DIMENSION THI(2),DG(10),VLG(10),DL(10),ANG(10),ANL(10),ANO(10),EP(
     110)
      DIMENSION SN(10),SLM(2)
      DIMENSION Y(8,37),SAVE(444,1)
      DIMENSION YE(10)
      DIMENSION EMVP(10),YV(10),TEXIT(10)
      DIMENSION XR(40),XS(40),ALS(10),YS(40),YC(50)
      DIMENSION DENS(10),VDEN(10),PEN(10),B(10),SL(10),WD(10),WT(10)
      DIMENSION ALE(10),ERA(10)
      COMMON/ALPH/AREL,IREL
      COMMON/REFL/IREF
      COMMON/NS/NIMP
      COMMON/DATE/AMA,BMA,DMA,DMB,W,TAA,TL,TA,TR,AD,VZ1,VZ2,QFS,N,NF,TW,
     1LQ,XS,ALS,KR,LP,LZ,HDIA,TTH,HOA,ANT,AUD,COV
      COMMON/HOLD/HOL(10)
      COMMON/RELV/RV(10)
      COMMON/HDIF/TPD(10),TPT(10)
      COMMON/SSL/ALLS(10)
      COMMON/DIST/IMARK,IDIST
      COMMON/FEED/XFS,E,FS,WS,DS,EMVP,YV,TEXIT,YE
      COMMON/NUM/IY
      COMMON/EPRS/ALE,ERA,EP,DL,DG,ANO,ANG,ANL,AB,ES,XR,WD,WT,VR,VS,DENS
     1,VDEN,B,PEN,CTP,ETP,TCOND,VD,VT,VC,VB,YS,YC,TBOIL,IC,IZ,HT
C
C     THIS ROUTINE CONTAINS ALL THE DIFFERENTIAL EQUATIONS
C
      WTM(1)=AMA
      WTM(2)=BMA
      DSD(1)=DMA
      DSD(2)=DMB
      IY=IY+1
      IF(IMARK.NE.2)IMARK=1
      IF(IY.NE.1)GO TO 20
```

```
      IF(IC.NE.1)GO TO 20
C
C     SET IMARK EQUAL TO 2 FOR NO CHANGE IN FLOWS.
C
C
C     STEP CHANGE MADE IN ONE OF THE PARAMETERS
C
C
C     INITIATE ANY CHANGES IN FLOW RATE
C     CHANGES IN THIS SECTION ARE FOR STEP CHANGES ONLY.
C
      READ(1,100)IMARK,NIMP,AIN,VVS,XFS
  100 FORMAT(2I0,3F0.0)
      IF(IREF.EQ.2)VS=VS+VVS/LQ
      IF(IREF.EQ.2)WS=ALS(2)-VS*LQ
   20 CONTINUE
      ALS(1)=WS
      IF(IV.NE.1)GO TO 2
      IF(IMARK.EQ.2)GO TO 2
      IF(IREF.EQ.2)GO TO 2
C
C     INITIALISE THE FLOW CHANGES AND THE STEADY-STATE VALUES THAT WILL
C     BE ACHIEVED EVENTUALLY
C
C
      IF(IC.NE.1)GO TO 1
      IF(NIMP EQ.NF)GO TO 3
      ALS(NIMP+1)=ALS(NIMP+1)+AIN
      DO 4 J1=1,N+1
    4 ALLS(J1)=ALS(J1)+AIN
      GO TO 2
    3 TST=T
      TAU=(WD(NF)+WT(NF))*3600.0/ALS(NF)
      ALSS=ALS(NF)
      DO 5 J1=1,NF
    5 ALLS(J1)=ALS(J1)+AIN
```

```
      FS=FS+AIN
    1 IF(IC.NE.2)GO TO 2
      IF(NIMP.NE.NF)GO TO 2
      TSI=(T-TST)/TAU
      IF(TSI.GE.20.0)TSI=20.0
      ALS(NF)=ALSS+AIN*(1.0-EXP(-TSI))
      IF(TSI.GE.20.0)ALS(NF)=ALSS+AIN
    2 CONTINUE
      IF(IY.EQ.1)CALL FLOW1(IMARK,IC,N,XR,WD,WT,VS,ALS,WS,LQ,VR,OFS,FS,D
     1S,XS,AMA,BMA,YS,DMA,DMB,DENS,TEXIT,VDEN,SL,VVB,VB,TL,TW,T,HT,TBOIL
     2,YV)
      IF(IY.EQ.1)CALL AICHE(IMARK,SL,YV,VB,EPP,WTM,DSD,CDD,VIS,THI,VLG,S
     1N,SLM,TEXIT)
C
C     REBOILER BALANCE
C
      XZ=Y(1,1)
      CALL EQUIDA(XZ,YZ,TEP)
      TBOIL=TEP
      YS(1)=YZ
      YV(1)=YS(1)
      TAUR=3600.*VB/ALS(2)
      SAVE(N2,1)=(Y(1,2)-(WS*Y(1,1)+VS*YS(1)*LQ)/ALS(2))/TAUR
      DO 12 I=1,KR
   12 YC(I)=YS(1)
      KT=KR
      DO 13 J1=2,N
      VV=VS
      IF(J1.GE.NF)VV=VR
      J=(J1-1)*(LQ+1)-(LQ+1)
      SUM3=0.0
      SUM2=0.0
      BETA=B(J1)
      ALPHA=B(J1)/(1+B(J1))
C
C     SET HOLDUPS
```

```
C
      VD=WD(J1)
      VT=WT(J1)
      TAUT=3600.*VT/ALS(J1)
      TAUD=3600*VD/ALS(J1)
      IF(IY.NE.1)GO TO 22
      IF(IMARK.EQ.1.AND.IC.LT.1)GO TO 27
      IF(IMARK.EQ.2)GO TO 27
      IF(ALLS(J1).EQ.ALS(J1))GO TO 27
      GO TO 26
   27 TVC=0.0
      TVD=0.0
      TPT(J1)=0.0
      TPD(J1)=0.0
      GO TO 28
   26 CONTINUE
      TVC=VT/((VT+VD)*LQ)
      TVD=VD/(VT+VD)
      TPT(J1)=TVC*(ALS(J1+1)/ALS(J1)-1.0)*LQ/TAUT
      TPD(J1)=TVD*(ALS(J1+1)/ALS(J1)-1.0)/TAUD
      IF(J1.NE.NF)GO TO 28
      TPT(J1)=TVC*((ALS(J1+1)+QFS*FS)/ALS(J1)-1.0)*LQ/TAUT
      TPD(J1)=TVD*((ALS(J1+1)+QFS*FS)/ALS(J1)-1.0)/TAUD
   28 CONTINUE
   22 CONTINUE
      E=EP(J1)
C
C     EXIT POOL
C
      KZ=(J+LQ+2-J1-LZ)/LP
      XZ=Y(1,J+2)
      CALL EQUIDA(XZ,YZ,TEP)
      YS(J+2)=YZ
      TEXIT(J1)=TEP
      SUM3=YS(J+2)
      YS(J+2)=E*YS(J+2)+(1-E)*YC(KZ)
```

```
      SAVE(N2+J+1,1)=((1+BETA)*Y(1,J+3)-(1+BETA)*Y(1,J+2)+VV*(YC(KZ)-YS(
     1J+2))/ALS(J1)-Y(1,J+2)*(ALS(J1+1)/ALS(J1)-1.0)*TVC)*LQ/TAUT
      IF(J1.EQ.NF)SAVE(N2+J+1,1)=((1+BETA)*Y(1,J+3)-(1+BETA)*Y(1,J+2)+(V
     1S*YC(KZ)-VR*YS(J+2))/ALS(J1)-Y(1,J+2)*((ALS(J1+1)+QFS*FS)/ALS(J1)-
     21.0)*TVC)*LQ/TAUT
      SUM2=SUM2+YS(J+2)
C
C     CENTRE POOLS ON THE TRAY
C
      DO 16 I=3,(LQ+1)
      XZ=Y(1,J+I)
      CALL EQUIDA(XZ,YZ,TEP)
      YS(J+I)=YZ
      KZ=(J+I+LQ-J1-LZ)/LP
      YS(J+I)=E*YS(J+I)+(1-E)*YC(KZ)
      SUM2=SUM2+YS(J+I)
      SAVE(N2+J+I-1,1)=((1+BETA)*Y(1,J+I+1)-(1+2*BETA)*Y(1,J+I)+BETA*Y(1
     1,J+I-1)+VV*(YC(KZ)-YS(J+I))/ALS(J1)-Y(1,J+I)*(ALS(J1+1)/ALS(J1)-1.
     20)*TVC)*LQ/TAUT
      IF(J1.NE.NF)GO TO 6
      SAVE(N2+J+I-1,1)=((1.0+BETA)*Y(1,J+I+1)-(1.0+2.0*BETA)*Y(1,J+I)+BE
     1TA*Y(1,J+I-1)+(VS*YC(KZ)-YS(J+I)*VR)/ALS(J1)-Y(1,J+I)*((ALS(J1+1)+
     2QFS*FS)/ALS(J1)-1.0)*TVC)*LQ/TAUT
    6 CONTINUE
   16 CONTINUE
C
C     DOWNCOMER AT BEGINNING OF TRAY
C
      SAVE(N2+J+LQ+1,1)=((ALS(J1+1)*Y(1,J+LQ+3)/ALS(J1)+BETA*Y(1,J+LQ+1)
     1)-(1.0+BETA)*Y(1,J+LQ+2)-Y(1,J+LQ+2)*(ALS(J1+1)/ALS(J1)-1.0)*TVD)/
     2TAUD
      IF(J1.EQ.NF)SAVE(N2+J+LQ+1,1)=((ALS(J1+1)*Y(1,J+LQ+3)+QFS*FS*XFS)/
     1ALS(J1)+BETA*Y(1,J+LQ+1)-((1+BETA)+((ALS(J1+1)+QFS*FS)/ALS(J1)-1.0
     2)*TVD)*Y(1,J+LQ+2))/TAUD
C
C     VAPOUR CELL CONCENTRATION
```

```
C
      YE(J1)=SUM3
      RV(J1)=YE(J1)*(1.0-Y(1,J+2))/(Y(1,J+2)*(1.0-YE(J1)))
      YV(J1)=SUM2/LQ
C
C     PLATE EFFICIENCY AND EFFICIENCY RATIO CALCULATIONS
C
      EMVP(J1)=(YV(J1)-YV(J1-1))/(YE(J1)-YV(J1-1))
      K2=J+LQ+2
      DO 17 I=1,KR
      SUM1=0.0
      DO 18 K1=1,LP
   18 SUM1=SUM1+YS(K2-K1)
      K2=K2-LP
      KT=KT+1
   17 YC(KT)=SUM1/LP
      ALE(J1)=EP(J1)*AB(J1)
      ERA(J1)=EMVP(J1)/EP(J1)
   13 CONTINUE
      SUM1=0.0
      DO 19 KP=1,KR
   19 SUM1=SUM1+YC(KT-KR+KP)
      SUM1=SUM1/KR
      YOUT=SUM1
C
C     CONDENSER BALANCE
C
      TAUC=3600*VC/(VV*LQ)
      SAVE(N2+LQ+J+2,1)=(YOUT-Y(1,J+LQ+3))/TAUC
      XZ=Y(1,J+LQ+3)
      CALL EQUIDA(XZ,YZ,TEP)
      TCOND=TEP
      DS=VR*LQ-ALS(N+1)
C
C     TOP PRODUCT COMPARISON.
C
```

```
CTP=DS*Y(1,J+LQ+3)
ETP=FS*XFS-WS*Y(1,1)
DIFF=CTP-ETP
RETURN
END
```

```
      SUBROUTINE FLOW1(IMARK,IC,N,XR,WD.WT,VS.ALS.WS,LQ,VR,QFS,FS,DS,XS,
     1AMA.BMA,YS,DMA,DMB,DENS,TEXIT.VDEN,SL.VVB,VB,TL,TW,T,HT,TBOIL.VV)
      DIMENSION XR(40),WD(10),WT(10),ALS(10),XS(40),YS(40),DENS(10),TEXI
     1T(10),VDEN(10).SL(10),YV(10)
C
C     THIS ROUTINE CALCULATES THE AVERAGE DENSITIES AND VOLUMETRIC FLOWS
C
C     CALCULATE THE AVERAGE DENSITY, TEMPERATURE AND VOLUMETRIC FLOW
C     ON EACH PLATE.
C
   25 XR(J)=WD(J)+WT(J)
      VS=(ALS(2)-WS)/LQ
      VR=VS+(1-QFS)*FS/LQ
C
C     CALLS FLOW2 IF FLOW CHANGES ARE INTRODUCED
C
      IF(IMARK.EQ.1.AND.IC.GE.1)CALL FLOW2(TEXIT,VS,VR,YS.WD,WT,DENS,VDE
     1N,YV,T,HT,FS,IMARK)
      DO 10 I=2,N
      J3=(I-1)*(LQ+1)-(LQ+1)+1
      ZT=0.0
      ZD=0.0
      ZV=0.0
      ZL=0.0
      DO 3 J=1,LQ+1
      J4=J3+J
      ZD=ZD+XS(J4)*AMA/(XS(J4)*AMA+(1-XS(J4))*BMA)
      ZL=ZL+XS(J4)*AMA+(1-XS(J4))*BMA
      XZ=XS(J4)
      IF(J.EQ.(LQ+1))GO TO 3
      CALL EQUIDA(XZ,YZ,TEP)
      YS(J3+J)=YZ
      ZT=ZT+TEP
      ZV=ZV+YS(J3+J)*AMA+(1-YS(J3+J))*BMA
    3 CONTINUE
      ZD=ZD/(LQ+1)
```

```
      ZL=ZL/(LQ+1)
      ZV=ZV/LQ
      ZT=ZT/LQ
    7 CONTINUE
      DENS(I)=ZD*DMA+(1-ZD)*DMB
      TEXIT(I)=ZT
      VV=VS
      IF(I.GE.NF)VV=VR
      VDEN(I)=VV*LQ*ZV
      SL(I)=ALS(I)*ZL
   10 CONTINUE
      IF(IMARK.EQ.2)GO TO 4
C
C     REBOILER BALANCE
C
      IF(VB.GT.0.0)GO TO 4
      IF(IC.GE.1)VVB=VB
      PL=DENS(2)*62.4
      VB=TL*TW*0.5*PL/(SL(2)/ALS(2))
    4 CONTINUE
      RETURN
      END
```

```
      SUBROUTINE FLOW2(TEXIT,VS,VR,YS,WD,WT,DENS,VDEN,YV,T,HT,FS,IMARK)
      DIMENSION Y(8,40),YMAX(40),SAVE(40,12),PSAVE(1600),ERROR(40),A(8),
     1PERTST(7,2,3),TK1(10),TK2(10)
      DIMENSION WW(10)
      DIMENSION ALS(10),XS(40),WD(10),WT(10),YS(40),SL(10),DENS(10),VDEN
     1(10),WTM(2),DSD(2),EPP(2),VIS(2),THI(2),CDD(2),YV(10),TWO(3)
      DIMENSION TEXIT(10)
      COMMON/DATE/AMA,BMA,DMA,DMB,W,TAA,TL,TA,TR,AD,VZ1,VZ2,QFS,N,NF,TW,
     1LQ,XS,ALS,KR,LP,LZ,HDIA,TTH,HOA,ANT,AUD,COV
      COMMON/PFS1/Y,SAVE,PSAVE,YMAX,ERROR,A,PERTST
      COMMON/SSL/ALLS(10)
      COMMON/LIQS/FK1(10),FK2(10)
      COMMON/REFL/IREF
      COMMON/NS/NIMP
C
C     THIS SUBROUTINE IS ONLY CALLED IF FLOW RATE CHANGES ARE MADE.
C     IT FINDS THE CHANGE IN HOLDUP WITH RESPECT TO FLOW AND COMPOSITION
C     CHANGES AND HENCE CALCULATES THE NEW FLOW RATES.
C
      NST=NIMP
      IF(NIMP.EQ.NF)NST=NST-1
      TK2(NST+1)=0.0
      DO 1 I=2,NST
      J=NST+2-I
      WRITE(2,202)WD(J),WT(J)
  202 FORMAT(//3X,2(F8.5,3X))
      TK1(J)=WT(J)+WD(J)
      IF(ABS(ALS(J)-ALLS(J)).LE.0.001)ALS(J)=ALLS(J)
      IF(ALS(J).EQ.ALLS(J).AND.IREF.NE.2)GO TO 1
      J3=(J-1)*(LQ+1)-(LQ+1)+1
      DO 2 ISET=1,3
      ZT=0.0
      ZD=0.0
      ZV=0.0
      IF(ISET.EQ.2)ALS(J)=ALS(J)*21.0/20.0
      DO 3 K=1,LQ+1
```

```
      J4=J3+K
      IF(ISET.EQ.3)XS(J4)=XS(J4)*21.0/20.0
      ZD=ZD+XS(J4)*AMA/(XS(J4)*AMA+(1-XS(J4))*BMA)
      ZL=ZL+XS(J4)*AMA+(1-XS(J4))*BMA
      XZ=XS(J4)
      IF(K.EQ.(LQ+1))GO TO 4
      CALL EQUIDA(XZ,YZ,TEP)
      YS(J3+K)=YZ
      ZT=ZT+TEP
      ZV=ZV+YS(J3+K)*AMA+(1-YS(J3+K))*BMA
    4 IF(ISET.EQ.3)XS(J4)=XS(J4)*20.0/21.0
    3 CONTINUE
      ZD=ZD/(LQ+1)
      ZL=ZL/(LQ+1)
      ZV=ZV/LQ
      ZT=ZT/LQ
      DENS(J)=ZD*DMA+(1-ZD)*DMB
      TEXIT(J)=ZT
      VV=VS
      IF(J.GE.NF)VV=VR
      VDEN(J)=VV*LQ*ZV
      SL(J)=ALS(J)*ZL
      ALC=SL(J)/(3600.0*DENS(J)*62.4)
      UG=VV*359.05*LQ*(TEXIT(J)+273)/(3600.0*TA*273)
      PV=ABS(VDEN(J)/(3600.0*UG*TA))
      PL=DENS(J)*62.4
      WI=W*12
      F=UG*PV**0.5
      FL=60*.2642*ALC*6.24/(.22*TW)
      HC=(103.0+11.8*WI-40.5*F+1.25*FL)/(PL*12)
      HF=(2.53*F*F+1.89*WI-1.6)/12.0
      UL=FL*.22/(6.24*.2642*60*HC)
      WT(J)=TA*HC*PL/(SL(J)/ALS(J))
      HVB=21*(TEXIT(J)+273.2)
      SIG=HVB*DENS(J)/364.0
      H2=(0.558*(ALC/AUD)**2)/12.0
```

```
      HSIG=0.06*SIG/(PL*HDIA*12)
      VH=UG*TA/HOA
      G4=SQRT(TA*4/3.1415926)
      WTM(1)=AMA
      WTM(2)=BMA
      DSD(1)=DMA
      DSD(2)=DMB
      TEXIT(J)=TEXIT(J)+273.2
      DO 5 L=1,2
      XZ=1.0
      IF(L.EQ.2)XZ=0.0
      CALL EQUIDA(XZ,YZ,TEP)
      TEP=TEP+273.2
      EPP(L)=1.15*TEP
      XXZ=TEXIT(J)/EPP(L)
      CALL IPNEU(XXZ,ANS)
      CDD(L)=1.18*((WTM(L)/DSD(L))**(1.0/3.0))
      VIS(L)=0.00646*ANS*SQRT(WTM(L)*TEXIT(J))/(CDD(L)**2)
    5 CONTINUE
      YY1=YV(J)
      YY2=1.0-YY1
      YY3=YY1/YY2
      DO 6 L=1,2
      JL=2
      IF(L.EQ.2)JL=1
      THI(L)=SQRT(2.0)*((1+(VIS(L)/VIS(JL))**0.5*(WTM(JL)/WTM(L))**0.25)
     1**2)/(4*SQRT(1+WTM(L)/WTM(JL)))
    6 CONTINUE
      VG=VIS(1)/(1+THI(1)/YY3)+VIS(2)/(1+THI(2)*YY3)
      G3=PV*UG*G4/(VG/3600.0)
      FF=0.014*G3**(-0.0055/9600.0)
      HD=12*COV*PV*(VH**2)*(0.4*(1.25-HOA/ANT)+4*TTA*FF/HDIA+(1-HOA/ANT)
     1**2)/(64*PL*12)
      TEXIT(J)=TEXIT(J)-273.2
      H3=HD+HC+H2+HSIG
      WD(J)=AD*H3*PL/(SL(J)/ALS(J))
```

```
      TWO(ISET)=WT(J)+WD(J)
      IF(ISET.EQ.1)WRITE(2,203) HC,H3,HF,WD(J),WT(J),F,UG,J
  203 FORMAT(/3X,7(E13.5,2X),I2)
      IF(ISET.EQ.1)WW(J)=WD(J)
      IF(ISET.EQ.2)ALS(J)=ALS(J)*20.0/21.0
    2 CONTINUE
      AX1=XS(J3+1)
      AX2=XS(J3+1)*21.0/20.0
      FK1(J)=(TWO(2)-TWO(1))/(ALS(J)*0.05)
      FK1(J)=3600.0*FK1(J)
      FK2(J)=(TWO(3)-TWO(1))/(AX2-AX1)
      IF(IREF.EQ.2.AND.IZZ.GE.(NST-1))GO TO 1
      IF(IREF.EQ.2)GO TO 9
      GO TO 10
    9 IZZ=IZZ+1
      IF(TWO(1).GT.TK1(J))GO TO 10
      TK2(J)=TK1(J)-TWO(1)+TK2(J+1)
      IF(J.EQ.NST)GO TO 8
      JJ4=1+(LQ+1)*(J-1)
      Y(1,JJ4)=(WW(J)*Y(1,JJ4)+TK2(J+1)*Y(1,JJ4+1))/(TK2(J+1)+WW(J))
      IF(J.NE.2)GO TO 8
      VB=TL*TW*0.5*PL/(SL(2)/ALS(2))
      Y(1,1)=(VB*Y(1,1)+TK2(2)*Y(1,2))/(TK2(2)+VB)
      GO TO 8
   10 CONTINUE
      IF(J.NE.NF)GO TO 7
      ALS(J)=(FK1(J)*ALS(J)+(T-HT)*(ALS(J+1)+OFS*FS))/(FK1(J)+(T-HT))
      GO TO 8
    7 CONTINUE
      ALS(J)=(FK1(J)*ALS(J)+(T-HT)*ALS(J+1))/(FK1(J)+(T-HT))
    8 CONTINUE
    1 CONTINUE
      IF(IREF.EQ.2)IMARK=2
      RETURN
      END
```

```
      SUBROUTINE EQUIDA(XZ,YZ,TEP)
      COMMON/ALPH/AREL,IREL
      COMMON/EQU/SLOPE,CONST,TB1,TB2
C
C     THIS SUBROUTINE CALCULATES THE VAPOUR COMPOSITION IN EQUILIBRIUM
C     WITH THE LIQUID AND THE TEMPERATURE.
C
      IF(IREL.EQ.2)GO TO 3
      YZ=AREL*XZ/(1+XZ*(AREL-1))
      GO TO 2
    3 YZ=SLOPE*XZ+CONST
    2 CONTINUE
      XXZ=1-XZ
      IF(ABS(XXZ).GT.20.0)XXZ=20.0
      TEP=TB1+EXP(XXZ)*(TB2-TB1)
      RETURN
      END
```

```
      SUBROUTINE AICHE(IMARK,SL,YV,Y8,EPP,WTM,DSD,CDD,VIS,THI,VLG,SN,SLM
     1,TEXIT)
      DIMENSION Y8(40),AB(10),ES(10),EPP(2),WTM(2),DSD(2),CDD(2),VIS(2)
      DIMENSION THI(2),DG(10),VLG(10),DL(10),ANG(10),ANL(10),EP(10),ANO(
     110),SN(10),SLM(2),YV(10),TEXIT(10),XS(40),XR(40),ALS(10),YS(40),YC
     2(50),DENS(10),VDEN(10),PEN(10),B(10),SL(10),WD(10),WT(10),ALE(10),
     3ERA(10)
      COMMON/HOLD/HOL(10)
      COMMON/SSL/ALLS(10)
      COMMON/DATE/AMA,BMA,DMA,DMB,W,TAA,TL,TA,TR,AD,VZ1,VZ2,QFS,N,NF,TW,
     1LQ,XS,ALS,KR,LP,LZ,HDIA,TTH,HOA,ANT,AUD,COV
      COMMON/EPRS/ALE,ERA,EP,DL,DG,ANO,ANG,ANL,AB,ES,XR,WD,WT,VR,VS,DENS
     1,VDEN,B,PEN,CTP,ETP,TCOND,VD,VT,VC,VB,YS,YC,TBOIL,IC,IZ,HT
      COMMON/REFL/IREF
C
C     THIS SUBROUTINE CALCULATES ALL THE PARAMETERS INVOLVED WITH PLATE
C     AND POINT EFFICIENCIES.
C
C
C     CALCULATES THE LIQUID AND FROTH HEIGHTS, AND FLOW RATE PARAMETERS.
C     CALCULATE PECLET NUMBER AND BACKFLOW RATIO
C
      DO 1 J1=2,N
      J=(J1-1)*(LQ+1)-(LQ+1)
      VV=VS
      IF(J1.GE.NF)VV=VR
      ALC=SL(J1)/(3600*DENS(J1)*62.4)
      UG=VV*359.05*LQ*(TEXIT(J1)+273)/(3600*TA*273)
      PV=ABS(VDEN(J1)/(3600*UG*TA))
      PL=DENS(J1)*62.4
      WI=W*12
      F=UG*PV**0.5
      FL=60*.2642*ALC*6.24/(.22*TW)
      HC=(103.0+11.8*WI-40.5*F+1.25*FL)/(PL*12)
      HF=(2.53*F*F+1.89*WI-1.6)/12.0
      UL=FL*.22/(6.24*.2642*60*HC)
```

```
      DE=(0.0609395*HC*UL**3)/((HC*UL/HF)**2.5664)
      PE=UL*TR/DE
      AN=LQ+1
      BETA=AN/PE-0.5
      IF(IREF.NE.2.AND.IMARK.EQ.2)GO TO 6
      IF(IC.EQ.2.AND.IMARK.EQ.2)GO TO 6
      IF(IREF.EQ.2.AND.IC.EQ.1)GO TO 2
      IF(IC.GE.1.AND.ALLS(J1).EQ.ALS(J1))GO TO 6
    2 CONTINUE
      PEN(J1)=PE
      B(J1)=BETA
      IF(IMARK.NE.2)GO TO 5
    6 BETA=B(J1)
      PL=DENS(J1)*62.4
    5 CONTINUE
C
C     THE CALCULATION OF THE POINT EFFICIENCY ACCORDING TO THE A.I.CH.E.
C     MANUAL. THESE CALCULATIONS INCLUDE NUMBER OF GAS STAGES ; THE
C     SLOPE OF THE EQUILIBRIUM LINE ; ABSORPTION COEFFICIENT.
C
      TEXIT(J1)=TEXIT(J1)+273.2
      ISJ=J+2
      ISI=J+2+LQ
      DO 8 ISK=ISJ,ISI,LQ
      XZ=XS(ISK)
      CALL EQUIDA(XZ,YZ,TEP)
    8 Y8(ISK)=YZ
      SUMA=XS(ISI)-XS(ISJ)
      ES(J1)=(Y8(ISI)-Y8(ISJ))/SUMA
      AB(J1)=ES(J1)*LQ*VV/ALS(J1)
      DO 24 I=1,2
      XZ=1.0
      IF(I.EQ.2)XZ=0.0
      CALL EQUIDA(XZ,YZ,TEP)
      TEP=TEP+273.2
      EPP(I)=1.15*TEP
```

```
      XXZ=TEXIT(J1)/EPP(I)
      CALL IPNEU(XXZ,ANS)
      CDD(I)=1.18*((WTM(I)/DSD(I))**(1.0/3.0))
      VIS(I)=0.00646*ANS*SQRT(WTM(I)*TEXIT(J1))/(CDD(I)**2)
   24 CONTINUE
      YY1=YV(J1)
      IF(YV(J1).LE.0.0)YY1=Y8(ISJ)
      YY2=1.0-YY1
      YY3=YY1/YY2
      DO 11 I=1,2
      JL=2
      IF(I.EQ.2)JL=1
      THI(I)=SQRT(2.0)*((1+(VIS(I)/VIS(JL))**0.5*(WTM(JL)/WTM(I))**0.25)
     1**2)/(4*SQRT(1+WTM(I)/WTM(JL)))
   11 CONTINUE
      VG=VIS(1)/(1+THI(1)/YY3)+VIS(2)/(1+THI(2)*YY3)
      BC1=(CDD(1)+CDD(2))/2
      BC2=SQRT(EPP(1)*EPP(2))
      BC3=TEXIT(J1)/BC2
      CALL IDNEU(BC3,AMS)
      BC4=SQRT((WTM(1)+WTM(2))/(WTM(1)*WTM(2)))
      BC5=(10.7-2.46*BC4)*1.0E-4
      DG(J1)=(BC5*BC4*TEXIT(J1)**(3.0/2))/(AMS*BC1**2)
      IF(DG(J1).LE.1.0E-10)DG(J1)=1.0E-10
      SN(J1)=VG/(PV*DG(J1))
      ANG(J1)=(0.776+0.11*WI-0.29*F+0.0217*FL)/SQRT(SN(J1))
      DO 14 I=1,2
      JL=2
      IF(I.EQ.2)JL=1
      SLM(I)=2.87E-7*WTM(JL)**0.5/((WTM(I)/DSD(I))**0.6)
   14 CONTINUE
      BC6=0.0
      DO 15 I=2,LQ+1
   15 BC6=BC6+XS(J+I)
      BC6=BC6/LQ
      IF(BC6.GE.1.0)BC6=1.0
```

```
      IF(BC6.LE.0.0)BC6=0.0
      BC7=(SLM(2)-SLM(1))+BC6+SLM(1)
      BC8=ABS(PL/62.4)
      VLG(J1)=BC8*EXP(BC6*ALOG(VZ1/DMA)+(1-BC6)*ALOG(VZ2/DMB))
      DL(J1)=BC7*TEXIT(J1)/VLG(J1)
      CTL=TAA/UL
      ANL(J1)=SQRT(1.065E4*DL(J1))*(0.15+0.26*F)*CTL
      BC9=1/ANG(J1)+AB(J1)/ANL(J1)
      ANO(J1)=ABS(1.0/BC9)
      IF(ANO(J1).LE.1.0E-3)ANO(J1)=1.0E-3
      IF(ANO(J1).GE.90.0)ANO(J1)=90.0
      EP(J1)=1.0-EXP(-ANO(J1))
      TEXIT(J1)=TEXIT(J1)-273.2
      IF(IREF.EQ.2.AND.IC.EQ.1)GO TO 3
      IF(IMARK.EQ.2)GO TO 1
C
C     CALCULATES MOLAR HOLDUPS ON THE TRAY AND THE DOWNCOMERS.
C
      IF(IC.GE.1.AND.ALLS(J1).EQ.ALS(J1))GO TO 1
    3 CONTINUE
      HOL(J1)=HC*PL
      WT(J1)=TA*HOL(J1)/(SL(J1)/ALS(J1))
      HVB=21*(TEXIT(J1)+273.2)
      SIG=HVB*DENS(J1)/364.0
      H2=(0.558*(ALC/AUD)**2)/12.0
      HSIG=0.06*SIG/(PL*HDIA*12)
      VH=UG*TA/HOA
      G4=SQRT(TA*4/3.1415926)
      G3=PV*UG*G4/(VG/3600.0)
      FF=0.014*G3**(-0.0055/9600.0)
      HD=12*COV*PV*(VH**2)*(0.4*(1.25-HOA/ANT)+4*TTA*FF/HDIA+(1-HOA/ANT)
     1**2)/(64*PL*12)
      H3=HD+HC+H2+HSIG
      WD(J1)=AD+H3*PL/(SL(J1)/ALS(J1))
      IF(IMARK.EQ.1)WRITE(2,203)HC,H3,HF,WD(J1),WT(J1),F,UG,J1
  203 FORMAT(/3X,7(E13.5,2X),I2)
```

```
      IF(VC.GT.0.0)GO TO 1
      IF(J1.NE.N)GO TO 1
      VC=2*1*2*PL/(SI(N)/ALS(N))
    1 CONTINUE
      RETURN
      END
```

```
      SUBROUTINE IPNEU(XXZ,ANS)
C
C     THIS SUBROUTINE CONTAINS A POLYNOMIAL EQUATION FOR CALCULATING
C     THE CORRECTED COLLISION INTEGRAL FOR VISCOSITY
C
      X=XXZ
      ANS=-.401220751E-3*X**7+.72353998E-2*X**6-.052798314*X**5+.1985071
     1*X**4-.39222949*X**3+.29797*X**2+.32825044*X+.243213819
      RETURN
      END
```

```
      SUBROUTINE IDNEU(BC3,AMS)
C
C     THIS SUBROUTINE CONTAINS A POLYNOMIAL EQUATION FOR CALCULATING
C     THE COLLISION INTEGRAL FOR DIFFUSION
C
      X=BC3
      AMS=-.00224977819*X**7+.0420454127*X**6-.32383476*X**5+1.33657855*
     1X**4-3.22085296*X**3+4.633577*X**2-3.90948762*X+2.162113535
      RETURN
      END
```

```
      SUBROUTINE WRITER(I,TIM,LQ,N)
      DIMENSION CX(100),RX(100)
      DIMENSION TIM(I)
      COMMON/CR/CX,RX
      COMMON/GRAF/EOMAX,EOMIN,ELMAX,ELMIN,AMAXE,AMINE
      COMMON/EPRS/BX(281),IC,IZ,HT
      COMMON/PFS1/AX(2530)
C
C     THIS ROUTINE READS THE COMPUTED DATA ON DISC AND WRITES
C
      REWIND 3
      DO 32 II=1,I
      WRITE(2,18)
      READ(3)BX(1)
      READ(3)(BX(J),J=1,6)
      WRITE(2,19)(BX(J),J=1,6)
      WRITE(2,20)
      READ(3)BX(7),BX(8)
      WRITE(2,21)BX(7),BX(8)
      READ(3)BX(9),BX(10)
      WRITE(2,22)BX(9),BX(10)
      RX(II)=BX(9)
      CX(II)=BX(10)
      READ(3)BX(11),BX(12)
      WRITE(2,23)BX(11),BX(12)
      READ(3)(BX(J),J=13,15)
      WRITE(2,24)(BX(J),J=13,15)
      WRITE(2,25)(I21,I21=1,N-1)
      WRITE(2,26)
      READ(3)(BX(J),J=16,(15+N-1))
      WRITE(2,17)(BX(J),J=16,(15+N-1))
      DO 33 IJ=1,N-1
      IK=(IJ-1)*I
   33 AX(II+IK)=BX(15+IJ)
      J1=15+N-1
      N2=N*2-2
```

```
      DO 31 JJ=1,LQ
      READ(3)(BX((JJ-1)*N2+J1+J),J=1,N2)
      WRITE(2,27)(BX((JJ-1)*N2+J1+J),J=1,N2)
   31 CONTINUE
      N3=15+N-1+LQ*2*(N-1)
      READ(3)(BX(N3+J),J=1,N-1)
      N4=N3+N-1
      WRITE(2,28)(BX(N3+J),J=1,N-1)
      READ(3)BX(N4+1),BX(N4+2)
      WRITE(2,29)BX(N4+1),BX(N4+2)
      N5=N4+2
      IZZ=I*(N-1)
      IJ2=(N-1)*2
      II1=N3-IJ2
      DO 34 IJ=1,N-1
      IK=(IJ-1)*I+IZZ
   34 AX(II+IK)=BX(II1+IJ+IJ-1)
      IST=2
      DO 56 IK=1,13
      JK=(IK-1)*(N-1)+N5
      READ(3)(BX(JK+JF),JF=1,N-1)
      GO TO (40,41,42,43,44,45,46,47,51,52,54,55,57),IK
   40 WRITE(2,1)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   41 WRITE(2,2)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   42 WRITE(2,3)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   43 WRITE(2,4)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   44 WRITE(2,5)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   45 WRITE(2,6)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   46 WRITE(2,7)(BX(JK+JF),JF=1,N-1)
      GO TO 30
```

```
   47 WRITE(2,8)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   51 WRITE(2,12)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   52 WRITE(2,13)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   54 WRITE(2,15)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   55 WRITE(2,16)(BX(JK+JF),JF=1,N-1)
      GO TO 30
   57 WRITE(2,36)(BX(JK+JF),JF=1,N-1)
   30 CONTINUE
      IF(IK.LE.8)GO TO 56
      IF(IK.GE.12)GO TO 56
      IZZ=I*(N-1)*IST
      IST=IST+1
      DO 35 IN1=1,N-1
      IN2=(IN1-1)*I
   35 AX(IZZ+II+IN2)=BX(JK+IN1)
   56 CONTINUE
   32 CONTINUE
C
C     REWRITE ONTO DISC THE INFORMATION NEEDED FOR PLOTTING.
C
      REWIND 3
      N9=(N-1)*IST*I
      WRITE(3)I,N,N9
      WRITE(3)(AX(J),J=1,N9)
      WRITE(3)(TIM(J),J=1,I)
      WRITE(3)(CX(J),J=1,I)
      WRITE(3)(RX(J),J=1,I)
      WRITE(3)EOMAX,EOMIN,ELMAX,ELMIN,AMAXE,AMINE
    1 FORMAT(15X,10HV ENTERING,5X,5(F8.5,8X))
    2 FORMAT(15X,10HV LEAVING.,5X,5(F8.5,8X))
    3 FORMAT(15X,13HV EQUILIBRIUM,2X,5(F8.5,8X))
    4 FORMAT(15X,13HBOILING TEMP.,2X,5(F8.3,8X))
```

```
 5 FORMAT(15X,9HPECLET NO,6X,5(F8.3,8X))
 6 FORMAT(15X,8HBACKFLOW,7X,5(F8.3,8X))
 7 FORMAT(15X,9HEQ. SLOPE,6X,5(F8.4,8X))
 8 FORMAT(15X,6HLAMBDA,9X,5(F8.4,8X))
12 FORMAT(15X,4HEOG.,11X,5(F8.4,8X))
13 FORMAT(15X,4HEMV.,11X,5(F8.4,8X))
15 FORMAT(15X,8HEMV/EOG.,7X,5(F8.4,8X))
16 FORMAT(15X,13HR.VOLATILITY.,2X,5(F8.4,8X))
17 FORMAT(15X,9HDOWNCOMER,1X,5(F8.5,8X))
18 FORMAT(1H1,/////////28X,5HTIME.,5X,4HFEED,6X,2HXF,6X,2HWS,7X,2HDS,
  15X,6HREFLUX)
19 FORMAT(27X,F7.3,3X,F7.2,3X,F4.2,2(3X,F6.2),3X,F6.3)
20 FORMAT(//32X,8HREBOILER,11X,9HCONDENSER,9X,12HTOP PRODUCTS)
21 FORMAT(15X,12HMOLAR HOLDUP.5X,F8.5,12X,F8.5)
22 FORMAT(15X,12HLIQUID COMP.,5X,F8.5,12X,F8.5)
23 FORMAT(15X,12HVAPOUR COMP.,5X,F8.5,26X,5HCAL.=,F9.4)
24 FORMAT(15X,12HBOILING TEMP,5X,F7.3,13X,F7.3,7X,5HEST.=,F9.4)
25 FORMAT(//15X,5HPLATE,5X,5(7X,I2,7X))
26 FORMAT(25X,5(4X,1HX,7X,1HY,3X))
27 FORMAT(25X,10F8.4)
28 FORMAT(15X,11HLIQUID FLOW,4X,5(F8.3,8X))
29 FORMAT(15X,11HVAPOUR FLOW,6X,9HSTRIPPER=,F8.3,10X,10HRECTIFIER=,F8
  1.3)
36 FORMAT(15X,13HMOLAR HOLDUPS,2X,5(F8.4,8X))
   RETURN
   END
```

```
      MASTER DRAWER
      DIMENSION RX(100),CX(100),TIM(100)
      COMMON/GRAF/EOMAX,EOMIN,ELMAX,ELMIN,AMAXE,AMINE
      COMMON/PFS1/AX(2530)
      COMMON/CR/CX,RX
C
C     THIS SUBROUTINE READS FROM THE DISC ALL THE DATA TO BE PLOTTED
C
      CALL UTPOP
      REWIND 3
      READ(3)I,N,N9
      READ(3)(AX(J),J=1,N9)
      READ(3)(TIM(J),J=1,I)
      READ(3)(CX(J),J=1,I)
      READ(3)(RX(J),J=1,I)
      READ(3)EOMAX,EOMIN,ELMAX,ELMIN,AMAXE,AMINE
      CALL PLOT1(I,TIM,N)
      CALL UTPCL
      STOP
      END
```

```
      SUBROUTINE PLOT1(I,TIM,N)
      DIMENSION RX(100),CX(100),TIM(100)
      COMMON/PFS1/AX(2530)
      COMMON/GRAF/EOMAX,EOMIN,ELMAX,ELMIN,AMAXE,AMINE
      COMMON/CR/CX,RX
C
C     THIS SUBROUTINE PLOTS THE DATA REQUIRED.
C
C
      CALL UTPOP
      TLIM=TIM(I)-TIM(1)
      XINS=5.0
      YINS=8.0
      YMIN=0.0
      YMAX=1.0
      XMIN=TIM(1)
      XMAX=TIM(I)
      X2=XINS*(-TIM(1)/TLIM)
      NX=2
      NY=3
      X4=0.1
      Y4=7.8
      ANG=0.0
      DO 2 J=1,5
      X1=8.0
      Y1=6.0
      CALL UTP2(X1,Y1,-1)
      GO TO (29,30,31,32,33),J
   29 CALL UTP4A(XMIN,XMAX,YMIN,YMAX,XINS,YINS,14HTIME. MINUTES.,NX,20HC
     1ONC. AT EXIT WEIRS.,NY)
      CALL UTP4B(TIM,CX,I,3)
      CALL UTP4B(TIM,RX,I,3)
      I1=1
      GO TO 8
   30 I1=0
      CALL UTP4A(XMIN,XMAX,YMIN,YMAX,XINS,YINS,14HTIME. MINUTES.,NX,20HC
```

```
     1ONC. AT DOWNCOMERS.,NY)
      GO TO 8
   31 I1=2
      YMAX=EOMAX
      YMIN=EOMIN
      NY=4
      CALL UTP4A(XMIN,XMAX,YMIN,YMAX,XINS,YINS,14HTIME. MINUTES.,NX,28HC
     1HANGES IN POINT EFFICIENCY.,NY)
      GO TO 8
   32 I1=3
      YMIN=AMINE
      YMAX=AMAXE
      NY=4
      CALL UTP4A(XMIN,XMAX,YMIN,YMAX,XINS,YINS,14HTIME. MINUTES.,NX,28HC
     1HANGES IN PLATE EFFICIENCY.,NY)
      GO TO 8
   33 I1=4
      YMIN=ELMIN
      YMAX=ELMAX
      NY=4
      CALL UTP4A(XMIN,XMAX,YMIN,YMAX,XINS,YINS,14HTIME. MINUTES.,NX,28HC
     1HANGES IN EFFICIENCY RATIO.,NY)
    8 DO 5 JI=1,N-1
      IK=(JI-1)*I+(N-1)*I*I1
      DO 6 JJ=1,I
    6 CX(JJ)=AX(IK+JJ)
    5 CALL UTP4B(TIM,CX,I,3)
      Y1=0.0
      Y2=8.0
      CALL UTP2(X2,Y1,1)
      CALL UTP2(X2,Y2,2)
      CALL UTP6(40HSTEADY-STATE.            UNSTEADY-STATE.,5,40,ANG,X4,
     1Y4)
      X1=8.0
      Y1=6.0
      CALL UTP2(X1,Y1,-1)
```

```
2 CONTINUE
  CALL UTPCL
  RETURN
  END
  FINISH
```

D.2 The Integration Routine

The subroutine integrates a set of N ordinary equations one step of length H, where H may be specified by the user, but is controlled by the subroutine to control the estimated error within a specified tolerance if possible.

A multistep predictor corrector method is used whose order is automatically chosen by the subroutine as the integration proceeds. Either an Adams method or methods suitable for stiff equations can be selected. The starting procedure is automatic and the information retained by the program about previous steps is stored in such a way as to make the interpolation to a non mesh point straightforward. The methods used are described from a mathematical point of view in the following papers (32,33,34).

The integration routine may call up to three subroutines:-

1) DIFFUN
2) PEDERV
3) MATINV

DIFFUN must always be provided and must evaluate the derivatives of the dependent variables Y with respect to the independent variable T.

MATINV is only called if stiff methods are requested and is a matrix inversion routine. The parameter J1 should be set by MATINV to +1 if the inversion is successful and -1 if the matrix is nearly singular.

PEDERV need only be provided if the method flag MF is set to 1, otherwise a dummy subroutine should be provided. If PEDERV is required, then the equations for evaluating the Jacobian matrix should be provided.

All floating point variables are taken to be double precision except for those whose name begin with P. These may be calculated in single precision if an increase in speed and reduced storage is required.

The parameters for the subroutine DIFSUB have the following meanings:-

N	the number of first order differential equations. N may be decreased on later calls if the number of active equations reduces, but it must not increase without setting JSTART = 1.
T	the independent variable.
Y	an 8 by N array containing the dependent variables and their scaled derivatives. Y(j+1,1) contains the jth derivative of Y(1) scaled by H**j/j! where H is the current step size.
SAVE	a block of atleast 12N floating point locations used by the subroutines.
H	the step size to be attempted on the next step. H may be increased or decreased by the program to achieve an economical integration.
HMIN	the minimum step size that will be used for the integration and must be set smaller than H.

HMAX	the maximum step length that is required.
EPS	the error test constant. Single step error estimates divided by YMAX(I) must be less than this in the euclidean norm. The step and/or order is adjusted to achieve this.
MF	this is the method flag:- 0 an Adams predictor corrector is used. 1 a multistep mehtod suitable for stiff equations is used. The subroutine PEDERV must be provided. 2 the same as for 1 except the partial derivatives are estimated by the numerical differencing of the derivatives and the subroutine PEDERV is not called.
YMAX	an array of N locations containing the maximum of each Y seen so far and should normally be set to 1 on starting.
ERROR	an array of N elements which contains the estimated one step error in each component.
KFLAG	a completion code with the following meanings:- +1 the step was successful. -1 the step was taken with H=HMIN, but the requested error was not achieved. -2 the maximum order specified was found to be too large. -3 corrector convergence could not be achieved for H greater than HMIN.

-4 the requested error is smaller than could be handled.

JSTART an input indicator with the following meanings:-

-1 repeat the last step with a new H.

0 perform the first step. The first step must be done with value of JSTART so that the subroutine can initialise itself.

+1 take a new step continuing from the last.

JSTART is set to NQ, the current order of the method at exit. NQ is also the order of the maximum derivative available.

MAXDER the maximum derivative that should be used in the method. The order is thus restricted as it is equal to the highest derivative used and must be less than 8 for the Adams method and 7 for stiff methods.

PSAVE a block of N*N floating point locations.

```
      SUBROUTINE DIFSUB(N,T,Y,SAVE,H,HMIN,HMAX,EPS,MF,YMAX,ERROR,KFLAG,
     1JSTART,MAXDER,PSAVE,N4)
      DIMENSION Y(8,N),YMAX(N),SAVE(N,12),ERROR(N),PSAVE(N4),
     1A(8),PERTST(7,2,3)
      DATA PERTST /2.0,4.5,7.333,10.42,13.7,17.15,1.0,
     12.0,12.0,24.0,37.89,53.33,70.08,87.97,3.0,6.0,9.167,12.5,15.98,
     11.0,1.0,12.0,24.0,37.89,53.33,70.08,87.97,1.0,1.,1.,0.5,0.1667,
     10.04133,0.008267,1.0,1.0,1.0,2.0,1.0,.3157,.07407,.0139/
      DATA A(2) /-1.0/
      IRET=1
      KFLAG=1
      IF(JSTART.LE.0)GO TO 140
  100 DO 110 I=1,N
      DO 110 J=1,K
  110 SAVE(I,J)=Y(J,I)
      HOLD=HNEW
  120 IF(H.EQ.HOLD)GO TO 130
      RACUM=H/HOLD
      IRET1=1
      GO TO 750
  130 CONTINUE
      NQOLD=NQ
      TOLD=T
      RACUM=1.0
      IF(JSTART.GT.0)GO TO 250
      GO TO 170
  140 IF(JSTART.EQ.-1)GO TO 160
      NQ=1
      N3=N
      N1=N*10
      N2=N1+1
      N4=N**2
      N5=N1+N
      N6=N5+1
      CALL DIFFUN(T,Y,SAVE,N2,N)
      DO 150 I=1,N
```

```
      ND=N1+I
  150 Y(2,I)=SAVE(ND,1)*H
      HNEW=H
      K=2
      GO TO 100
  160 IF(NQ.EQ.NQOLD)JSTART=1
      T=TOLD
      NQ=NQOLD
      K=NQ+1
      GO TO 120
  170 IF(MF.EQ.0)GO TO 180
      IF(NQ.GT.6)GO TO 190
      GO TO (221,222,223,224,225,226),NQ
  180 IF(NQ.GT.7)GO TO 190
      GO TO (211,212,213,214,215,216,217),NQ
  190 KFLAG=-2
      RETURN
  211 A(1)=-1.0
      GO TO 230
  212 A(1)=-0.500000000
      A(3)=-0.500000000
      GO TO 230
  213 A(1)=-0.4166666666666667
      A(3)=-0.750000000
      A(4)=-0 1666666666666667
      GO TO 230
  214 A(1)=-0.375000000
      A(3)=-0.9166666666666667
      A(4)=-0.3333333333333333
      A(5)=-0.04166666666666667
      GO TO 230
  215 A(1)=-0.3486111111111111
      A(3)=-1.0416666666666667
      A(4)=-0.4861111111111111
      A(5)=-0.1041666666666667
      A(6)=-0.008333333333333333
```

```
      GO TO 230
  216 A(1)=-0.3298611111111111
      A(3)=-1.1416666666666667
      A(4)=-0.625000000
      A(5)=-0.1770833333333333
      A(6)=-0.0250000000
      A(7)=-0.001388888888888889
      GO TO 230
  217 A(1)=-0.3155919312169312
      A(3)=-1.235000000
      A(4)=-0.7518518518518519
      A(5)=-0.2552083333333333
      A(6)=-0.04861111111111111
      A(7)=-0.004861111111111111
      A(8)=-0.0001984126984126984
      GO TO 230
  221 A(1)=-1.000000000
      GO TO 230
  222 A(1)=-0.6666666666666667
      A(3)=-0.3333333333333333
      GO TO 230
  223 A(1)=-0.5454545454545455
      A(3)=A(1)
      A(4)=-0.09090909090909091
      GO TO 230
  224 A(1)=-0.480000000
      A(3)=-0.700000000
      A(4)=-0.200000000
      A(5)=-0.020000000
      GO TO 230
  225 A(1)=-0.437956204379562
      A(3)=-0.8211678832116788
      A(4)=-0.3102189781021898
      A(5)=-0.05474452554744526
      A(6)=-0.0036496350364963504
      GO TO 230
```

```
226 A(1)=-0.4081632653061225
    A(3)=-0.9206349206349206
    A(4)=-0.4166666666666667
    A(5)=-0.0992063492063492
    A(6)=-0.0119047619047619
    A(7)=-0.000566893424036282
230 K=NQ+1
    IDOUB=K
    MTYP=(4-MF)/2
    ENQ1=0.5/FLOAT(NQ)
    ENQ2=.5/FLOAT(NQ+1)
    ENQ3=.5/FLOAT(NQ+2)
    PEPSH=EPS
    EUP=(PERTST(NQ,MTYP,2)*PEPSH)**2
    E=(PERTST(NQ,MTYP,1)*PEPSH)**2
    EDWN=(PERTST(NQ,MTYP,3)*PEPSH)**2
    IF(EDWN.EQ.0)GO TO 780
    BND=EPS*ENQ3/FLOAT(N)
240 IWEVAL=MF
    GO TO (250,680),IRET
250 T=T+H
    DO 260 J=2,K
    DO 260 J1=J,K
    J2=K-J1+J-1
    DO 260 I=1,N
260 Y(J2,I)=Y(J2,I)+Y(J2+1,I)
    DO 270 I=1,N
270 ERROR(I)=0.0
    DO 430 L=1,3
    CALL DIFFUN(T,Y,SAVE,N2,N)
    IF(IWEVAL.LT.1)GO TO 350
    IF(MF.EQ.2)GO TO 310
    CALL PEDERV(T,Y,PSAVE,N3)
    R=A(1)*H
    DO 280 I=1,N4
280 PSAVE(I)=PSAVE(I)*R
```

```
  290 DO 300 I=1,N
      ND=I*(N3+1)-N3
  300 PSAVE(ND)=1.+PSAVE(ND)
      IWEVAL=-1
      CALL MATINV(PSAVE,N,J1,N3,N4)
      IF(J1.GT.0)GO TO 350
      GO TO 440
310   CCC=EPS
      DO 320 I=1,N
 320  SAVE(I,9)=Y(1,I)                                          6
      DO 340 J=1,N
      R=EPS*DMAX1(EPS,DABS(SAVE(J,9)))                           601
      Y(1,J)=Y(1,J)+R
      D=A(1)*H/R
      CALL DIFFUN(T,Y,SAVE,N6,N)
      DO 330 I=1,N
      ND=I+(J-1)*N3
      NDD=N5+I
      NDDD=N1+I
  330 PSAVE(ND)=(SAVE(NDD,1)-SAVE(NDDD,1))*D
  340 Y(1,J)=SAVE(J,9)                                          7
      EPS=CCC
      GO TO 290
  350 IF(MF.NE.0)GO TO 370
      DO 360 I=1,N
      ND=N1+I
  360 SAVE(I,9)=Y(2,I)-SAVE(ND,1)*H
      GO TO 410
  370 DO 380 I=1,N
      NDD=N1+I
      ND=N5+I
  380 SAVE(ND,1)=Y(2,I)-SAVE(NDD,1)*H
      DO 400 I=1,N
      D=0.0
      DO 390 J=1,N
      ND=I+(J-1)*N3
```

```
      NDD=N5+J
  390 D=D+PSAVE(ND)*SAVE(NDD,1)
  400 SAVE(I,9)=D                                                    799
  410 NT=N
      DO 420 I=1,N
      Y(1,I)=Y(1,I)+A(1)*SAVE(I,9)                                   801
      Y(2,I)=Y(2,I)-SAVE(I,9)                                        802
      ERROR(I)=ERROR(I)+SAVE(I,9)                                    803
      IF(DABS(SAVE(I,9)).LE.(BND*YMAX(I)))NT=NT-1                    9
  420 CONTINUE
      IF(NT.LE.0)GO TO 490
  430 CONTINUE
  440 T=T-H
      IF((H.LE.HMIN*1.00001).AND.((IWEVAL-MTYP).LT.-1))GO TO 460
      IF((MF.EQ.0).OR.(IWEVAL.NE.0))RACUM=RACUM*.25
      IWEVAL=MF
      IRET1=2
      GO TO 750
  460 KFLAG=-3
  470 DO 480 I=1,N
      DO 480 J=1,K
  480 Y(J,I)=SAVE(I,J)                                               10
      H=HOLD
      NQ=NQOLD
      JSTART=NQ
      RETURN
  490 D=0.0
      DO 500 I=1,N
  500 D=D+(ERROR(I)/YMAX(I))**2
      IWEVAL=0
      IF(D.GT.E)GO TO 540
      IF(K.LT.3)GO TO 520
      DO 510 J=3,K
      DO 510 I=1,N
  510 Y(J,I)=Y(J,I)+A(J)*ERROR(I)
  520 KFLAG=1
```

```
      HNEW=H
      IF(IDOUB.LE.1)GO TO 550
      IDOUB=IDOUB-1
      IF(IDOUB.GT.1)GO TO 700
      DO 530 I=1,N
  530 SAVE(I,10)=ERROR(I)                                                 11
      GO TO 700
  540 KFLAG=KFLAG-2
      IF(H.LE.(HMIN*1.00001))GO TO 740
      T=TOLD
      IF(KFLAG.LE.-5)GO TO 720
  550 PR2=(D/E)**ENQ2*1.2
      PR3=1.E+20
      IF((NQ.GE.MAXDER).OR.(KFLAG.LE.-1))GO TO 570
      D=0.0
      DO 560 I=1,N
  560 D=D+((ERROR(I)-SAVE(I,10))/YMAX(I))**2                              12
      PR3=(D/EUP)**ENQ3*1.4
  570 PR1=1.E+20
      IF(NQ.LE.1)GO TO 590
      D=0.0
      DO 580 I=1,N
580   D=D+(Y(K,I)/YMAX(I))**2
      PR1=(D/EDWN)**ENQ1*1.3
  590 CONTINUE
      IF(PR2.LE.PR3)GO TO 650
      IF(PR3.LT.PR1)GO TO 660
  600 R=1.0/AMAX1(PR1,1.E-4)
      NEWQ=NQ-1
  610 IDOUB=10
      IF((KFLAG.EQ.1).AND.(R.LT.(1.1)))GO TO 700
      IF(NEWQ.LE.NQ)GO TO 630
      DO 620 I=1,N
  620 Y(NEWQ+1,I)=ERROR(I)*A(K)/FLOAT(K)
  630 K=NEWQ+1
      IF(KFLAG.EQ.1)GO TO 670
```

```
      IRET1=3
      RACUM=RACUM*R
      GO TO 750
  640 CONTINUE
      IF(NEWQ.EQ.NQ)GO TO 250
      NQ=NEWQ
      GO TO 170
  650 IF(PR2.GT.PR1)GO TO 600
      NEWQ=NQ
      R=1.0/AMAX1(PR2,1.E-4)
      GO TO 610
  660 R=1.0/AMAX1(PR3,1.E-4)
      NEWQ=NQ+1
      GO TO 610
  670 IRET=2
      R=DMIN1(R,HMAX/DABS(H))
      H=H*R
      HNEW=H
      IF(NQ.EQ.NEWQ)GO TO 680
      NQ=NEWQ
      GO TO 170
  680 R1=1.0
      DO 690 J=2,K
      R1=R1*R
      DO 690 I=1,N
  690 Y(J,I)=Y(J,I)*R1
      IDOUB=K
700    DO 710 I=1,N
  710 YMAX(I)=DMAX1(YMAX(I),DABS(Y(1,I)))
      JSTART=NQ
      RETURN
  720 IF(NQ.EQ.1)GO TO 780
      CALL DIFFUN(T,Y,SAVE,N2,N)
      R=H/HOLD
      DO 730 I=1,N
      Y(1,I)=SAVE(I,1)
```

```
      ND=N1+I
      SAVE(I,2)=HOLD*SAVE(ND,1)
  730 Y(2,I)=SAVE(I,2)*R                                                15
      NQ=1
      KFLAG=1
      GO TO 170
  740 KFLAG=-1
      HNEW=H
      JSTART=NQ
      RETURN
  750 R2=DMAX1(DABS(HMIN/HOLD),RACUM)
      R2=DMIN1(R2,DABS(HMAX/HOLD))
      R1=1.0
      DO 760 J=2,K
      R1=R1*R2
      DO 760 I=1,N
  760 Y(J,I)=SAVE(I,J)*R1                                               16
      RACUM=R2
      H=HOLD*RACUM
      DO 770 I=1,N
  770 Y(1,I)=SAVE(I,1)                                                  17
      IDOUB=K
      GO TO (130,250,640),IRET1
  780 KFLAG=-4
      GO TO 470
      END
```

```
      SUBROUTINE MATINV(A,N,J1,NN,N4)
      DIMENSION A(N4),L(100),M(100)
C
C     THIS SUBROUTINE CARRIES OUT THE MATRIX INVERSION FOR THE
C     INTEGRATION PROGRAM.
C
      D=1.0
      NK=-NN
      DO 80 K=1,N
      NK=NK+NN
      L(K)=K
      M(K)=K
      KK=NK+K
      BIGA=A(KK)
      DO 20 J=K,N
      IZ=NN*(J-1)
      DO 20 I=K,N
      IJ=IZ+I
10     IF(ABS(BIGA)-ABS(A(IJ)))15,20,20
15     BIGA=A(IJ)
      L(K)=I
      M(K)=J
20     CONTINUE
      J=L(K)
      IF(J-K)35,35,25
25     KI=K-NN
      DO 30 I=1,N
      KI=KI+NN
      HOLD=-A(KI)
      JI=KI-K+J
      A(KI)=A(JI)
30     A(JI)=HOLD
35    I=M(K)
      IF(I-K)45,45,38
38     JP=NN*(I-1)
      DO 40 J=1,N
```

```
       JK=NK+J
       JI=JP+J
       HOLD=-A(JK)
       A(JK)=A(JI)
 40     A(JI)=HOLD
 45     IF(BIGA)48,46,48
 46     D=0.0
       J1=-1
       RETURN
 48      DO 55 I=1,N
       IF(I-K)50,55,50
 50     IK=NK+I
       A(IK)=A(IK)/(-BIGA)
 55     CONTINUE
       DO 65 I=1,N
       IK=NK+I
       IJ=I-NN
       DO 65 J=1,N
       IJ=IJ+NN
       IF(I-K)60,65,60
 60     IF(J-K)62,65,62
 62     KJ=IJ-I+K
       A(IJ)=A(IK)*A(KJ)+A(IJ)
 65     CONTINUE
       KJ=K-NN
       DO 75 J=1,N
       KJ=KJ+NN
       IF(J-K)70,75,70
 70     A(KJ)=A(KJ)/BIGA
 75     CONTINUE
       D=D*BIGA
       A(KK)=1.0/BIGA
 80     CONTINUE
       K=N
 100     K=(K-1)
       IF(K)150,150,105
```

```
 105     I=L(K)
        IF(I-K)120,120,108
 108     JQ=NN*(K-1)
        JR=NN*(I-1)
        DO 110 J=1,N
        JK=JQ+J
        HOLD=A(JK)
        JI=JR+J
        A(JK)=-A(JI)
 110     A(JI)=HOLD
 120    J=M(K)
        IF(J-K)100,100,125
 125    KI=K-NN
        DO 130 I=1,N
        KI=KI+NN
        HOLD=A(KI)
        JI=KI-K+J
        A(KI)=-A(JI)
 130    A(JI)=HOLD
        GO TO 100
 150    J1=1
        RETURN
        END
```

Appendix E

Multicomponent Simulation Program

The multicomponent simulation program is similar to that of the binary simulation program. The subroutines for the integration and the matrix inversion DIFSUB and MATINV are exactly the same as those preduced in Appendix D. The subroutines for calculating the collision integrals IPNEU and IDNEU are also the same as those for the binary case in Appendix D.

In this section only the subroutines which are different to the binary case are produced. These subroutines are:-

Master Ternary:- this subroutine estimates the column dimensions after first reading in the required data. The data and the order inwhich it is read in is given in detail before the subroutine listings. The nomenclature used for the program is roughly the same as that for the binary system.

The subroutines SYSTEM, EQUIDA, AICHE, WRITER, DIFFUN, MASTER DRAWER and PLOT 1 all have the same function as those described for the binary system. The only difference in the subroutines is that the multicomponent simulation program is not set to handle flow changes and that the DO LOOPS are more nested to handle the extra equations due to more than one component being operated on.

The only major difference between the binary subroutines and the multicomponent subroutines is that:-

SUBROUTINE AICHE combines all the work needed for estimating average liquid and vapour densities, Peclet number and the parameters needed to estimate the point efficiency. Further, the assumption of one point efficiency per plate is replaced by one point efficiency per pool. In this way the point efficiency can vary along the plate for any one component due to the interaction of the other components on it.

SUBROUTINE EQUIDA no longer estimates the linear equilibrium relationships but requires constant relative volatilities for using in the equations derived by Wood (116).

Data to be read in:-

N,NF,QFS,W,TL,NK

AMA,DSD,TBO,AREL,VZ1   (1 data card per component)

XFS(1 to NK)

XXS(1)

XXS(2 to total)

YA(1)

YA(2 to total)

TCAN,STEP1,STEP2,HMAX1,HMAX3

IMARK,XFS(1 to NK)

Nomenclature:-

n   number of plates counting the reboiler as plate 1.

NF	feed plate number.
QFS	q-line feed condition.
W	weir height (inches).
TL	column diameter (ft).
NK	number of components.
AMA	molecular weight.
DSD	density of the liquid (gm/cc).
TBO	boiling point of liquid °c.
AREL	relative volatility
VZ1	viscosity (centipoise).
XFS	feed composition (mf).
XXS(1 to NK)	reboiler compositions at steady-state (mf).
XXS(NK to total)	steady-state liquid compositions working up the column (mf).
YA(1 to NK)	average vapour composition leaving the reboiler.
YA(NK+1 to NxNK)	average vapour composition leaving the plates (mf).
XFS(1 to NK)	the new unsteady-state values of the feed compositions (mf).

```
      MASTER TERNARY
      DIMENSION ALS(10),AMA(5),DSD(5),TBO(5),AREL(5),VZ1(5),XFS(5)
      DIMENSION XXS(150),XS(150)
      DIMENSION YS(150),YA(150),YC(150),EMVP(30),ALE(30),YV(50)
      DIMENSION TEXIT(10)
      COMMON/AIS/XXS,XS,ALS,XFS,YS,YA,YC,EMVP,ALE,YV,FS,WS,DS,VS,VR,RR,T
     1COND,TBOIL,TEXIT
      COMMON/WEIR/W
      COMMON/DATE/TW,TL,TR,AD,TW1,TA,TAA,HDIA,TTH,HOA,ANT,AUD,COV
      COMMON/PHYS/AMA,TBO,DSD,AREL,VZ1
      COMMON/PARS/N,NF,NK,LQ,QFS
C
C
C     THIS ROUTINE READS IN COLUMN AND SYSTEM PARAMETERS, THEN
C     CALCULATES THE COLUMN AND TRAY AREAS.
C
C
C
C     NF=FEED PLATE        N=NUMBER OF PLATES WHERE THE REBOILER IS NO. 1
C     QFS=Q LINE FEED CONDITION, ASSUMED = OR > 1.
C     COLUMN DIMENSIONS:-  TW=WIDTH;  TAA=ACTIVE LENGTH;  TL=DIAMETER
C     PHYSICAL PROPERTIES:-    DENSITIES         DSD
C                              MOLE WEIGHTS      AMA
C                              VISCOSITIES       VZ1
C                              BOILING TEMP.     TBO
C                              RELATIVE VOLS.    AREL
C
C     WEIR HT. INS.=W      ACTIVE AREA=TA    POINT EFFICIENCY=EP
C     NUMBER OF POOLS FOR ACTIVE AREA=LQ
C     REFLUX RATIO=RR       BACKFLOW RATIO=BETA,B
C     LIQUID FLOW=ALS       VAPOUR FLOW=VV,VS,VR
C     MOLAL HOLDUPS=VD,VT,VR,VC FOR DOWNCOMER,TRAY,REBOILER,CONDENSER.
C     FEED RATE=FS     FEED COMPOSITION =XFS
C     BOTTOM RATE =WS        TOP RATE =DS
C     ESTIMATED TOP PRODUCTS=ETP    CALCULATED TOP PRODUCTS=CTP
C     COMPOSITIONS:-   LIQUID=XS ;Y    VAPOUR=YA   VAPOUR CELL=YC
```

```
C     PECLET NO.=PE   EDDY DIFFUSION COEFFICIENT=DE
C     HEIGHTS :- CLEAR LIQUID=HC     FROTH=HF
C     VELOCITIES :-  FROTH=UL  SUPERFICIAL GAS=UG
C     DENSITIES LB/CU.FT   LIQUID=PL     VAPOUR=PV
C     SPACE TIMES SECS.    REBOILER=TAUR      TEMPERATURES TBOIL
C                         CONDENSER=TAUC                   TCOND
C                         DOWNCOMER=TAUD                   TEXIT
C                             TRAYS=TAUT                   TEXIT
C
C
C     --------------------------------------------------------------------
C
C
C     SET COLUMN AND SYSTEM PARAMETERS
C
      READ(1,101)N,NF,QFS,W,TL,NK
  101 FORMAT(2I0,3F0.0,I0)
      W=W/12
      DO 1 I=1,NK
    1 READ(1,100)AMA(I),DSD(I),TBO(I),AREL(I),V21(I)
  100 FORMAT(5F0.0)
      DO 5 I=1,NK
    5 AREL(I)=AREL(I)/AREL(NK)
      READ(1,102)AKI,FS,WS,LP,LQ
  102 FORMAT(3F0.0,2I0)
      READ(1,103)(XFS(I),I=1,NK)
  103 FORMAT(5F0.0)
      TW=.885*TL
      TA1=SQRT((TL/2)**2-(.385*TL)**2)
      TD1=TL/2-TA1
      TW1=.77*TL
      AD2=TW1*TD1
      THETA=ASIN(TW1/TL)
      TTA=3.1415926*(TL**2)/4
      AS=TTA*THETA/3.1415926
      AD=AS-0.5*TA1*TW1
```

```
      TD2=TD1*AD/AD2
      TR=TD2+2*TA1
      TAA=2*TA1
      TA=TTA-2*AD
      HDIA=.1875
      TTH=0.1
      HOA=0.1*TTA
      ANT=TTA-AD
      AUD=TW1*1.0/12.0
      COV=1.09*(HDIA/TTH)**0.25
      W2=W*12.0
      APR=1.0
      PIT=0.5
      WRITE(2,200)
  200 FORMAT(1H1,//////////28X,31HDISTILLATION COLUMN DIMENSIONS.)
      WRITE(2,202)
  202 FORMAT(/30X,21H  WEIR AND DOWNCOMER.)
      WRITE(2,201)
  201 FORMAT(/15X,6HDIA.FT,4X,38HHT.INS     LENG.FT     SEC.AREA     APRON)
      WRITE(2,203)TL,W2,TW1,AD,APR
  203 FORMAT(17X,F3.1,6X,F4.2,6X,F6.4,5X,F6.3,8X,F3.1)
      WRITE(2,204)
  204 FORMAT(//21X,35HPERFORATIONS                TRAY AREAS.)
      WRITE(2,205)
  205 FORMAT(15X,48HDIA.INS   PITCH   AREA     NET    ACTIVE   GROSS)
      WRITE(2,206)HDIA,PIT,HOA,ANT,TA,TTA
  206 FORMAT(15X,F6.4,5X,F3.1,3X,F7.4,2X,F6.3,2X,F6.3,2X,F6.3)
C
C     READ IN INITIAL ESTIMATES OF TRAY CONCENTRATIONS
C     SET FEED RATE AND COMPOSITION
C
      NPOOL=(N-1)*(LQ+1)+2
      NLIM=NPOOL*NK
      READ(1,104)(XXS(I),I=1,NK)
      READ(1,104)(XXS(I),I=(NK+1),NLIM)
      READ(1,104)(YA(I),I=1,NK)
```

```
    READ(1,104)(YA(I),I=(NK+1),(NLIM-NK))
104 FORMAT(12F0.0)
    ALS(2)=AKI
    KR=LQ/LP
    LZ=LQ-LP
    WRITE(2,207)
207 FORMAT(//17X,4HDENS,4X,6HMOL WT,4X,4HVISC,4X,6HB.TEMP,5X,5HR.VOL)
    DO 2 J=1,NK
  2 WRITE(2,208)DSP(J),AMA(J),VZ1(J),TBO(J),AREL(J)
208 FORMAT(16X,2(F6.3,3X,F6.2,3X),F6.2)
    CALL SYSTEM
    STOP
    END
```

```fortran
      SUBROUTINE EQUIDA(J4,J3,NK,XXS,XS,YS,AREL,TBO,TBOIL,TCOND,LQ,N,TEP
     1)
      DIMENSION AREL(5),YS(150),XXS(150),XS(150),ES(150),TBO(5),EP(150)
      COMMON/LAM/ES
C
C     THIS SUBROUTINE CALCULATE THE EQUILIBRIUM VALUE OF THE VAPOUR
C     PHASE USING CONSTANT RELATIVE VOLATILITIES AND THE THEORY OF WOODS
C     THE TEMPERATURES ARE ALSO ESTIMATED.
C
      NLIM=((N-1)*(LQ+1)+2)*NK
      DO 6 I=1,NLIM
    6 XXS(I)=XS(I)
      SUM1=0.0
      SUM6=0.0
      DO 1 I=1,NK
      II=I+J3+J4
      SUM6=SUM6+TBO(I)*XS(II)
      SUM1=SUM1+AREL(I)*XS(II)
    1 CONTINUE
      SUM3=SUM1**2
      TEP=SUM6
      DO 2 II=1,NK
      I=II+J3+J4
      YS(I)=AREL(II)*XS(I)/SUM1
      SUM5=0.0
      SUM8=0.0
      SUM2=0.0
      DO 3 JJ=1,NK
      J=JJ+J3+J4
      IF(JJ.EQ.II)GO TO 4
      SUM2=SUM2+AREL(JJ)*XS(J)
      SUM8=SUM8-AREL(II)*AREL(JJ)*XS(J)
      GO TO 3
    4 CONTINUE
    3 CONTINUE
      SUM5=(-SUM8+SUM2*AREL(II))/SUM3
```

```
      ES(I)=SUM5
    2 CONTINUE
      NLIM=NK+N*(LQ+1)-NK*LQ
      SUM2=0.0
      SUM1=0.0
      DO 5 I=1,NK
      SUM2=SUM2+XS(NLIM+I)*TBO(I)
      SUM1=SUM1+XS(I)*TBO(I)
    5 CONTINUE
      TBOIL=SUM1
      TCOND=SUM2
      RETURN
      END
```

```
      SUBROUTINE SYSTEM
      DIMENSION YA(150),YS(150),YC(150),XXS(150),XS(150),TEXIT(10),ALS(1
     10),AREL(5),AMA(5),TBO(5),DSD(5),VZ1(5),XR(10),WD(10),WT(10),B(10)
      DIMENSION EMAXE(5),EMINE(5)
      DIMENSION ALE(30)
      DIMENSION PEN(10),EP(150),XFS(5),TIM(100),EMVP(30),YV(50)
      DIMENSION Y(8,80),SAVE(80,12),PSAVE(6400),YMAX(80),ERROR(80),A(8)
      DIMENSION PERTST(7,2,3)
      DIMENSION ETP(5),CTP(5)
      COMMON/TOPS/ETP,CTP
      COMMON/MAS/WT,WD,XR,B,PEN,EP,VB,VC
      COMMON/GRAF/PSAVE,SAVE,Y,EMAXE,EMINE
      COMMON/PHYS/AMA,TBO,DSD,AREL,VZ1
      COMMON/METH/MF
      COMMON/CTER/IY,IC,IMARK
      COMMON/AIS/XXS,XS,ALS,XFS,YS,YA,YC,EMVP,ALE,YV,FS,WS,DS,VS,VR,RR,T
     1COND,TBOIL,TEXIT
      COMMON/PARS/N,NF,NK,LQ,QFS
C
C     THIS SUBROUTINE IS THE CENTRE OF THE PROGRAM.
C     IT INITIALISES THE INTEGRATION ROUTINE PARAMETERS, CALLS THE
C     INTEGRATION ROUTINE THEN WRITES THE GENERATED DATA ONTO DISC.
C
      WRITE(2,201)
  201 FORMAT(1H1)
      READ(1,100)TCAN,STEP1,STEP2,HMAX1,HMAX3
  100 FORMAT(5F0.0)
      IZ=0
      IC=0
C
C     CALCULATE FLOW RATES OF STREAMS
C     SET VALUES FOR THE BOTTOM RATE
C
      DS=FS-WS
      DO 2 I=2,NF-1
    2 ALS(I+1)=ALS(I)
```

```
      DO 8 I=(NF+1),(N+1)
    8 ALS(I)=ALS(NF)-FS+QFS
      VS=(ALS(2)-WS)/LQ
      VR=(QFS*FS+ALS(NF+1)-WS)/LQ
      EPS=1.0E-3
      MF=2
      NL=NK-1
      NLIM=(N-1)*(LQ+1)+NL+2*NL
      DO 3 I=1,NLIM
      ERROR(I)=0.0
      IF(((I/2)*2).EQ.I)GO TO 10
      YMAX(I)=0.1
      YMAX(I+1)=0.3
   10 CONTINUE
      DO 3 J=1,NLIM
    3 PSAVE(J+(I-1)*NLIM)=0.0
      DO 4 I=1,8
      DO 4 J=1,NLIM
    4 V(I,J)=0.0
      DO 5 I=1,12
      DO 5 J=1,NLIM
    5 SAVE(J,I)=0.0
      DO 6 I=1,8
    6 A(I)=0.0
      DO 7 I=1,7
      DO 7 J=1,2
      DO 7 K=1,3
    7 PERTST(I,J,K)=0.0
C
C     SET THE MAXIMUM TIME TO RUN:- BIGT
C     SET THE MINIMUM, MAXIMUM AND INITIAL INTEGRATION STEP LENGTHS
C     SET THE INITIAL LIMITS FOR PLOTTING PURPOSES.
C
      T=0.0
      H=1.0E-7
      HMIN=1.0E-20
```

```
      BIGT=4*TCAN
      STEP=STEP1
      HMAX=HMAX1
      JSTART=0
      MAXDER=7
      MLIM=((N-1)*(LO+1)+2)*NK
      II=0
      DO 9 I=1,MLIM
      XS(I)=XXS(I)
      IF(((I/NK)*NK).EQ.I)GO TO 9
      II=II+1
      Y(1,II)=XXS(I)
    9 CONTINUE
      NN=NLIM
      N4=NLIM*NLIM
      TMAX=0.0
      DO 23 I=1,NK
      EMAXE(I)=0.0
   23 EMINE(I)=4.0
      T=0
   20 CONTINUE
      IY=0
      IF(T.GE.1.AND.KFLAG.LE.0)IY=1
      CALL DIFSUB(NN,T,Y,SAVE,H,HMIN,HMAX,EPS,MF,YMAX,ERROR,KFLAG,JSTART
     1,MAXDER,PSAVE,N4)
      IF(KFLAG)11,11,12
   11 WRITE(2,13)KFLAG
      EPS=EPS*10
   13 FORMAT(3X,6HKFLAG=,I5)
      GO TO 20
   12 CONTINUE
      HT=T
      JJ=0
      DO 1 JX=1,NLIM
      JJ=JJ+1
      XS(JJ)=Y(1,JX)
```

```
      IF(((JX/NL)*NL).NE.JX)GO TO 1
      JJ=JJ+1
      XS(JJ)=1.0-XS(JJ-1)-XS(JJ-2)
    1 CONTINUE
      IZ=IZ+1
      IF(T.LT.TMAX)GO TO 20
      TMAX=TMAX+STEP
      IF(I.NE.0)GO TO 32
      WRITE(2,202)
  202 FORMAT(1H1,////5X,6HLAMBDA,6X,5HSLOPE,7X,6HPT EFF,6X,5HC.P.E,7X,4H
     1EMVP,8X,5HC.E.R,6X,8HEMVP/EOG)
      DO 33 J=2,N
      JG=(LQ+1)*NK
      JF=(J-2)*JG+NK
      I=I+1
      IF(IC.LT.1)TST=T
      TACT=T-TST
      RR=ALS(N+1)/DS
      TIM(I)=T/60.0
      WRITE(2,200)T
  200 FORMAT(3X,F10.5)
      MLIM=((N-1)*(LQ+1)+2)*NK
      WRITE(3)TACT,FS,WS,DS,RR,VB,VC
      DO 15 J=1,NK
   15 WRITE(3)ARFL(J),XFS(J),XS(J),XS(MLIM-NK+J),ETP(J),CTP(J)
      DO 16 J=1,N-1
      JJ=J+1
      VF=VS*LQ
   16 WRITE(3)XR(JJ),TEXIT(JJ),PFN(JJ),B(JJ),ALS(JJ),VF
      DO 17 JI=1,NK
      DO 17 J=1,N-1
      MX1=((J-1)*(LQ+1)+1)*NK
      MX2=(J*(LQ+1)+1)*NK
      JS=MX2-NK+JI
      JV=J*NK+JI
      WRITE(3)(XS(JS-(JY-1)*NK),JY=1,LQ+1),(YA(JS-(JY-1)*NK),JY=2,LQ+1),
```

```
     1YV(JV),EMVP(JV-NK)
C
C     SET NEW LIMITS FOR PLOTTING.
C
      IF(EMINE(JI).GE.EMVP(JV-NK))EMINE(JI)=EMVP(JV-NK)
      IF(EMAXE(JI).LE.EMVP(JV-NK))EMAXE(JI)=EMVP(JV-NK)
   17 CONTINUE
      IF(IC.EQ.1)IC=2
      IF(T.GE.(2*TCAN))HMAX3=HMAX4
      IF((T-TST-4*TAU).GE.0.0.AND.IC.EQ.2)HMAX=HMAX3
      IF(IC.EQ.0.AND.T.GE.TCAN)IC=1
      IF(IC.NE.1)GO TO 22
      TAU=3600*XR(NF)/ALS(NF)
      HMAX2=TAU/2.0
      HMAX=HMAX2
      STEP=HMAX2/2.0
      HMAX3=(BIGT-TST-8.0*TAU)/20.0
      IF(HMAX3.LE.HMAX2)HMAX3=HMAX2
      HMAX4=4*HMAX3
      H=0.02
      TMAX=T
      JSTART=0
   22 CONTINUE
      IF(I.GE.60)GO TO 31
C
C     CALCULATED TRUE TIME IN MINUTES:- TIM(J)
C
      IF(T.LT.BIGT)GO TO 20
   31 CONTINUE
      DO 18 J=1,I
   18 TIM(J)=TIM(J)-TST/60.0
C
C     SET FINAL LIMITS FOR PLOTTING
C
      DO 24 J=1,NK
      EMAXE(J)=(INT(EMAXE(J)*10.0)+2.0)/10.0
```

```
      IF(EMINE(J).LT.0.0)GO TO 25
      EMINE(J)=(INT(EMINE(J)*10.0))/10.0
      GO TO 24
   25 EMINE(J)=(INT(EMINE(J)*10.0)-2.0)/10.0
   24 CONTINUE
      CALL WRITER(I,N,NK,LQ,TIM)
      RETURN
```

```
      SUBROUTINE WRITER(I,N,NK,LQ,TIM)
      DIMENSION TIM(100)
      DIMENSION EMAXE(5),EMINE(5)
      COMMON/AIS/AX(893)
      COMMON/GRAF/BX(8000),EMAXE,EMINE
C
C     THIS ROUTINE READS THE COMPUTED DATA ON DISC AND WRITES
C
      NLIM=I*NK*2*N
      NX1=I*(N+1)
      NX5=I*2*N
      REWIND 3
      DO 1 II=1,I
      WRITE(2,204)
      READ(3)(AX(J),J=1,7)
      WRITE(2,200)(AX(J),J=1,7)
      WRITE(2,205)
      DO 2 K=1,NK
      NX2=(K-1)*NX5
      NX3=NX2+N*I
      N2=7+(K-1)*6
      READ(3)(AX(N2+J),J=1,6)
      BX(NX2+II)=AX(10+(K-1)*6)
      BX(NX3+II)=AX(11+(K-1)*6)
    2 WRITE(2,208)K,(AX(N2+J),J=1,6)
      N3=N2+6
      NZ=6*(N-1)+7+6*NK
      WRITE(2,206)
      DO 3 K=1,N-1
      N4=N3+(K-1)*6
      READ(3)(AX(N4+J),J=1,6)
      WRITE(2,201)K,(AX(N4+J),J=1,6)
    3 CONTINUE
      N5=N4+6
      DO 4 JI=1,NK
      WRITE(2,202)JI
```

```
      WRITE(2,207)(J,J=1,LQ),(K,K=1,LQ)
      DO 4 JK=1,N-1
      NX4=(JI-1)*NX5+JK*I
      NX6=NX4+NX1-I
      N6=N5+((JI-1)*(N-1)+(JK-1))*(3+2*LQ)
      READ(3)(AX(N6+J),J=1,(2*LQ+3))
      WRITE(2,203)JK,(AX(N6+J),J=1,(2*LQ+3))
      NNX=NZ+LQ+1+((JK-1)+(JI-1)*(N-1))*(2*LQ+3)
      NNY=NZ+(JK+(JI-1)*(N-1))*(2*LQ+3)
      BX(NX4+II)=AX(NNX)
      BX(NX6+II)=AX(NNY)
    4 CONTINUE
    1 CONTINUE
C
C     REWRITE ONTO DISC THE INFORMATION NEEDED FOR PLOTTING.
C
      REWIND 3
      WRITE(3)NLIM,N,LQ,NK,I
      WRITE(3)(BX(J),J=1,NLIM)
      WRITE(3)(TIM(J),J=1,I)
      WRITE(3)(EMINE(J),EMAXE(J),J=1,NK)
  204 FORMAT(1H1,/////////18X,59HTIME    FEED    BOTTOMS    TOPS    R.RATIO
     1 RB.HOLDUP  C.HOLDUP)
  205 FORMAT(16X,45HCOMPONENT  REL.VOL  FEED COMP.  BOILER  COND.,4X,12H
     1ETP.    CTP.)
  206 FORMAT(12X,68HPLATE  HOLDUP  TEMP.  PECLET NO.  BACKFLOW  LIQUID F
     1LOW  VAPOUR FLOW)
  208 FORMAT(19X,I1,2X,2(6X,F4.2),3X,2(2X,F6.4),1X,2(F7.3,2X))
  200 FORMAT(16X,F7.3,2X,F5.1,2X,F6.1,3X,F6.1,3X,F6.3,2(3X,F7.4))
  201 FORMAT(14X,I1,4X,F6.4,2X,2(F6.2,3X),2X,F7.4,2(5X,F7.2,1X))
  202 FORMAT(12X,10HCOMPONENT=,I2,8X,18HLIQUID COMPOSITION,22X,18HVAPOUR
     1 COMPOSITION,9X,10HMEAN  EMV.)
  207 FORMAT(14X,10HPLATE   DC,12(4X,I1,1X))
  203 FORMAT(16X,I1,2X,14(1X,F5.4),F7.4)
      RETURN
      END
```

```
      SUBROUTINE AICHE
      DIMENSION AB(150)
      DIMENSION FS(150)
      DIMENSION FXS(150),EYA(150)
      DIMENSION ALS(10),AREL(5),VZ1(5),AMA(5),DSD(5),TEXIT(10),B(10),PEN
     1(10),WD(10),WT(10),XR(10),HOL(10),SL(10),DENS(10),VDEN(10)
      DIMENSION EPP(5),CDD(5),VIS(5),DG(5),ANG(150)
      DIMENSION XXS(150),XS(150),YS(150),YC(150),YA(150)
      DIMENSION EP(150),TBO(5)
      DIMENSION XFS(5)
      DIMENSION EMVP(30),ALE(30),YV(50)
      COMMON/LAM/ES
      COMMON/AIS/XXS,XS,ALS,XFS,YS,YA,YC,EMVP,ALF,YV,FS,WS,DS,VS,VR,RR,T
     1COND,TBOIL,TEXIT
      COMMON/WEIR/W
      COMMON/CTER/IY,IC,IMARK
      COMMON/PHYS/AMA,TBO,DSD,AREL,VZ1
      COMMON/DATE/TW,TL,TR,AD,TW1,TA,TAA,HDIA,TTH,HOA,ANT,AUD,COV
      COMMON/PARS/N,NF,NK,LQ,QFS
      COMMON/MAS/WT,WD,XR,B,PEN,EP,VB,VC
C
C     THIS SUBROUTINE CALCULATES ALL THE PARAMETERS INVOLVED WITH PLATE
C     AND POINT EFFICIENCIES.
C
      DO 1 J=2,N
    1 XR(J)=WD(J)+WT(J)
      VS=(ALS(2)-WS)/LQ
      VR=VS+(1-QFS)*FS/LQ
      IF(IY.NE.1)GO TO 20
      NLIM=((N-1)*(LQ+1)+2)*NK
   20 CONTINUE
      SUM1=0.0
      DO 23 J=1,NK
   23 SUM1=SUM1+AREL(J)*XS(J)
      SUM2=1.0/SUM1
      DO 19 J=1,NK
```

```
      YS(J)=AREL(J)*XS(J)/SUM1
      DO 19 I=1,LQ+1
      IF(I.EQ.(LQ+1))GO TO 19
      NL=J+I*NK
      YC(NL)=YS(J)
   19 CONTINUE
      DO 2 J=2,N
      J3=(J-2)*(LQ+1)*NK+NK
      ZT=0.0
      ZD=0.0
      ZV=0.0
      ZL=0.0
      DO 3 I=1,LQ+1
      J4=(I-1)*NK
      ZS=0.0
      DO 4 IK=1,NK
    4 ZS=ZS+XS(J3+J4+IK)*AMA(IK)
      DO 5 IK=1,NK
      ZD=ZD+XS(J3+J4+IK)*AMA(IK)*DSD(IK)/ZS
    5 ZL=ZL+XS(J3+J4+IK)*AMA(IK)
      IF(I.EQ.(LQ+1))GO TO 3
      CALL EQUIDA(J4,J3,NK,XXS,XS,YS,AREL,TBO,TBOIL,TCOND,LQ,N,TEP)
      ZT=ZT+TEP
      DO 6 IK=1,NK
    6 ZV=ZV+YS(J3+J4+IK)*AMA(IK)
    3 CONTINUE
      ZD=ZD/(LQ+1)
      ZL=ZL/(LQ+1)
      ZT=ZT/LQ
      ZV=ZV/LQ
      DENS(J)=ZD
      TEXIT(J)=ZT
      VV=VS
      IF(J.GE.NF)VV=VR
      SL(J)=ALS(J)*ZL
      VDEN(J)=VV*LQ*ZV
```

```
    2 CONTINUE
      IF(IMARK.EQ.2)GO TO 7
C
C     REBOILER BALANCE
C
      IF(IC.GE.1)VVB=VB
      PL=DENS(2)*62.4
      VB=TL*TW*0.5*PL/(SL(2)/ALS(2))
    7 CONTINUE
C
C     CALCULATES THE LIQUID AND FROTH HEIGHTS, AND FLOW RATE PARAMETERS,
C     CALCULATE PECLET NUMBER AND BACKFLOW RATIO
C
      DO 8 J1=2,N
      J3=(J1-2)*(LQ+1)*NK+NK
      VV=VS
      IF(J1.GE.NF)VV=VR
      ALC=SL(J1)/(3600*DENS(J1)*62.4)
      UG=VV*359.05*LQ*(TEXIT(J1)+273)/(3600*TA*273)
      PV=VDEN(J1)/(3600*UG*TA)
      PL=DENS(J1)*62.4
      WI=W*12
      F=UG*PV**0.5
      FL=60*.2642*ALC*6.24/(.22*TW)
      HC=(103.0+11.8*WI-40.5*F+1.25*FL)/(PL*12)
      HF=(2.53*F*F+1.89*WI-1.6)/12.0
      UL=FL*.22/(6.24*.2642*60*HC)
      DE=(0.0609395*HC*UL**3)/((HC*UL/HF)**2.5664)
      PE=UL*TR/DE
      AN=LQ+1
      BETA=AN/PE-0.5
      IF(IMARK.EQ.2)GO TO 9
      PEN(J1)=PE
      B(J1)=BETA
    9 CONTINUE
C
```

```
C     THE CALCULATION OF THE POINT EFFICIENCY ACCORDING TO THE A.I.CH.E.
C     MANUAL. THESE CALCULATIONS INCLUDE NUMBER OF GAS STAGES ; THE
C     SLOPE OF THE EQUILIBRIUM LINE ; ABSORPTION COEFFICIENT.
C
      TEXIT(J1)=TEXIT(J1)+273.2
      IXI=(LQ+1)*NK*2
      DO 10 JK=1,LQ
      IJK=(JK-1)*NK+J3
      DO 11 II=1,NK
   11 AB(IJK+II)=ES(IJK+II)*LQ*VV/ALS(J1)
   10 CONTINUE
      DO 14 JK=1,LQ
      IJK=(JK-1)*NK+J3
      LK=JK*NK
      DO 12 II=1,NK
      EPP(II)=1.15*(TBO(II)+273.2)
      XXZ=TEXIT(J1)/EPP(II)
      CALL IPNEU(XXZ,ANS)
      CDD(II)=1.18*((AMA(II)/DSD(II))**(1.0/3.0))
      VIS(II)=0.00646*ANS*SQRT(AMA(II)*TEXIT(J1))/(CDD(II)**2)
   12 CONTINUE
      ICO=0
      SUM2=0.0
      DO 15 JI=1,NK
      SUM1=0.0
      SUM3=0.0
      DO 16 JJ=1,NK
      IF(YA(IJK+JJ).LE.0.0)YA(IJK+JJ)=YS(IJK+JJ)
      IF(JJ.EQ.JI)GO TO 16
      ICO=ICO+1
      BC1=(CDD(JJ)+CDD(JI))/2.0
      BC2=SQRT(EPP(JJ)*EPP(JI))
      BC3=TEXIT(J1)/BC2
      CALL IDNEU(BC3,AMS)
      BC4=SQRT((AMA(JI)+AMA(JJ))/(AMA(JI)*AMA(JJ)))
      BC5=(10.7-2.46*BC4)*1.0E-4
```

```
      DGG=(BC5*BC4*TEXIT(J1)**(3.0/2.0))/(AMS*BC1**2)
      SUM1=SUM1+YA(IJK+JJ)/DGG
      THI=SQRT(2.0)*((1+(VIS(JI)/VIS(JJ))**0.5*(AMA(JJ)/AMA(JI))**0.25)*
     1*2)/(4*SQRT(1+AMA(JI)/AMA(JJ)))
      SUM3=SUM3+YA(IJK+JJ)*THI
   16 CONTINUE
      DG(JI)=(1.0-YA(IJK+JI))/SUM1
      SUM2=SUM2+VIS(JI)/(1+SUM3/YA(IJK+JI))
   15 CONTINUE
      VG=SUM2
      SUM1=0.0
      DO 17 JI=1,NK
      SN=VG/(PV*DG(JI))
      ANG(JI+IJK)=(0.776+0.11*WI-0.29*F+0.0217*FL)/SQRT(SN)
      EP(JI+IJK)=1.0-EXP(-ANG(JI+IJK))
      DG(JI)=(1-EP(JI+IJK))*YC(JI+IJK)+EP(JI+IJK)*YS(IJK+JI)
      SUM1=SUM1+DG(JI)
   17 CONTINUE
      DO 18 JI=1,NK
      YA(JI+IJK)=DG(JI)+YS(JI+IJK)*(1.0-SUM1)
   18 YC(IXI+J3-LK-NK+JI)=YA(JI+IJK)
   14 CONTINUE
      TEXIT(J1)=TEXIT(J1)-273.2
      IF(IMARK.EQ.2)GO TO 8
C
C     CALCULATES MOLAR HOLOUPS ON THE TRAY AND THE DOWNCOMERS.
C
      HOL(J1)=HC*PL
      WT(J1)=TA*HOL(J1)/(SL(J1)/ALS(J1))
      HVB=21*(TEXIT(J1)+273.2)
      SIG=HVB*DENS(J1)/364.0
      H2=(0.558*(ALC/AUD)**2)/12.0
      HSIG=0.06*SIG/(PL*HDIA*12)
      VH=UG*TA/HOA
      G4=SORT(TA*4/3.1415926)
      G3=PV*UG*G4/(VG/3600.0)
```

```
      FF=0.014*G3**(-0.0055/9600.0)
      HD=12*COV*PV*(VH**2)*(0.4*(1.25-HOA/ANT)+4*TTA*FF/HDIA+(1-HOA/ANT)
     1**2)/(64*PL*12)
      H3=HD+HC+H2+HSIG
      WD(J1)=AD*H3*PL/(SL(J1)/ALS(J1))
      IF(J1.NE.N)GO TO 8
      VC=2*1*2*PL/(SL(N)/ALS(N))
    8 CONTINUE
      IF(IY.NE.1)GO TO 21
   21 CONTINUE
      RETURN
      END
```

```
      SUBROUTINE DIFFUN(T,Y,SAVE,N2,NN)
      DIMENSION PWD(10).PWT(10)
      DIMENSION XXS(150),XS(150),YS(150),YC(150),YA(150),YOUT(5),XR(10)
      DIMENSION ALS(10),YV(50),B(10),TEXIT(10),Y(8,NN)
      DIMENSION SAVE(888,1),PEN(10)
      DIMENSION WD(10),WT(10),SUM2(5),EP(150).ALE(30).EMVP(30),XFS(5)
      DIMENSION ETP(5),CTP(5)
      COMMON/CTER/IY,IC,IMARK
      COMMON/AIS/XXS,XS,ALS,XFS,YS,YA,YC.EMVP,ALE,YV,FS,WS,DS,VS,VR,R9,T
     1COND,TBOIL,TEXIT
      COMMON/TOPS/ETP,CTP
      COMMON/PARS/N,NF,NK,LQ,QFS
      COMMON/MAS/WT,WD,XR,B,PEN,EP,VB,VC
      COMMON/HDIF/TPD(10),TPT(10)
C
C     THIS ROUTINE CONTAINS ALL THE DIFFERENTIAL EQUATIONS
C
      NL=NK-1
      ALS(1)=WS
      IY=IY+1
      IF(IMARK.NE.2)IMARK=1
      IF(IY.NE.1)GO TO 1
      IF(IC.NE.1)GO TO 1
C
C     SET IMARK EQUAL TO 2 FOR NO CHANGE IN FLOWS.
C
C     INSERT THE STEP CHANGE TO BE MADE.
      READ(1,100)IMARK,(XFS(I),I=1,NK)
  100 FORMAT(I0,5F0.0)
C
    1 CONTINUE
      IF(IC.GE.1.AND.IY.EQ.1)WRITE(2,200)IMARK,(XFS(I),I=1,NK),T
  200 FORMAT(3X,I4,3X,4F10.4)
      IF(IY.EQ.1)CALL AICHE
C
C     REBOILER BALANCE
```

```
C
      TAUR=3600.0*VB/ALS(2)
      DO 3 I=1,NK-1
      SAVE(I+N2-1,1)=(Y(1,I+NL)-(WS*Y(1,I)+VS*LQ*YS(I))/ALS(2))/TAUR
      YV(I)=YC(I)
    3 CONTINUE
      YV(NK)=YC(NK)
      DO 4 J1=2,N
      VV=VS
      IF(J1.GE.NF)VV=VR
      VR=VS+(1-QFS)*FS/LQ
      J=(J1-2)*(LQ+1)*(NK-1)+(NK-1)
      J2=(J1-2)*(LQ+1)*NK+NK
      DO 5 I=1,NK-1
    5 SUM2(I)=0.0
      SUM2(NK)=LQ
      ALPHA=B(J1)/(1+B(J1))
      BETA=B(J1)
C
C     SET HOLDUPS
C
      VT=WT(J1)
      VD=WD(J1)
      IF(IV.NE.1)GO TO 22
      TVD=PWD(J1)-WD(J1)
      TVC=PWT(J1)-WT(J1)
      PWD(J1)=WD(J1)
      PWT(J1)=WT(J1)
      IF(IMARK.EQ.2)GO TO 16
      IF((T-HT).LE.0.0)GO TO 16
      TVD=TVD/(T-HT)
      TVC=TVC/((T-HT)*LQ)
      TPD(J1)=TVD
      TPT(J1)=TVC
      GO TO 17
   16 TVD=0.0
```

```
      TVC=0.0
      TPD(J1)=TVD
      TPT(J1)=TVC
   17 CONTINUE
      IF(IMARK.EQ.2)GO TO 22
      IF(XR(J1).LE.2.0)GO TO 22
      IF(IC.LT.1)GO TO 22
   27 CONTINUE
   22 CONTINUE
C
C     EXIT POOL
C
      TAUT=3600.0*VT/ALS(J1)
      DO 6 I=1,NL
      SAVE(J+N2-1+I,1)=((1+BETA)*(Y(1,J+NL+I)-Y(1,J+I))+VV*(YC(J2+I)-YA(
     1J2+I))/ALS(J1)-Y(1,J+I)*TVC/ALS(J1))*LQ/TAUT
      SUM2(I)=SUM2(I)+YA(J2+I)
      SUM2(NK)=SUM2(NK)-YA(J2+I)
    6 CONTINUE
C     CENTRE POOLS ON THE TRAY
C
C
      DO 7 IK=2,LQ
      JK=(IK-1)*NL+J
      JL=(IK-1)*NK+J2
      DO 8 I=1,NL
      SUM2(I)=SUM2(I)+YA(JL+I)
      SUM2(NK)=SUM2(NK)-YA(JL+I)
      SAVE(JK+N2-1+I,1)=((1+BETA)*Y(1,JK+I+NL)-(1+2*BETA)*Y(1,JK+I)+BETA
     1*Y(1,JK+I-NL)+VV*(YC(JL+I)-YA(JL+I))/ALS(J1)-Y(1,JK+I)*TVC/ALS(J1)
     2)*LQ/TAUT
      IF(J1.NE.NF)GO TO 8
      SAVE(JK+N2-1+I,1)=((1+BETA)*Y(1,JK+I+NL)-(1+2*BETA)*Y(1,JK+I)+BETA
     1*Y(1,JK+I-NL)+(VS*YC(JL+I)-YA(JL+I)*VR)/ALS(J1)-Y(1,JK+I)*TVC/ALS(
     2J1))*LQ/TAUT
    8 CONTINUE
```

```
    7 CONTINUE
C       DOWNCOMER AT BEGINNING OF TRAY
C
C
        TAUD=VD*3600.0/ALS(J1)
        DO 10 I=1,NL
        NLL=LQ*NL+J
        SAVE(NLL+N2-1+I,1)=(ALS(J1+1)*Y(1,NLL+NL+I)/ALS(J1)+BETA*Y(1,NLL+I
       1-NL)-(1+BETA+TVD/ALS(J1))*Y(1,NLL+I))/TAUD
        IF(J1.EQ.NF)SAVE(N2-1+I+NLL,1)=((ALS(J1+1)*Y(1,NLL+NL+I))/ALS(J1)+
       1BETA*Y(1,NLL+I-NL)-(1+BETA+TVD/ALS(J1))*Y(1,NLL+I)+QFS+FS*XFS(I)/A
       2LS(J1))/TAUD
   10 CONTINUE
C
C       VAPOUR CELL CONCENTRATION
C
C
C       PLATE EFFICIENCY AND EFFICIENCY RATIO CALCULATIONS
C
        DO 11 I=1,NK
        JN=(J1-1)*NK
        YV(JN+I)=SUM2(I)/LQ
        EMVP(JN+I-NK)=(YV(JN+I)-YV(JN-NK+I))/(YS(J2+I)-YV(JN-NK+I))
        ALF(JN+I-NK)=EMVP(JN-NK+I)/EP(J2+I)
   11 CONTINUE
    4 CONTINUE
C
C       CONDENSER BALANCE
C
        TAUC=3600*VC/(VV*LQ)
        DO 12 I=1,NK
        SUM1=0.0
        DO 13 JJ=1,LQ
        JI=JJ*NK-NK
   13 SUM1=SUM1+YA(J2+JI+I)
        YOUT(I)=SUM1/LQ
```

```
      JI=(N-1)*(LQ+1)*NL+NL
      IF(I.EQ.NK)GO TO 14
      SAVE(JI+N2-1+I,1)=(YOUT(I)-Y(1,JI+I))/TAUC
   14 CONTINUE
   12 CONTINUE
      JI=(N-1)*(LQ+1)*NK+NK
C
C     TOP PRODUCT COMPARISON.
C
      DO 15 I=1,NK
      CTP(I)=DS*XS(JI+I)
   15 ETP(I)=FS*XFS(I)-WS*XS(I)
      RETURN
      END
```

```
MASTER DRAWER
DIMENSION TIM(100)
DIMENSION EMAXE(5),EMINE(5)
COMMON/GRAF/BX(8000),EMAXE,EMINE
CALL UTPOP
REWIND 3
READ(3)NLIM,N,LQ,NK,I
READ(3)(BX(J),J=1,NLIM)
READ(3)(TIM(J),J=1,I)
READ(3)(EMINE(J),EMAXE(J),J=1,NK)
CALL PLOT1(NLIM,N,LQ,NK,I,TIM)
CALL UTPCL
STOP
END
```

```
      SUBROUTINE PLOT1(NLIM,N,LQ,NK,I,TIM)
      DIMENSION TIM(100),C(100)
      DIMENSION EMAXE(5),EMINE(5)
      COMMON/GRAF/BX(8000),EMAXE,EMINE
C
C     THIS SUBROUTINE PLOTS THE DATA REQUIRED.
C
      XINS=5.0
      YINS=7.0
      XMAX=TIM(I)
      XMIN=TIM(1)
      X1=6.0
      Y1=2.0
      Y2=4.0
      Y3=0.0
      Y4=7.0
      X2=(-TIM(1)/(TIM(I)-TIM(1)))*XINS
      DO 1 JK=1,NK
      CALL UTP2(X1,Y1,-1)
      YMAX=1.0
      IF(JK.EQ.2)YMAX=0.35
      YMIN=0.0
      NY=5
      NX=2
      CALL UTP4A(XMIN,XMAX,YMIN,YMAX,XINS,YINS,13HTIME MINUTES.,NX,39HLI
     1QUID CONCENTRATION LEAVING THE PLATES,NY)
      N2=(JK-1)*2*N*I
      N3=N2+I*(N+1)
      DO 2 J=1,N+1
      N1=(J-1)*I+N2
      DO 3 K=1,I
    3 C(K)=BX(N1+K)
    2 CALL UTP4B(TIM,C,I,3)
      CALL UTP2(X2,Y3,1)
      CALL UTP2(X2,Y4,2)
      NY=2
```

```
      YMAX=EMAXE(JK)
      YMIN=EMINE(JK)
      CALL UTP2(X1,Y2,-1)
      CALL UTP4A(XMIN,XMAX,YMIN,YMAX,XINS,YINS,13HTIME MINUTES.,NX,16HPL
     1ATE EFFICIENCY,NY)
      DO 4 J=1,N-1
      N4=N3+(J-1)*I
      DO 5 K=1,I
    5 C(K)=BX(N4+K)
    4 CALL UTP4B(TIM,C,I,3)
      CALL UTP2(X2,Y3,1)
      CALL UTP2(X2,Y4,2)
      CALL UTP2(X1,Y1,-1)
    1 CONTINUE
      RETURN
      END
      FINISH
```

Appendix F

Unsteady-State Simulation Results for a Feed Composition Change: Binary System

The simulation results of a step change in the feed composition for a binary system are given. The operating conditions are also given in the results. The step change made was a step change in the feed composition from 0.5 to 0.4.

Not all the simulation results are given as there were too many, but results at suitable intervals were selected to give a comprehensive list showing the trends of the responses.

```
            DISTILLATION COLUMN DIMENSIONS.

                 WEIR AND DOWNCOMER.

DIA.FT     HT.INS     LENG.FT     SEC.AREA     APRON
  7.0       2.25      5.3900       4.747         1.0


     PERFORATIONS                 TRAY AREAS.
DIA.INS    PITCH    AREA      NET     ACTIVE    GROSS
0.1875      0.5    3.8485   33.737   28.990   38.485


               DATA USED FOR COMPUTATION.
M.WT.A    M.WT.B    ROE A    ROE B    VIS A    VIS B    REL.V.
 78.00     92.00    0.680    0.776     0.26     0.27     1.10
```

```
                     TIME.     FEED      XF      WS       DS     REFLUX
                     0.000    300.00    0.50   150.00   150.00    4.667


                      REBOILER          CONDENSER         TOP PRODUCTS
MOLAR HOLDUP         11.64460            2.15163
LIQUID COMP.          0.45967            0.54033
VAPOUR COMP.          0.48342                            CAL.=  81.0497
BOILING TEMP         94.331             91.671           EST.=  81.0490


PLATE                  1                  2                 3                  4                  5
                X       Y          X       Y          X       Y          X       Y          X       Y
DOWNCOMER   0.49004            0.49997            0.51106            0.52389            0.53827
              0.4879  0.4988   0.4979  0.5092   0.5088  0.5193   0.5213  0.5301   0.5354  0.5424
              0.4860  0.4977   0.4961  0.5085   0.5067  0.5184   0.5190  0.5290   0.5328  0.5412
              0.4841  0.4967   0.4945  0.5080   0.5049  0.5176   0.5170  0.5283   0.5305  0.5402
              0.4825  0.4958   0.4930  0.5076   0.5033  0.5170   0.5152  0.5277   0.5285  0.5396
              0.4810  0.4950   0.4918  0.5074   0.5020  0.5167   0.5137  0.5274   0.5268  0.5392
              0.4799  0.4944   0.4909  0.5074   0.5011  0.5167   0.5127  0.5275   0.5258  0.5393
LIQUID FLOW     1000.000          1000.000           700.000           700.000            700.000
VAPOUR FLOW       STRIPPER= 850.000               RECTIFIER= 850.000
V ENTERING       0.48342           0.49640           0.50802           0.51762            0.52834
V LEAVING.       0.49640           0.50802           0.51762           0.52834            0.54033
V EQUILIBRIUM    0.50367           0.51471           0.52490           0.53651            0.54946
BOILING TEMP.    93.645            93.276            92.938            92.557             92.136
PECLET NO         7.781             7.834             5.619             5.664              5.714
BACKFLOW          0.400             0.394             0.746             0.736              0.725
EQ. SLOPE        1.0006            0.9986            0.9966            0.9943             0.9917
LAMBDA           0.8505            0.8488            1.2101            1.2073             1.2042
EOG.             0.5419            0.5418            0.4742            0.4742             0.4741
EMV.             0.6409            0.6345            0.5684            0.5674             0.5678
EMV/EOG          1.1826            1.1711            1.1986            1.1967             1.1978
R.VOLATILITY.    1.1000            1.1000            1.1000            1.1000             1.1000
MOLAR HOLDUPS    4.0128            4.0211            3.4837            3.4927             3.5029
```

```
                    TIME.     FEED      XF      WS        DS      REFLUX
                    0.020    300.00    0.40   150.00    150.00    4.667


                 REBOILER              CONDENSER           TOP PRODUCTS
MOLAR HOLDUP     11.64460               2.15163
LIQUID COMP.      0.45967               0.54033
VAPOUR COMP.      0.48342                                CAL.=  81.0497
BOILING TEMP     94.331                91.671            EST.=  51.0490


PLATE               1                    2                    3                    4                    5
               X         Y          X         Y          X         Y          X         Y          X         Y
DOWNCOMER  0.49004             0.49996              0.51106              0.52389              0.53827
            0.4879    0.4988    0.4979    0.5092    0.5088    0.5193    0.5213    0.5301    0.5354    0.5424
            0.4860    0.4977    0.4961    0.5085    0.5067    0.5184    0.5190    0.5290    0.5328    0.5412
            0.4841    0.4967    0.4944    0.5080    0.5049    0.5176    0.5169    0.5283    0.5305    0.5402
            0.4825    0.4958    0.4930    0.5076    0.5033    0.5170    0.5152    0.5277    0.5285    0.5396
            0.4810    0.4950    0.4918    0.5074    0.5020    0.5167    0.5137    0.5274    0.5268    0.5392
            0.4799    0.4944    0.4909    0.5074    0.5011    0.5167    0.5127    0.5275    0.5258    0.5393
LIQUID FLOW      1000.000            1000.000              700.000              700.000              700.000
VAPOUR FLOW       STRIPPER= 850.000                   RECTIFIER= 850.000
V ENTERING        0.48342             0.49640              0.50802              0.51762              0.52834
V LEAVING.        0.49640             0.50802              0.51762              0.52834              0.54033
V EQUILIBRIUM     0.50367             0.51471              0.52490              0.53651              0.54946
BOILING TEMP.      93.645              93.276               92.938               92.557               92.136
PECLET NO           7.781               7.834                5.619                5.664                5.714
BACKFLOW            0.400               0.394                0.746                0.736                0.725
EQ. SLOPE          1.0006              0.9986               0.9966               0.9943               0.9917
LAMBDA             0.8505              0.8488               1.2101               1.2073               1.2042
EOG.               0.5419              0.5418               0.4742               0.4742               0.4741
EMV.               0.6409              0.6345               0.5684               0.5674               0.5679
EMV/EOG.           1.1826              1.1711               1.1985               1.1967               1.1978
R.VOLATILITY.      1.1000              1.1000               1.1000               1.1000               1.1000
MOLAR HOLDUPS      4.0128              4.0211               3.4837               3.4927               3.5029
```

```
                TIME.      FEED       XF       WS        DS       REFLUX
                10.737     300.00     0.40     150.00    150.00    4.667


                  REBOILER               CONDENSER              TOP PRODUCTS
MOLAR HOLDUP     11.64460                 2.15163
LIQUID COMP.      0.45950                 0.53946
VAPOUR COMP.      0.48325                                  CAL.=  80.9186
BOILING TEMP      94.337                 91.699            EST.=  51.0745


PLATE                  1                        2                        3                        4                        5
                 X          Y            X          Y            X          Y            X          Y            X          Y
DOWNCOMER   0.48517                 0.47450                 0.51037                 0.52324                 0.53770
              0.4842     0.4966     0.4767     0.4974     0.5080     0.5163     0.5204     0.5259     0.5349     0.5411
              0.4830     0.4960     0.4738     0.4988     0.5057     0.5150     0.5180     0.5254     0.5323     0.5397
              0.4819     0.4954     0.4806     0.5000     0.5037     0.5136     0.5159     0.5252     0.5299     0.5386
              0.4807     0.4948     0.4819     0.5010     0.5018     0.5121     0.5141     0.5251     0.5278     0.5377
              0.4796     0.4942     0.4829     0.5018     0.5002     0.5108     0.5127     0.5252     0.5261     0.5370
              0.4787     0.4937     0.4835     0.5024     0.4991     0.5095     0.5118     0.5255     0.5250     0.5367
LIQUID FLOW       1000.000               1000.000                700.000                700.000                700.000
VAPOUR FLOW         STRIPPER= 850.000                    RECTIFIER= 850.000
V ENTERING         0.48325                0.49513                0.50024                0.51288                0.52538
V LEAVING.         0.49513                0.50024                0.51288                0.52538                0.53847
V EQUILIBRIUM      0.50256                0.50735                0.52295                0.53559                0.54868
BOILING TEMP.       93.683                 93.522                 93.003                 92.587                 92.161
PECLET NO            7.781                  7.834                  5.619                  5.664                  5.714
BACKFLOW             0.400                  0.394                  0.746                  0.736                  0.725
EQ. SLOPE           1.0006                 1.0002                 0.9966                 0.9943                 0.9917
LAMBDA              0.8505                 0.8502                 1.2102                 1.2073                 1.2042
EOG.                0.5419                 0.5418                 0.4742                 0.4742                 0.4741
EMV.                0.6153                 0.4181                 0.5568                 0.5505                 0.5619
EMV/EOG.            1.1355                 0.7716                 1.1742                 1.1611                 1.1852
R.VOLATILITY.       1.1000                 1.1000                 1.1000                 1.1000                 1.1000
MOLAR HOLDUPS       4.0128                 4.0211                 3.4837                 3.4927                 3.5029
```

```
                    TIME.      FEED       XF       WS        DS       REFLUX
                    18.037     300.00     0.40     150.00    150.00    4.667

                    REBOILER               CONDENSER          TOP PRODUCTS
MOLAR HOLDUP        11.64460               2.15163
LIQUID COMP.         0.45901               0.53831
VAPOUR COMP.         0.48275                                  CAL.=  80.7472
BOILING TEMP        94.354                91.735              EST.=  51.1482

PLATE                  1                   2                   3                   4                   5
                   X         Y         X         Y         X         Y         X         Y         X         Y
DOWNCOMER      0.47949             0.47039             0.50921             0.52236             0.53676
                 0.4792    0.4937    0.4715    0.4939    0.5066    0.5132    0.5195    0.5240    0.5339    0.5395
                 0.4787    0.4935    0.4728    0.4947    0.5043    0.5118    0.5169    0.5234    0.5313    0.5381
                 0.4782    0.4932    0.4740    0.4955    0.5021    0.5104    0.5147    0.5230    0.5290    0.5370
                 0.4775    0.4928    0.4752    0.4964    0.5001    0.5090    0.5129    0.5228    0.5269    0.5361
                 0.4769    0.4925    0.4763    0.4971    0.4985    0.5078    0.5114    0.5229    0.5252    0.5355
                 0.4764    0.4922    0.4771    0.4976    0.4973    0.5068    0.5105    0.5232    0.5240    0.5353
LIQUID FLOW        1000.000            1000.000             700.000             700.000             700.000
VAPOUR FLOW          STRIPPER= 850.000                   RECTIFIER= 850.000
V ENTERING          0.48275             0.49298             0.49586             0.50982             0.52321
V LEAVING.          0.49298             0.49586             0.50982             0.52321             0.53689
V EQUILIBRIUM       0.50018             0.50091             0.52111             0.53428             0.54774
BOILING TEMP.       93.763              93.738              93.064              92.630              92.191
PECLET NO            7.781               7.834               5.619               5.664               5.714
BACKFLOW             0.400               0.394               0.746               0.736               0.725
EQ. SLOPE            1.0012              1.0017              0.9968              0.9944              0.9918
LAMBDA               0.8510              0.8515              1.2104              1.2075              1.2043
EOG.                 0.5419              0.5419              0.4742              0.4742              0.4741
EMV.                 0.5866              0.3637              0.5528              0.5476              0.5576
EMV/EOG.             1.0825              0.6712              1.1656              1.1549              1.1763
R.VOLATILITY.        1.1000              1.1000              1.1000              1.1000              1.1000
MOLAR HOLDUPS        4.0128              4.0211              3.4837              3.4927              3.5029
```

```
                  TIME.       FEED       XF        WS         DS         REFLUX
                  23.337      300.00     0.40    150.00     150.00       4.667


                  REBOILER              CONDENSER              TOP PRODUCTS
MOLAR HOLDUP      11.64460               2.15163
LIQUID COMP.       0.45813               0.53696
VAPOUR COMP.       0.48187                                  CAL.=  80.5440
BOILING TEMP       94.384                91.778             EST.=  51.2807


PLATE                1                     2                      3                    4                     5
                X          Y          X          Y          X          Y          X          Y          X          Y
DOWNCOMER    0.47445               0.46804               0.50777               0.52126               0.53555
              0.4744     0.4907     0.4686     0.4914     0.5051     0.5103     0.5183     0.5223     0.5327     0.5377
              0.4743     0.4906     0.4693     0.4918     0.5026     0.5090     0.5157     0.5215     0.5301     0.5364
              0.4741     0.4905     0.4701     0.4923     0.5003     0.5076     0.5134     0.5210     0.5278     0.5353
              0.4738     0.4904     0.4709     0.4928     0.4983     0.5064     0.5115     0.5207     0.5257     0.5345
              0.4735     0.4902     0.4716     0.4932     0.4966     0.5053     0.5100     0.5207     0.5240     0.5339
              0.4732     0.4901     0.4722     0.4936     0.4954     0.5046     0.5090     0.5210     0.5229     0.5338
LIQUID FLOW      1000.000             1000.000               700.000              700.000               700.000
VAPOUR FLOW        STRIPPER= 850.000                    RECTIFIER= 850.000
V ENTERING       0.48187              0.49043               0.49251               0.50721               0.52121
V LEAVING.       0.49043              0.49251               0.50721               0.52121               0.53527
V EQUILIBRIUM    0.49704              0.49604               0.51919               0.53280               0.54656
BOILING TEMP.     93.869               93.903                93.127                92.679                92.229
PECLET NO          7.781                7.834                 5.619                 5.664                 5.714
BACKFLOW           0.400                0.394                 0.746                 0.736                 0.725
EQ. SLOPE         1.0019               1.0027                0.9971                0.9946                0.9920
LAMBDA            0.8516               0.8523                1.2108                1.2077                1.2045
EOG.              0.5420               0.5420                0.4743                0.4742                0.4741
EMV.              0.5648               0.3700                0.5509                0.5473                0.5545
EMV/EOG.          1.0421               0.6826                1.1616                1.1541                1.1697
R.VOLATILITY.     1.1000               1.1000                1.1000                1.1000                1.1000
MOLAR HOLDUPS     4.0128               4.0211                3.4837                3.4927                3.5029
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                   TIME.      FEED      XF       WS        DS      REFLUX
                   32.637     300.00    0.40   150.00    150.00     4.667


                     REBOILER              CONDENSER          TOP PRODUCTS
MOLAR HOLDUP        11.64460               2.15163
LIQUID COMP.         0.45689               0.53548
VAPOUR COMP.         0.43062                              CAL.=  80.3227
BOILING TEMP        94.427                91.825          EST.=  51.4665


PLATE                 1                     2                     3                     4                     5
                   X         Y         X         Y         X         Y         X         Y         X         Y
DOWNCOMER     0.47051             0.46631             0.50620             0.51999             0.53417
                0.4705    0.4880    0.4666    0.4892    0.5034    0.5079    0.5170    0.5206    0.5313    0.5360
                0.4704    0.4880    0.4670    0.4894    0.5008    0.5066    0.5143    0.5197    0.5287    0.5347
                0.4703    0.4879    0.4675    0.4897    0.4985    0.5053    0.5120    0.5192    0.5264    0.5336
                0.4702    0.4879    0.4679    0.4900    0.4964    0.5042    0.5100    0.5188    0.5243    0.5328
                0.4701    0.4878    0.4684    0.4903    0.4946    0.5032    0.5085    0.5187    0.5226    0.5323
                0.4699    0.4877    0.4688    0.4905    0.4935    0.5025    0.5075    0.5190    0.5215    0.5323
LIQUID FLOW      1000.000            1000.000             700.000             700.000             700.000
VAPOUR FLOW         STRIPPER= 850.000             RECTIFIER= 850.000
V ENTERING       0.48062             0.48788             0.48984             0.50494             0.51934
V LEAVING.       0.48788             0.48984             0.50494             0.51934             0.53364
V EQUILIBRIUM    0.49374             0.49257             0.51728             0.53125             0.54523
BOILING TEMP.    93.931              94.020              93.191              92.729              92.273
PECLET NO         7.781               7.834               5.619               5.664               5.714
BACKFLOW          0.400               0.394               0.746               0.736               0.725
EQ. SLOPE        1.0027              1.0034              0.9974              0.9949              0.9922
LAMBDA           0.8523              0.8529              1.2112              1.2080              1.2048
EOG.             0.5420              0.5421              0.4743              0.4742              0.4741
EMV.             0.5538              0.4167              0.5504              0.5474              0.5524
EMV/EOG.         1.0218              0.7688              1.1606              1.1545              1.1653
R.VOLATILITY.    1.1000              1.1000              1.1000              1.1000              1.1000
MOLAR HOLDUPS    4.0128              4.0211              3.4837              3.4927              3.5029
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                    TIME.      FEED      XF      WS        DS      REFLUX
                   39.937     300.00    0.40   150.00    150.00     4.667


                   REBOILER              CONDENSER          TOP PRODUCTS
MOLAR HOLDUP       11.64460               2.15163
LIQUID COMP.        0.45538               0.53394
VAPOUR COMP.        0.47910                                CAL.=  80.0913
BOILING TEMP       94.479                91.874            EST.=  51.6926


PLATE                1                  2                    3                   4                   5
                 X          Y        X         Y         X         Y         X         Y         X         Y
DOWNCOMER   0.46751              0.46484             0.50460             0.51858             0.53268
               0.4674    0.4856   0.4650    0.4872    0.5017    0.5058    0.5156    0.5190    0.5298    0.5344
               0.4673    0.4856   0.4652    0.4873    0.4991    0.5045    0.5129    0.5180    0.5272    0.5330
               0.4672    0.4855   0.4654    0.4875    0.4967    0.5033    0.5105    0.5174    0.5249    0.5320
               0.4671    0.4855   0.4657    0.4877    0.4946    0.5022    0.5085    0.5170    0.5228    0.5312
               0.4670    0.4854   0.4660    0.4878    0.4928    0.5012    0.5069    0.5169    0.5212    0.5308
               0.4669    0.4854   0.4662    0.4880    0.4916    0.5006    0.5059    0.5171    0.5201    0.5307
LIQUID FLOW       1000.000          1000.000            700.000             700.000             700.000
VAPOUR FLOW        STRIPPER= 850.000                 RECTIFIER= 850.000
V ENTERING        0.47910           0.48550             0.48760             0.50292             0.51757
V LEAVING.        0.48550           0.48760             0.50292             0.51757             0.53202
V EQUILIBRIUM     0.49067           0.49000             0.51541             0.52968             0.54378
BOILING TEMP.     949084             94.107              93.253              92.781              92.320
PECLET NO          7.781              7.834               5.619               5.664               5.714
BACKFLOW           0.400              0.394               0.746               0.736               0.725
EQ. SLOPE          1.0034             1.0039              0.9978              0.9951              0.9925
LAMBDA             0.8529             0.8533              1.2116              1.2084              1.2051
EOG.               0.5421             0.5421              0.4743              0.4742              0.4741
EMV.               0.5529             0.4668              0.5508              0.5476              0.5512
EMV/EOG.           1.0199             0.8610              1.1613              1.1548              1.1626
R.VOLATILITY.      1.1000             1.1000              1.1000              1.1000              1.1000
MOLAR HOLDUPS      4.0128             4.0211              3.4837              3.4927              3.5029
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                TIME.      FEED       XF       WS         DS       REFLUX
                54.537     300.00    0.40    150.00     150.00      4.667


                      REBOILER            CONDENSER           TOP PRODUCTS
MOLAR HOLDUP          11.64460             2.15163
LIQUID COMP.           0.45193             0.53077
VAPOUR COMP.           0.47372                                CAL.=  79.6183
BOILING TEMP          94.595              91.975              EST.=  52.1967


PLATE                   1                      2                     3                     4                     5
                 X            Y         X           Y         X           Y         X           Y         X           Y
DOWNCOMER     0.46317                 0.46220               0.50138               0.51555               0.52956
               0.4629     0.4814     0.4622     0.4836     0.4984     0.5020     0.5125     0.5156     0.5267     0.5311
               0.4626     0.4813     0.4622     0.4836     0.4957     0.5007     0.5098     0.5147     0.5241     0.5298
               0.4624     0.4811     0.4622     0.4836     0.4933     0.4995     0.5074     0.5140     0.5218     0.5287
               0.4622     0.4810     0.4623     0.4837     0.4911     0.4984     0.5054     0.5136     0.5197     0.5280
               0.4620     0.4810     0.4624     0.4838     0.4893     0.4976     0.5038     0.5135     0.5181     0.5275
               0.4619     0.4809     0.4624     0.4838     0.4880     0.4970     0.5027     0.5137     0.5170     0.5275
LIQUID FLOW      1000.000              1000.000             700.000               700.000               700.000
VAPOUR FLOW        STRIPPER= 850.000                    RECTIFIER= 850.000
V ENTERING        0.47572              0.48112              0.48368               0.49921               0.51417
V LEAVING.        0.48112              0.48368              0.49921               0.51417               0.52878
V EQUILIBRIUM     0.48528              0.48589              0.51180               0.52652               0.54074
BOILING TEMP.     94.268               94.247               93.373                92.885                92.419
PECLET NO          7.781                7.834                5.619                 5.664                 5.714
BACKFLOW           0.400                0.394                0.746                 0.736                 0.725
FR. SLOPE         1.0045               1.0046               0.9984                0.9957                0.9930
LAMBDA            0.8538               0.8539               1.2124                1.2091                1.2058
EOG.              0.5421               0.5421               0.4743                0.4742                0.4741
EMV.              0.5648               0.5364               0.5520                0.5477                0.5498
EMV/EOG.          1.0417               0.9894               1.1639                1.1549                1.1596
R.VOLATILITY.     1.1000               1.1000               1.1000                1.1000                1.1000
MOLAR HOLDUPS     4.0128               4.0211               3.4837                3.4927                3.5029
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                 TIME.      FEED       XF       VS        DS      REFLUX
                 69.137     300.00    0.40    150.00    150.00     4.667

                     REBOILER               CONDENSER          TOP PRODUCTS
MOLAR HOLDUP        11.64460                 2.15163
LIQUID COMP.         0.44829                 0.52759
VAPOUR COMP.         0.47197                                   CAL.=  79.1382
BOILING TEMP        94.724                  92.077             EST.=  52.7559

PLATE                 1                        2                        3                        4                        5
                  X        Y               X        Y               X        Y               X        Y               X        Y
DOWNCOMER      0.45989                  0.45979                  0.49818                  0.51241                  0.52639
                0.4595   0.4780        0.4596   0.4805        0.4952   0.4989        0.5094   0.5125        0.5235   0.5279
                0.4591   0.4778        0.4595   0.4805        0.4925   0.4975        0.5067   0.5115        0.5209   0.5266
                0.4587   0.4776        0.4595   0.4805        0.4900   0.4963        0.5043   0.5108        0.5186   0.5256
                0.4584   0.4774        0.4594   0.4805        0.4878   0.4953        0.5022   0.5104        0.5166   0.5248
                0.4581   0.4772        0.4594   0.4806        0.4860   0.4944        0.5006   0.5103        0.5149   0.5244
                0.4580   0.4771        0.4594   0.4807        0.4847   0.4938        0.4995   0.5105        0.5138   0.5244
LIQUID FLOW       1000.000                 1000.000                 700.000                  700.000                  700.000
VAPOUR FLOW        STRIPPER= 850.000                        RECTIFIER= 850.000
V ENTERING        0.47197                  0.47750                  0.48055                  0.49602                  0.51099
V LEAVING.        0.47750                  0.48055                  0.49602                  0.51099                  0.52560
V EQUILIBRIUM     0.48149                  0.48300                  0.50852                  0.52335                  0.53756
BOILING TEMP.     94.397                   94.346                   93.483                   92.990                   92.523
PECLET NO          7.781                    7.834                    5.619                    5.664                    5.714
BACKFLOW           0.400                    0.394                    0.746                    0.736                    0.725
EQ. SLOPE          1.0053                   1.0052                   0.9991                   0.9963                   0.9936
LAMBDA             0.8545                   0.8544                   1.2131                   1.2098                   1.2065
EOG.               0.5422                   0.5422                   0.4743                   0.4742                   0.4741
EMV.               0.5816                   0.5546                   0.5532                   0.5477                   0.5498
EMV/EOG.           1.0727                   1.0229                   1.1662                   1.1550                   1.1596
R.VOLATILITY.      1.1000                   1.1000                   1.1000                   1.1000                   1.1000
MOLAR HOLDUPS      4.0128                   4.0211                   3.4837                   3.4927                   3.5029
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                  TIME.     FEED      XF       WS        DS       REFLUX
                  63.737    300.00    0.40    150.00    150.00     4.667


                     REBOILER            CONDENSER          TOP PRODUCTS
MOLAR HOLDUP       11.64460               2.15163
LIQUID COMP.        0.44471               0.52446
VAPOUR COMP.        0.46832                                CAL.=  78.6662
BOILING TEMP       94.850                92.178            EST.=  53.2980


PLATE                  1                     2                     3                     4                     5
                   X          Y          X          Y          X          Y          X          Y          X          Y
DOWNCOMER      0.45703               0.45752               0.49503               0.50926               0.52324
                0.4565     0.4747     0.4573     0.4776     0.4921     0.4958     0.5062     0.5094     0.5204     0.5248
                0.4560     0.4745     0.4571     0.4776     0.4893     0.4945     0.5035     0.5084     0.5178     0.5235
                0.4556     0.4742     0.4569     0.4776     0.4869     0.4933     0.5011     0.5077     0.5155     0.5224
                0.4552     0.4740     0.4568     0.4776     0.4847     0.4922     0.4991     0.5073     0.5134     0.5217
                0.4549     0.4738     0.4567     0.4777     0.4828     0.4914     0.4975     0.5072     0.5117     0.5213
                0.4546     0.4737     0.4566     0.4777     0.4816     0.4908     0.4964     0.5074     0.5106     0.5213
LIQUID FLOW        1000.000             1000.000              700.000              700.000              700.000
VAPOUR FLOW          STRIPPER= 850.000               RECTIFIER= 850.000
V ENTERING         0.46832              0.47416              0.47764              0.49300              0.50791
V LEAVING.         0.47416              0.47764              0.49300              0.50791              0.52249
V EQUILIBRIUM      0.47822              0.48029              0.50538              0.52021              0.53441
BOILING TEMP.      94.509               94.438               93.588               93.094               92.626
PECLET NO           7.781                7.834                5.619                5.664                5.714
BACKFLOW            0.400                0.394                0.746                0.736                0.725
EQ. SLOPE          1.0059               1.0057               0.9997               0.9969               0.9942
LAMBDA             0.8550               0.8548               1.2139               1.2105               1.2073
EOG.               0.5422               0.5422               0.4743               0.4742               0.4742
EMV.               0.5900               0.5672               0.5537               0.5480               0.5501
EMV/EOG.           1.0882               1.0460               1.1674               1.1556               1.1602
R.VOLATILITY.      1.1000               1.1000               1.1000               1.1000               1.1000
MOLAR HOLDUPS      4.0128               4.0211               3.4837               3.4927               3.5029
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                   TIME.      FEED      XF       WS        DS       REFLUX
                  98.337     300.00    0.40    150.00    150.00     4.667


                      REBOILER                 CONDENSER              TOP PRODUCTS
MOLAR HOLDUP          11.64460                  2.15163
LIQUID COMP.          0.44127                   0.52138
VAPOUR COMP.          0.46484                                      CAL.=  78.2046
BOILING TEMP          94.971                    92.277             EST.=  53.8160


PLATE                    1                   2                   3                   4                   5
                    X         Y         X         Y         X         Y         X         Y         X         Y
DOWNCOMER       0.45438             0.45535             0.49194             0.50615             0.52014
                  0.4533    0.4717    0.4550    0.4749    0.4890    0.4929    0.5031    0.5064    0.5173    0.5218
                  0.4532    0.4714    0.4547    0.4749    0.4863    0.4916    0.5004    0.5054    0.5147    0.5204
                  0.4527    0.4711    0.4545    0.4748    0.4838    0.4904    0.4981    0.5048    0.5124    0.5194
                  0.4523    0.4708    0.4543    0.4748    0.4816    0.4893    0.4960    0.5043    0.5103    0.5187
                  0.4519    0.4706    0.4541    0.4749    0.4798    0.4885    0.4944    0.5042    0.5086    0.5182
                  0.4516    0.4705    0.4540    0.4750    0.4786    0.4879    0.4933    0.5044    0.5075    0.5182
LIQUID FLOW         1000.000            1000.000             700.000             700.000             700.000
VAPOUR FLOW          STRIPPER= 850.000                   RECTIFIER= 850.000
V ENTERING          0.46484             0.47101             0.47488             0.49010             0.50493
V LEAVING.          0.47101             0.47488             0.49010             0.50493             0.51945
V EQUILIBRIUM       0.47520             0.47772             0.50236             0.51713             0.53132
BOILING TEMP.       94.613              94.526              93.690              93.196              92.727
PECLET NO           7.781               7.834               5.619               5.664               5.714
BACKFLOW            0.400               0.394               0.746               0.736               0.725
EQ. SLOPE           1.0065              1.0062              1.0003              0.9975              0.9948
LAMBDA              0.8555              0.8552              1.2146              1.2112              1.2080
EOG.                0.5422              0.5422              0.4744              0.4743              0.4742
EMV.                0.5955              0.5761              0.5541              0.5484              0.5505
EMV/EOG.            1.0983              1.0626              1.1681              1.1564              1.1609
R.VOLATILITY.       1.1000              1.1000              1.1000              1.1000              1.1000
MOLAR HOLDUPS       4.0128              4.0211              3.4837              3.4927              3.5029
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               TIME.      FEED      XF       WS        DS      REFLUX
              120.236    300.00    0.40    150.00    150.00    4.667


                  REBOILER              CONDENSER            TOP PRODUCTS
MOLAR HOLDUP      11.64460               2.15163
LIQUID COMP.       0.43639               0.51692
VAPOUR COMP.       0.45992                                  CAL.=  77.5344
BOILING TEMP       95.141                92.422             EST.=  54.5467


PLATE                  1                     2                     3                     4                     5
                  X          Y          X          Y          X          Y          X          Y          X          Y
DOWNCOMER      0.45066               0.45223               0.48747               0.50163               0.51565
               0.4500     0.4673     0.4518     0.4711     0.4845     0.4887     0.4986     0.5021     0.5128     0.5174
               0.4493     0.4670     0.4514     0.4710     0.4818     0.4874     0.4959     0.5011     0.5102     0.5160
               0.4487     0.4667     0.4511     0.4709     0.4794     0.4862     0.4936     0.5005     0.5079     0.5150
               0.4482     0.4664     0.4508     0.4709     0.4772     0.4852     0.4915     0.5000     0.5058     0.5143
               0.4477     0.4661     0.4506     0.4709     0.4754     0.4844     0.4899     0.4999     0.5041     0.5138
               0.4473     0.4659     0.4504     0.4710     0.4742     0.4839     0.4889     0.5001     0.5030     0.5138
LIQUID FLOW        1000.000              1000.000               700.000               700.000               700.000
VAPOUR FLOW         STRIPPER= 850.000                     RECTIFIER= 850.000
V ENTERING         0.45992               0.46656               0.47096               0.48596               0.50062
V LEAVING.         0.46656               0.47096               0.48596               0.50062               0.51506
V EQUILIBRIUM      0.47095               0.47407               0.49800               0.51267               0.52682
BOILING TEMP.      94.759                94.652                93.836                93.344                92.875
PECLET NO           7.781                 7.834                 5.619                 5.464                 5.714
BACKFLOW            0.400                 0.394                 0.746                 0.736                 0.725
EQ. SLOPE          1.0072                1.0068                1.0011                0.9984                0.9957
LAMBDA             0.8562                0.8558                1.2156                1.2123                1.2090
EOG.               0.5423                0.5423                0.4744                0.4743                0.4742
EMV.               0.6016                0.5857                0.5547                0.5491                0.5511
EMV/EOG.           1.1094                1.0801                1.1692                1.1578                1.1623
R.VOLATILITY.      1.1000                1.1000                1.1000                1.1000                1.1000
MOLAR HOLDUPS      4.0128                4.0211                3.4837                3.4927                3.5029
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                 TIME.       FEED       XF       WS        DS      REFLUX
                136.736     300.00    0.40    150.00    150.00     4.667


                  REBOILER            CONDENSER          TOP PRODUCTS
MOLAR HOLDUP     11.64460              2.15163
LIQUID COMP.      0.42900              0.50989
VAPOUR COMP.      0.45249                                CAL.=  76.5600
BOILING TEMP     95.401               92.633             EST.=  55.6501


PLATE               1                    2                    3                    4                    5
                X        Y          X        Y          X        Y          X        Y          X        Y
DOWNCOMER  0.44495             0.44737             0.48042             0.49450             0.50856
             0.4441   0.4608   0.4468   0.4652   0.4775   0.4822   0.4915   0.4955   0.5057   0.5105
             0.4433   0.4603   0.4462   0.4650   0.4749   0.4809   0.4889   0.4945   0.5031   0.5092
             0.4426   0.4599   0.4458   0.4649   0.4725   0.4798   0.4865   0.4938   0.5008   0.5081
             0.4419   0.4596   0.4454   0.4649   0.4704   0.4788   0.4845   0.4933   0.4987   0.5074
             0.4413   0.4592   0.4450   0.4649   0.4686   0.4780   0.4829   0.4932   0.4970   0.5070
             0.4409   0.4590   0.4448   0.4649   0.4674   0.4775   0.4818   0.4933   0.4959   0.5070
LIQUID FLOW      1000.000           1000.000            700.000            700.000            700.000
VAPOUR FLOW        STRIPPER= 850.000                  RECTIFIER= 850.000
V ENTERING        0.45249            0.45980            0.46496            0.47954            0.49392
V LEAVING.        0.45980            0.46496            0.47954            0.49392            0.50819
V EQUILIBRIUM     0.46450            0.46845            0.49120            0.50567            0.51976
BOILING TEMP.     94.982             94.846             94.066             93.578             93.108
PECLET NO          7.781              7.834              5.619              5.664              5.714
BACKFLOW           0.400              0.394              0.746              0.736              0.725
EQ. SLOPE         1.0084             1.0079             1.0024             0.9997             0.9970
LAMBDA            0.8572             0.8567             1.2173             1.2139             1.2107
EOG.              0.5424             0.5423             0.4744             0.4743             0.4743
EMV.              0.6089             0.5964             0.5556             0.5503             0.5524
EMV/EOG.          1.1226             1.0997             1.1710             1.1602             1.1647
R.VOLATILITY.     1.1000             1.1000             1.1000             1.1000             1.1000
MOLAR HOLDUPS     4.0128             4.0211             3.4837             3.4927             3.5029
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                TIME.      FEED       XF       VS        DS      REFLUX
               229.736    300.00    0.40    150.00    150.00     4.667


                 REBOILER               CONDENSER          TOP PRODUCTS
MOLAR HOLDUP     11.64460                2.15163
LIQUID COMP.      0.41640                0.49739
VAPOUR COMP.      0.43973                                CAL.=  74.6081
BOILING TEMP     95.850                 93.061           EST.=  57.5402


PLATE                 1                    2                    3                    4                    5
                X          Y          X          Y          X          Y          X          Y          X          Y
DOWNCOMER  0.43499               0.43881               0.46793               0.48181               0.49593
             0.4339     0.4493     0.4379     0.4549     0.4651     0.4708     0.4789     0.4836     0.4931     0.4982
             0.4329     0.4488     0.4372     0.4546     0.4626     0.4695     0.4763     0.4826     0.4905     0.4969
             0.4320     0.4483     0.4365     0.4544     0.4603     0.4684     0.4740     0.4819     0.4881     0.4959
             0.4311     0.4478     0.4359     0.4543     0.4583     0.4675     0.4720     0.4814     0.4861     0.4952
             0.4303     0.4474     0.4354     0.4543     0.4566     0.4668     0.4704     0.4812     0.4844     0.4948
             0.4298     0.4471     0.4350     0.4544     0.4554     0.4664     0.4694     0.4814     0.4833     0.4948
LIQUID FLOW       1000.000             1000.000              700.000              700.000              700.000
VAPOUR FLOW        STRIPPER= 850.000               RECTIFIER= 850.000
V ENTERING        0.43973              0.44811              0.45449              0.46824              0.48203
V LEAVING.        0.44811              0.45449              0.46824              0.48203              0.49596
V EQUILIBRIUM     0.45328              0.45858              0.47916              0.49318              0.50714
BOILING TEMP.     95.373               95.188               94.477               93.999               93.529
PECLET NO          7.781                7.834                5.619                5.664                5.714
BACKFLOW           0.400                0.394                0.746                0.736                0.725
EQ. SLOPE         1.0099               1.0092               1.0042               1.0015               0.9988
LAMBDA            0.8584               0.8578               1.2193               1.2161               1.2128
EMG.              0.5424               0.5424               0.4745               0.4744               0.4743
EMV.              0.6185               0.6093               0.5574               0.5528               0.5548
EMV/EMG.          1.1402               1.1234               1.1747               1.1653               1.1698
R.VOLATILITY.     1.1000               1.1000               1.1000               1.1000               1.1000
MOLAR HOLDUPS     4.0128               4.0211               3.4837               3.4927               3.5029
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                TIME.     FEED.      XF      US        DS       REFLUX
               346.536   300.00    0.40    150.00    150.00     4.667


                 REBOILER            CONDENSER          TOP PRODUCTS
MOLAR HOLDUP     11.64460             2.15163
LIQUID COMP.      0.40113             0.48160
VAPOUR COMP.      0.42423                               CAL.=  72.2406
BOILING TEMP     96.401              93.587             EST.=  59.8299


PLATE               1                   2                   3                   4                   5
               X         Y         X         Y         X         Y         X         Y         X         Y
DOWNCOMER  0.42265             0.42810             0.45221             0.46578             0.47999
             0.4213    0.4353    0.4269    0.4422    0.4496    0.4565    0.4630    0.4688    0.4771    0.4828
             0.4200    0.4346    0.4259    0.4418    0.4472    0.4553    0.4605    0.4677    0.4745    0.4815
             0.4189    0.4340    0.4249    0.4415    0.4450    0.4543    0.4582    0.4670    0.4722    0.4805
             0.4178    0.4334    0.4241    0.4413    0.4431    0.4535    0.4563    0.4665    0.4701    0.4798
             0.4168    0.4329    0.4234    0.4412    0.4415    0.4529    0.4547    0.4663    0.4685    0.4794
             0.4161    0.4325    0.4229    0.4413    0.4404    0.4526    0.4537    0.4664    0.4674    0.4794
LIQUID FLOW     1000.000           1000.000            700.000             700.000             700.000
VAPOUR FLOW        STRIPPER= 850.000               RECTIFIER= 850.000
V ENTERING       0.42423            0.43379            0.44156             0.45416             0.46712
V LEAVING.       0.43379            0.44156            0.45416             0.46712             0.48055
V EQUILIBRIUM    0.43947            0.44632            0.46405             0.47744             0.49117
BOILING TEMP.    95.850             95.617             94.978              94.536              94.067
PECLET NO         7.781              7.834              5.619               5.664               5.714
BACKFLOW          0.400              0.394              0.746               0.736               0.725
EQ. SLOPE         1.0127             1.0116             1.0074              1.0047              1.0021
LAMBDA            0.8603             0.8598             1.2232              1.2201              1.2158
EMG.              0.5426             0.5425             0.4746              0.4745              0.4744
EMV.              0.6275             0.6203             0.5602              0.5566              0.5585
EMV/EMG.          1.1563             1.1433             1.1803              1.1730              1.1773
R.VOLATILITY.     1.1000             1.1000             1.1000              1.1000              1.1000
MOLAR HOLDUPS     4.0128             4.0211             3.4837              3.4927              3.5029
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                 TIME.      FEED      XF       WS       DS     REFLUX
                434.136    300.00    0.40    150.00   150.00    4.667

                   REBOILER            CONDENSER        TOP PRODUCTS
MOLAR HOLDUP       11.64460             2.15163
LIQUID COMP.        0.39256             0.47259
VAPOUR COMP.        0.41551                             CAL.=  70.8881
BOILING TEMP       96.714              93.891           EST.=  61.1158


PLATE                1                   2                   3                   4                   5
               X          Y        X          Y        X          Y        X          Y        X          Y
DOWNCOMER  0.41567            0.42202            0.44326            0.45664            0.47088
            0.4142    0.4274   0.4207    0.4350   0.4407    0.4483   0.4539    0.4603   0.4680    0.4740
            0.4123    0.4266   0.4194    0.4345   0.4384    0.4472   0.4514    0.4593   0.4654    0.4727
            0.4115    0.4259   0.4184    0.4342   0.4363    0.4463   0.4492    0.4585   0.4631    0.4717
            0.4103    0.4253   0.4174    0.4340   0.4345    0.4455   0.4473    0.4580   0.4610    0.4710
            0.4092    0.4247   0.4166    0.4339   0.4329    0.4449   0.4458    0.4578   0.4594    0.4706
            0.4084    0.4243   0.4160    0.4339   0.4319    0.4447   0.4448    0.4579   0.4583    0.4706
LIQUID FLOW     1000.000          1000.000            700.000            700.000            700.000
VAPOUR FLOW       STRIPPER= 850.000                RECTIFIER= 850.000
V ENTERING       0.41551           0.42571            0.43424            0.44616            0.45862
V LEAVING.       0.42571           0.43424            0.44616            0.45862            0.47175
V EQUILIBRIUM    0.43165           0.43936            0.45544            0.46845            0.48204
BOILING TEMP.    96.136            95.863             95.297             94.845             94.378
PECLET NO         7.781             7.834              5.619              5.664              5.714
BACKFLOW          0.400             0.394              0.746              0.736              0.725
EQ. SLOPE        1.0142            1.0129             1.0092             1.0066             1.0040
LAMBDA           0.8621            0.8610             1.2254             1.2223             1.2191
EOG.             0.5427            0.5426             0.4747             0.4746             0.4745
EMV.             0.6318            0.6251             0.5620             0.5590             0.5608
EMV/EOG.         1.1642            1.1519             1.1841             1.1779             1.1819
R.VOLATILITY.    1.1000            1.1000             1.1000             1.1000             1.1000
MOLAR HOLDUPS    4.0128            4.0211             3.4837             3.4927             3.5029
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                 TIME.      FEED       XF        WS         DS      REFLUX
                550.935    300.00    0.40    150.00    150.00     4.667


                     REBOILER              CONDENSER           TOP PRODUCTS
MOLAR HOLDUP         11.64460               2.15163
LIQUID COMP.          0.38393               0.46343
VAPOUR COMP.          0.40671                                CAL.=  69.5143
BOILING TEMP          97.033               94.203            EST.=  62.4105


PLATE                  1                    2                    3                    4                    5
                   X         Y          X         Y          X         Y          X         Y          X         Y
DOWNCOMER      0.40360              0.41586              0.43419              0.44737              0.46162
                 0.4069    0.4194    0.4143    0.4277    0.4317    0.4401    0.4447    0.4517    0.4588    0.4650
                 0.4054    0.4186    0.4130    0.4272    0.4295    0.4390    0.4423    0.4507    0.4562    0.4637
                 0.4040    0.4178    0.4117    0.4268    0.4275    0.4381    0.4401    0.4499    0.4538    0.4627
                 0.4027    0.4171    0.4107    0.4265    0.4257    0.4374    0.4382    0.4494    0.4518    0.4620
                 0.4015    0.4164    0.4097    0.4264    0.4243    0.4369    0.4367    0.4491    0.4502    0.4617
                 0.4007    0.4160    0.4091    0.4264    0.4233    0.4367    0.4358    0.4492    0.4491    0.4617
LIQUID FLOW        1000.000             1000.000             700.000              700.000              700.000
VAPOUR FLOW          STRIPPER= 850.000                  RECTIFIER= 850.000
V ENTERING          0.40671              0.41754              0.42684              0.43806              0.45000
V LEAVING.          0.41754              0.42684              0.43806              0.45000              0.46282
V EQUILIBRIUM       0.42375              0.43232              0.44672              0.45933              0.47276
BOILING TEMP.        96.418               96.112               95.603               95.162               94.697
PECLET NO             7.781                7.834                5.619                5.664                5.714
BACKFLOW              0.400                0.394                0.746                0.736                0.725
EQ. SLOPE            1.0158               1.0143               1.0110               1.0085               1.0059
LAMBDA               0.8634               0.8621               1.2277               1.2247               1.2214
EOG.                 0.5428               0.5427               0.4747               0.4746               0.4745
EMV.                 0.6357               0.6292               0.5641               0.5617               0.5633
EMV/EOG.             1.1712               1.1594               1.1883               1.1834               1.1870
R.VOLATILITY.        1.1000               1.1000               1.1000               1.1000               1.1000
MOLAR HOLDUPS        4.0128               4.0211               3.4837               3.4927               3.5029
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               TIME.      FEED      XF      WS        DS      REFLUX
              667.735    306.00    0.40   150.00    150.00    4.667


                     REBOILER              CONDENSER          TOP PRODUCTS
MOLAR HOLDUP         11.64460               2.15163
LIQUID COMP.          0.37767               0.45674
VAPOUR COMP.          0.40032                                 CAL.=  68.5117
BOILING TEMP         97.265                94.432             EST.=  63.3490


PLATE                 1                     2                     3                     4                     5
                  X         Y          X         Y          X         Y          X         Y          X         Y
DOWNCOMER     0.40347              0.41138              0.42759              0.44061              0.45487
              0.4617    0.4136    0.4097    0.4224    0.4252    0.4341    0.4380    0.4455    0.4520    0.4585
              0.4001    0.4127    0.4082    0.4218    0.4230    0.4331    0.4356    0.4444    0.4494    0.4572
              0.3986    0.4119    0.4069    0.4214    0.4211    0.4322    0.4335    0.4436    0.4471    0.4562
              0.3972    0.4111    0.4057    0.4211    0.4194    0.4315    0.4316    0.4431    0.4451    0.4555
              0.3959    0.4104    0.4048    0.4210    0.4180    0.4311    0.4301    0.4428    0.4434    0.4552
              0.3950    0.4099    0.4041    0.4210    0.4170    0.4309    0.4292    0.4429    0.4424    0.4552
LIQUID FLOW      1000.000             1000.000              700.000              700.000              700.000
VAPOUR FLOW         STRIPPER= 850.000              RECTIFIER= 850.000
V ENTERING        0.40032              0.41161              0.42146              0.43216              0.44372
V LEAVING.        0.41161              0.42146              0.43216              0.44372              0.45630
V EQUILIBRIUM     0.41801              0.42720              0.44037              0.45267              0.46598
BOILING TEMP.     96.624               96.295               95.827               95.394               94.931
PECLET NO          7.781                7.834                5.619                5.664                5.714
BACKFLOW           0.400                0.394                0.746                0.736                0.725
EQ. SLOPE          1.0169               1.0153               1.0124               1.0099               1.0073
LAMBDA             0.8644               0.8630               1.2293               1.2264               1.2231
EOG.               0.5428               0.5427               0.4748               0.4747               0.4746
EMV.               0.6383               0.6319               0.5658               0.5638               0.5652
EMV/EOG.           1.1759               1.1642               1.1917               1.1878               1.1908
R.VOLATILITY.      1.1000               1.1000               1.1000               1.1000               1.1000
MOLAR HOLDUPS      4.0128               4.0211               3.4837               3.4927               3.5029
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              TIME.       FEED       XF       WS        DS       REFLUX
             755.335    300.00     0.40    150.00    150.00      4.667


                     REBOILER              CONDENSER          TOP PRODUCTS
MOLAR HOLDUP        11.64460                2.15163
LIQUID COMP.         0.37414                0.45295
VAPOUR COMP.         0.39671                                CAL.=  67.9428
BOILING TEMP        97.397                 94.563           EST.=  63.8793


PLATE                   1                     2                     3                     4                     5
                 X            Y         X          Y         X          Y          X          Y          X          Y
DOWNCOMER  0.40056              0.40884              0.42385              0.43677              0.45104
             0.3987    0.4103    0.4071    0.4194    0.4215    0.4307    0.4342    0.4419    0.4482    0.4548
             0.3971    0.4094    0.4056    0.4188    0.4194    0.4297    0.4318    0.4408    0.4456    0.4535
             0.3955    0.4085    0.4042    0.4184    0.4174    0.4288    0.4297    0.4401    0.4433    0.4525
             0.3940    0.4077    0.4030    0.4181    0.4158    0.4282    0.4279    0.4395    0.4413    0.4518
             0.3928    0.4070    0.4019    0.4179    0.4144    0.4278    0.4264    0.4393    0.4396    0.4515
             0.3918    0.4065    0.4012    0.4179    0.4135    0.4277    0.4254    0.4393    0.4385    0.4515
LIQUID FLOW       1000.000            1000.000             700.000              700.000              700.000
VAPOUR FLOW         STRIPPER= 850.000                 RECTIFIER= 850.000
V ENTERING          0.39671             0.40826             0.41841             0.42881             0.44016
V LEAVING.          0.40826             0.41841             0.42881             0.44016             0.45260
V EQUILIBRIUM       0.41476             0.42429             0.43677             0.44890             0.46213
BOILING TEMP.       96.741              96.399              95.955              95.527              95.065
PECLET NO            7.781               7.834               5.619               5.664               5.714
BACKFLOW             0.400               0.394               0.746               0.736               0.725
EQ. SLOPE           1.0176              1.0159              1.0132              1.0107              1.0081
LAMBDA              0.8649              0.8635              1.2303              1.2273              1.2241
EOG.                0.5429              0.5428              0.4748              0.4747              0.4746
EMV.                0.6397              0.6333              0.5668              0.5651              0.5663
EMV/EOG.            1.1784              1.1667              1.1938              1.1904              1.1931
R.VOLATILITY.       1.1000              1.1000              1.1000              1.1000              1.1000
MOLAR HOLDUPS       4.0128              4.0211              3.4837              3.4927              3.5029
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Appendix G

Unsteady-State Simulation Results for a Feed Composition Change: Ternary System

The simulation results for a step change in the feed composition is given. The operating conditions also given in the results. The step change made was:-

Component	XFS (SS)	XFS (US)
1	0.35	0.45
2	0.35	0.3
3	0.3	0.25

Not all the simulation results are given as there were too many, but results at suitable intervals were selected to give a comprehensive list showing the trends of the responses.

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          DISTILLATION COLUMN DIMENSIONS.

                WEIR AND DOWNCOMER.

DIA.FT     HT.INS     LENG.FT     SEC.AREA     APRON
  7.0        2.25      5.3900       4.747         1.0


      PERFORATIONS               TRAY AREAS.
DIA.INS    PITCH     AREA      NET     ACTIVE    GROSS
0.1875      0.5     3.8485   33.737   28.990   38.485


  DENS     MOL WT     VISC     B.TEMP      R.VOL
  0.700     60.00     0.270     80.00      3.00
  0.900     90.00     0.270    100.00      2.00
  1.100    120.00     0.270    120.00      1.00
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      TIME   FEED   BOTTOMS    TOPS   R.RATIO  RB.HOLDUP  C.HOLDUP
      0.000  300.0   150.0    150.0    4.667   13.9660    2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.     CTP.
       1         3.00       0.35     0.1325  0.5675  85.129   85.126
       2         2.00       0.35     0.3476  0.3524  52.862   52.865
       3         1.00       0.30     0.5199  0.0801  12.009   12.009
PLATE   HOLDUP  TEMP.   PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1     3.4561  101.71     4.95      0.9143     1000.00       850.00
  2     3.6088   99.25     5.52      0.7690     1000.00       850.00
  3     3.3529   96.50     4.49      1.0574      700.00       850.00
  4     3.5219   94.09     5.00      0.9006      700.00       850.00
  5     3.6982   91.79     5.54      0.7636      700.00       850.00
COMPONENT= 1         LIQUID COMPOSITION                    VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2820 .2704 .2596 .2499 .2412 .2341 .2294 .3348 .3285 .3226 .3173 .3129 .3099 .3210 0.8758
    2    .3438 .3321 .3214 .3119 .3037 .2970 .2927 .3988 .3943 .3912 .3891 .3882 .3888 .3917 0.6055
    3    .4083 .3959 .3844 .3742 .3654 .3584 .3541 .4560 .4508 .4469 .4441 .4427 .4432 .4473 0.5338
    4    .4812 .4677 .4552 .4441 .4344 .4265 .4215 .5149 .5096 .5058 .5033 .5022 .5029 .5064 0.5300
    5    .5568 .5427 .5297 .5179 .5075 .4989 .4933 .5763 .5709 .5670 .5643 .5630 .5637 .5675 0.5271
COMPONENT= 2         LIQUID COMPOSITION                    VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .4205 .4206 .4205 .4202 .4196 .4190 .4186 .4279 .4304 .4326 .4346 .4361 .4372 .4331 0.9713
    2    .4110 .4137 .4159 .4177 .4190 .4198 .4203 .4151 .4177 .4198 .4212 .4221 .4221 .4197 0.9595
    3    .4230 .4262 .4289 .4309 .4325 .4335 .4341 .4034 .4062 .4082 .4095 .4100 .4096 .4078 0.8232
    4    .3967 .4023 .4074 .4117 .4153 .4180 .4197 .3785 .3821 .3849 .3868 .3878 .3877 .3846 0.6287
    5    .3584 .3662 .3732 .3794 .3847 .3889 .3915 .3461 .3499 .3528 .3547 .3557 .3553 .3524 0.5818
COMPONENT= 3         LIQUID COMPOSITION                    VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2975 .3090 .3198 .3300 .3392 .3469 .3521 .2373 .2412 .2448 .2481 .2510 .2529 .2459 1.3111
    2    .2452 .2542 .2627 .2704 .2773 .2831 .2870 .1862 .1879 .1890 .1896 .1898 .1892 .1886 0.5572
    3    .1686 .1779 .1867 .1948 .2021 .2080 .2119 .1406 .1430 .1449 .1464 .1473 .1472 .1449 0.4873
    4    .1222 .1300 .1374 .1442 .1504 .1555 .1588 .1067 .1083 .1093 .1099 .1100 .1094 .1089 0.4813
    5    .0848 .0911 .0970 .1027 .1078 .1122 .1151 .0776 .0791 .0803 .0810 .0813 .0810 .0801 0.4771
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      TIME    FEED   BOTTOMS    TOPS   R.RATIO  RB.HOLDUP  C.HOLDUP
      3.498  300.0    150.0    150.0    4.667    13.9660    2.7425
     COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.     CTP.
           1      3.00      0.45     0.1325  0.5675 115.129   85.129
           2      2.00      0.30     0.3476  0.3524  37.862   52.863
           3      1.00      0.25     0.5199  0.0800  -2.991   12.008
  PLATE   HOLDUP  TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
    1     3.4561  101.71    4.95      0.9143     1000.00       850.00
    2     3.6088   99.16    5.52      0.7690     1000.00       850.00
    3     3.3529   96.50    4.49      1.0574      700.00       850.00
    4     3.5219   94.09    5.00      0.9006      700.00       850.00
    5     3.6982   91.79    5.54      0.7636      700.00       850.00
  COMPONENT= 1        LIQUID COMPOSITION                    VAPOUR COMPOSITION          MEAN  EMV.
   PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
     1    .2824 .2706 .2598 .2499 .2412 .2341 .2294 .3348 .3285 .3226 .3173 .3129 .3099 .3210 0.8758
     2    .3609 .3439 .3292 .3168 .3067 .2988 .2939 .4034 .3971 .3927 .3899 .3886 .3890 .3935 0.6176
     3    .4083 .3959 .3845 .3743 .3655 .3586 .3543 .4562 .4510 .4473 .4450 .4442 .4457 .4482 0.5346
     4    .4812 .4678 .4553 .4441 .4344 .4266 .4216 .5163 .5104 .5062 .5035 .5023 .5030 .5070 0.5306
     5    .5568 .5427 .5297 .5179 .5075 .4990 .4934 .5763 .5710 .5671 .5645 .5634 .5644 .5678 0.5273
  COMPONENT= 2        LIQUID COMPOSITION                    VAPOUR COMPOSITION          MEAN  EMV.
   PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
     1    .4203 .4205 .4205 .4201 .4196 .4190 .4186 .4278 .4304 .4326 .4346 .4361 .4372 .4331 0.9712
     2    .4024 .4078 .4121 .4153 .4175 .4190 .4197 .4117 .4158 .4187 .4207 .4217 .4219 .4184 1.0228
     3    .4230 .4262 .4288 .4309 .4324 .4334 .4339 .4033 .4060 .4079 .4089 .4090 .4078 .4071 0.8522
     4    .3966 .4023 .4073 .4116 .4152 .4180 .4197 .3775 .3815 .3846 .3866 .3877 .3876 .3843 0.6318
     5    .3584 .3662 .3732 .3794 .3847 .3889 .3915 .3461 .3499 .3527 .3546 .3554 .3548 .3522 0.5824
  COMPONENT= 3        LIQUID COMPOSITION                    VAPOUR COMPOSITION          MEAN  EMV.
   PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
     1    .2973 .3089 .3198 .3299 .3392 .3469 .3521 .2373 .2412 .2448 .2481 .2510 .2529 .2459 1.3111
     2    .2367 .2482 .2587 .2679 .2758 .2822 .2864 .1848 .1871 .1886 .1894 .1897 .1891 .1881 0.5611
     3    .1686 .1779 .1867 .1948 .2021 .2080 .2118 .1405 .1429 .1448 .1461 .1468 .1465 .1446 0.4875
     4    .1222 .1300 .1374 .1442 .1504 .1555 .1588 .1062 .1080 .1092 .1098 .1100 .1094 .1088 0.4815
     5    .0848 .0911 .0970 .1027 .1078 .1122 .1151 .0776 .0791 .0802 .0809 .0812 .0808 .0800 0.4771
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    TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
   12.134  300.0    150.0     150.0     4.667   13.9660    2.7425
  COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.     ETP.     CTP.
      1        3.00      0.45      0.1326  0.5682 115.120   85.187
      2        2.00      0.30      0.3475  0.3520  37.866   52.823
      3        1.00      0.25      0.5199  0.0799  -2.986   11.991
PLATE  HOLDUP  TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561  101.64     4.95      0.9143     1000.00      850.00
  2    3.6088   98.80     5.52      0.7690     1000.00      850.00
  3    3.3529   96.47     4.49      1.0574      700.00      850.00
  4    3.5219   94.07     5.00      0.9006      700.00      850.00
  5    3.6982   91.78     5.54      0.7636      700.00      850.00
COMPONENT= 1         LIQUID COMPOSITION                          VAPOUR COMPOSITION         MEAN   EMV.
 PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .2898 .2766 .2646 .2538 .2443 .2365 .2314 .3373 .3304 .3241 .3184 .3137 .3106 .3224 0.8766
   2    .3711 .3563 .3427 .3304 .3196 .3107 .3049 .4114 .4054 .4007 .3973 .3953 .3953 .4009 0.6235
   3    .4090 .3967 .3855 .3755 .3670 .3603 .3561 .4597 .4549 .4517 .4497 .4493 .4507 .4527 0.5365
   4    .4818 .4685 .4563 .4451 .4354 .4275 .4225 .5192 .5134 .5091 .5061 .5046 .5052 .5096 0.5336
   5    .5572 .5431 .5301 .5184 .5080 .4996 .4940 .5776 .5723 .5686 .5662 .5653 .5662 .5694 0.5282
COMPONENT= 2         LIQUID COMPOSITION                          VAPOUR COMPOSITION         MEAN   EMV.
 PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .4169 .4178 .4183 .4184 .4183 .4180 .4177 .4261 .4291 .4316 .4338 .4356 .4367 .4321 0.9709
   2    .3973 .4018 .4057 .4089 .4116 .4136 .4148 .4061 .4100 .4131 .4155 .4171 .4176 .4132 0.9406
   3    .4226 .4256 .4281 .4300 .4314 .4322 .4326 .4008 .4033 .4049 .4055 .4054 .4042 .4040 0.9935
   4    .3962 .4017 .4066 .4109 .4145 .4173 .4190 .3754 .3794 .3825 .3848 .3861 .3861 .3824 0.6458
   5    .3581 .3659 .3729 .3791 .3843 .3884 .3911 .3452 .3489 .3516 .3534 .3541 .3535 .3511 0.5849
COMPONENT= 3         LIQUID COMPOSITION                          VAPOUR COMPOSITION         MEAN   EMV.
 PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .2933 .3056 .3171 .3278 .3375 .3455 .3509 .2365 .2405 .2443 .2478 .2507 .2527 .2454 1.3119
   2    .2316 .2419 .2516 .2606 .2688 .2757 .2803 .1825 .1846 .1862 .1872 .1876 .1872 .1859 0.5632
   3    .1684 .1777 .1864 .1944 .2016 .2075 .2113 .1394 .1418 .1435 .1447 .1453 .1451 .1433 0.4880
   4    .1220 .1297 .1371 .1439 .1501 .1552 .1585 .1054 .1072 .1084 .1091 .1093 .1087 .1080 0.4822
   5    .0847 .0909 .0969 .1025 .1077 .1120 .1149 .0772 .0787 .0798 .0804 .0807 .0803 .0795 0.4775
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      TIME   FEED   BOTTOMS   TOPS   R.RATIO  RB.HOLDUP  C.HOLDUP
     19.036  300.0   150.0    150.0    4.667   13.9660    2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.    CTP.
       1        3.00      0.45     0.1331  0.5690 115.077   85.282
       2        2.00      0.30     0.3473  0.3514  37.884   52.755
       3        1.00      0.25     0.5196  0.0796  -2.961   11.963
PLATE  HOLDUP  TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561  101.52     4.95      0.9143     1000.00       850.00
  2    3.6088   98.62     5.52      0.7690     1000.00       850.00
  3    3.3529   96.44     4.49      1.0574      700.00       850.00
  4    3.5219   94.05     5.00      0.9006      700.00       850.00
  5    3.6982   91.76     5.54      0.7636      700.00       850.00
COMPONENT= 1        LIQUID COMPOSITION                      VAPOUR COMPOSITION         MEAN  EMV.
  PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
    1   .2966 .2829 .2704 .2590 .2489 .2406 .2352 .3409 .3336 .3269 .3209 .3159 .3126 .3252 0.8757
    2   .3741 .3600 .3472 .3356 .3255 .3171 .3116 .4146 .4092 .4052 .4023 .4007 .4010 .4055 0.6136
    3   .4102 .3981 .3871 .3772 .3688 .3621 .3580 .4633 .4584 .4550 .4529 .4521 .4532 .4558 0.5376
    4   .4825 .4694 .4572 .4462 .4366 .4288 .4238 .5208 .5153 .5112 .5084 .5070 .5076 .5117 0.5344
    5   .5578 .5438 .5309 .5191 .5088 .5003 .4948 .5791 .5738 .5701 .5676 .5665 .5674 .5707 0.5290
COMPONENT= 2        LIQUID COMPOSITION                      VAPOUR COMPOSITION         MEAN  EMV.
  PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
    1   .4140 .4152 .4159 .4163 .4164 .4163 .4162 .4237 .4269 .4297 .4322 .4341 .4354 .4303 0.9717
    2   .3956 .3999 .4035 .4065 .4089 .4108 .4118 .4039 .4073 .4101 .4122 .4135 .4137 .4101 0.8618
    3   .4218 .4247 .4270 .4289 .4302 .4310 .4313 .3984 .4009 .4025 .4034 .4034 .4024 .4018 1.1028
    4   .3957 .4011 .4060 .4102 .4137 .4164 .4181 .3742 .3780 .3810 .3832 .3844 .3844 .3809 0.6514
    5   .3577 .3654 .3724 .3786 .3838 .3879 .3905 .3441 .3479 .3506 .3524 .3531 .3526 .3501 0.5869
COMPONENT= 3        LIQUID COMPOSITION                      VAPOUR COMPOSITION         MEAN  EMV.
  PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
    1   .2894 .3019 .3137 .3247 .3347 .3430 .3486 .2353 .2394 .2433 .2469 .2499 .2520 .2445 1.3159
    2   .2303 .2401 .2493 .2578 .2656 .2721 .2765 .1815 .1834 .1847 .1855 .1858 .1852 .1844 0.5593
    3   .1680 .1772 .1859 .1939 .2010 .2069 .2107 .1383 .1407 .1425 .1438 .1445 .1444 .1423 0.4883
    4   .1218 .1295 .1368 .1436 .1497 .1548 .1581 .1050 .1067 .1078 .1084 .1086 .1080 .1074 0.4825
    5   .0845 .0907 .0967 .1023 .1074 .1118 .1147 .0768 .0783 .0794 .0801 .0804 .0800 .0791 0.4777
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     TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
    27.874  300.0    150.0     150.0     4.667    13.9660    2.7425
   COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.     CTP.
       1        3.00      0.45     0.1344  0.5704 114.968   85.434
       2        2.00      0.30     0.3468  0.3503  37.926   52.645
       3        1.00      0.25     0.5187  0.0792  -2.894   11.921
 PLATE  HOLDUP  TEMP.   PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
   1    3.4561  101.35     4.95      0.9143     1000.00      850.00
   2    3.6088   98.50     5.52      0.7690     1000.00      850.00
   3    3.3529   96.39     4.49      1.0574      700.00      850.00
   4    3.5219   94.02     5.00      0.9006      700.00      850.00
   5    3.6982   91.74     5.54      0.7636      700.00      850.00
 COMPONENT= 1         LIQUID COMPOSITION                       VAPOUR COMPOSITION             MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3023 .2887 .2762 .2648 .2547 .2463 .2408 .3449 .3376 .3308 .3246 .3195 .3161 .3289 0.8728
    2    .3763 .3625 .3499 .3387 .3290 .3211 .3159 .4176 .4125 .4089 .4064 .4052 .4058 .4094 0.6064
    3    .4123 .4004 .3896 .3799 .3715 .3649 .3608 .4667 .4617 .4582 .4558 .4549 .4558 .4589 0.5380
    4    .4838 .4708 .4588 .4479 .4384 .4307 .4258 .5227 .5173 .5134 .5107 .5094 .5101 .5139 0.5342
    5    .5591 .5452 .5323 .5205 .5102 .5017 .4962 .5808 .5755 .5717 .5691 .5680 .5687 .5723 0.5298
 COMPONENT= 2         LIQUID COMPOSITION                       VAPOUR COMPOSITION             MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .4116 .4128 .4136 .4141 .4143 .4143 .4142 .4212 .4244 .4273 .4299 .4319 .4333 .4280 0.9738
    2    .3943 .3985 .4020 .4049 .4072 .4089 .4099 .4019 .4052 .4077 .4095 .4105 .4106 .4075 0.8256
    3    .4203 .4231 .4254 .4271 .4284 .4292 .4295 .3961 .3987 .4004 .4013 .4014 .4005 .3997 1.1634
    4    .3948 .4002 .4049 .4090 .4125 .4151 .4168 .3728 .3766 .3795 .3815 .3827 .3826 .3793 0.6531
    5    .3567 .3645 .3715 .3776 .3828 .3869 .3895 .3429 .3466 .3494 .3512 .3520 .3516 .3489 0.5888
 COMPONENT= 3         LIQUID COMPOSITION                       VAPOUR COMPOSITION             MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2862 .2985 .3102 .3211 .3310 .3394 .3450 .2339 .2380 .2419 .2455 .2486 .2507 .2431 1.3237
    2    .2294 .2390 .2480 .2563 .2638 .2700 .2742 .1805 .1823 .1835 .1841 .1842 .1836 .1830 0.5562
    3    .1674 .1764 .1850 .1930 .2001 .2060 .2097 .1372 .1396 .1415 .1429 .1437 .1437 .1414 0.4885
    4    .1214 .1291 .1363 .1431 .1491 .1542 .1574 .1045 .1061 .1072 .1078 .1079 .1073 .1068 0.4825
    5    .0842 .0903 .0963 .1019 .1070 .1113 .1143 .0763 .0779 .0789 .0797 .0800 .0797 .0787 0.4780
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    TIME    FEED   BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
   38.647  300.0    150.0    150.0     4.667    13.9660    2.7425
   COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.     ETP.     CTP.
       1        3.00      0.45     0.1367  0.5724 114.674   85.706
       2        2.00      0.30     0.3461  0.3488  38.028   52.445
       3        1.00      0.25     0.5172  0.0787  -2.702   11.850
PLATE  HOLDUP   TEMP.   PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561  101.17      4.95      0.9143     1000.00       850.00
  2    3.6088   98.41      5.52      0.7690     1000.00       850.00
  3    3.3529   96.30      4.49      1.0574      700.00       850.00
  4    3.5219   93.96      5.00      0.9006      700.00       850.00
  5    3.6982   91.69      5.54      0.7636      700.00       850.00
COMPONENT= 1        LIQUID COMPOSITION                          VAPOUR COMPOSITION          MEAN  EMV.
 PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .3060 .2926 .2803 .2690 .2590 .2508 .2453 .3496 .3424 .3358 .3298 .3247 .3214 .3339 0.8691
   2    .3787 .3651 .3527 .3415 .3319 .3241 .3190 .4216 .4166 .4131 .4108 .4097 .4103 .4137 0.6037
   3    .4150 .4033 .3926 .3830 .3747 .3682 .3641 .4704 .4655 .4620 .4597 .4588 .4597 .4627 0.5369
   4    .4859 .4729 .4610 .4502 .4408 .4332 .4283 .5256 .5203 .5164 .5138 .5126 .5132 .5170 0.5337
   5    .5610 .5472 .5343 .5226 .5123 .5039 .4983 .5832 .5780 .5741 .5716 .5704 .5711 .5747 0.5303
COMPONENT= 2        LIQUID COMPOSITION                          VAPOUR COMPOSITION          MEAN  EMV.
 PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .4098 .4111 .4119 .4124 .4127 .4127 .4127 .4183 .4214 .4243 .4268 .4288 .4301 .4249 0.9767
   2    .3928 .3969 .4004 .4033 .4056 .4072 .4082 .3993 .4025 .4049 .4066 .4076 .4076 .4048 0.8163
   3    .4185 .4213 .4235 .4252 .4264 .4271 .4275 .3934 .3960 .3976 .3985 .3987 .3978 .3970 1.1394
   4    .3934 .3987 .4034 .4075 .4109 .4135 .4151 .3707 .3744 .3772 .3792 .3804 .3803 .3770 0.6534
   5    .3553 .3630 .3700 .3761 .3813 .3854 .3880 .3410 .3447 .3475 .3493 .3502 .3498 .3471 0.5902
COMPONENT= 3        LIQUID COMPOSITION                          VAPOUR COMPOSITION          MEAN  EMV.
 PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .2842 .2963 .3078 .3185 .3283 .3365 .3420 .2322 .2361 .2399 .2435 .2465 .2485 .2411 1.3345
   2    .2285 .2380 .2469 .2552 .2625 .2687 .2728 .1791 .1809 .1820 .1826 .1827 .1821 .1816 0.5548
   3    .1664 .1754 .1840 .1919 .1989 .2047 .2085 .1361 .1385 .1404 .1418 .1426 .1425 .1403 0.4883
   4    .1208 .1284 .1356 .1423 .1484 .1534 .1566 .1037 .1054 .1064 .1070 .1071 .1065 .1060 0.4823
   5    .0836 .0898 .0957 .1013 .1064 .1107 .1136 .0757 .0773 .0784 .0791 .0794 .0791 .0782 0.4781
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      TIME    FEED   BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
     51.639  300.0   150.0     150.0     4.667   13.9660     2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.     ETP.     CTP.
        1        3.00       0.45     0.1395  0.5750 114.285   86.051
        2        2.00       0.30     0.3453  0.3469  38.146   52.183
        3        1.00       0.25     0.5152  0.0781  -2.431   11.766
 PLATE  HOLDUP  TEMP.   PECLET NO.   BACKFLOW  LIQUID FLOW  VAPOUR FLOW
   1    3.4561  101.05     4.95       0.9143     1000.00        850.00
   2    3.6088   98.32     5.52       0.7690     1000.00        850.00
   3    3.3529   96.21     4.49       1.0574      700.00        850.00
   4    3.5219   93.89     5.00       0.9006      700.00        850.00
   5    3.6982   91.63     5.54       0.7636      700.00        850.00
 COMPONENT= 1          LIQUID COMPOSITION                         VAPOUR COMPOSITION          MEAN  EMV.
   PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
     1   .3093 .2960 .2837 .2725 .2625 .2543 .2489 .3536 .3465 .3400 .3340 .3291 .3258 .3382 0.8692
     2   .3814 .3679 .3556 .3446 .3351 .3273 .3222 .4252 .4203 .4168 .4145 .4134 .4140 .4174 0.6029
     3   .4183 .4066 .3960 .3864 .3782 .3717 .3677 .4739 .4690 .4656 .4633 .4624 .4633 .4662 0.5354
     4   .4887 .4759 .4640 .4533 .4439 .4363 .4315 .5286 .5234 .5195 .5169 .5157 .5163 .5201 0.5334
     5   .5636 .5498 .5370 .5254 .5151 .5067 .5012 .5859 .5806 .5768 .5743 .5731 .5738 .5774 0.5303
 COMPONENT= 2          LIQUID COMPOSITION                         VAPOUR COMPOSITION          MEAN  EMV.
   PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
     1   .4080 .4093 .4102 .4108 .4111 .4112 .4112 .4158 .4190 .4218 .4243 .4263 .4276 .4225 0.9775
     2   .3912 .3952 .3987 .4015 .4038 .4054 .4064 .3970 .4001 .4025 .4042 .4052 .4052 .4023 0.8110
     3   .4164 .4191 .4213 .4230 .4242 .4249 .4253 .3910 .3934 .3951 .3960 .3961 .3952 .3945 1.0851
     4   .3914 .3967 .4013 .4054 .4088 .4114 .4130 .3684 .3721 .3749 .3769 .3780 .3779 .3747 0.6527
     5   .3534 .3611 .3680 .3741 .3793 .3834 .3860 .3389 .3426 .3454 .3472 .3481 .3477 .3450 0.5902
 COMPONENT= 3          LIQUID COMPOSITION                         VAPOUR COMPOSITION          MEAN  EMV.
   PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
     1   .2827 .2947 .3061 .3167 .3264 .3345 .3399 .2306 .2345 .2382 .2417 .2446 .2466 .2394 1.3388
     2   .2274 .2368 .2457 .2538 .2612 .2673 .2714 .1778 .1796 .1807 .1813 .1814 .1808 .1803 0.5545
     3   .1653 .1743 .1827 .1905 .1976 .2033 .2070 .1352 .1375 .1393 .1407 .1415 .1415 .1393 0.4879
     4   .1199 .1275 .1347 .1414 .1473 .1523 .1555 .1030 .1046 .1056 .1062 .1063 .1057 .1052 0.4822
     5   .0830 .0891 .0950 .1005 .1056 .1099 .1128 .0752 .0767 .0778 .0785 .0788 .0785 .0776 0.4780
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      TIME   FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
     64.631  300.0    150.0    150.0     4.667    13.9660    2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.     CTP.
        1        3.00      0.45     0.1422  0.5778 113.868   86.457
        2        2.00      0.30     0.3446  0.3447  38.259   51.868
        3        1.00      0.25     0.5132  0.0775  -2.128   11.675
 PLATE  HOLDUP  TEMP.   PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
   1    3.4561  100.95     4.95      0.9143     1000.00       850.00
   2    3.6088   98.24     5.52      0.7690     1000.00       850.00
   3    3.3529   96.12     4.49      1.0574      700.00       850.00
   4    3.5219   93.80     5.00      0.9006      700.00       850.00
   5    3.6982   91.56     5.54      0.7636      700.00       850.00
 COMPONENT= 1         LIQUID COMPOSITION                      VAPOUR COMPOSITION           MEAN  EMV.
   PLATE   DC    1     2     3     4     5     6     1     2     3     4     5     6
     1    .3124 .2991 .2868 .2757 .2657 .2575 .2521 .3576 .3506 .3441 .3382 .3333 .3300 .3423 0.8705
     2    .3839 .3705 .3583 .3474 .3379 .3302 .3252 .4287 .4239 .4204 .4182 .4171 .4177 .4210 0.6020
     3    .4215 .4099 .3993 .3898 .3816 .3751 .3711 .4774 .4726 .4691 .4669 .4660 .4668 .4698 0.5344
     4    .4918 .4790 .4671 .4564 .4471 .4396 .4347 .5318 .5266 .5228 .5203 .5191 .5197 .5234 0.5329
     5    .5664 .5527 .5399 .5283 .5181 .5097 .5042 .5888 .5836 .5798 .5773 .5761 .5768 .5804 0.5300
 COMPONENT= 2         LIQUID COMPOSITION                      VAPOUR COMPOSITION           MEAN  EMV.
   PLATE   DC    1     2     3     4     5     6     1     2     3     4     5     6
     1    .4064 .4077 .4087 .4093 .4096 .4097 .4097 .4134 .4166 .4194 .4218 .4239 .4252 .4200 0.9781
     2    .3897 .3938 .3972 .4000 .4022 .4039 .4048 .3947 .3978 .4001 .4018 .4027 .4027 .4000 0.8051
     3    .4143 .4170 .4192 .4210 .4222 .4229 .4233 .3885 .3909 .3925 .3934 .3936 .3928 .3919 1.0445
     4    .3892 .3945 .3992 .4032 .4066 .4092 .4108 .3660 .3696 .3724 .3744 .3755 .3754 .3722 0.6507
     5    .3512 .3589 .3658 .3719 .3771 .3812 .3838 .3366 .3403 .3430 .3448 .3457 .3453 .3426 0.5898
 COMPONENT= 3         LIQUID COMPOSITION                      VAPOUR COMPOSITION           MEAN  EMV.
   PLATE   DC    1     2     3     4     5     6     1     2     3     4     5     6
     1    .2812 .2932 .3045 .3151 .3247 .3328 .3382 .2290 .2329 .2365 .2400 .2429 .2448 .2377 1.3401
     2    .2264 .2357 .2445 .2526 .2599 .2659 .2700 .1766 .1784 .1794 .1800 .1802 .1796 .1790 0.5541
     3    .1642 .1731 .1815 .1893 .1962 .2020 .2056 .1342 .1365 .1383 .1397 .1405 .1404 .1382 0.4876
     4    .1190 .1265 .1337 .1403 .1463 .1512 .1544 .1022 .1038 .1048 .1054 .1055 .1049 .1044 0.4820
     5    .0824 .0884 .0943 .0998 .1048 .1091 .1119 .0746 .0761 .0772 .0779 .0782 .0779 .0770 0.4779
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     TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
    84.118  300.0    150.0     150.0     4.667    13.9660    2.7425
   COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.     ETP.     CTP.
        1       3.00       0.45     0.1458  0.5822 113.309   87.111
        2       2.00       0.30     0.3438  0.3412  38.391   51.353
        3       1.00       0.25     0.5104  0.0766  -1.700   11.537
PLATE  HOLDUP   TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561  100.81     4.95      0.9143     1000.00        850.00
  2    3.6088   98.12     5.52      0.7690     1000.00        850.00
  3    3.3529   95.98     4.49      1.0574      700.00        850.00
  4    3.5219   93.68     5.00      0.9006      700.00        850.00
  5    3.6982   91.44     5.54      0.7636      700.00        850.00
COMPONENT= 1         LIQUID COMPOSITION                        VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3166 .3034 .2912 .2801 .2702 .2620 .2567 .3629 .3560 .3496 .3438 .3390 .3357 .3479 0.8719
    2    .3875 .3742 .3621 .3513 .3419 .3343 .3293 .4334 .4287 .4254 .4231 .4221 .4227 .4259 0.6008
    3    .4264 .4148 .4042 .3947 .3865 .3800 .3760 .4823 .4776 .4741 .4719 .4709 .4718 .4748 0.5335
    4    .4965 .4837 .4719 .4613 .4520 .4444 .4396 .5365 .5314 .5276 .5251 .5239 .5245 .5282 0.5321
    5    .5709 .5572 .5445 .5329 .5228 .5144 .5089 .5932 .5880 .5843 .5818 .5806 .5812 .5848 0.5296
COMPONENT= 2         LIQUID COMPOSITION                        VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .4042 .4056 .4066 .4072 .4076 .4077 .4078 .4103 .4134 .4162 .4186 .4206 .4219 .4168 0.9792
    2    .3877 .3917 .3951 .3979 .4001 .4017 .4027 .3917 .3947 .3970 .3986 .3995 .3995 .3968 0.7983
    3    .4111 .4139 .4162 .4179 .4192 .4200 .4204 .3849 .3874 .3890 .3899 .3901 .3893 .3885 0.9875
    4    .3859 .3912 .3959 .3999 .4033 .4060 .4076 .3625 .3660 .3688 .3707 .3718 .3718 .3686 0.6465
    5    .3478 .3554 .3624 .3685 .3736 .3778 .3804 .3331 .3368 .3395 .3413 .3422 .3418 .3391 0.5885
COMPONENT= 3         LIQUID COMPOSITION                        VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2791 .2910 .3022 .3127 .3222 .3302 .3356 .2267 .2306 .2342 .2376 .2404 .2424 .2353 1.3420
    2    .2249 .2341 .2428 .2508 .2580 .2640 .2680 .1749 .1766 .1777 .1783 .1784 .1778 .1773 0.5536
    3    .1624 .1713 .1796 .1874 .1943 .2000 .2036 .1327 .1350 .1368 .1382 .1390 .1389 .1368 0.4872
    4    .1177 .1251 .1322 .1388 .1447 .1496 .1528 .1010 .1026 .1036 .1042 .1043 .1037 .1032 0.4817
    5    .0814 .0874 .0932 .0986 .1036 .1078 .1107 .0737 .0752 .0763 .0770 .0773 .0770 .0761 0.4776
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      TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
    103.605  300.0    150.0     150.0     4.667    13.9660    2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.     CTP.
        1        3.00      0.45     0.1491  0.5866 112.794   87.773
        2        2.00      0.30     0.3431  0.3377  38.498   50.826
        3        1.00      0.25     0.5077  0.0757  -1.292   11.401
PLATE   HOLDUP   TEMP.   PECLET NO.   BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1     3.4561  100.68      4.95       0.9143     1000.00       850.00
  2     3.6088   98.01      5.52       0.7690     1000.00       850.00
  3     3.3529   95.84      4.49       1.0574      700.00       850.00
  4     3.5219   93.55      5.00       0.9006      700.00       850.00
  5     3.6982   91.33      5.54       0.7636      700.00       850.00
COMPONENT= 1         LIQUID COMPOSITION                     VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3206 .3075 .2954 .2843 .2744 .2663 .2609 .3679 .3611 .3547 .3490 .3442 .3410 .3530 0.8730
    2    .3909 .3778 .3658 .3551 .3458 .3382 .3332 .4379 .4332 .4299 .4277 .4267 .4273 .4305 0.5998
    3    .4313 .4197 .4090 .3995 .3913 .3848 .3807 .4870 .4823 .4789 .4766 .4757 .4765 .4795 0.5326
    4    .5012 .4885 .4767 .4661 .4568 .4492 .4444 .5410 .5360 .5323 .5298 .5286 .5292 .5328 0.5313
    5    .5753 .5617 .5490 .5376 .5274 .5191 .5136 .5975 .5924 .5887 .5862 .5850 .5857 .5892 0.5290
COMPONENT= 2         LIQUID COMPOSITION                     VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .4022 .4036 .4047 .4053 .4057 .4059 .4060 .4075 .4105 .4133 .4157 .4177 .4190 .4139 0.9803
    2    .3857 .3897 .3930 .3958 .3981 .3997 .4007 .3889 .3918 .3941 .3957 .3966 .3966 .3939 0.7914
    3    .4080 .4108 .4132 .4150 .4163 .4172 .4176 .3816 .3841 .3857 .3866 .3868 .3861 .3851 0.9364
    4    .3824 .3878 .3925 .3966 .4001 .4027 .4044 .3590 .3626 .3653 .3672 .3683 .3682 .3651 0.6420
    5    .3442 .3519 .3588 .3649 .3702 .3743 .3769 .3296 .3333 .3359 .3377 .3386 .3382 .3356 0.5870
COMPONENT= 3         LIQUID COMPOSITION                     VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2771 .2889 .3000 .3104 .3198 .3278 .3331 .2246 .2284 .2320 .2353 .2381 .2400 .2331 1.3441
    2    .2234 .2326 .2411 .2491 .2562 .2621 .2661 .1733 .1749 .1760 .1766 .1767 .1761 .1756 0.5531
    3    .1608 .1695 .1778 .1855 .1924 .1980 .2016 .1314 .1336 .1354 .1367 .1375 .1375 .1354 0.4867
    4    .1164 .1238 .1308 .1373 .1432 .1480 .1512 .0999 .1015 .1025 .1030 .1031 .1025 .1021 0.4813
    5    .0804 .0864 .0921 .0975 .1024 .1066 .1094 .0729 .0743 .0754 .0761 .0764 .0761 .0752 0.4773
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     TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
   123.092  300.0    150.0     150.0     4.667    13.9660    2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.     ETP.     CTP.
        1        3.00      0.45     0.1522  0.5910 112.319   88.431
        2        2.00      0.30     0.3426  0.3342  38.583   50.300
        3        1.00      0.25     0.5052  0.0748  -0.902   11.269
PLATE   HOLDUP   TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1     3.4561  100.56     4.95      0.9143     1000.00      850.00
  2     3.6088   97.90     5.52      0.7690     1000.00      850.00
  3     3.3529   95.71     4.49      1.0574      700.00      850.00
  4     3.5219   93.43     5.00      0.9006      700.00      850.00
  5     3.6982   91.21     5.54      0.7636      700.00      850.00
COMPONENT= 1          LIQUID COMPOSITION                      VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3244 .3113 .2992 .2882 .2784 .2702 .2649 .3725 .3657 .3594 .3538 .3490 .3458 .3577 0.8740
    2    .3942 .3812 .3693 .3587 .3494 .3419 .3369 .4420 .4374 .4342 .4321 .4311 .4317 .4347 0.5988
    3    .4360 .4244 .4137 .4042 .3960 .3894 .3853 .4915 .4868 .4834 .4811 .4801 .4809 .4840 0.5317
    4    .5059 .4932 .4814 .4708 .4615 .4540 .4492 .5454 .5404 .5368 .5343 .5331 .5338 .5373 0.5305
    5    .5797 .5662 .5536 .5421 .5320 .5237 .5183 .6018 .5967 .5930 .5905 .5893 .5900 .5936 0.5284
COMPONENT= 2          LIQUID COMPOSITION                      VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .4004 .4018 .4029 .4036 .4041 .4043 .4044 .4049 .4080 .4107 .4131 .4151 .4163 .4113 0.9814
    2    .3838 .3878 .3911 .3939 .3962 .3978 .3988 .3863 .3892 .3914 .3930 .3939 .3938 .3912 0.7846
    3    .4049 .4078 .4102 .4121 .4135 .4145 .4149 .3784 .3809 .3825 .3835 .3837 .3830 .3820 0.8922
    4    .3790 .3844 .3892 .3934 .3968 .3995 .4012 .3557 .3592 .3619 .3638 .3649 .3648 .3617 0.6374
    5    .3407 .3484 .3553 .3615 .3667 .3708 .3735 .3262 .3298 .3325 .3343 .3352 .3348 .3321 0.5852
COMPONENT= 3          LIQUID COMPOSITION                      VAPOUR COMPOSITION         MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2752 .2868 .2979 .3082 .3176 .3255 .3307 .2226 .2263 .2299 .2331 .2359 .2378 .2310 1.3460
    2    .2220 .2310 .2395 .2474 .2544 .2603 .2643 .1717 .1734 .1744 .1750 .1751 .1745 .1740 0.5527
    3    .1591 .1678 .1760 .1837 .1905 .1961 .1997 .1300 .1323 .1341 .1354 .1362 .1361 .1340 0.4863
    4    .1151 .1224 .1294 .1359 .1417 .1465 .1496 .0989 .1004 .1014 .1019 .1020 .1014 .1010 0.4810
    5    .0795 .0854 .0911 .0964 .1013 .1054 .1082 .0720 .0735 .0745 .0752 .0755 .0752 .0743 0.4770
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     TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
   194.346  300.0    150.0     150.0     4.667    13.9660    2.7425
   COMPONENT   REL.VOL  FEED COMP.  BOILER   COND.     ETP.     CTP.
       1         3.00      0.45      0.1645  0.6008 111.096   89.520
       2         2.00      0.30      0.3398  0.3263  38.845   49.423
       3         1.00      0.25      0.4957  0.0730   0.059   11.058
PLATE  HOLDUP    TEMP.   PECLET NO.   BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561   100.29      4.95       0.9143     1000.00       850.00
  2    3.6088    97.67      5.52       0.7690     1000.00       850.00
  3    3.3529    95.43      4.49       1.0574      700.00       850.00
  4    3.5219    93.17      5.00       0.9006      700.00       850.00
  5    3.6982    91.00      5.54       0.7636      700.00       850.00
COMPONENT= 1          LIQUID COMPOSITION                  VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3365 .3237 .3118 .3009 .2912 .2832 .2779 .3833 .3767 .3706 .3651 .3605 .3574 .3689 0.8777
    2    .4047 .3920 .3804 .3700 .3610 .3536 .3488 .4516 .4471 .4440 .4419 .4409 .4415 .4445 0.5973
    3    .4504 .4388 .4282 .4187 .4105 .4040 .3999 .5015 .4968 .4934 .4912 .4902 .4909 .4940 0.5295
    4    .5189 .5064 .4949 .4844 .4753 .4680 .4633 .5549 .5500 .5464 .5441 .5430 .5436 .5470 0.5284
    5    .5902 .5771 .5649 .5539 .5441 .5360 .5308 .6103 .6054 .6018 .5994 .5982 .5989 .6023 0.5266
COMPONENT= 2          LIQUID COMPOSITION                  VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3938 .3953 .3965 .3973 .3978 .3981 .3983 .3987 .4016 .4043 .4067 .4086 .4098 .4050 0.9833
    2    .3773 .3812 .3845 .3873 .3895 .3912 .3921 .3801 .3830 .3851 .3866 .3875 .3875 .3850 0.7748
    3    .3950 .3980 .4006 .4026 .4041 .4052 .4057 .3712 .3737 .3754 .3763 .3766 .3759 .3748 0.8194
    4    .3693 .3746 .3793 .3835 .3869 .3896 .3913 .3484 .3518 .3544 .3563 .3573 .3572 .3543 0.6279
    5    .3323 .3398 .3464 .3523 .3574 .3614 .3640 .3193 .3228 .3254 .3271 .3280 .3276 .3250 0.5812
COMPONENT= 3          LIQUID COMPOSITION                  VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2697 .2810 .2917 .3018 .3109 .3187 .3238 .2180 .2216 .2251 .2282 .2310 .2328 .2261 1.3475
    2    .2180 .2268 .2351 .2427 .2495 .2553 .2591 .1683 .1699 .1709 .1715 .1716 .1710 .1705 0.5518
    3    .1547 .1632 .1712 .1787 .1854 .1909 .1944 .1273 .1295 .1312 .1325 .1333 .1332 .1312 0.4854
    4    .1119 .1190 .1258 .1321 .1377 .1424 .1455 .0967 .0982 .0991 .0996 .0997 .0992 .0987 0.4802
    5    .0774 .0831 .0886 .0938 .0985 .1025 .1053 .0704 .0718 .0728 .0735 .0738 .0735 .0727 0.4764
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     TIME    FEED   BOTTOMS    TOPS   R.RATIO  RB.HOLDUP  C.HOLDUP
   267.287  300.0   150.0    150.0    4.667   13.9660    2.7425
   COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.    CTP.
       1        3.00       0.45    0.1744  0.6105 109.710   90.807
       2        2.00       0.30    0.3378  0.3184  39.136   48.382
       3        1.00       0.25    0.4877  0.0711   1.154   10.811
PLATE  HOLDUP  TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561  99.97     4.95      0.9143     1000.00       850.00
  2    3.6088  97.41     5.52      0.7690     1000.00       850.00
  3    3.3529  95.13     4.49      1.0574      700.00       850.00
  4    3.5219  92.90     5.00      0.9006      700.00       850.00
  5    3.6982  90.76     5.54      0.7636      700.00       850.00
COMPONENT= 1          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3473 .3346 .3229 .3122 .3027 .2948 .2895 .3956 .3892 .3833 .3780 .3735 .3705 .3817 0.8810
    2    .4139 .4015 .3902 .3800 .3712 .3640 .3593 .4624 .4581 .4551 .4531 .4522 .4528 .4556 0.5952
    3    .4628 .4513 .4408 .4313 .4231 .4166 .4125 .5126 .5080 .5047 .5024 .5014 .5020 .5052 0.5273
    4    .5304 .5181 .5067 .4964 .4874 .4802 .4755 .5653 .5605 .5571 .5548 .5538 .5544 .5576 0.5265
    5    .6002 .5874 .5755 .5647 .5551 .5472 .5420 .6197 .6149 .6114 .6091 .6080 .6086 .6120 0.5251
COMPONENT= 2          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3881 .3897 .3909 .3918 .3925 .3929 .3931 .3916 .3945 .3971 .3994 .4012 .4024 .3977 0.9859
    2    .3717 .3755 .3788 .3815 .3837 .3854 .3864 .3732 .3759 .3780 .3794 .3802 .3802 .3778 0.7617
    3    .3864 .3896 .3923 .3944 .3961 .3972 .3978 .3632 .3657 .3673 .3683 .3686 .3680 .3669 0.7662
    4    .3606 .3660 .3707 .3748 .3783 .3810 .3827 .3405 .3438 .3463 .3481 .3490 .3489 .3461 0.6191
    5    .3243 .3316 .3381 .3439 .3489 .3529 .3554 .3117 .3151 .3176 .3193 .3201 .3197 .3172 0.5778
COMPONENT= 3          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2647 .2757 .2862 .2960 .3049 .3124 .3174 .2129 .2163 .2196 .2227 .2253 .2270 .2206 1.3502
    2    .2145 .2230 .2310 .2384 .2451 .2507 .2544 .1645 .1660 .1670 .1675 .1676 .1671 .1666 0.5508
    3    .1508 .1591 .1670 .1743 .1808 .1862 .1897 .1242 .1263 .1280 .1293 .1300 .1300 .1280 0.4845
    4    .1090 .1160 .1226 .1287 .1342 .1388 .1418 .0943 .0957 .0966 .0971 .0972 .0967 .0963 0.4794
    5    .0754 .0810 .0864 .0914 .0960 .0999 .1026 .0686 .0700 .0710 .0716 .0719 .0717 .0708 0.4759
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      TIME    FEED   BOTTOMS   TOPS   R.RATIO  RB.HOLDUP  C.HOLDUP
    356.820  300.0   150.0    150.0    4.667    13.9660    2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.    ETP.    CTP.
        1        3.00      0.45     0.1834  0.6221 108.161   92.436
        2        2.00      0.30     0.3367  0.3090  39.417   47.064
        3        1.00      0.25     0.4800  0.0689   2.422   10.500
PLATE  HOLDUP   TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561   99.60     4.95      0.9143     1000.00       850.00
  2    3.6088   97.10     5.52      0.7690     1000.00       850.00
  3    3.3529   94.76     4.49      1.0574      700.00       850.00
  4    3.5219   92.56     5.00      0.9006      700.00       850.00
  5    3.6982   90.47     5.54      0.7636      700.00       850.00
COMPONENT= 1         LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3585 .3460 .3344 .3239 .3144 .3066 .3014 .4093 .4032 .3975 .3923 .3880 .3851 .3959 0.8841
    2    .4235 .4115 .4004 .3905 .3819 .3748 .3702 .4747 .4706 .4677 .4658 .4650 .4655 .4682 0.5926
    3    .4763 .4648 .4543 .4448 .4366 .4300 .4259 .5254 .5210 .5177 .5154 .5143 .5149 .5181 0.5248
    4    .5433 .5311 .5199 .5097 .5008 .4936 .4890 .5775 .5730 .5697 .5675 .5665 .5672 .5702 0.5241
    5    .6120 .5994 .5878 .5771 .5677 .5599 .5548 .6311 .6264 .6231 .6208 .6197 .6203 .6236 0.5232
COMPONENT= 2         LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3823 .3841 .3855 .3865 .3873 .3878 .3881 .3838 .3866 .3891 .3914 .3932 .3944 .3897 0.9895
    2    .3659 .3697 .3729 .3757 .3779 .3795 .3805 .3653 .3679 .3699 .3713 .3720 .3720 .3697 0.7449
    3    .3772 .3806 .3834 .3858 .3876 .3889 .3896 .3540 .3564 .3581 .3591 .3594 .3589 .3576 0.7164
    4    .3509 .3563 .3611 .3653 .3689 .3716 .3733 .3310 .3342 .3366 .3383 .3393 .3391 .3364 0.6086
    5    .3149 .3220 .3285 .3343 .3392 .3432 .3457 .3024 .3057 .3082 .3098 .3106 .3103 .3078 0.5734
COMPONENT= 3         LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE   DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2592 .2699 .2801 .2896 .2983 .3056 .3105 .2069 .2102 .2134 .2163 .2188 .2205 .2144 1.3541
    2    .2106 .2189 .2266 .2338 .2402 .2456 .2493 .1600 .1615 .1624 .1629 .1630 .1625 .1620 0.5496
    3    .1465 .1546 .1623 .1694 .1758 .1811 .1845 .1206 .1227 .1243 .1255 .1262 .1262 .1242 0.4834
    4    .1057 .1125 .1190 .1250 .1303 .1348 .1377 .0914 .0928 .0937 .0942 .0942 .0937 .0933 0.4785
    5    .0731 .0785 .0837 .0886 .0931 .0969 .0995 .0665 .0678 .0688 .0694 .0697 .0694 .0686 0.4752
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   TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP   C.HOLDUP
 460.494  300.0    150.0     150.0     4.667    13.9660     2.7425
  COMPONENT  REL.VOL  FEED COMP.  BOILER   COND.    ETP.     CTP.
      1        3.00      0.45     0.1914  0.6336 106.949    94.190
      2        2.00      0.30     0.3359  0.2997  39.534    45.642
      3        1.00      0.25     0.4727  0.0667   3.517    10.168
PLATE  HOLDUP   TEMP.  PECLET NO.  BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561   99.27     4.95      0.9143     1000.00      850.00
  2    3.6088   96.81     5.52      0.7690     1000.00      850.00
  3    3.3529   94.40     4.49      1.0574      700.00      850.00
  4    3.5219   92.22     5.00      0.9006      700.00      850.00
  5    3.6982   90.16     5.54      0.7636      700.00      850.00
COMPONENT= 1          LIQUID COMPOSITION                        VAPOUR COMPOSITION         MEAN  EMV.
 PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .3683 .3558 .3443 .3338 .3243 .3165 .3114 .4204 .4144 .4089 .4038 .3995 .3967 .4073 0.8857
   2    .4324 .4206 .4097 .3999 .3914 .3844 .3798 .4851 .4812 .4784 .4766 .4758 .4764 .4789 0.5901
   3    .4892 .4777 .4670 .4574 .4491 .4425 .4383 .5370 .5325 .5292 .5270 .5258 .5263 .5297 0.5226
   4    .5560 .5439 .5327 .5225 .5137 .5065 .5018 .5890 .5845 .5814 .5792 .5783 .5790 .5819 0.5219
   5    .6238 .6114 .5999 .5894 .5801 .5724 .5674 .6421 .6376 .6343 .6321 .6310 .6316 .6348 0.5213
COMPONENT= 2          LIQUID COMPOSITION                        VAPOUR COMPOSITION         MEAN  EMV.
 PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .3775 .3795 .3811 .3823 .3833 .3839 .3843 .3777 .3805 .3831 .3853 .3872 .3884 .3837 0.9931
   2    .3606 .3645 .3678 .3706 .3729 .3746 .3756 .3588 .3614 .3633 .3646 .3653 .3652 .3631 0.7269
   3    .3685 .3721 .3752 .3778 .3798 .3813 .3822 .3457 .3482 .3499 .3510 .3514 .3509 .3495 0.6778
   4    .3415 .3470 .3519 .3562 .3598 .3627 .3644 .3223 .3254 .3277 .3294 .3302 .3301 .3275 0.5980
   5    .3054 .3126 .3190 .3248 .3297 .3337 .3362 .2934 .2966 .2990 .3007 .3014 .3011 .2987 0.5684
COMPONENT= 3          LIQUID COMPOSITION                        VAPOUR COMPOSITION         MEAN  EMV.
 PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1    .2542 .2647 .2746 .2839 .2924 .2996 .3043 .2018 .2050 .2081 .2109 .2133 .2149 .2090 1.3594
   2    .2070 .2150 .2225 .2295 .2357 .2410 .2445 .1561 .1575 .1583 .1588 .1589 .1584 .1580 0.5484
   3    .1423 .1502 .1578 .1648 .1711 .1762 .1796 .1173 .1193 .1208 .1220 .1227 .1227 .1208 0.4823
   4    .1025 .1091 .1154 .1213 .1265 .1309 .1337 .0888 .0901 .0909 .0914 .0915 .0910 .0906 0.4776
   5    .0708 .0761 .0811 .0859 .0902 .0939 .0964 .0645 .0657 .0666 .0672 .0675 .0673 .0665 0.4744
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      TIME    FEED    BOTTOMS    TOPS    R.RATIO  RB.HOLDUP  C.HOLDUP
   574.053  300.0    150.0     150.0     4.667    13.9660     2.7425
   COMPONENT   REL.VOL   FEED COMP.   BOILER   COND.    ETP.     CTP.
       1         3.00        0.45      0.1977  0.6442 105.783   95.871
       2         2.00        0.30      0.3358  0.2911  39.634   44.280
       3         1.00        0.25      0.4666  0.0647   4.584    9.848
PLATE  HOLDUP    TEMP.   PECLET NO.   BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1    3.4561    98.96      4.95       0.9143     1000.00      850.00
  2    3.6088    96.55      5.52       0.7690     1000.00      850.00
  3    3.3529    94.07      4.49       1.0574      700.00      850.00
  4    3.5219    91.90      5.00       0.9006      700.00      850.00
  5    3.6982    89.87      5.54       0.7636      700.00      850.00
COMPONENT= 1         LIQUID COMPOSITION                         VAPOUR COMPOSITION           MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3767 .3644 .3529 .3424 .3330 .3252 .3201 .4307 .4248 .4193 .4144 .4102 .4074 .4178 0.8875
    2    .4401 .4284 .4177 .4081 .3996 .3927 .3882 .4946 .4908 .4881 .4864 .4856 .4862 .4886 0.5883
    3    .5005 .4888 .4781 .4684 .4600 .4532 .4490 .5475 .5431 .5398 .5375 .5363 .5367 .5402 0.5209
    4    .5673 .5552 .5440 .5339 .5250 .5177 .5131 .5996 .5952 .5921 .5901 .5892 .5899 .5927 0.5200
    5    .6345 .6223 .6109 .6005 .5912 .5836 .5786 .6525 .6480 .6448 .6426 .6416 .6421 .6453 0.5196
COMPONENT= 2         LIQUID COMPOSITION                         VAPOUR COMPOSITION           MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3736 .3757 .3774 .3788 .3799 .3807 .3812 .3723 .3751 .3776 .3799 .3817 .3830 .3783 0.9963
    2    .3562 .3601 .3635 .3664 .3687 .3704 .3715 .3530 .3555 .3573 .3586 .3593 .3592 .3571 0.7140
    3    .3610 .3649 .3682 .3710 .3732 .3749 .3759 .3383 .3408 .3426 .3437 .3442 .3438 .3422 0.6527
    4    .3331 .3387 .3438 .3482 .3520 .3549 .3568 .3142 .3173 .3196 .3212 .3220 .3218 .3193 0.5893
    5    .2968 .3039 .3104 .3162 .3212 .3252 .3278 .2850 .2882 .2906 .2922 .2929 .2926 .2903 0.5638
COMPONENT= 3         LIQUID COMPOSITION                         VAPOUR COMPOSITION           MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2497 .2599 .2697 .2788 .2871 .2941 .2987 .1970 .2001 .2030 .2057 .2081 .2096 .2039 1.3637
    2    .2037 .2115 .2188 .2256 .2317 .2368 .2403 .1524 .1537 .1545 .1550 .1551 .1546 .1542 0.5474
    3    .1385 .1463 .1537 .1606 .1668 .1719 .1751 .1142 .1161 .1177 .1188 .1195 .1195 .1176 0.4813
    4    .0996 .1061 .1122 .1179 .1231 .1273 .1301 .0863 .0875 .0883 .0888 .0888 .0883 .0880 0.4767
    5    .0687 .0738 .0787 .0833 .0876 .0912 .0937 .0625 .0638 .0646 .0652 .0655 .0653 .0645 0.4737
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    TIME     FEED    BOTTOMS    TOPS    R.RATIO   RB.HOLDUP   C.HOLDUP
  727.234   300.0    150.0     150.0     4.667    13.9660     2.7425
  COMPONENT   REL.VOL   FEED COMP.   BOILER   COND.     ETP.      CTP.
      1         3.00       0.45      0.2060  0.6534  104.970    97.323
      2         2.00       0.30      0.3347  0.2837   39.632    43.108
      3         1.00       0.25      0.4594  0.0629    5.399     9.569
PLATE   HOLDUP   TEMP.   PECLET NO.   BACKFLOW   LIQUID FLOW   VAPOUR FLOW
  1     3.4561   98.72      4.95       0.9143      1000.00       850.00
  2     3.6088   96.33      5.52       0.7690      1000.00       850.00
  3     3.3529   93.80      4.49       1.0574       700.00       850.00
  4     3.5219   91.63      5.00       0.9006       700.00       850.00
  5     3.6982   89.62      5.54       0.7636       700.00       850.00
COMPONENT= 1          LIQUID COMPOSITION                         VAPOUR COMPOSITION          MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1     .3847 .3724 .3610 .3505 .3411 .3333 .3282 .4380 .4322 .4268 .4219 .4177 .4150 .4253 0.8883
   2     .4476 .4360 .4254 .4158 .4074 .4006 .3961 .5017 .4980 .4954 .4937 .4929 .4935 .4959 0.5868
   3     .5112 .4995 .4887 .4790 .4705 .4637 .4594 .5556 .5512 .5479 .5455 .5443 .5447 .5482 0.5197
   4     .5778 .5657 .5546 .5445 .5356 .5284 .5238 .6079 .6036 .6005 .5985 .5977 .5984 .6011 0.5186
   5     .6440 .6320 .6207 .6105 .6013 .5938 .5889 .6608 .6564 .6533 .6511 .6501 .6506 .6537 0.5182
COMPONENT= 2          LIQUID COMPOSITION                         VAPOUR COMPOSITION          MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1     .3697 .3720 .3739 .3754 .3767 .3776 .3781 .3686 .3714 .3740 .3763 .3782 .3794 .3747 0.9990
   2     .3519 .3558 .3593 .3622 .3646 .3664 .3676 .3488 .3512 .3531 .3543 .3550 .3548 .3529 0.7032
   3     .3538 .3579 .3614 .3644 .3668 .3686 .3697 .3326 .3352 .3370 .3383 .3388 .3384 .3367 0.6360
   4     .3253 .3311 .3362 .3407 .3446 .3476 .3495 .3079 .3109 .3132 .3148 .3156 .3154 .3130 0.5827
   5     .2892 .2963 .3028 .3085 .3135 .3175 .3201 .2783 .2814 .2838 .2854 .2861 .2858 .2835 0.5601
COMPONENT= 3          LIQUID COMPOSITION                         VAPOUR COMPOSITION          MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
   1     .2456 .2556 .2651 .2741 .2822 .2891 .2937 .1933 .1963 .1992 .2018 .2041 .2056 .2001 1.3681
   2     .2006 .2082 .2154 .2220 .2280 .2330 .2363 .1495 .1508 .1516 .1520 .1521 .1516 .1513 0.5465
   3     .1349 .1426 .1499 .1566 .1627 .1677 .1709 .1117 .1136 .1151 .1162 .1169 .1169 .1151 0.4804
   4     .0968 .1032 .1092 .1148 .1198 .1240 .1267 .0842 .0855 .0862 .0867 .0867 .0862 .0859 0.4759
   5     .0667 .0717 .0765 .0810 .0852 .0887 .0911 .0609 .0621 .0630 .0636 .0638 .0636 .0628 0.4730
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    TIME    FEED    BOTTOMS    TOPS    R.RATIO   RB.HOLDUP   C.HOLDUP
  835.481  300.0    150.0     150.0     4.667    13.9660      2.7425
  COMPONENT   REL.VOL  FEED COMP.  BOILER   COND.     ETP.     CTP.
      1         3.00      0.45     0.2074  0.6611 103.993    98.658
      2         2.00      0.30     0.3359  0.2775  39.692    42.035
      3         1.00      0.25     0.4567  0.0614   6.315     9.307
PLATE   HOLDUP   TEMP.   PECLET NO.   BACKFLOW   LIQUID FLOW   VAPOUR FLOW
  1     3.4561   98.45     4.95        0.9143      1000.00        850.00
  2     3.6088   96.11     5.52        0.7690      1000.00        850.00
  3     3.3529   93.53     4.49        1.0574       700.00        850.00
  4     3.5219   91.38     5.00        0.9006       700.00        850.00
  5     3.6982   89.39     5.54        0.7636       700.00        850.00
COMPONENT= 1          LIQUID COMPOSITION                            VAPOUR COMPOSITION           MEAN   EMV.
  PLATE    0C     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3898 .3776 .3661 .3557 .3463 .3384 .3333 .4465 .4408 .4355 .4308 .4268 .4241 .4341 0.8887
    2    .4521 .4407 .4302 .4207 .4124 .4056 .4012 .5096 .5059 .5034 .5017 .5010 .5016 .5039 0.5851
    3    .5181 .5063 .4954 .4855 .4770 .4701 .4657 .5640 .5596 .5563 .5539 .5527 .5530 .5566 0.5187
    4    .5851 .5730 .5618 .5516 .5427 .5354 .5308 .6162 .6120 .6090 .6070 .6062 .6069 .6096 0.5173
    5    .6516 .6395 .6283 .6180 .6088 .6013 .5963 .6689 .6646 .6614 .6593 .6583 .6588 .6619 0.5170
COMPONENT= 2          LIQUID COMPOSITION                            VAPOUR COMPOSITION           MEAN   EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3677 .3701 .3721 .3738 .3752 .3762 .3769 .3642 .3670 .3695 .3717 .3736 .3748 .3701 1.0030
    2    .3495 .3535 .3570 .3599 .3624 .3643 .3654 .3441 .3465 .3482 .3494 .3501 .3500 .3480 0.6941
    3    .3495 .3538 .3575 .3606 .3632 .3651 .3663 .3268 .3294 .3313 .3325 .3331 .3327 .3310 0.6246
    4    .3201 .3260 .3312 .3359 .3399 .3430 .3449 .3016 .3046 .3069 .3084 .3092 .3090 .3066 0.5776
    5    .2832 .2904 .2970 .3028 .3079 .3120 .3146 .2718 .2749 .2772 .2788 .2796 .2792 .2769 0.5570
COMPONENT= 3          LIQUID COMPOSITION                            VAPOUR COMPOSITION           MEAN   EMV.
  PLATE    0C     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2425 .2524 .2617 .2705 .2786 .2853 .2898 .1893 .1922 .1950 .1975 .1997 .2011 .1958 1.3735
    2    .1984 .2058 .2128 .2193 .2252 .2301 .2334 .1464 .1476 .1484 .1488 .1489 .1484 .1481 0.5455
    3    .1324 .1400 .1472 .1538 .1598 .1648 .1680 .1092 .1110 .1125 .1136 .1142 .1143 .1125 0.4797
    4    .0948 .1011 .1070 .1125 .1175 .1216 .1243 .0822 .0834 .0841 .0845 .0846 .0841 .0838 0.4752
    5    .0652 .0701 .0748 .0792 .0833 .0868 .0891 .0594 .0605 .0613 .0619 .0622 .0619 .0612 0.4725
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      TIME    FEED    BOTTOMS    TOPS    R.RATIO   RB.HOLDUP   C.HOLDUP
    945.297  300.0    150.0     150.0     4.667    13.9660     2.7425
    COMPONENT   REL.VOL  FEED COMP.   BOILER   COND.     ETP.     CTP.
       1          3.00      0.45      0.2099  0.6660  103.734    99.562
       2          2.00      0.30      0.3361  0.2737   39.588    41.314
       3          1.00      0.25      0.4540  0.0604    6.678     9.125
 PLATE  HOLDUP    TEMP.  PECLET NO.   BACKFLOW   LIQUID FLOW  VAPOUR FLOW
   1    3.4561    98.36     4.95       0.9143      1000.00       850.00
   2    3.6088    96.01     5.52       0.7690      1000.00       850.00
   3    3.3529    93.38     4.49       1.0574       700.00       850.00
   4    3.5219    91.23     5.00       0.9006       700.00       850.00
   5    3.6982    89.24     5.54       0.7636       700.00       850.00
 COMPONENT= 1          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN   EMV.
   PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
     1    .3926 .3803 .3688 .3583 .3489 .3411 .3359 .4489 .4432 .4378 .4330 .4289 .4261 .4363 0.8896
     2    .4550 .4435 .4330 .4235 .4152 .4084 .4039 .5122 .5086 .5060 .5044 .5037 .5043 .5065 0.5848
     3    .5228 .5109 .4998 .4899 .4812 .4742 .4699 .5677 .5632 .5598 .5574 .5562 .5565 .5601 0.5180
     4    .5901 .5780 .5667 .5564 .5475 .5402 .5355 .6203 .6161 .6131 .6112 .6104 .6111 .6137 0.5166
     5    .6566 .6445 .6333 .6230 .6138 .6063 .6013 .6734 .6691 .6659 .6638 .6628 .6633 .6664 0.5163
 COMPONENT= 2          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN   EMV.
   PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
     1    .3667 .3693 .3714 .3732 .3746 .3757 .3764 .3634 .3662 .3689 .3712 .3731 .3744 .3695 1.0032
     2    .3481 .3522 .3558 .3588 .3613 .3632 .3644 .3427 .3451 .3469 .3481 .3488 .3486 .3467 0.6893
     3    .3466 .3510 .3549 .3582 .3609 .3629 .3641 .3244 .3270 .3290 .3303 .3309 .3306 .3287 0.6166
     4    .3165 .3225 .3279 .3326 .3367 .3399 .3419 .2986 .3016 .3038 .3054 .3062 .3059 .3036 0.5740
     5    .2793 .2865 .2931 .2990 .3041 .3083 .3109 .2682 .2713 .2736 .2752 .2760 .2757 .2733 0.5549
 COMPONENT= 3          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN   EMV.
   PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
     1    .2407 .2505 .2598 .2685 .2765 .2832 .2877 .1877 .1906 .1933 .1959 .1980 .1995 .1942 1.3753
     2    .1969 .2043 .2113 .2177 .2235 .2284 .2317 .1450 .1463 .1471 .1475 .1475 .1471 .1467 0.5451
     3    .1306 .1381 .1453 .1519 .1579 .1628 .1660 .1079 .1098 .1112 .1123 .1130 .1130 .1112 0.4792
     4    .0934 .0996 .1055 .1109 .1158 .1199 .1226 .0811 .0823 .0830 .0834 .0835 .0830 .0827 0.4748
     5    .0641 .0689 .0736 .0780 .0820 .0855 .0878 .0585 .0596 .0604 .0610 .0612 .0610 .0603 0.4720
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      TIME    FEED   BOTTOMS    TOPS   R.RATIO  RB.HOLDUP  C.HOLDUP
    *1115.755  300.0    150.0     150.0     4.667   13.9660     2.7425
    COMPONENT  REL.VOL  FEED COMP.  BOILER  COND.     ETP.     CTP.
        1        3.00       0.45      0.2137  0.6696 103.326  100.179
        2        2.00       0.30      0.3358  0.2708  39.576   40.824
        3        1.00       0.25      0.4505  0.0596   7.098    8.997
PLATE   HOLDUP   TEMP.  PECLET NO.   BACKFLOW  LIQUID FLOW  VAPOUR FLOW
  1     3.4561   98.26      4.95       0.9143     1000.00       850.00
  2     3.6088   95.92      5.52       0.7690     1000.00       850.00
  3     3.3529   93.27      4.49       1.0574      700.00       850.00
  4     3.5219   91.11      5.00       0.9006      700.00       850.00
  5     3.6982   89.13      5.54       0.7636      700.00       850.00
COMPONENT= 1          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3959 .3836 .3722 .3618 .3524 .3446 .3395 .4521 .4464 .4411 .4363 .4322 .4295 .4396 0.8903
    2    .4579 .4465 .4360 .4266 .4183 .4116 .4071 .5151 .5115 .5090 .5074 .5067 .5073 .5095 0.5845
    3    .5269 .5150 .5040 .4940 .4853 .4783 .4739 .5709 .5665 .5631 .5606 .5593 .5596 .5633 0.5178
    4    .5942 .5820 .5708 .5606 .5516 .5443 .5397 .6237 .6195 .6165 .6146 .6138 .6145 .6171 0.5162
    5    .6604 .6484 .6372 .6269 .6178 .6103 .6053 .6768 .6725 .6694 .6672 .6662 .6668 .6698 0.5158
COMPONENT= 2          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .3653 .3679 .3700 .3719 .3733 .3745 .3752 .3619 .3648 .3674 .3697 .3716 .3729 .3681 1.0042
    2    .3465 .3506 .3542 .3573 .3598 .3618 .3630 .3411 .3435 .3453 .3465 .3471 .3470 .3451 0.6869
    3    .3440 .3485 .3524 .3558 .3586 .3607 .3619 .3222 .3249 .3269 .3282 .3288 .3285 .3266 0.6132
    4    .3136 .3196 .3251 .3299 .3340 .3372 .3392 .2961 .2991 .3014 .3029 .3037 .3034 .3011 0.5721
    5    .2763 .2836 .2902 .2961 .3012 .3054 .3080 .2655 .2686 .2709 .2725 .2733 .2729 .2706 0.5535
COMPONENT= 3          LIQUID COMPOSITION                        VAPOUR COMPOSITION          MEAN  EMV.
  PLATE    DC     1     2     3     4     5     6     1     2     3     4     5     6
    1    .2388 .2485 .2577 .2664 .2743 .2809 .2854 .1860 .1888 .1915 .1940 .1961 .1976 .1923 1.3765
    2    .1956 .2029 .2097 .2161 .2219 .2267 .2299 .1437 .1449 .1457 .1461 .1462 .1457 .1454 0.5448
    3    .1290 .1365 .1436 .1502 .1561 .1610 .1642 .1068 .1086 .1101 .1111 .1118 .1118 .1101 0.4789
    4    .0922 .0983 .1042 .1096 .1144 .1185 .1211 .0802 .0814 .0821 .0825 .0825 .0821 .0818 0.4744
    5    .0633 .0681 .0726 .0770 .0810 .0844 .0867 .0578 .0589 .0597 .0602 .0605 .0603 .0596 0.4717
```

Appendix H

Unsteady-State Vaporisation Efficiency Results

The vaporisation efficiency responses for step changes in the multicomponent system investigated by Holland (48) are listed. The five component system was C_3H_8, n-C_4H_{10}, n-C_5H_{12}, n-C_6H_{14} and n-C_7H_{16}. The initial steady-state and unsteady-state feed inputs were:-

Component	$F.X_i$ (SS) moles/min	$F.X_i$ (US) moles/min
1) C_3H_8	10	35
2) n-C_4H_{10}	25	25
3) n-C_5H_{12}	30	20
4) n-C_6H_{14}	25	15
5) n-C_7H_{16}	10	5

It can be seen from the listings that the vaporisation efficiencies were the same on every plate for each component and if this was true then they should converge to a constant value for each plate at the final steady-state which is shown to be untrue.

```
VAPORISATION EFFICIENCIES.
REBOILER.                                           COMPONENT
TIME       1          2          3          4          5
 0.0    1.05000    1.02000    1.05000    0.99000    0.96000
 0.1    1.04154    1.02589    1.04627    0.98650    0.95760
 0.2    1.03476    1.03205    1.04374    0.98415    0.95629
 0.3    1.02948    1.03854    1.04229    0.98298    0.95599
 0.4    1.02457    1.04440    1.04086    0.98227    0.95572
 0.5    1.01908    1.04876    1.03850    0.98132    0.95460
 0.6    1.01232    1.05099    1.03454    0.97944    0.95202
 0.7    1.00399    1.05083    1.02869    0.97622    0.94771
 0.8    0.99420    1.04843    1.02107    0.97160    0.94177
 0.9    0.98332    1.04421    1.01210    0.96582    0.93457
 1.0    0.97184    1.03875    1.00228    0.95929    0.92659
 1.1    0.96024    1.03260    0.99212    0.95240    0.91829
 1.2    0.94889    1.02621    0.98202    0.94550    0.91004
 1.3    0.93804    1.01989    0.97227    0.93884    0.90210
 1.4    0.92783    1.01385    0.96302    0.93255    0.89461
 1.5    0.91832    1.00820    0.95436    0.92673    0.88767
 1.6    0.90955    1.00302    0.94632    0.92139    0.88130
 1.7    0.90148    0.99830    0.93889    0.91654    0.87550
 1.8    0.89407    0.99404    0.93205    0.91216    0.87023
 1.9    0.88729    0.99022    0.92574    0.90822    0.86547
 2.0    0.88107    0.98680    0.91994    0.90467    0.86118
 2.2    0.87017    0.98107    0.90968    0.89867    0.85384
 2.4    0.86101    0.97660    0.90093    0.89389    0.84794
 2.6    0.85330    0.97317    0.89343    0.89011    0.84321
 2.8    0.84679    0.97061    0.88695    0.88717    0.83946
 3.0    0.84129    0.96876    0.88131    0.88492    0.83653
 3.2    0.83664    0.96751    0.87637    0.88325    0.83427
 3.4    0.83271    0.96676    0.87201    0.88206    0.83259
 3.6    0.82939    0.96642    0.86814    0.88127    0.83138
 3.8    0.82659    0.96643    0.86466    0.88082    0.83058
 4.0    0.82424    0.96673    0.86153    0.88066    0.83012
 4.2    0.82227    0.96728    0.85867    0.88075    0.82995
 4.4    0.82063    0.96803    0.85604    0.88103    0.83002
 4.6    0.81927    0.96894    0.85361    0.88148    0.83029
 4.8    0.81816    0.97000    0.85134    0.88207    0.83073
 5.0    0.81725    0.97116    0.84920    0.88278    0.83132
```

```
VAPORISATION EFFICIENCIES.
PLATE= 1                                                 COMPONENT
TIME        1             2             3             4             5
 0.0     1.05000       1.02000       1.05000       0.99000       0.96000
 0.1     1.04154       1.02589       1.04627       0.98650       0.95760
 0.2     1.03476       1.03205       1.04374       0.98415       0.95629
 0.3     1.02948       1.03854       1.04229       0.98298       0.95599
 0.4     1.02457       1.04440       1.04086       0.98227       0.95572
 0.5     1.01908       1.04876       1.03850       0.98132       0.95460
 0.6     1.01232       1.05099       1.03454       0.97944       0.95202
 0.7     1.00399       1.05083       1.02869       0.97622       0.94771
 0.8     0.99420       1.04843       1.02107       0.97160       0.94177
 0.9     0.98332       1.04421       1.01210       0.96582       0.93457
 1.0     0.97184       1.03875       1.00228       0.95929       0.92659
 1.1     0.96024       1.03260       0.99212       0.95240       0.91829
 1.2     0.94889       1.02621       0.98202       0.94550       0.91004
 1.3     0.93804       1.01989       0.97227       0.93884       0.90210
 1.4     0.92783       1.01385       0.96302       0.93255       0.89461
 1.5     0.91832       1.00820       0.95436       0.92673       0.88767
 1.6     0.90955       1.00302       0.94632       0.92139       0.88130
 1.7     0.90148       0.99830       0.93889       0.91654       0.87550
 1.8     0.89407       0.99404       0.93205       0.91216       0.87023
 1.9     0.88729       0.99022       0.92574       0.90822       0.86547
 2.0     0.88107       0.98680       0.91994       0.90467       0.86118
 2.2     0.87017       0.98107       0.90968       0.89867       0.85384
 2.4     0.86101       0.97660       0.90093       0.89389       0.84794
 2.6     0.85330       0.97317       0.89343       0.89011       0.84321
 2.8     0.84679       0.97061       0.88695       0.88717       0.83946
 3.0     0.84129       0.96876       0.88131       0.88492       0.83653
 3.2     0.83664       0.96751       0.87637       0.88325       0.83427
 3.4     0.83271       0.96676       0.87201       0.88206       0.83259
 3.6     0.82939       0.96642       0.86814       0.88127       0.83138
 3.8     0.82659       0.96643       0.86466       0.88082       0.83058
 4.0     0.82424       0.96673       0.86153       0.88066       0.83012
 4.2     0.82227       0.96728       0.85867       0.88075       0.82995
 4.4     0.82063       0.96803       0.85604       0.88103       0.83002
 4.6     0.81927       0.96894       0.85361       0.88148       0.83029
 4.8     0.81816       0.97000       0.85134       0.88207       0.83073
 5.0     0.81725       0.97116       0.84920       0.88278       0.83132
```

```
VAPORISATION EFFICIENCIES.
PLATE= 2                                              COMPONENT
TIME         1           2           3           4           5
 0.0     1.05000     1.02000     1.05000     0.99000     0.96000
 0.1     1.04419     1.02849     1.04894     0.98901     0.96003
 0.2     1.03988     1.03715     1.04890     0.98902     0.96102
 0.3     1.03596     1.04508     1.04885     0.98917     0.96201
 0.4     1.03238     1.05237     1.04880     0.98976     0.96301
 0.5     1.02911     1.05908     1.04872     0.99097     0.96400
 0.6     1.02609     1.06529     1.04861     0.99276     0.96497
 0.7     1.02330     1.07104     1.04847     0.99499     0.96593
 0.8     1.02070     1.07637     1.04829     0.99750     0.96688
 0.9     1.01828     1.08134     1.04809     1.00017     0.96781
 1.0     1.01602     1.08598     1.04784     1.00290     0.96872
 1.1     1.01391     1.09032     1.04757     1.00563     0.96962
 1.2     1.01193     1.09438     1.04726     1.00832     0.97050
 1.3     1.01006     1.09820     1.04692     1.01093     0.97136
 1.4     1.00831     1.10179     1.04655     1.01345     0.97221
 1.5     1.00665     1.10518     1.04615     1.01587     0.97305
 1.6     1.00509     1.10837     1.04572     1.01818     0.97387
 1.7     1.00361     1.11140     1.04526     1.02038     0.97468
 1.8     1.00220     1.11426     1.04477     1.02248     0.97548
 1.9     1.00087     1.11698     1.04425     1.02448     0.97627
 2.0     0.99961     1.11956     1.04371     1.02638     0.97704
 2.2     0.99725     1.12435     1.04253     1.02991     0.97854
 2.4     0.99512     1.12871     1.04125     1.03311     0.98000
 2.6     0.99316     1.13268     1.03986     1.03601     0.98142
 2.8     0.99137     1.13633     1.03838     1.03864     0.98279
 3.0     0.98972     1.13968     1.03680     1.04105     0.98411
 3.2     0.98820     1.14277     1.03512     1.04325     0.98540
 3.4     0.98679     1.14563     1.03336     1.04526     0.98664
 3.6     0.98548     1.14829     1.03152     1.04712     0.98784
 3.8     0.98425     1.15077     1.02959     1.04884     0.98901
 4.0     0.98312     1.15308     1.02759     1.05042     0.99013
 4.2     0.98205     1.15524     1.02553     1.05189     0.99122
 4.4     0.98105     1.15726     1.02339     1.05326     0.99228
 4.6     0.98011     1.15917     1.02119     1.05453     0.99330
 4.8     0.97923     1.16096     1.01894     1.05573     0.99428
 5.0     0.97839     1.16265     1.01664     1.05684     0.99523
```

```
VAPORISATION EFFICIENCIES.
PLATE= 3                                           COMPONENT
TIME       1           2           3           4           5
 0.0    1.05000     1.02000     1.05000     0.99000     0.96000
 0.1    1.04006     1.02443     1.04479     0.98510     0.95624
 0.2    1.03090     1.02820     1.03984     0.98049     0.95272
 0.3    1.02269     1.03169     1.03542     0.97650     0.94969
 0.4    1.01552     1.03517     1.03166     0.97359     0.94727
 0.5    1.00940     1.03880     1.02863     0.97199     0.94553
 0.6    1.00427     1.04264     1.02631     0.97165     0.94445
 0.7    1.00002     1.04667     1.02462     0.97235     0.94396
 0.8    0.99652     1.05087     1.02346     0.97386     0.94397
 0.9    0.99363     1.05517     1.02272     0.97595     0.94438
 1.0    0.99125     1.05950     1.02229     0.97845     0.94510
 1.1    0.98927     1.06382     1.02211     0.98119     0.94605
 1.2    0.98760     1.06807     1.02209     0.98408     0.94717
 1.3    0.98619     1.07224     1.02218     0.98703     0.94840
 1.4    0.98497     1.07628     1.02233     0.98999     0.94971
 1.5    0.98390     1.08020     1.02251     0.99291     0.95106
 1.6    0.98296     1.08397     1.02270     0.99576     0.95243
 1.7    0.98211     1.08760     1.02288     0.99853     0.95381
 1.8    0.98135     1.09107     1.02305     1.00121     0.95518
 1.9    0.98065     1.09441     1.02315     1.00378     0.95654
 2.0    0.98000     1.09760     1.02324     1.00625     0.95787
 2.2    0.97883     1.10358     1.02327     1.01088     0.96046
 2.4    0.97778     1.10905     1.02311     1.01511     0.96293
 2.6    0.97683     1.11406     1.02277     1.01898     0.96528
 2.8    0.97596     1.11867     1.02224     1.02250     0.96751
 3.0    0.97515     1.12290     1.02155     1.02572     0.96962
 3.2    0.97439     1.12681     1.02066     1.02867     0.97163
 3.4    0.97368     1.13042     1.01965     1.03138     0.97353
 3.6    0.97300     1.13376     1.01846     1.03387     0.97534
 3.8    0.97237     1.13687     1.01716     1.03617     0.97706
 4.0    0.97176     1.13976     1.01575     1.03829     0.97870
 4.2    0.97119     1.14247     1.01419     1.04026     0.98026
 4.4    0.97065     1.14499     1.01254     1.04209     0.98175
 4.6    0.97013     1.14736     1.01079     1.04379     0.98318
 4.8    0.96963     1.14958     1.00896     1.04538     0.98454
 5.0    0.96916     1.15168     1.00704     1.04687     0.98584
```

```
VAPORISATION EFFICIENCIES.
FEED PLATE                                        COMPONENT
TIME       1          2          3          4          5
 0.0    1,05000    1.02000    1,05000    0,99000    0.96000
 0.1    1,05533    1.03947    1,06013    0,99956    0.97028
 0.2    1,06033    1.05755    1,06953    1.00848    0.97992
 0.3    1,06499    1.07437    1,07825    1,01689    0.98897
 0.4    1,06931    1.09001    1,08631    1,02517    0.99745
 0.5    1,07330    1.10456    1,09375    1,03352    1.00539
 0.6    1,07696    1.11810    1,10059    1,04198    1.01281
 0.7    1,08030    1.13070    1,10688    1,05041    1.01974
 0.8    1,08334    1.14243    1,11262    1,05871    1.02621
 0.9    1,08607    1.15333    1,11786    1,06675    1.03224
 1.0    1,08853    1.16347    1,12262    1,07446    1.03785
 1.1    1,09070    1.17290    1,12691    1,08180    1.04305
 1.2    1,09262    1.18165    1,13077    1,08872    1.04789
 1.3    1,09428    1.18977    1,13422    1,09522    1.05236
 1.4    1,09571    1.19730    1,13727    1,10130    1.05649
 1.5    1,09691    1.20427    1,13996    1,10695    1.06030
 1.6    1,09790    1.21072    1,14229    1,11220    1.06580
 1.7    1,09868    1.21668    1,14428    1,11705    1.06702
 1.8    1,09927    1.22218    1,14596    1,12151    1.06996
 1.9    1,09968    1.22725    1,14734    1,12562    1.07264
 2.0    1,09991    1.23190    1,14844    1,12938    1.07508
 2.2    1,09991    1.24009    1,14985    1,13593    1.07928
 2.4    1,09934    1.24692    1,15031    1,14131    1.08264
 2.6    1,09827    1.25255    1,14991    1,14565    1.08528
 2.8    1,09676    1.25712    1,14876    1.14906    1.08726
 3.0    1,09487    1.26076    1,14695    1.15165    1.08867
 3.2    1,09266    1.26358    1,14455    1.15353    1.08957
 3.4    1,09018    1.26567    1,14163    1.15478    1.09002
 3.6    1,08746    1.26713    1,13827    1.15549    1.09008
 3.8    1,08456    1.26804    1,13452    1.15572    1.08979
 4.0    1,08150    1.26847    1,13045    1.15554    1.08922
 4.2    1,07832    1.26848    1,12605    1.15500    1.08839
 4.4    1,07504    1.26814    1,12143    1.15417    1.08734
 4.6    1,07169    1.26748    1,11661    1.15307    1.08611
 4.8    1,06830    1.26656    1,11165    1.15176    1.08472
 5.0    1,06488    1.26542    1.10650    1.15026    1.08321
```

```
VAPORISATION EFFICIENCIES.
PLATE= 5                                               COMPONENT
TIME       1           2           3           4           5
 0.0    1.05000     1.02000     1.05000     0.99000     0.96000
 0.1    1.05022     1.03443     1.05498     0.99471     0.96557
 0.2    1.05037     1.04762     1.05948     0.99900     0.97072
 0.3    1.05048     1.05973     1.06356     1.00304     0.97550
 0.4    1.05056     1.07089     1.06726     1.00719     0.97996
 0.5    1.05060     1.08120     1.07062     1.01167     0.98413
 0.6    1.05062     1.09076     1.07368     1.01650     0.98804
 0.7    1.05062     1.09964     1.07647     1.02156     0.99173
 0.8    1.05061     1.10792     1.07902     1.02673     0.99522
 0.9    1.05059     1.11565     1.08134     1.03190     0.99851
 1.0    1.05056     1.12289     1.08346     1.03698     1.00164
 1.1    1.05052     1.12968     1.08539     1.04194     1.00462
 1.2    1.05047     1.13606     1.08715     1.04672     1.00746
 1.3    1.05042     1.14207     1.08875     1.05132     1.01017
 1.4    1.05036     1.14774     1.09020     1.05572     1.01276
 1.5    1.05031     1.15310     1.09152     1.05992     1.01525
 1.6    1.05025     1.15817     1.09271     1.06392     1.01763
 1.7    1.05019     1.16298     1.09377     1.06774     1.01992
 1.8    1.05012     1.16754     1.09473     1.07137     1.02212
 1.9    1.05006     1.17188     1.09558     1.07483     1.02425
 2.0    1.05000     1.17600     1.09632     1.07813     1.02629
 2.2    1.04988     1.18368     1.09754     1.08426     1.03018
 2.4    1.04975     1.19068     1.09842     1.08983     1.03381
 2.6    1.04963     1.19709     1.09899     1.09491     1.03722
 2.8    1.04951     1.20297     1.09928     1.09956     1.04043
 3.0    1.04940     1.20840     1.09932     1.10382     1.04345
 3.2    1.04929     1.21342     1.09912     1.10774     1.04632
 3.4    1.04918     1.21808     1.09870     1.11136     1.04903
 3.6    1.04908     1.22241     1.09809     1.11471     1.05160
 3.8    1.04898     1.22645     1.09730     1.11781     1.05405
 4.0    1.04889     1.23022     1.09634     1.12070     1.05638
 4.2    1.04880     1.23376     1.09525     1.12339     1.05859
 4.4    1.04871     1.23708     1.09397     1.12590     1.06071
 4.6    1.04863     1.24020     1.09258     1.12825     1.06273
 4.8    1.04855     1.24314     1.09107     1.13046     1.06466
 5.0    1.04847     1.24592     1.08945     1.13253     1.06651
```

```
VAPORISATION EFFICIENCIES.
PLATE= 6                                          COMPONENT
TIME        1          2          3          4          5
 0.0     1.05000    1.02000    1.05000    0.99000    0.96000
 0.1     1.04523    1.02952    1.04998    0.99000    0.96099
 0.2     1.04089    1.03816    1.04992    0.98998    0.96195
 0.3     1.03691    1.04604    1.04982    0.99008    0.96290
 0.4     1.03325    1.05325    1.04968    0.99059    0.96381
 0.5     1.02987    1.05987    1.04949    0.99171    0.96471
 0.6     1.02674    1.06596    1.04927    0.99339    0.96558
 0.7     1.02382    1.07159    1.04901    0.99550    0.96643
 0.8     1.02110    1.07680    1.04871    0.99789    0.96726
 0.9     1.01855    1.08163    1.04837    1.00043    0.96807
 1.0     1.01616    1.08613    1.04799    1.00303    0.96885
 1.1     1.01391    1.09031    1.04757    1.00563    0.96961
 1.2     1.01178    1.09422    1.04711    1.00817    0.97035
 1.3     1.00976    1.09787    1.04661    1.01062    0.97107
 1.4     1.00784    1.10128    1.04607    1.01298    0.97177
 1.5     1.00602    1.10448    1.04550    1.01523    0.97244
 1.6     1.00428    1.10749    1.04489    1.01736    0.97310
 1.7     1.00262    1.11031    1.04424    1.01938    0.97373
 1.8     1.00103    1.11296    1.04355    1.02129    0.97434
 1.9     0.99951    1.11545    1.04283    1.02308    0.97493
 2.0     0.99804    1.11780    1.04207    1.02477    0.97551
 2.2     0.99526    1.12210    1.04045    1.02786    0.97659
 2.4     0.99267    1.12593    1.03869    1.03057    0.97759
 2.6     0.99023    1.12934    1.03680    1.03295    0.97852
 2.8     0.98793    1.13238    1.03477    1.03504    0.97937
 3.0     0.98574    1.13509    1.03263    1.03686    0.98015
 3.2     0.98365    1.13751    1.03036    1.03845    0.98086
 3.4     0.98165    1.13967    1.02798    1.03982    0.98150
 3.6     0.97972    1.14159    1.02550    1.04101    0.98208
 3.8     0.97786    1.14330    1.02291    1.04202    0.98259
 4.0     0.97606    1.14481    1.02022    1.04289    0.98303
 4.2     0.97432    1.14614    1.01745    1.04361    0.98342
 4.4     0.97261    1.14731    1.01459    1.04420    0.98374
 4.6     0.97096    1.14834    1.01165    1.04468    0.98402
 4.8     0.96933    1.14923    1.00865    1.04506    0.98423
 5.0     0.96775    1.15000    1.00557    1.04534    0.98440
```

```
VAPORISATION EFFICIENCIES.
PLATE= 7                                          COMPONENT
TIME       1          2          3          4          5
 0.0    1.05000    1.02000    1.05000    0.99000    0.96000
 0.1    1.03480    1.01924    1.03950    0.98011    0.95139
 0.2    1.02026    1.01758    1.02911    0.97036    0.94289
 0.3    1.00667    1.01553    1.01920    0.96120    0.93481
 0.4    0.99449    1.01373    1.01030    0.95343    0.92765
 0.5    0.98422    1.01289    1.00298    0.94775    0.92195
 0.6    0.97625    1.01355    0.99768    0.94454    0.91810
 0.7    0.97070    1.01598    0.99458    0.94384    0.91629
 0.8    0.96738    1.02015    0.99354    0.94539    0.91638
 0.9    0.96593    1.02574    0.99420    0.94874    0.91805
 1.0    0.96583    1.03233    0.99608    0.95336    0.92087
 1.1    0.96664    1.03948    0.99873    0.95875    0.92441
 1.2    0.96796    1.04683    1.00176    0.96451    0.92833
 1.3    0.96951    1.05411    1.00489    0.97034    0.93236
 1.4    0.97110    1.06113    1.00793    0.97605    0.93634
 1.5    0.97261    1.06780    1.01078    0.98151    0.94015
 1.6    0.97398    1.07407    1.01336    0.98667    0.94374
 1.7    0.97519    1.07992    1.01566    0.99149    0.94708
 1.8    0.97622    1.08557    1.01768    0.99597    0.95019
 1.9    0.97707    1.09042    1.01943    1.00012    0.95305
 2.0    0.97778    1.09511    1.02092    1.00397    0.95570
 2.2    0.97878    1.10352    1.02322    1.01083    0.96041
 2.4    0.97934    1.11081    1.02474    1.01673    0.96447
 2.6    0.97957    1.11718    1.02564    1.02183    0.96799
 2.8    0.97957    1.12280    1.02602    1.02628    0.97109
 3.0    0.97939    1.12779    1.02598    1.03018    0.97384
 3.2    0.97909    1.13224    1.02559    1.03364    0.97632
 3.4    0.97871    1.13626    1.02491    1.03671    0.97856
 3.6    0.97827    1.13989    1.02397    1.03946    0.98062
 3.8    0.97779    1.14321    1.02285    1.04194    0.98251
 4.0    0.97728    1.14624    1.02150    1.04419    0.98426
 4.2    0.97676    1.14902    1.02000    1.04623    0.98589
 4.4    0.97624    1.15159    1.01837    1.04810    0.98741
 4.6    0.97572    1.15397    1.01662    1.04981    0.98884
 4.8    0.97520    1.15618    1.01475    1.05138    0.99019
 5.0    0.97469    1.15824    1.01279    1.05284    0.99146
```

```
VAPORISATION EFFICIENCIES.
PLATE= 8                                         COMPONENT
TIME        1          2          3          4          5
 0.0    1.05000    1.02000    1.05000    0.99000    0.96000
 0.1    1.04525    1.02953    1.05000    0.99001    0.96100
 0.2    1.04095    1.03822    1.04998    0.99004    0.96201
 0.3    1.03705    1.04618    1.04996    0.99022    0.96303
 0.4    1.03350    1.05350    1.04993    0.99083    0.96405
 0.5    1.03026    1.06026    1.04989    0.99208    0.96507
 0.6    1.02729    1.06654    1.04984    0.99392    0.96610
 0.7    1.02457    1.07238    1.04978    0.99623    0.96714
 0.8    1.02208    1.07783    1.04971    0.99885    0.96819
 0.9    1.01979    1.08294    1.04964    1.00164    0.96924
 1.0    1.01768    1.08775    1.04955    1.00453    0.97030
 1.1    1.01574    1.09228    1.04946    1.00744    0.97136
 1.2    1.01395    1.09657    1.04935    1.01033    0.97243
 1.3    1.01230    1.10063    1.04924    1.01317    0.97351
 1.4    1.01078    1.10449    1.04912    1.01593    0.97460
 1.5    1.00938    1.10817    1.04899    1.01862    0.97569
 1.6    1.00810    1.11169    1.04885    1.02123    0.97679
 1.7    1.00691    1.11506    1.04871    1.02374    0.97790
 1.8    1.00583    1.11829    1.04855    1.02618    0.97901
 1.9    1.00483    1.12140    1.04839    1.02854    0.98013
 2.0    1.00392    1.12439    1.04821    1.03081    0.98125
 2.2    1.00233    1.13008    1.04784    1.03516    0.98353
 2.4    1.00103    1.13541    1.04743    1.03924    0.98582
 2.6    0.99997    1.14045    1.04699    1.04311    0.98815
 2.8    0.99914    1.14524    1.04652    1.04679    0.99049
 3.0    0.99852    1.14981    1.04602    1.05030    0.99286
 3.2    0.99808    1.15420    1.04548    1.05368    0.99525
 3.4    0.99781    1.15844    1.04491    1.05694    0.99766
 3.6    0.99770    1.16253    1.04431    1.06011    1.00009
 3.8    0.99772    1.16651    1.04368    1.06318    1.00254
 4.0    0.99787    1.17039    1.04302    1.06619    1.00500
 4.2    0.99814    1.17417    1.04233    1.06913    1.00747
 4.4    0.99852    1.17787    1.04161    1.07202    1.00995
 4.6    0.99900    1.18150    1.04087    1.07486    1.01244
 4.8    0.99957    1.18507    1.04010    1.07765    1.01493
 5.0    1.00022    1.18858    1.03931    1.08041    1.01743
```

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VAPORISATION EFFICIENCIES.
CONDENSER.                                  COMPONENT
TIME       1          2          3          4          5
 0.0    1.05000    1.02000    1.05000    0.99000    0.96000
 0.1    1.05474    1.03888    1.05953    0.99900    0.96973
 0.2    1.05826    1.05548    1.06744    1.00650    0.97801
 0.3    1.06089    1.07023    1.07409    1.01298    0.98516
 0.4    1.06286    1.08343    1.07975    1.01898    0.99143
 0.5    1.06433    1.09533    1.08461    1.02489    0.99699
 0.6    1.06543    1.10613    1.08882    1.03083    1.00197
 0.7    1.06624    1.11599    1.09247    1.03674    1.00647
 0.8    1.06683    1.12502    1.09567    1.04257    1.01057
 0.9    1.06723    1.13332    1.09847    1.04824    1.01433
 1.0    1.06750    1.14100    1.10095    1.05371    1.01780
 1.1    1.06766    1.14811    1.10310    1.05894    1.02102
 1.2    1.06773    1.15473    1.10501    1.06592    1.02401
 1.3    1.06773    1.16089    1.10669    1.06864    1.02682
 1.4    1.06767    1.16666    1.10817    1.07511    1.02945
 1.5    1.06757    1.17206    1.10946    1.07734    1.03194
 1.6    1.06744    1.17713    1.11059    1.08134    1.03429
 1.7    1.06727    1.18190    1.11157    1.08511    1.03651
 1.8    1.06709    1.18640    1.11241    1.08868    1.03863
 1.9    1.06688    1.19065    1.11315    1.09205    1.04065
 2.0    1.06667    1.19467    1.11375    1.09524    1.04258
 2.2    1.06621    1.20209    1.11461    1.10112    1.04620
 2.4    1.06572    1.20879    1.11513    1.10641    1.04954
 2.6    1.06523    1.21487    1.11532    1.11118    1.05263
 2.8    1.06474    1.22042    1.11522    1.11551    1.05552
 3.0    1.06425    1.22550    1.11487    1.11944    1.05822
 3.2    1.06377    1.23017    1.11429    1.12303    1.06076
 3.4    1.06331    1.23447    1.11350    1.12632    1.06315
 3.6    1.06286    1.23846    1.11251    1.12934    1.06541
 3.8    1.06242    1.24216    1.11136    1.13213    1.06755
 4.0    1.06200    1.24560    1.11005    1.13471    1.06958
 4.2    1.06159    1.24881    1.10859    1.13709    1.07151
 4.4    1.06120    1.25181    1.10700    1.13931    1.07335
 4.6    1.06083    1.25463    1.10529    1.14138    1.07510
 4.8    1.06047    1.25727    1.10347    1.14531    1.07677
 5.0    1.06012    1.25976    1.10156    1.14512    1.07836
```