

AN EXPERIMENTAL STUDY OF THE EFFICIENCY
OF DISTILLATION PROCESSES.

A Thesis submitted for the
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-by-
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ABSTRACT.

In distillation processes it is necessary that there always be heat as well as mass transfer. In this work the effect of the heat transfer process upon the transfer of components between phases and on the performance of equipment has been studied.

Two distillation units have been built and used: a short packed column, designed only to measure the way in which the efficiency of its separation of binary mixtures changes with composition; and a second unit, designed to give those measurements needed for understanding the heat transfer process.

Theory is developed in which it is postulated that condensation or evaporation in the bulk phases or at the interface can take place. It is shown that by taking account of this thermal distillation it is possible to explain experimental results obtained not only in this work but by many other workers: namely that the performance of distillation equipment varies with composition and has a maximum value.

PART I

INTRODUCTION AND LITERATURE SURVEY

INTRODUCTION.

For many years distillation has been studied as a straightforward process of mass transfer by diffusion and the fact that experimental results obtained by many workers gave values inconsistent with this view has been explained in several ways. For instance, modifications of the simple diffusional theory, such as those depending upon surface renewal or interface resistance have been proposed; or the difference between real and theoretical behaviour has been ascribed to the complex motion of the phases through the equipment.

It is unfortunate that distillation is one of the most difficult of the mass transfer processes to study experimentally; it involves a delicate and self-regulating balance between mass and heat transfer, and any disturbance of this balance markedly affects the operation of the equipment. For this reason accurate observations of the detail of its internal mechanism has not been achieved, and only external measurements of the result of the process, such as the composition gradient of the bulk phases along a column, have been made. Even so, by studying experimental results obtained from packed, wetted wall and plate distillation columns, it seems improbable that their

observed behaviour can be accounted for by those factors considered hitherto [1,2].

Essential to the process of distillation is the transfer of heat simultaneous with the transfer of mass, whereas heat transfer is incidental to most other processes of mass transfer: its effect on all these processes has been largely ignored. In distillation it has either not been taken into account at all, as in the diffusional theories, or it has been assumed that the rate of heat transfer is equivalent to the rate of mass transfer, so as to keep vapour and liquid phases at their dew and boiling points respectively [3]. For reasons which will be detailed later it is this assumption which is being questioned in the present work.

The principal evidence that performance does not conform with theory is provided by the fact that the efficiency of distillation equipment varies widely according to the composition of the mixture being separated. In this work we shall be mainly concerned with continuous counter current distillation columns, for which it is found that the height of a transfer unit, as defined by Chilton and Colburn [4] or the height equivalent to a theoretical plate [5] varies with composition and has a maximum value. This has

been noted by many workers either directly or by implication.

As an example, Keyes and Byman [6] found a sharp increase in the plate efficiency of a bubble-cap column from a value of below 5% to approximately 85% over a range of 10 mole per cent in composition of an ethanol-water mixture. They also found a decrease in efficiency near the azeotropic composition. A rapid decrease in the Murphree efficiency of a bubble-cap plate near the azeotrope of isopropanol-water has also been reported by Langdon and Keyes [7]. Chakvikilian [8] found a linear variation of the H.E.T.P. of a wetted wall column for the system carbon tetrachloride-benzene. In these investigations the slope of the equilibrium curve, the liquid viscosity, the surface tension, the volumetric liquid velocity and the time of contact varied markedly at different compositions. For this reason it was thought that the reported influence of concentration might partly be due to the variation of these properties. It may be noted that a small error in the location of the equilibrium curve could produce inaccuracies in the calculation of the plate efficiencies, especially near the azeotropic composition. Wijk and Thyssen [9], however, as a result of a careful study of the relationship between plate efficiency and liquid composition in a sieve-plate

column, using the system n-heptane-methylcyclohexane, concluded that such a wide variation in performance could not be attributed either to experimental error or to inaccurate poor equilibrium data. Shilling, Beyer and Watson [10] concluded that the rapid fall-off of efficiency when separating dilute mixtures is caused by the rapid increase in the effective resistance of the liquid film as the slope of the equilibrium curve increases. The variation and decrease of the efficiency of distillation equipment at extremes of the concentration range has also been noted by many others [1,2,10,11,12,13].

Sawitowski [2] pointed out that since the same shape of curve is obtained when efficiency is plotted against composition for many different types of equipment, the hydrodynamic behaviour of the phases, so different in packed, plate or wetted-wall columns, cannot be the main reason why there is a variation of performance with composition. He suggested three possible explanations for this phenomenon : namely (1) that the driving force is not the molar concentration difference of the bulk phases, (2) that there is a finite resistance to transfer at the interface where equilibrium conditions are not maintained, and (3) that there may be a change in interfacial area which depends on

concentration, owing to a change in physical properties, such as surface tension. He later rejected all these postulations on the grounds that none of them could account for the magnitude of the variation in performance observed experimentally.

The direct application of the two-film theory (values for film coefficients being taken from absorption work) and several other mechanisms [4,14] have been examined both theoretically and experimentally for distillation in a packed column by Smith [1], who also concluded that none of them could represent the process satisfactorily. It was suggested that the reason for this failure is in part owing to the fact that the flow of the liquid phase occurs as the combination of a large number of small transient pulses, and yet an undisturbed steady state flow in the column is usually assumed. Smith also pointed out that besides mass transfer by contact distillation (that is, under control of a concentration gradient across the phases) there is a possibility that there will occur thermal distillation: that is, direct condensation or evaporation of the vapour or liquid phases. Since a quantitative assessment of the effects of thermal distillation on the mass transfer process is the principal concern of this thesis, we shall outline

here the way in which it arises and the way in which it affects the separation.

The first to introduce a theory based upon heat rather than mass transfer was Kirschbaum [3]. He assumed that :

(1) the heat transfer coefficient between the phases is independent of the temperature difference,

(2) heat is transferred across the interface without any transfer of mass,

(3) the temperatures of both phases at the interface is the same, that is, there are equilibrium conditions, and (4) there is no concentration gradient in the vapour.

The mass transfer is purely by condensation of vapour and evaporation of liquid in the bulk phases respectively and is thus controlled by the overall temperature difference and heat transfer coefficient. He obtained an expression for the efficiency of separation as follows :

$$\int_{y_B}^{y_T} \frac{dy}{(y_e - x_e) \theta} = \frac{U a S Z}{G} \quad (1)$$

where θ = the overall temperature difference between the bulk phases,

U = the overall heat transfer coefficient,

G = molal rate of flow of vapour per unit cross-sectional area of column,

Z = height of packing,

a = interfacial area per unit volume of column,

S = cross-sectional area of column

subscript e = equilibrium value.

In effect Kirschbaum's theory does not differ from the mass transfer theories, for it assumes implicitly that the vapour and liquid are at their dew and boiling points respectively. The one to one correspondence between these temperatures and composition means that either side of equation (1) can be converted to the number of transfer units. (Kirschbaum has recently modified this theory [69]).

In fact the temperatures of the bulk phases must be dependent upon the rate of heat transfer and it is unlikely that the rate of heat transfer should correspond to the rate of mass transfer in the way assumed by Kirschbaum. If the rate of heat transfer is small, the temperature difference between the phases will not be able to remove all the heat needed to keep the vapour at its condensation point or to supply all the heat needed to keep the liquid at its boiling point: then the bulk vapour will superheat while the bulk liquid will subcool. Under these circumstances a distillation process by contact transfer will take place between the two phases not at their saturation temperatures. The more likely case will occur when the heat transfer

coefficient is large. Then the bulk phases will be at their condensation and boiling points, and excess transfer of heat beyond that needed to ensure this will cause condensation in the bulk vapour and evaporation in the bulk liquid. This is thermal distillation, which will occur as well as contact distillation.

A net transfer of mass by thermal distillation may also occur at the interface. Thus, if the rate of heat transfer through the vapour phase is smaller than that through the liquid phase by reason of the higher conductivity of the liquid there will be a net condensation.

Two distillation units have been built and used in the course of this work. The first has a short length of packed column and was used solely to measure values of the performance as a function of composition. The second, called the pear column by reason of its shape, is designed to measure the temperature of both phases at points along its length. Facilities were added so that the change in flow rate along the column could also be measured. This column was also used to give performance figures, but was designed in the first place to check quantitatively the theory of thermal distillation developed during the course of the work.

CHAPTER 1.

A CRITICAL SURVEY OF PREVIOUS WORK.

In the Introduction we have referred briefly to those aspects of the theory of distillation with which this thesis is concerned. In this chapter those parts of mass transfer theory which are related to the distillation process are discussed. It is envisaged that a mechanism of thermal distillation is superposed on one or other of these processes, and accounts only in part for the observed behaviour of distillation columns.

Basically theories of mass transfer assume diffusion through stagnant layers of fluid adjacent to the interface, the bulk of the phases being in turbulent motion. These conditions are possibly fulfilled in the case of liquid-solid and gas-solid systems, but are unlikely in a gas-liquid system, owing to the higher mobility of the phases. The application of these postulates to distillation results is even less successful than for absorption. The empirical correlation of experimental values of performance of packed columns is usually made in terms of the H.E.T.P. (height equivalent to a theoretical plate) first introduced by Peters [5] and involves the application of a stagewise

process to the apparently continuous process of a packed column. The use of the height of a transfer unit (H.T.U.) although more acceptable theoretically has no practical advantage over the H.E.T.P. as far as the empirical treatment of results is concerned, since it is found that both vary widely, not only from one piece of equipment to another, but within the same binary system. Apart from the theory of Kirschbaum which has already been discussed, no attempt has been made until recently to deal with distillation other than in terms of the two-film theory and its extensions.

A. THE TWO-FILM THEORY.

Whitman [14] proposed that the rate of transfer of a component between phases was limited by diffusion. He assumed :

- (1) laminar flow in a boundary layer adjacent to the interface and turbulent motion in the bulk phases when the composition was uniform at a given cross-section,
- (2) that the interface offers no resistance to the transfer of diffusing solute, that is, the concentrations at the interface are in equilibrium,

and (3) that the effective thickness of the boundary films is so small that the solute content of the films themselves may be considered to be negligible.

With these conditions the rate of transfer in each phase is proportional to the difference between the bulk concentration and the concentration at the interface (Fig.1.1), and in a steady state the rates of transfer in the two phases are equal.

$$N_A = k_g (y_i - y) = k_l (x - x_i) \quad (1.1)$$

where N_A = rate of mass transfer of component A in moles per unit area and time,

k_g, k_l = film coefficient of mass transfer in gas and liquid respectively,

y, x = mole fraction of the diffusing component in vapour and liquid respectively.

subscript i represents the interface.

This basic theory has undergone considerable development (v.i.)

B. J FACTORS IN MASS TRANSFER.

The idea of the transfer unit as a measure of the performance of continuous countercurrent equipment is due to Chilton and Colburn [4]. For distillation they applied the two-film theory and assumed constant molar overflow, that is, equi-molar latent heats of vaporisation of the

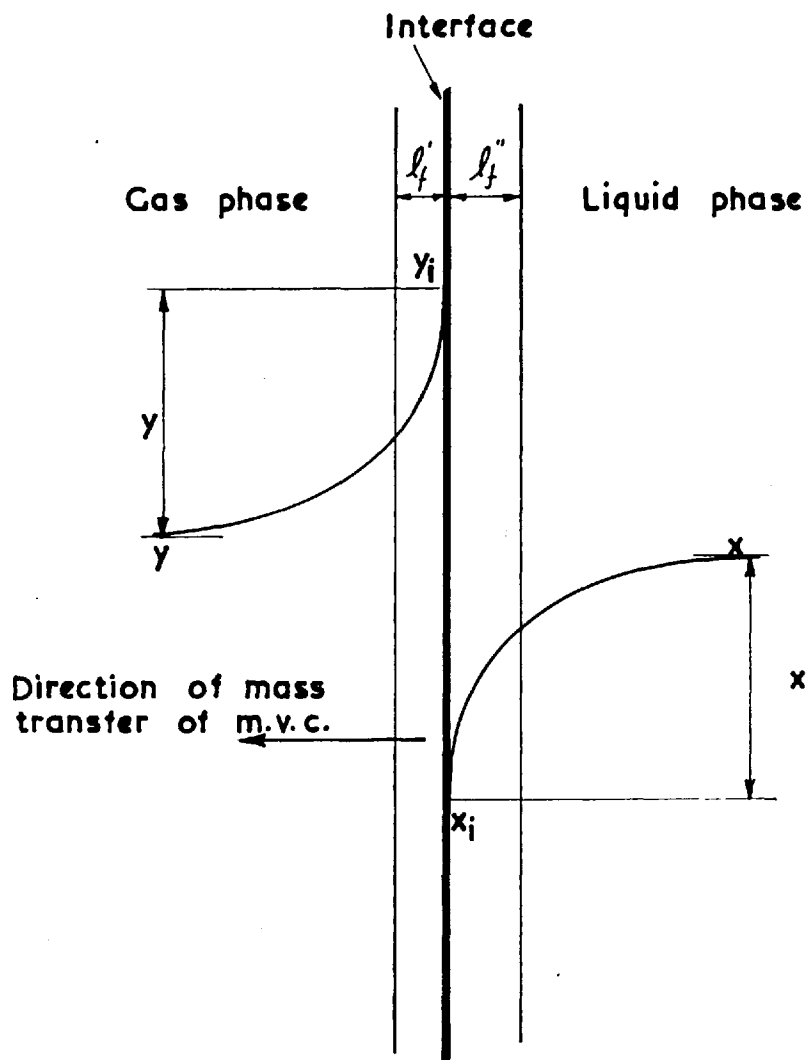


FIG. 1-1. CONCENTRATION GRADIENTS
BETWEEN TWO BULK PHASES

components.

In a differential height dZ of the vapour phase :

$$dN_g = k_g aS (y_i - y) dZ = G dy \quad (1.2)$$

In the liquid phase :

$$dN_l = k_l aS (x - x_i) dZ = L dx \quad (1.3)$$

where a = interfacial area per unit volume of packing,
 S = cross-sectional area of column,
 Z = height of packing.

The number of transfer units, N_G and N_L in the vapour and liquid phase, respectively, are then given by the relationships :

$$N_G = \int \frac{dy}{y_i - y} = \frac{k_g aS Z}{G} \quad (1.4)$$

$$N_L = \int \frac{dx}{x - x_i} = \frac{k_l aSZ}{L} \quad (1.5)$$

The height of a transfer unit (H_G and H_L in vapour and liquid phase) is that height of packing which gives a separation corresponding to one transfer unit.

$$H_G = \frac{Z}{N_G} = \frac{G}{k_g aS} \quad (1.6)$$

$$H_L = \frac{Z}{N_L} = \frac{L}{k_l aS} \quad (1.7)$$

Experimentally it is difficult to measure the interface concentrations. To avoid this an overall mass-transfer coefficient was introduced:

$$N_{OG} = \int \frac{dy}{y_e - y} = \frac{K_G aSZ}{G} \quad (1.8)$$

$$N_{OL} = \int \frac{dx}{x - x_e} = \frac{K_L aSZ}{L} \quad (1.9)$$

$$\text{and } H_{OG} = \frac{G}{K_G aS} \quad (1.10)$$

$$H_{OL} = \frac{L}{K_L aS} \quad (1.11)$$

where K_G , K_L = overall mass transfer coefficient in gas and liquid respectively,

H_{OG} , H_{OL} = overall height of transfer unit in gas and liquid respectively,

and N_{OG} , N_{OL} = overall values of number of transfer unit.

subscript e indicates the equilibrium value of the composition of the other phase.

Overall values and film values are related as follows :

(See Fig.1.2, points B and C).

$$H_{OG} = H_G + m \frac{G}{L} H_L \quad (1.12)$$

$$H_{OL} = H_L + \frac{L}{m G} H_G \quad (1.13)$$

where m is the slope of the equilibrium curve.

In distillation it is usually thought that the vapour film offers the major resistance.

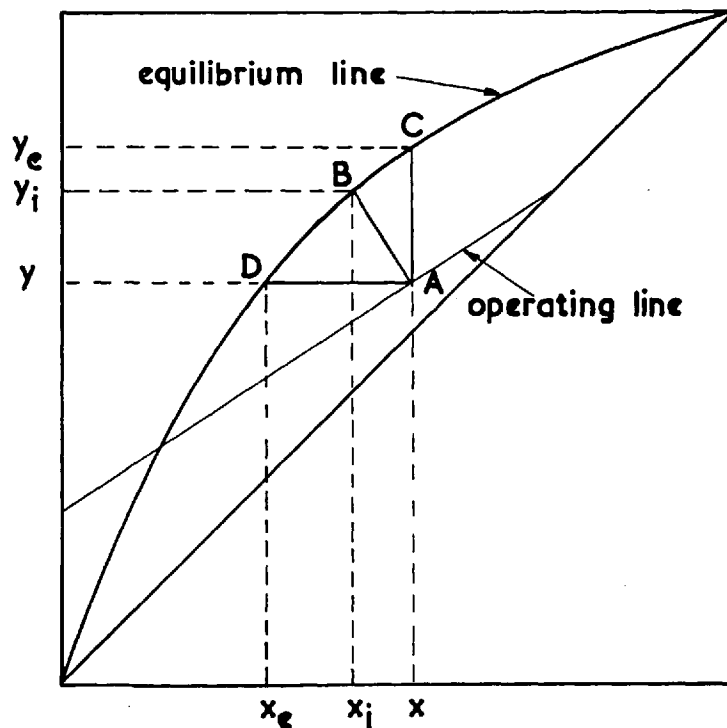


FIG.1-2. EQUILIBRIUM DISTRIBUTION
CURVE FOR BINARY SYSTEM

From the analogy between heat and mass transfer, Chilton and Colburn [15] suggested that mass-transfer data could be expressed in terms of mass-transfer factors

$$j = \frac{K_{OG} S}{G} \left(\frac{\mu}{\rho D} \right)^{2/3} = f(Re) \quad (1.14)$$

$$\text{or } j = \frac{K_{OG} S}{G} (Sc)^{2/3} = f(Re) \quad (1.15)$$

where μ = viscosity taken at film conditions, lb/hr/ft.

ρ = density, lb/ft³,

D = diffusivity, ft²/hr,

and Sc = Schmidt number,

$$\therefore H_{OG} = \frac{(Sc)^{2/3}}{j a} \quad (1.16)$$

For cases of flow inside tubes, across tubes and parallel to plane surfaces, they correlated j factors for mass transfer with Reynolds number and found they corresponded to j factors for heat transfer across surfaces 100 per cent effective.

The correlation was less successful for packed columns, owing partly to the fact that the area of contact varies with the liquid rate. For instance, Gamson, Thodos and Hougen [16] in their investigation of heat, mass and momentum transfer in the flow of gases through granular solids

reported the following relationships between the j factors and Reynolds number for turbulent flow ($Re > 350$).

$$j_D = \frac{k_g M_m p_{gf}}{G} \left(\frac{\mu}{\rho D_v} \right)_f^{2/3} = 0.989 \left(\frac{D_p G}{\mu} \right)^{-0.41} \quad (1.17)$$

where k_g = film coefficient on the gas phase,

M_m = average molecular weight,

p_{gf} = partial pressure of diffusing solute, atm.,

G = mass flow rate of vapour, lb/hr/ft²,

μ = absolute viscosity, lb/hr/ft.,

ρ = density, lb/ft³,

D_v = diffusivity of vapour, ft²/hr,

and D_p = average diameter of packing particles.

subscript f refers to gas film.

Also Wilke and Hougen [17] proposed the following relationship for laminar flow ($Re < 350$) :

$$j_D = \frac{k_g M_m p_{gf}}{G} \left(\frac{\mu}{\rho D_v} \right)_f^{2/3} = 1.82 \left(\frac{D_p G}{\mu} \right)^{-0.51} \quad (1.18)$$

From equations (1.17) and (1.18), the gas film coefficient can be calculated (calculating the Reynolds number using an equivalent diameter when open packings are used). Similar correlations have been developed for the liquid film resistance.

We note that according to equation (1.12), when H_{OG} is plotted vs. m at total reflux ($G/L = 1$), a straight

line should be obtained, the slope of which should equal H_L and the intercept on the ordinate axis H_G (v.i). This straight line, however, is not obtained [1,2,18]. Negative slopes and intercepts are found. Even allowing for the fact that the film coefficients vary with composition, this result implies negative coefficients of mass transfer, which is meaningless.

C. THE SURFACE RENEWAL THEORIES.

The first theory to take account of the complex motion of the phases is due to Higbie [19]. He derived what is called the penetration theory in 1935, although von Wroblewski [20], Stefan [21], Lederer [22], and Tammann and Jessen [23] had developed similar theories previously. In the penetration theory, Higbie assumed that the liquid film is in laminar flow and has a large thickness compared with the depth of penetration of diffusing material. Mixing occurs regularly at discontinuities in the packing, after which fresh surface is presented to the vapour. Each element of surface is exposed to the vapour for the same length of time, during which an unsteady mass-transfer process takes place by molecular diffusion.

The amount of gas absorbed up to the time of exposure t_e is

$$S = 2 (C_e - C_o) \sqrt{\frac{D t_e}{\pi}} \quad (1.19)$$

where S = gas absorbed or diffused in time t_e ,
 C_o = initial concentration of the solution,
 C_e = concentration of the solution corresponding to partial pressure of solute gas,
 D = diffusivity.

Thus, the liquid film coefficient in concentration units is

$$k_1 = 2 \sqrt{\frac{D}{\pi t_e}} \quad (1.20)$$

that is, it varies inversely as the square root of the time of exposure. The penetration theory gave a more realistic picture of the hydrodynamic conditions, at the interface, it is not valid when the time of contact is short (e.g. in packed columns). Marshall and Pigford [24] showed that equation (1.20) is applicable for the case of absorption of a pure gas, when the concentration at the interface of the liquid remains constant during the time of exposure, t_e .

Danckwerts [25] modified the penetration theory and assumed

- (1) constant mean rate of surface renewal which is independent of the time of exposure,
- (2) laminar flow of liquid film which has very large thickness compared with the depth of penetration of diffusing material,

and (3) regular mixing at discontinuities in the packing. He obtained the rate of mass transfer in terms of the concentration driving force and the mean rate of surface renewal. The liquid film mass transfer coefficient was given by

$$k_1 = \sqrt{DS} \quad (1.21)$$

where S = mean rate of liquid surface renewal per unit total surface.

With an interface resistance or gas-phase resistance, the usual equation (1.12) for the addition of resistance is still applicable. Kishinevskii [26] criticised the assumption of molecular diffusion in all these theories and proposed a convective process. Horne [27] found that if the rate of surface renewal is high and the average time of exposure of the surface is small in comparison with the total time of contact, then the error, which depends on the rate of surface renewal, produced by the use of Danckwerts' equation (1.21) will be negligible.

In fact, although these theories correspond more realistically to conditions in equipment, they still do not explain the form nor account for the magnitude of the variation in performance in a distillation mixture [1,2].

D. SURFACE TENSION.

Zuiderweg and Harmens [12] pointed out that surface tension plays an important role in determining the rate of transfer in distillation, through its effect on the wetting of the packing and hence on the interfacial area. They found that liquid films are stabilised when the surface tension of the reflux increases as it passes down the column: otherwise the liquid stream breaks up into narrow rivulets or droplets. Thus the interfacial area of the contact surface is increased or decreased depending upon the way in which the surface tension of a mixture varies with composition. In addition to this effect Sawistowski [28] pointed out that a variation of surface tension with composition may induce mixing and surface renewal in the liquid phase.

The effect of surface tension on the rate of mass transfer has been investigated by many workers in absorption and liquid-liquid extraction, usually by adding surface-active agents. It is found that a surface-active agent retards the rate of diffusion owing to interactions between the solute and the surface film [29]. The increase in interfacial area of contact caused by the lowering of interfacial tension by the surface-active agent decreases

the H.T.U. in liquid-liquid extraction [30]. Bond and Donald [31] found that the surface tension between the gas and liquid plays an important role in breaking the surface of the film.

Several other workers [32, 33, 34, 35, 36] have found similar results, but mainly this work is not relevant to distillation, or at least it is not the major factor affecting the rate of transfer.

That there may be a resistance at the interface in distillation process has been suggested in the past by many workers [2,12]. There appears to be no evidence for this, and it has been shown that in the absence of foreign material at the interface there is no resistance in absorption processes.

PART II

EXPERIMENTAL SECTION

Two distillation units have been built and used during the course of this work. The first unit was designed to measure the efficiency of separation (H_{OG}) over a small concentration interval and was a short packed column. The results obtained from this column were used to confirm and to extend data obtained by other workers [1,2,11,12,37]. The second unit was specifically designed to give measurements needed for understanding thermal distillation (v.i.); in particular, the change in flow rate along the height of the column and the temperatures of the two phases. This column consisted of 12 pear-shaped bulbs and is known as the pear column.

CHAPTER 2.

PACKED COLUMN DISTILLATION UNIT.

A. DESIGN AND DESCRIPTION OF THE EQUIPMENT.

In the design of this equipment (see Fig.2.1), several points were borne in mind: namely, that the distillation process was to be operated under total reflux; that it was important to approach adiabatic conditions closely; and that point values of the efficiency were to be measured (here found by measuring the separation over a small concentration interval). The unit consists of three main parts: the still, the column and the condenser and reflux equipment. All sections are made of glass.

(i) Still. The still is a 2-litre capacity, 3-neck pyrex flask. The central neck is connected to the vapour line, one of the side necks is attached to a flowmeter which measures the flow of liquid returned to the still, and the other side neck is used for filling and emptying the still.

(ii) Column. The column is made of a 1" i.d. glass tube of height 10" in which there is an effective packed length of 6". A packing support of wire gauze rests on four indentations near the lower end. A sidearm just below the packing support is fitted to hold the lower sampling device.

FIG. 2.3

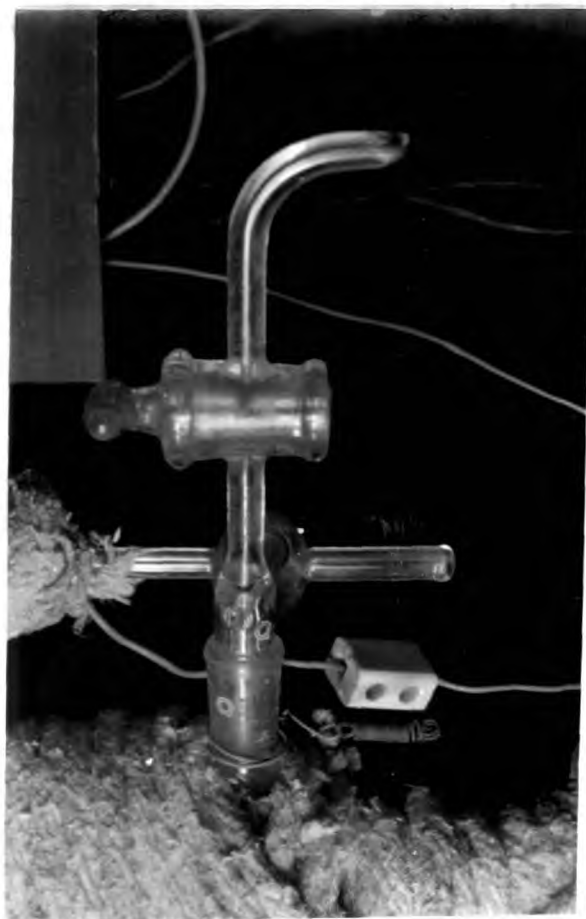
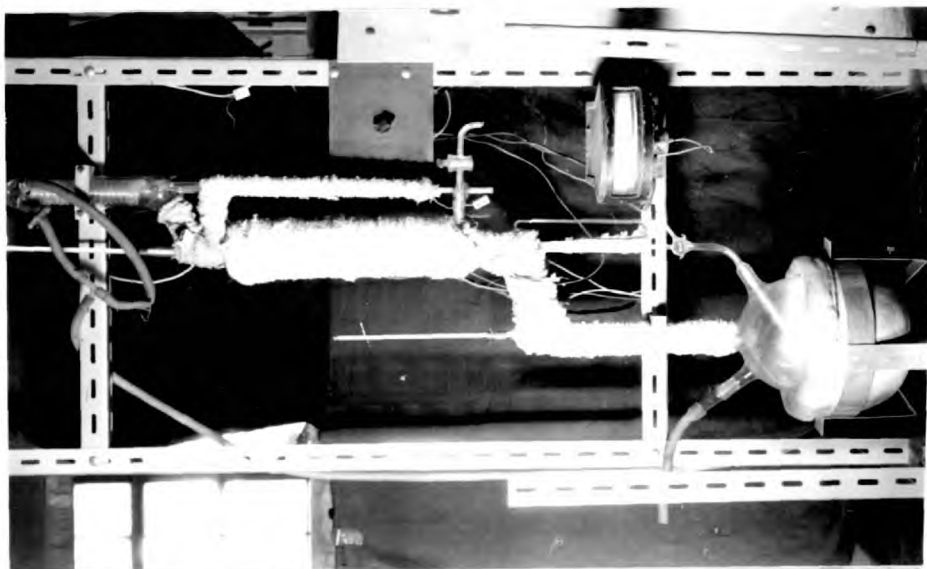


FIG. 2.2



At first results obtained from the 6" height column were not reproducible and several modifications to the apparatus were made. In particular the height of packing was increased from 6" to 10" to give a bigger separation of the components and to reduce the end effects in the column.

(iii) Condenser and reflux return line. A large double-coiled condenser was used to ensure total condensation of the vapour. The reflux was fed to the top of the column through a U-tube, having one arm wound with heating wire in order to raise the temperature of the condensate to its boiling point before returning it to the column.

(iv) Lagging. It is pointed out by Rene Romanet [38] and other workers that non-adiabatic operation of a distillation column will greatly influence the efficiency of separation. In this work we are interested especially in the effects of internal heat transfer on the distillation process and a heavily lagged column is therefore essential. With small columns it is necessary, in addition, to obtain adiabatic conditions by supplying heat to compensate for any heat loss to the surroundings.

The column is lagged with asbestos as follows:

asbestos cloth is wrapped round the glass tube followed by asbestos rope, paper, cloth and finally rope. Heating wire is wound between the asbestos paper and cloth. That adiabatic conditions have been attained is checked by thermocouples (v.i.).

The vapour line from the top of the column to the double-coiled condenser is heavily lagged and also wound with Brightray coiled wire to ensure that the vapour here is superheated and will not condense: this prevents rectification occurring along the line.

The vapour line from the still is lagged with two layers of asbestos rope in order to conserve heat. The reflux line is also lagged with asbestos rope. Temporary lagging is provided at joints for easy dismantling.

(v) Electrical heating.

(a) Still. The still is heated electrically with a 2L-isomantle of output power 435 watts. The supply is controlled by a variac.

(b) Column. It is known from experience that 100 watts is more than sufficient to compensate for heat losses from a column of this surface area. At 230 volts supply, a current 0.435 amps. is required (resistance of wire required 529 ohms). 36 S.W.G. Brightray wire is used,

resistance 11.4 ohms/ft. The maximum permissible current at a temperature of 300°C. is 0.7 amps. With a variac connected in the circuit only a 20 ft. length of heating wire is used (maximum current 0.66 amps.). It is wound round the internal lagging at uniform pitch.

(c) Vapour line to the condenser. It was found that 50 watts is sufficient to prevent any condensation along the vapour line. Therefore at 230 V. the current required is 0.217 amps. and the resistance 1060 ohms. Coiled wire is used for this section, its resistance when cold being 300 ohms/yd. A 6 ft. length of wire is used and a rheostat is connected in series in the circuit.

(d) Reflux line. The heating wire required in this section was calculated as follows. The maximum value of the product of specific heat and specific gravity of mixtures to be used in this equipment is 0.5 cal/cc.°C. If the maximum subcooling of condensed liquid is 100°C., and the reflux flow rate is 30 cc/min., then,

$$\begin{aligned}\text{heat energy required} &= 30/60 \times 0.5 \times 100 \text{ cal./sec.} \\ &= 104.8 \text{ watts.}\end{aligned}$$

In order to allow for heat loss to the surroundings, 150 watts of energy was supplied. A 4 ft. length of coiled wire of 300 ohms/yd. resistance is used and wound in uniform pitch. A rheostat is connected in series.

(vi) Thermocouples. Two copper-constantan thermocouples previously tested to give, at the same temperature, the same reading on a galvanometer, are used for checking the adiabatic condition of the column. They are installed in the positions shown in Fig.2.4. It is important that the outer thermocouple should be situated just inside but not in contact with the heating wire [1].

(vii) Sampling device. Samples of liquid are taken from two places in the equipment. A sample giving the composition at the top of the packing is taken from the lower end of the U-tube reflux line, since there is total condensation in the condenser and no rectification takes place in the vapour line. The bottom sample is taken through a tube inserted ^{through} the sidearm near the base of the packing (Fig.2.1). This tube has its top half cut away to form a trough to collect liquid from the centre of the packing support. The liquid is forced through a capillary tube by the pressure inside the column which is slightly higher than atmospheric. When not in use, the sampling tube is turned so that its trough faces downwards and collects no liquid. By cutting the trough in the sampling tube so that it lay at the centre of the column, it was ensured that liquid was removed from the centre of the packing and not from the wall of the column.

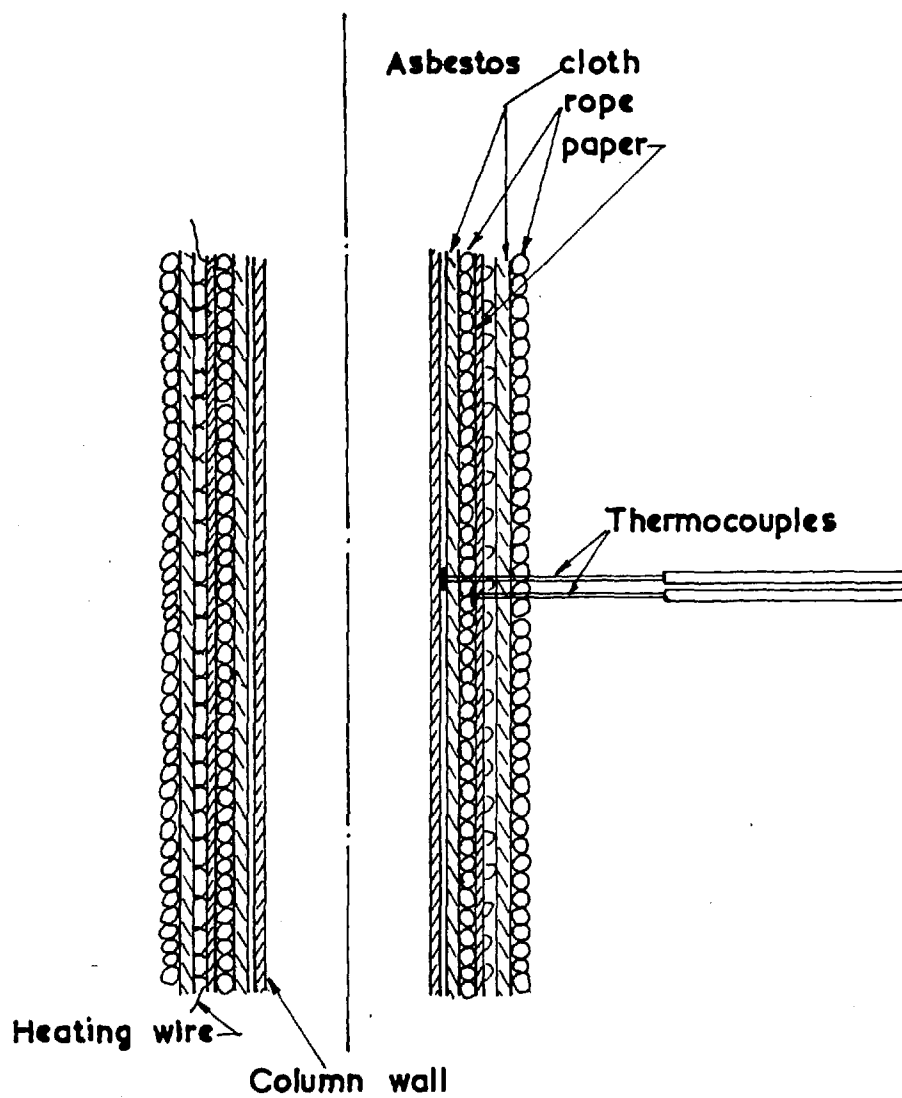


FIG. 2·4.

(viii) Packing. Fenske helices made of single-turns of glass are used as the packing material. The packing is supported by a stainless steel wire gauze, cone-shaped, resting on the indentations at the lower end of the column.

(ix) Liquid distributor. In order to give uniform wetting of the packing and to prevent liquid from flowing along the wall, two liquid distributors are installed between the packings at equal intervals. They are made from stainless steel wire gauze and are cone-shaped.

(x) Flowmeter. A capillary-type flowmeter [1] (Fig.2.5) is used. It is placed between the column and the still and measures the liquid flow at the bottom of the column. At steady flow, the head in the flowmeter is balanced by the head losses owing to friction in the movement of liquid through capillary, sudden contraction and sudden enlargement, giving [1]

$$Q = -\frac{A}{2B} \gamma + \sqrt{\left(\frac{A}{2B}\right)^2 + \frac{h+C}{B}} \quad (2.1)$$

where h = flowmeter reading,

Q = volume of liquid flow per unit time, cc/min.

and γ = kinematic viscosity in centistoke.

A, B , and C are the flowmeter constants.

The flow rate of liquid through the capillary depends on the

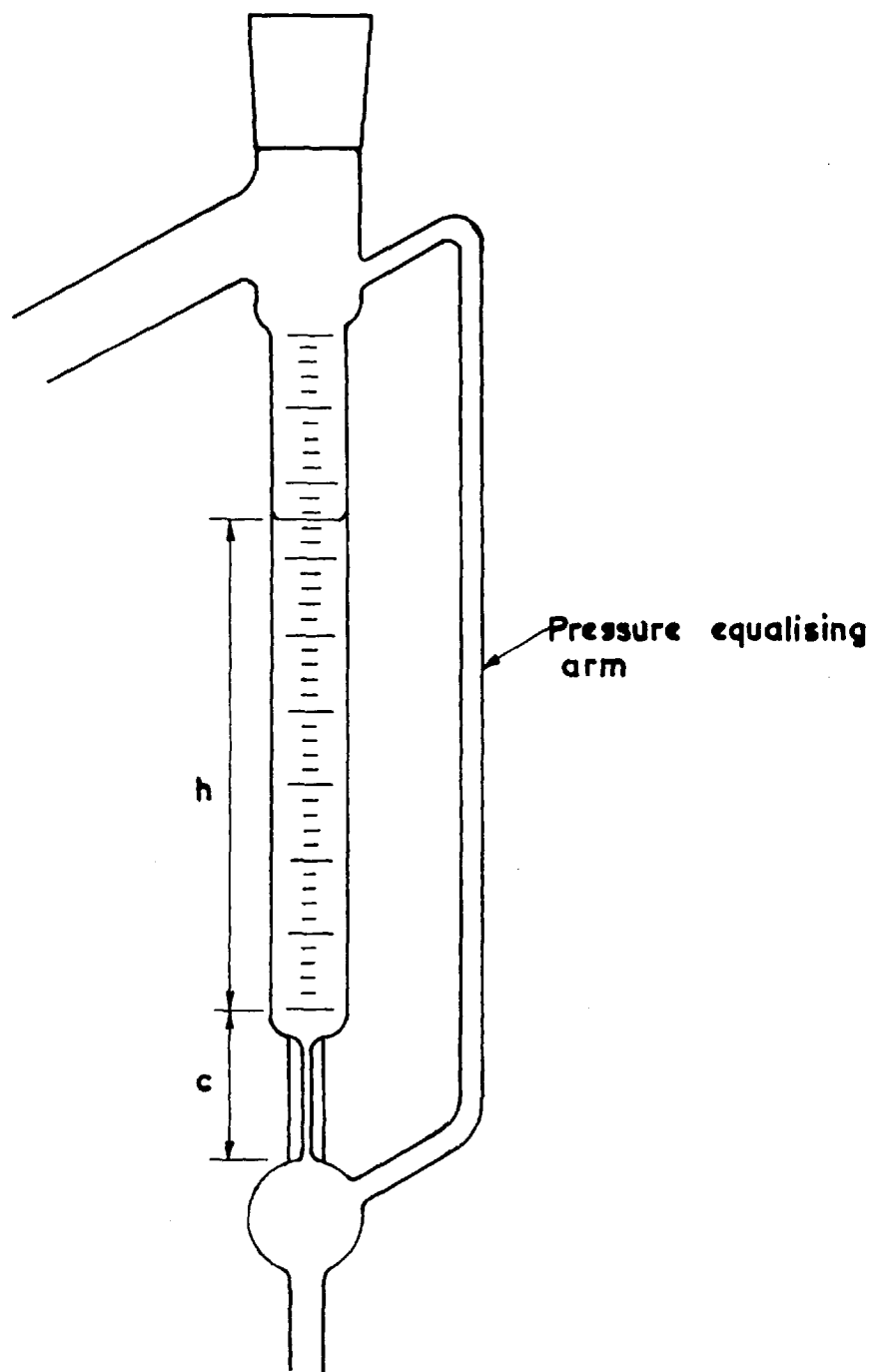


FIG. 2.5. GLASS CAPILLARY
FLOWMETER

liquid head and its kinematic viscosity. The flowmeter constants are independent of the liquid flow rate. The kinematic viscosity, can be obtained by calibrating the flowmeter with liquids of known viscosity. Three liquids: acetone, benzene and chloroform, were used for the calibration and the average values found for the constants were :

TABLE 2.1.

A	B	C	A/B
0.79	0.0198	4.5	3.99

B. PROCEDURE.

The binary systems which were used in the short packed column were carbon tetrachloride-toluene, acetone-benzene and acetone-chloroform. The separation of these mixtures by the column was measured with different concentrations of these mixtures in the still, under total reflux, and at a constant flow rate.

An equipment was started with about 1500 cc. of mixture in the still at a high mole fraction of m.v.c. The column was, at first, packed with 6" of packing in which two cone-shaped wire gauzes were introduced as liquid distributors. The still was heated at full load to make flooding occur inside the column and to ensure a proper

wetting of the packing. The heat load was then adjusted to give a flow rate of 25 cc./min. At the same time heat was supplied to the column and adjusted so that the two thermocouple readings were the same. Sufficient heat was supplied to the vapour line to prevent any condensation. Liquid reflux was heated to a temperature about 5°C. below its boiling point before it was returned to the top of the column. When a steady state was reached (after two hours) the top and bottom samples were collected in two small sampling tubes, the first few cc. of the liquids being discarded before sampling. The still concentration was then altered by adding a quantity of the less volatile component and the heat load was again adjusted to give the same flow rate. The compositions of these samples were determined with an Abbé refractometer kept at a constant temperature; 20°C. for carbon tetrachloride-toluene system, 16°C. for acetone-benzene and 25°C. for acetone-chloroform, according to the data available. For the carbon-tetrachloride-toluene mixture, two types of packings were used: 1/8" dia. Raschig rings and 5/32" dia. (approx.) Fenske helices. For the other two systems, the experiments were carried out using only Fenske helices. A 10" height of packing was used with one binary mixture, acetone-chloroform. This

reduced the experimental error, but introduced to a greater extent the influence of the concentration gradient within the column on the results obtained.

The number of transfer units was calculated by graphical integration

$$N_{OG} = \int_{y_B}^{y_T} \frac{dy}{y_e - y} \quad ,$$

unless the mixture was dilute in either component when it is more accurate to use the following formula (Appendix I):

$$N = \frac{y_T - y_B}{\Delta y_m} \quad (2.2)$$

where

$$\Delta y_m = \frac{(y_{eT} - y_T) - (y_{eB} - y_B)}{\ln \frac{(y_{eT} - y_T)}{(y_{eB} - y_B)}} \quad (2.3)$$

CHAPTER 3.PEAR COLUMN DISTILLATION UNIT.

A. DESIGN AND DESCRIPTION OF THE EQUIPMENT.

This second apparatus was designed specifically to measure certain effects expected as a result of thermal distillation. It was desired to know accurately the temperature of the vapour and liquid streams as well as the composition of the liquid stream at several points in the column. Also a knowledge of the change in flow rate of the vapour or liquid stream along the column was required. Since the theoretical work developed deals with continuous-contact distillation, a continuous and not a stepwise contact between the phases is desirable, but in order to facilitate sampling it is necessary to collect the liquid at several points in the column. By having alternate accumulation and streaming of the liquid the mass transfer process is more rapid, owing to frequent mixing, and gives more accurate results than would, for instance, a wetted-wall column. An apparatus incorporating these features was designed and built, and is described below. The equipment consists of four parts : still, column, vapour line adaptor and condenser. (Fig.3.1).

(i) Still (Fig.3.1). The still is a 3-litre capacity, 2-necked pyrex flask. The central neck is connected to the vapour line, while the side neck is joined to a flowmeter through which liquid leaving the column returns to the still. A thermometer pocket and a condenser used when it is desired to stop the distillation quickly are fitted to the vapour line, as shown.

(ii) Column (Fig.3.1). This is the main part of the equipment. It consists of twelve pear-shaped bulbs, each forming a trough to hold a small volume of liquid for sampling. The column is 3 ft. long and has six sections, each of two bulbs and each section having two side arms. The upper arm is used for the insertion of thermocouples for measuring liquid and vapour phase temperatures and the lower arm is for withdrawing liquid samples.

(iii) Vapour line adaptor. This is placed between the reflux condenser and the column. It acts both as vapour line and as reflux-return line. It is electrically heated to raise the temperature of the reflux condensate to its boiling point before re-entering the column. In the lower part of the line is a trough of small capacity connected to a tap which is used to divert all condensate from the column during the sampling period (Fig.3.1).



FIG. 3.2

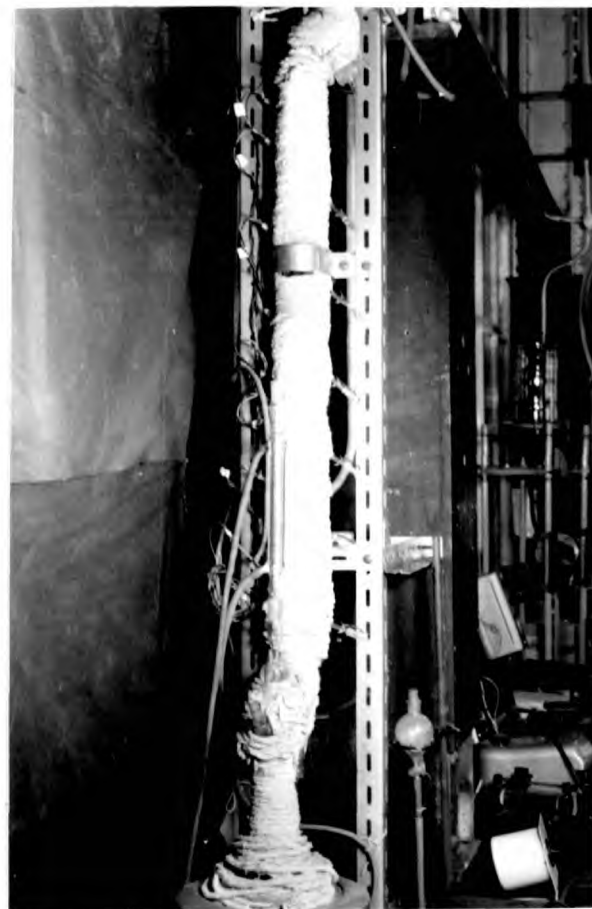


FIG. 3.3



FIG. 3.4



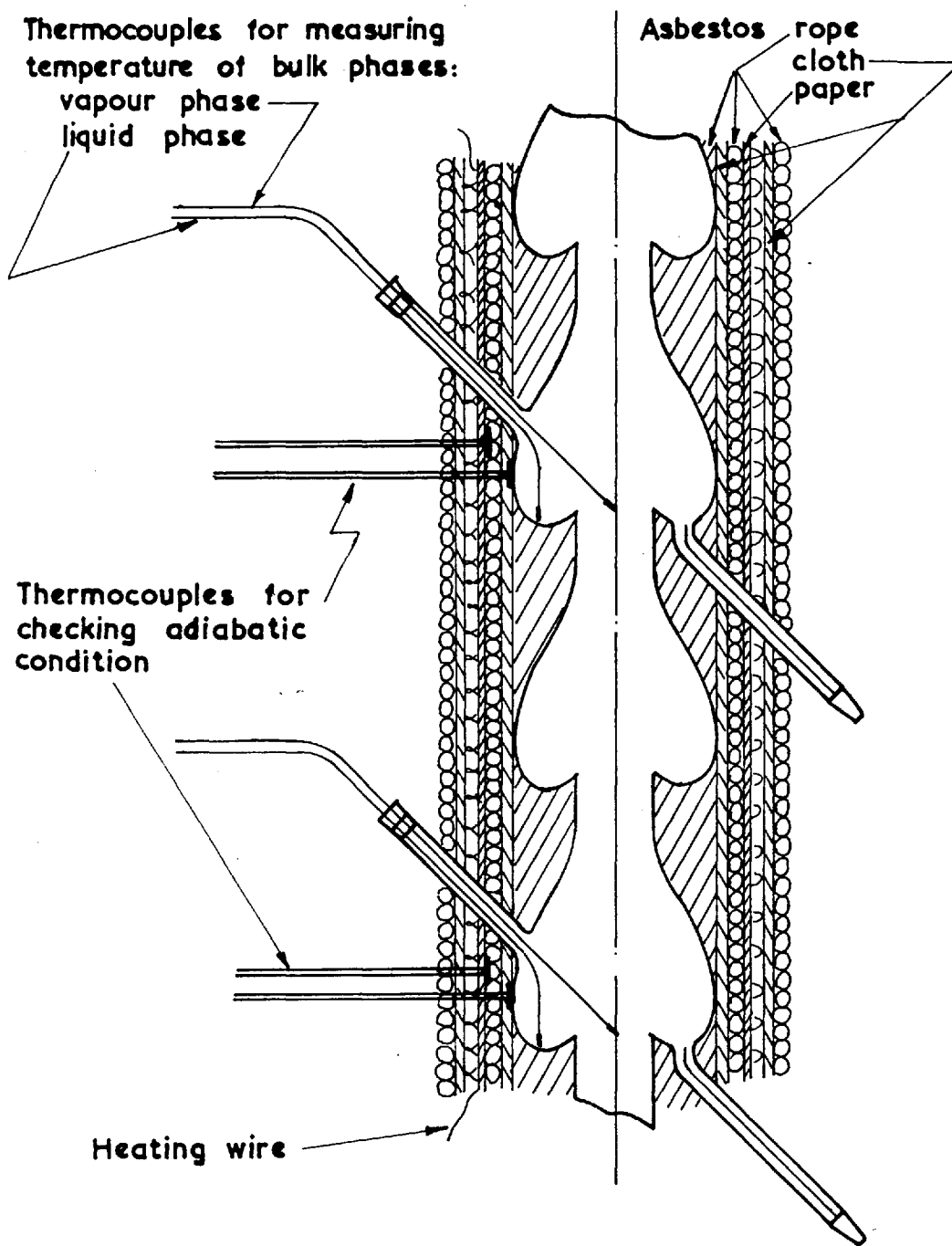
FIG. 3.5

(iv) Reflux condenser. Two water condensers, placed one on top of the other, are used for total condensation of the vapour. By using two condensers, each of which is fed with water separately, some control of the temperature of the returning condenser is obtained.

(v) Lagging. (Fig.3.6). The narrow neck of each pear bulb of the column was first wound with a single string of asbestos rope in order to make more uniform the diameter of the surface on which lagging is placed. The column is lagged as follows: asbestos cloth, rope, paper, heating wire, cloth and finally rope. The column is lagged in six sections, each containing two bulbs, one sampling device and one side arm for thermocouples. As with the packed column, in order to check that adiabatic conditions have been attained, a thermocouple is placed against the wall of the column and a second thermocouple is situated under the asbestos paper inside the electrical winding. The vapour lines from the still and the vapour line adaptor at the top of the column are lagged with asbestos rope.

(vi) Electrical heating.

(a) Still. The still is heated electrically with a 2L-isomantle of maximum output energy 435 watts. The power input is controlled by an on-off automatic switch.

FIG. 3-6.

(b) Column. Since the column is long and the temperature difference between the top and the bottom may be considerable, it is difficult to control adiabatic conditions by using only a single heating wire. Thus the column is divided into six sections and each section is separately heated. The height of each section is approx. 6", the maximum diameter of column being $1\frac{1}{2}$ " with an average thickness of 1" lagging.

Calculation of the amount of heat required for the column is based on the distillation of a mixture containing water, maximum temperature 100°C . Let the room temperature be 18°C ., then we have:

Thermal conductivities of glass and asbestos 0.6 and 0.114 BTU/hr./ft²/($^{\circ}\text{F}$ /ft.) respectively.

Log mean radius of column, with lagging material

$$= \frac{7/4 - 3/4}{\ln 7/3} = 1.178".$$

Heat conducted from each section of the column

$$= \frac{0.6 \times \pi \cdot \frac{3}{2 \times 12} \times \frac{6}{12} (100-18)}{\frac{1}{16 \times 12}} + \frac{0.114 \times \pi \cdot \frac{6}{12} \times \frac{1.178 \times 2}{12} (100-18)}{\frac{1}{12}}$$

$$= 220.1 \text{ BTU/hr} = 64.5 \text{ watts.}$$

36 S.W.G. Brightray wire of 11.4 ohms/ft resistance is used. The allowable current for the wire to maintain a temperature of 300°C . is 0.7 amps approximately.

$I = \frac{64.5}{230} = 0.281$ amps. and therefore $R = \frac{230}{0.281} = 819$ ohms. In fact a 20 ft. length of wire is used, resistance 228 ohms and current 0.53 amps. six of these windings are placed on the column and six variacs are used to control the power supply.

(c) Reflux line. It has been found by experiment that the heat energy required to raise the temperature of the condensate to its boiling point on returning into the column is 100 watts, including losses. Two yards of coiled Brightray wire of 300 ohms/yd. resistance is used, the current being controlled by a variac.

(vii) Thermocouples. Thermocouples are used for the measurement of liquid and vapour phase temperatures and to check that adiabatic conditions are maintained in the column.

The thermocouples used to measure the temperature of the liquid and vapour phases are inserted in pairs through the six side arms of the column, one in the centre of the vapour space and the other in the trough containing the liquid (Fig.3.6). All twelve thermocouples are connected through a multipoint Pye switch to a common cold junction at 0°C. and then to a galvanometer.

It was suspected that the temperature of the vapour was not given accurately by those thermocouples exposed to

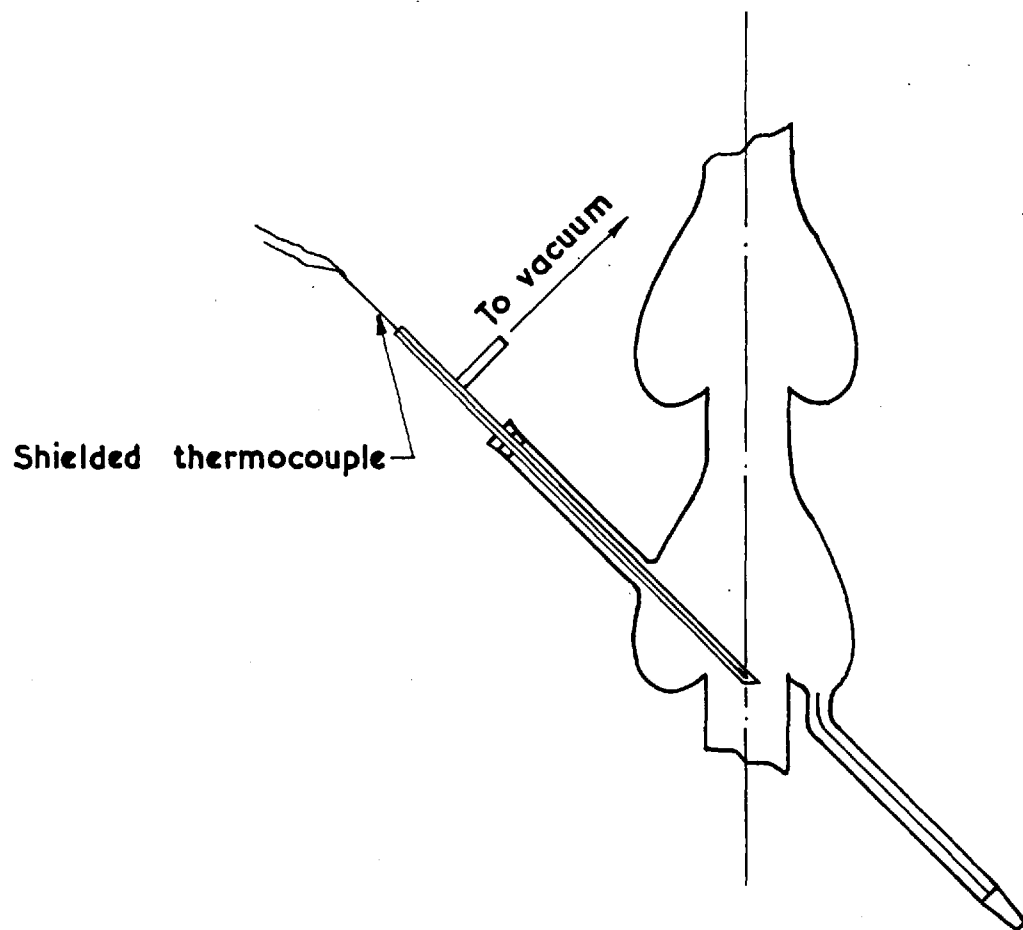


FIG. 3-7. SHIELDED THERMOCOUPLE
IN POSITION



FIG. 3-8. DETAIL OF
SHIELDED THERMOCOUPLE

the vapour and at a later stage in the experimental work a shielded thermocouple was used (Fig.3.8). The shield is designed to deflect falling liquid drops from the thermocouple, vapour being sucked slowly through the sleeve and over the thermocouple by applying a slight vacuum to the outer end.

In order to ensure that adiabatic conditions are maintained in the column, six pairs of thermocouples are installed as shown in Fig.3.6; one pair to each section of the column. When these thermocouples indicate the same temperature there can be no heat flow out of or into the column. They are also connected through a multipoint switch to a galvanometer and have a common cold junction at 0°C.

All thermocouples for measuring the vapour and liquid temperatures were accurately calibrated against a potentiometer with a cold junction at 0°C. The thermocouple wires were extended without a break to the multipoint switch and were calibrated with the current passing through the switch. The results obtained were the same as those given in the literature [39] for copper-constantan with reference junction at 0°C.

(viii) Sampling device. Six side arms at 45° from the vertical are fixed to the lower ends of every alternate trough. They are made of capillary tubes with cone socket ends. Through these sidearms samples are withdrawn from the column.

(ix) Bottom condenser (Fig.3.1). This is placed immediately above the vapour line from the still and is normally isolated by a glass tap. It is used only during the sampling period to cut off vapour from the column, and stop the distillation.

(x) Liquid distributor. The first bulb at the top of the column serves to spread the liquid evenly over the walls of the column. A pair of thermocouples are inserted in the liquid and vapour spaces of this bulb in order to ensure that the reflux is at its boiling point on entering the second bulb.

(xi) Flowmeter. The flow rate is measured by collecting liquid from the bottom of the column for a measured time. The liquid is collected in a calibrated tube by closing a mercury-seal valve (Fig.3.9).

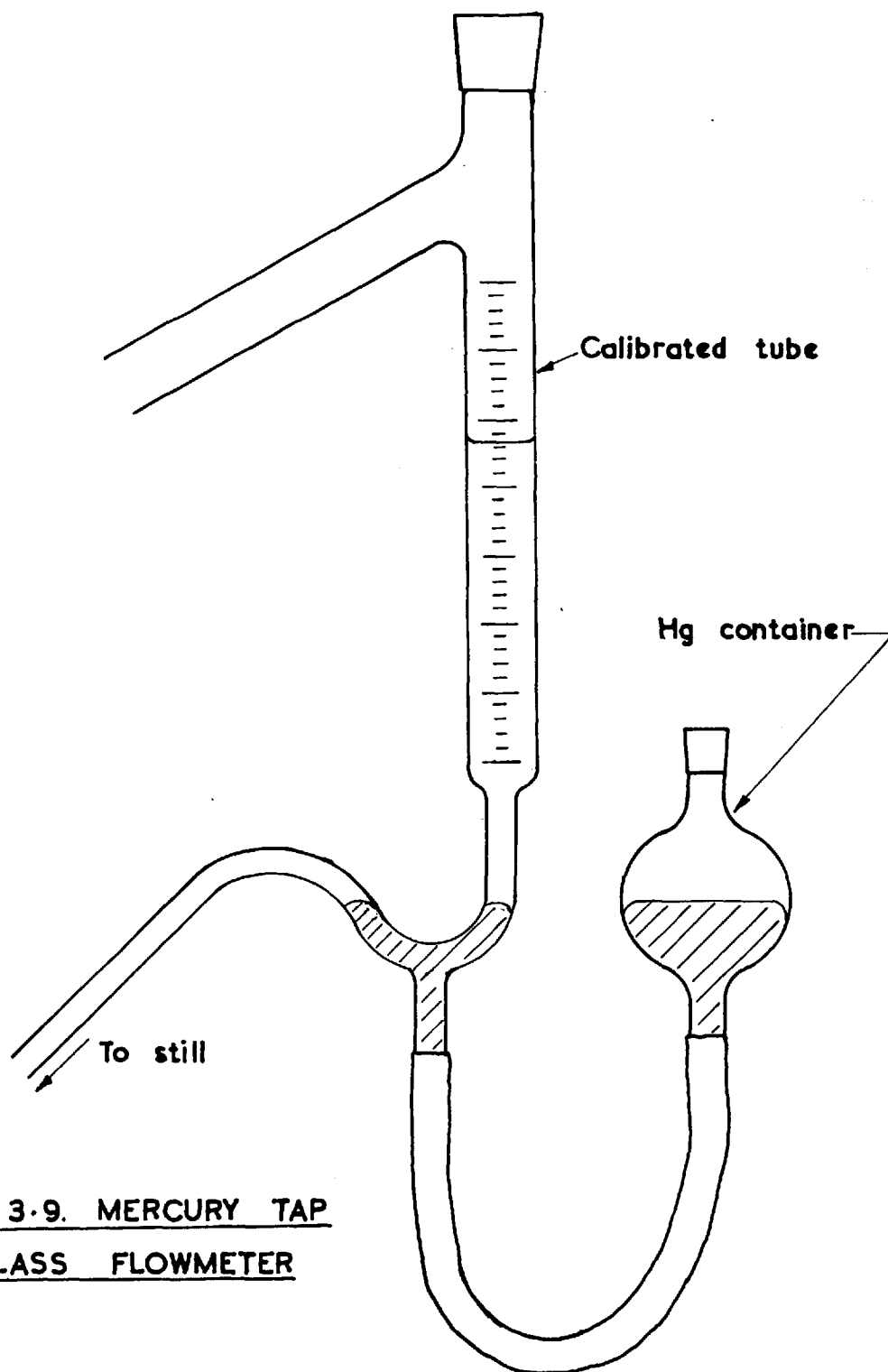


FIG. 3-9. MERCURY TAP
GLASS FLOWMETER

B. PROCEDURE.

The following observations were made with different concentrations in the still of two binary mixtures :

- (1) the separation and the number of transfer units under total reflux,
- (2) the temperature in the bulk of the vapour and liquid phases along the depth of the column, and
- (3) the change in the molal flow rate of the liquid phase.

The binary systems used were methanol-water and cyclohexane-toluene.

- (i) The number of transfer units at different concentrations in the still and the measurement of the temperature of the bulk phases.

About 1500 cc. of the more volatile component and about 50 cc. of the less volatile component was charged to the still. The still was heated and the isomantle controller was adjusted to give a liquid flow rate at total reflux approximately 12.5 cc./min. for methanol-water system and 20 cc./min. for cyclohexane-toluene system. Heat was supplied to the column and the voltage across each heating element was adjusted with the variac until each pair of the six pairs of thermocouples in the lagging read the same.

Heat was supplied to the vapour line in order to raise the reflux-liquid temperature to as close to its boiling point as possible. When a steady state condition was reached (after about two hours), the thermocouple readings for the liquid and vapour temperatures were taken. The column was then isolated by suddenly stopping the distillation. This was done by switching off the main electrical supply, opening the tap admitting vapour from the still to the condenser at the bottom of the column, and opening the tap at the vapour line adaptor to divert the reflux liquid from the column. Samples at each section were taken, with the first few cc. being discarded. The composition of each sample was determined by measuring its refractive index. The still concentration was then altered and the experiment repeated, the whole range of concentration being eventually covered.

The advantage of stopping the distillation suddenly and 'freezing' the concentration gradient in the column is that samples of the liquid in the troughs of the column can then be taken slowly, and taking the samples does not affect the liquid elsewhere in the column. At one stage of the work it was suspected that the stopping of the distillation was not effected quickly enough. A range of

experiments was therefore repeated, but the column was not isolated during sampling. It was found that this made little difference to the results. At the same time the temperature of the vapour was measured by a shielded thermocouple.

- (ii) The measurement of small changes in the molal flow rates of the phases by introducing concentrated salt solution into the reflux liquid.

The distillation was started and was carried out as described above. After $1\frac{1}{2}$ hours, concentrated sodium chloride solution (10% by weight) was introduced at the top of the condenser into the column at a constant flow rate of 0.625 cc/min. This small and uniform rate was obtained by electrolysing water, using platinum and nickel for the positive and negative electrodes respectively and adding a few cc. of sodium hydroxide to the electrolysis cell (Fig.3.10). The products from the electrolysis of water were used to build up a pressure inside a container and force the concentrated salt solution from this container into the column. The rate was determined entirely by the current used for electrolysis and thus maintained a constant small flow of concentrated salt solution into the column. When a steady state in the column was again reached (after about

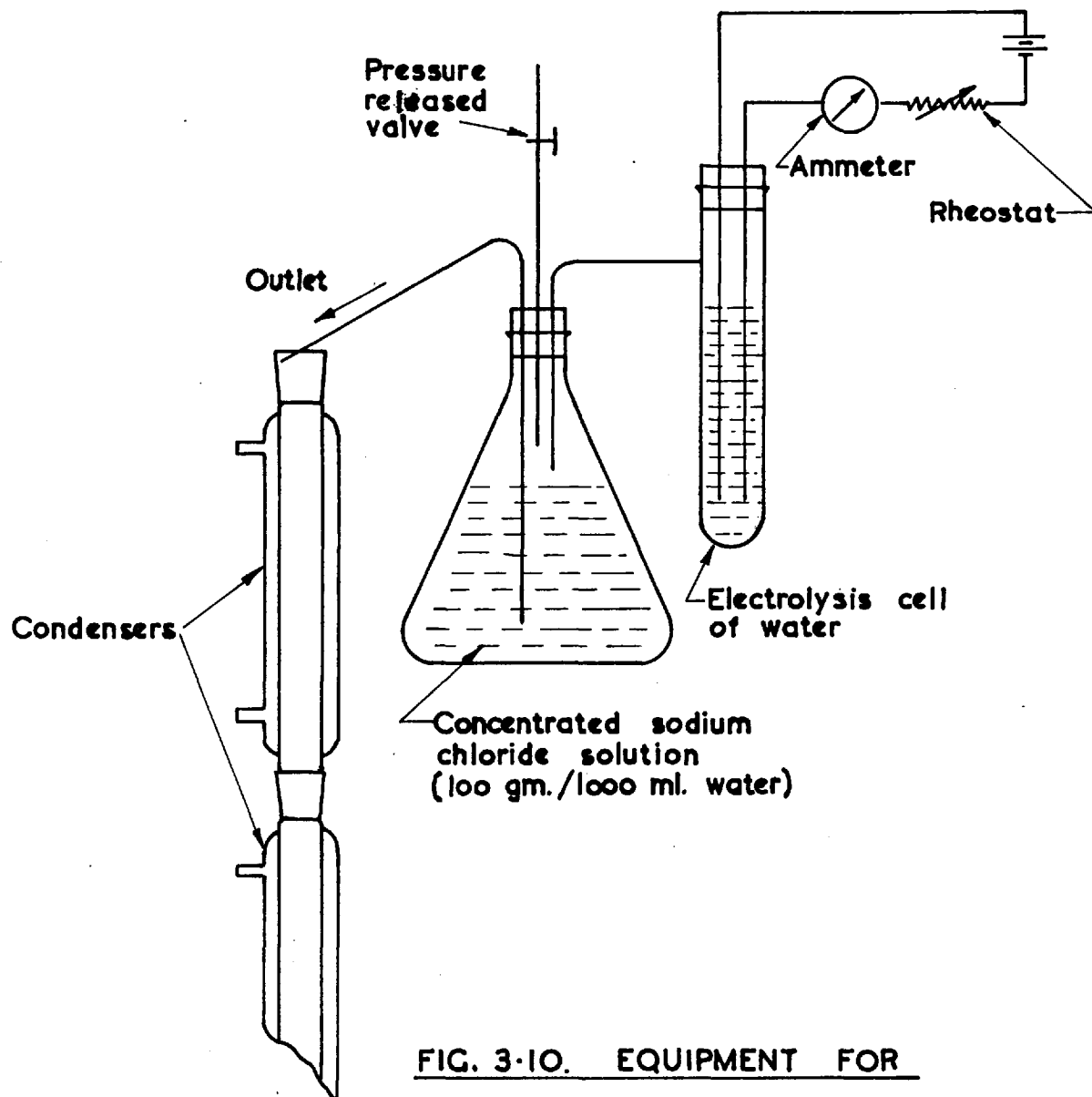


FIG. 3-10. EQUIPMENT FOR
INTRODUCING SALT SOLUTION
AT LOW FLOW RATE

1½ hours), the distillation process was suddenly stopped, the column isolated and samples taken. The refractive indices and the conductivities of each sample were measured and from these measurements the compositions of the samples are known (v.i.). From the measurements of percentage of salt in the samples, the change in molal flow rate of the liquid along the column was calculated (see Appendix I).

The change in liquid flow rate measured was unexpectedly high and so the experiments were repeated, both with and without the sudden stoppage of distillation process for sampling; and the percentage of salt in each sample was determined instead by titration with 0.01N of silver nitrate solution, using potassium chromate as an indicator. Results obtained for the change in flow rate along the column were again high.

(iii) Measurement of the heat transfer through column wall and lagging.

To make sure that the column was kept under adiabatic conditions and in order to be able to calculate what maximum error in the change of liquid flow rate could be caused by a slight but detectable difference in the thermocouple readings, the value of the heat transfer coefficient for the column wall and lagging was determined. It was

measured by running a cold methanol-water mixture of known composition into the top of the pear column at a rate approximately equal to that used during distillation, 12 cc./min. The adiabatic heaters were switched on so that there was a small temperature difference between the two sets of thermocouples set in the lagging. When steady, conditions had been reached, the temperature of the column wall and that of the inlet and outlet mixture were measured. The heat transfer coefficient was then calculated as follows.

Methanol-water mixture 0.5 m.f. was used: density and specific heat are 0.8765 gm./cc. and 0.83 cal/°C/gm. respectively.

Heat transferred into the liquid

$$\begin{aligned} Q &= 12 \times 0.8765 \times 0.83 \times \Delta t \quad \text{cal/min.} \\ &= 2.08 (t_T - t_B) \quad \text{Btu/hr.} \end{aligned} \quad (3.1)$$

where $(t_T - t_B)$ is the rise in temperature of the liquid as it flows through the column in °C.

Mean diameter of lagging material = 2"

Height of column = 25"

Surface area for the transfer of heat

$$A = \pi \times \frac{2}{12} \times \frac{25}{12} \text{ ft}^2 = 1.09 \text{ ft}^2$$

$$\begin{aligned} \text{Hence the heat transfer coefficient is } U &= \frac{Q}{A \cdot \Delta t} \\ &= \frac{2.08 (t_T - t_B)}{1.09 \Delta t} \end{aligned} \quad (3.2)$$

where Δt is the temperature difference across the lagging in °F.

The measurements show that for a log mean temperature difference of 6°C. across the lagging along the whole column, the rise in temperature of the liquid (0.5 m.f. methanol-water mixture) is 50°C.; hence U is 8.83 BTU/hr/ft²/°F. Thus a temperature difference of 0.6°C., which is equivalent to 0.02 millivolts of the potentiometer reading, easily detectable across the lagging, would transfer 0.17 BTU/min. heat energy and cause a change in flow rate of

$$\frac{0.17 \times 25}{8820 \times 454} \text{ gm/min.} = 0.12 \text{ cc./min.}$$

(The latent heat of vaporisation and the molecular weight of 0.5 m.f. methanol-water are 8820 BTU/lb-mol and 25 lb. respectively).

Thus percentage evaporation owing to external heat supply at a liquid flow rate 12 cc./min. is 1%, which is negligible compared with 10%, or above, total evaporation measured in experiments.

(iv) Detection of entrainment.

It was thought that some anomalous measurements were due to entrainment. This was detected as follows.

The experiment was carried out in exactly the same way as described in (ii) but concentrated salt solution was introduced through the side arm at position (4) (Fig.3.11).

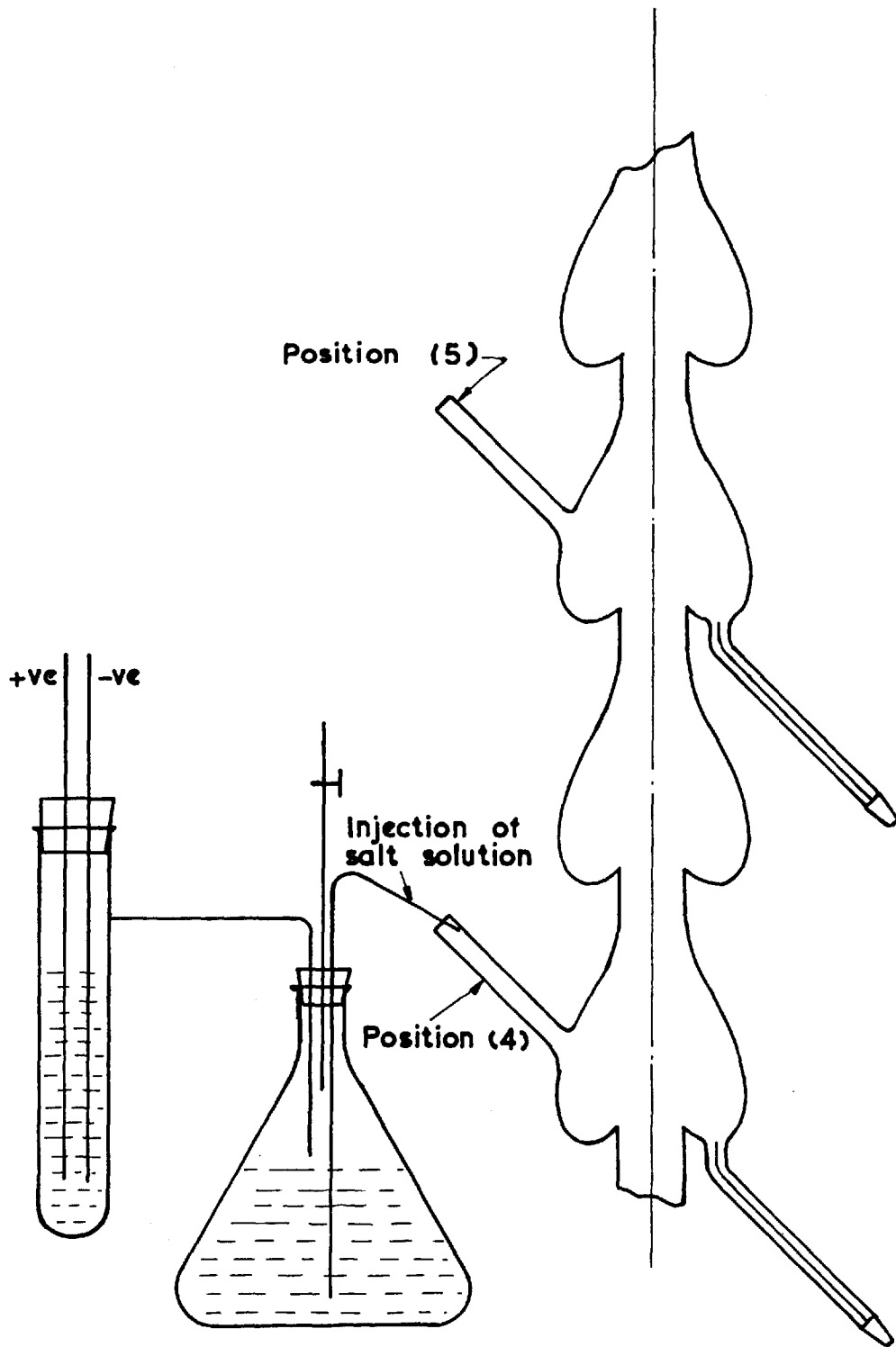


FIG. 3·11. INJECTION OF SALT
SOLUTION FOR DETECTING ENTRAINMENT

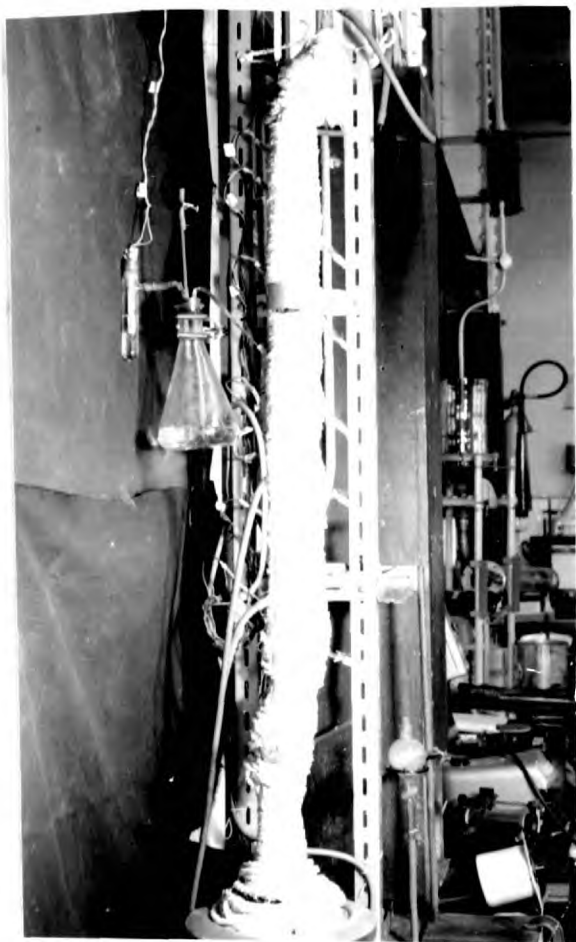


FIG. 3.12

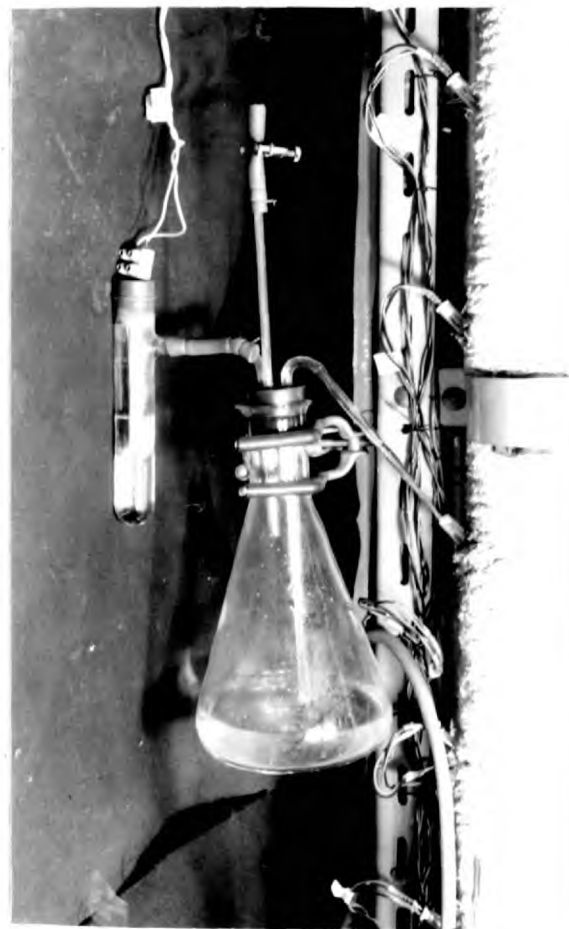


FIG. 3.13

When a steady state had been reached, liquid samples were withdrawn from the reflux-return line and at the positions above and below position (4). The conductivity of the sample was measured and it was titrated with silver nitrate solution. Both readings showed without doubt that there was salt in the condenser system above the point at which it was admitted. The experiment was repeated several times.

Since the salt concentration in the sample above position (4) was very weak, it was not possible to determine accurately the percentage of salt either from the previously calibrated conductivity-composition curve or by the titration method. In order to determine the maximum extent of entrainment, that concentration of salt which in solution just cannot be traced by titration with the standard silver nitrate solution was measured. This is the worst possible case. The results and calculations showed that the entrainment is less than 1% of the total liquid flow.

CHAPTER 4.

EXPERIMENTAL DATA.

A. EQUILIBRIUM DATA.

The liquid-vapour equilibrium data for the system carbon tetrachloride-toluene was taken from Ju Chin Chu [40], Kirschbaum [41], and Perry [42], for the system acetone-benzene from Griswold [43], for the system acetone-chloroform from Rosanoff and Easeley [44] and Perry [42], for the system methanol-water from Perry [42] and others [45,46,47, 48,49], and for the system cyclohexane-toluene from Sieg and Jost [50] and others [45,51,52]. The boiling point diagram for the system carbon tetrachloride-toluene was measured experimentally.

B. LIQUID PROPERTIES.

Few of the physical properties of both liquid and vapour mixtures were available in the literature and the values of these properties were calculated as follows.

(i) Density ρ_L (Graph 4.3). The density of mixtures of carbon tetrachloride-toluene at the boiling point is given by Sawistowski [2]. For other systems, the density was calculated from the equation [1] ;

$$\rho_L = \frac{\rho_A \rho_B M_{av}}{M_A x_A \rho_B + M_B x_B \rho_A} \quad (4.1)$$

where ρ_A, ρ_B = densities of pure components at the temperature of the boiling point of the mixture,

M_A, M_B = molecular weights of pure components,

M_{av} = average molecular weight of the mixture.

Densities of pure components were obtained from Timmerman[53].

(ii) Viscosity η_L (Graph 4.4). For the system of carbon tetrachloride-toluene, the viscosity of mixtures at their boiling points is given by Sawistowski [2]. For other systems, the viscosity was calculated as follows :

$$\log \eta_i = x_A \log \eta_A + x_B \log \eta_B \quad (4.2)$$

where η_A, η_B = viscosities of pure components at temperature t , concerned, this was obtained from the literature [54].

η_i = viscosity of mixture at temperature t_1 .

x_A, x_B = mole fractions of components in the mixture.

η_i was then extrapolated to the boiling point of the mixture by means of the relation proposed by Arrhenius and de Gusman [55]

$$\eta = A e^{B/RT} \quad (4.3)$$

where $A, B = \text{constants,}$

$R = \text{gas constant} = 1.9872 \text{ /gm.mol.}$

$T = \text{absolute temperature } ^\circ\text{K.}$

$$\text{therefore } \frac{\eta_L}{\eta_i} = \frac{e^{B/RT_b}}{e^{B/RT_i}} \quad (4.4)$$

$$\text{Or, } \ln \eta_L - \ln \eta_i = \frac{B}{R} \left(\frac{1}{T_b} - \frac{1}{T_i} \right) \quad (4.5)$$

where $T_b = \text{boiling point temperature of mixture.}$

(iii) Kinematic viscosity γ_L (Graph 4.5)

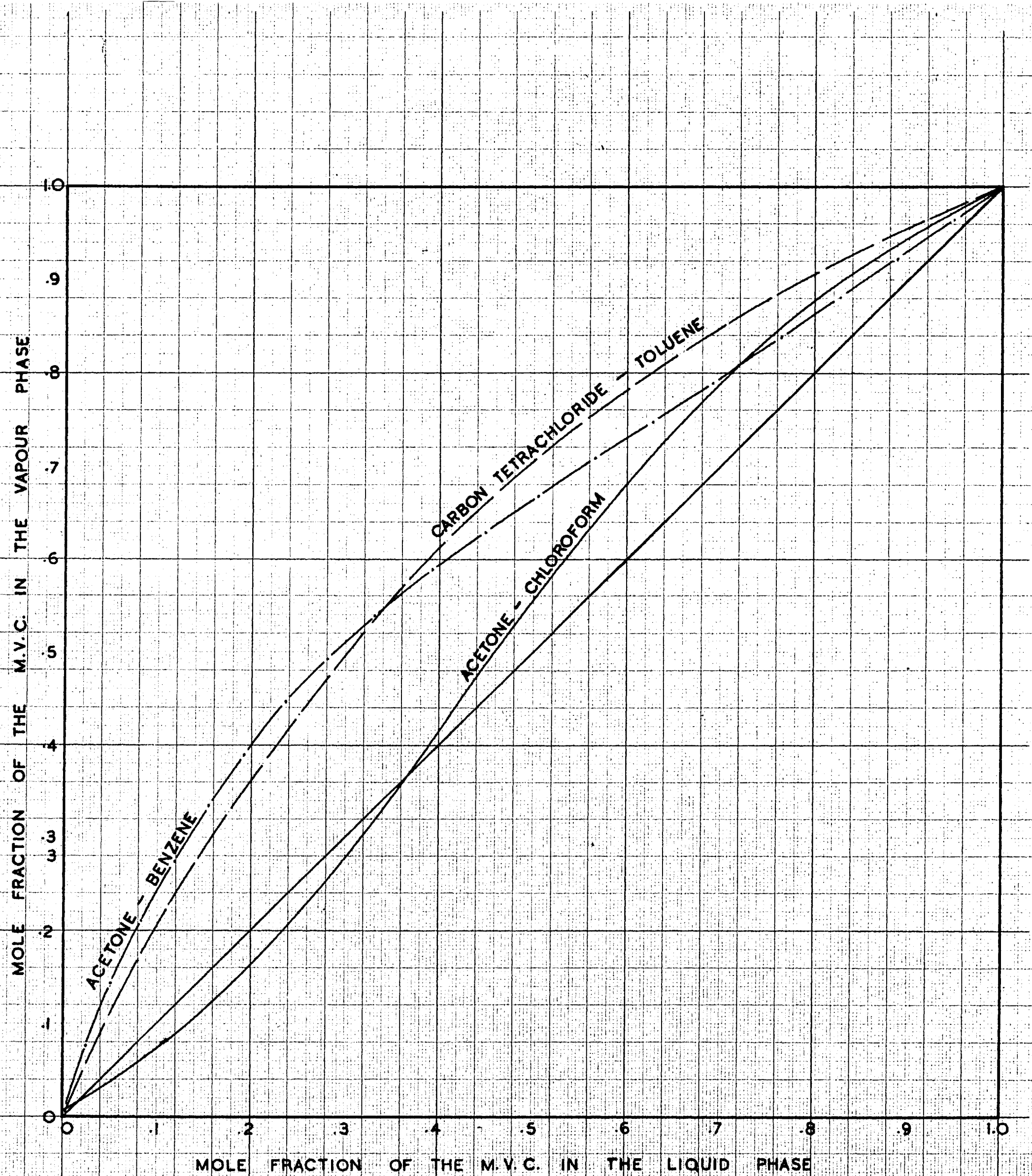
Kinematic viscosities of mixtures at the boiling point were required for measuring the liquid flow rate by the capillary flowmeter in the packed column and are calculated directly from the above values.

(iv) Refractive index n_D . The refractive indices of the following binary mixtures were obtained from the literature: carbon tetrachloride-toluene [2,54] (Graph 4.6); acetone-benzene [54,55,56] (Graph 4.7); acetone-chloroform [54,55] (Graph 4.7); methanol-water [54] (Graph 4.8) and cyclohexane-toluene [2] (Graph 4.6). These were checked using the Abbé refractometer. The refractive index-composition curves for methanol-water system, with different percentages of sodium chloride present, were measured at 25°C (Graph 4.9). This data was needed so that the compositions

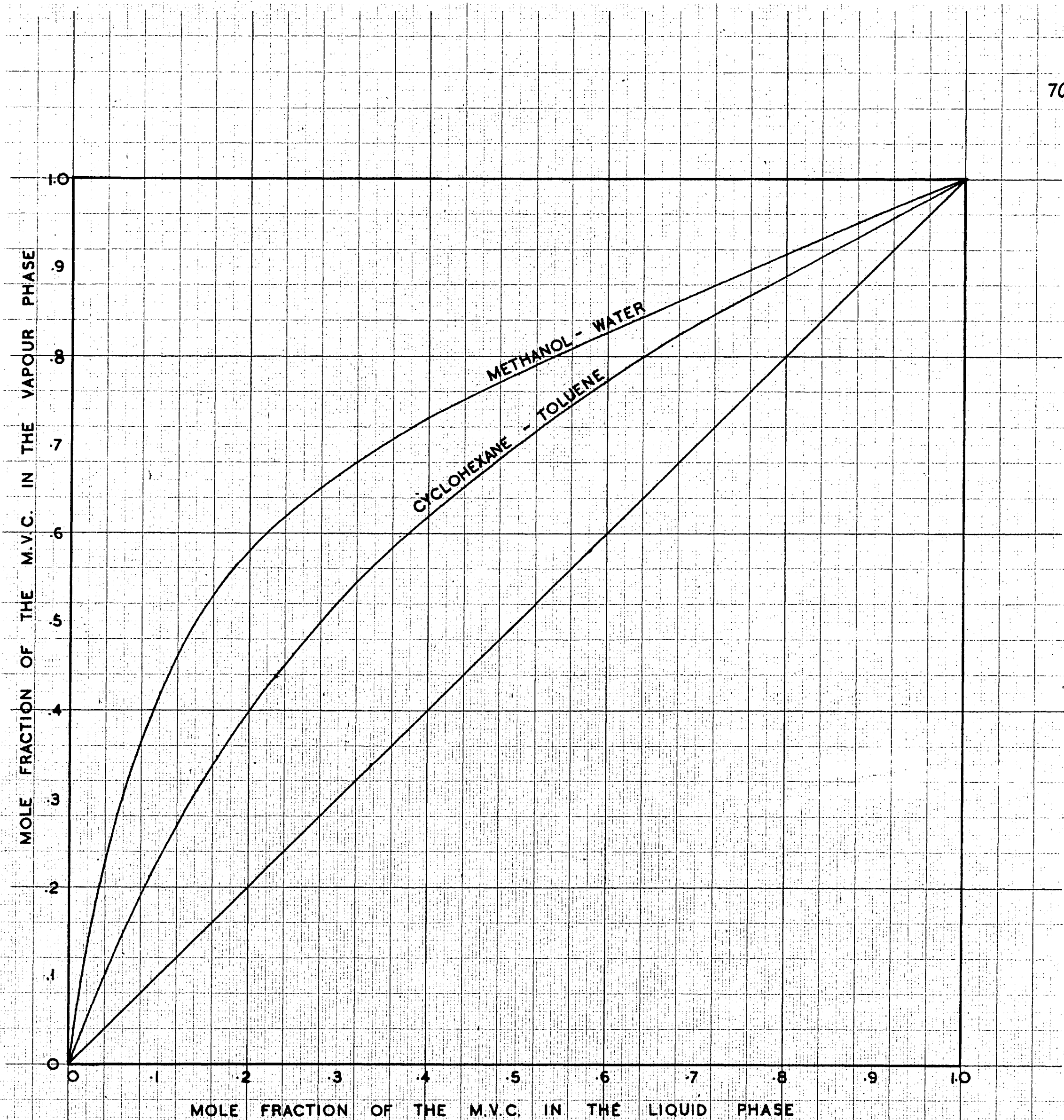
of the mixtures containing salt, used in some of the experiments in the pear column, could be determined.

(v) Conductivity.

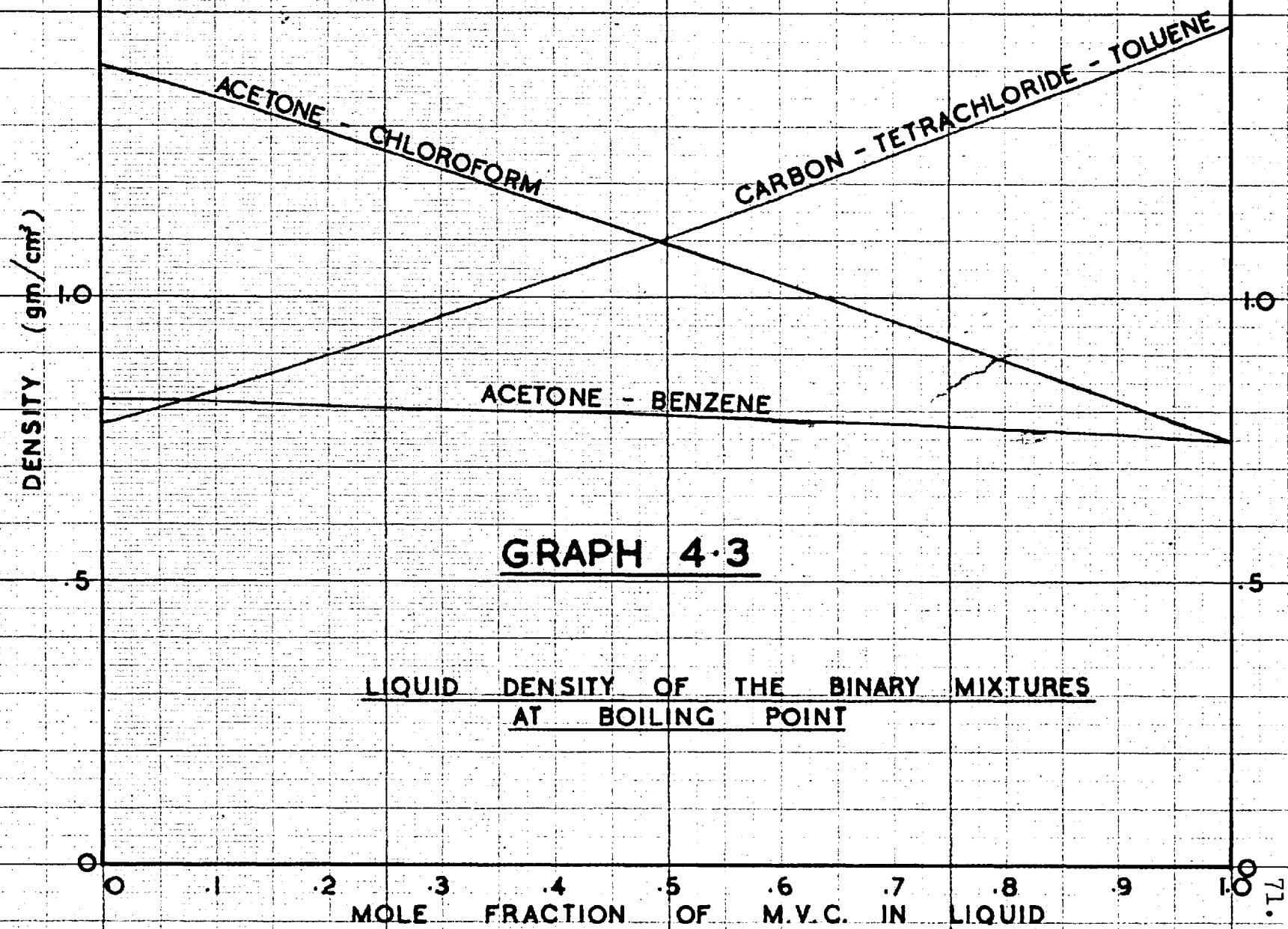
The relationship between conductivity and mole fraction of methanol in water with different percentages of sodium chloride present was determined at 25°C. by means of a Mullard conductivity bridge and conductivity cell of 1 cc. capacity (Graph 4.10).



GRAPH 4.1 EQUILIBRIUM CURVES ON M_CCABE - THIELE DIAGRAM



GRAPH 4.2 EQUILIBRIUM CURVES ON McCABE THIELE DIAGRAM

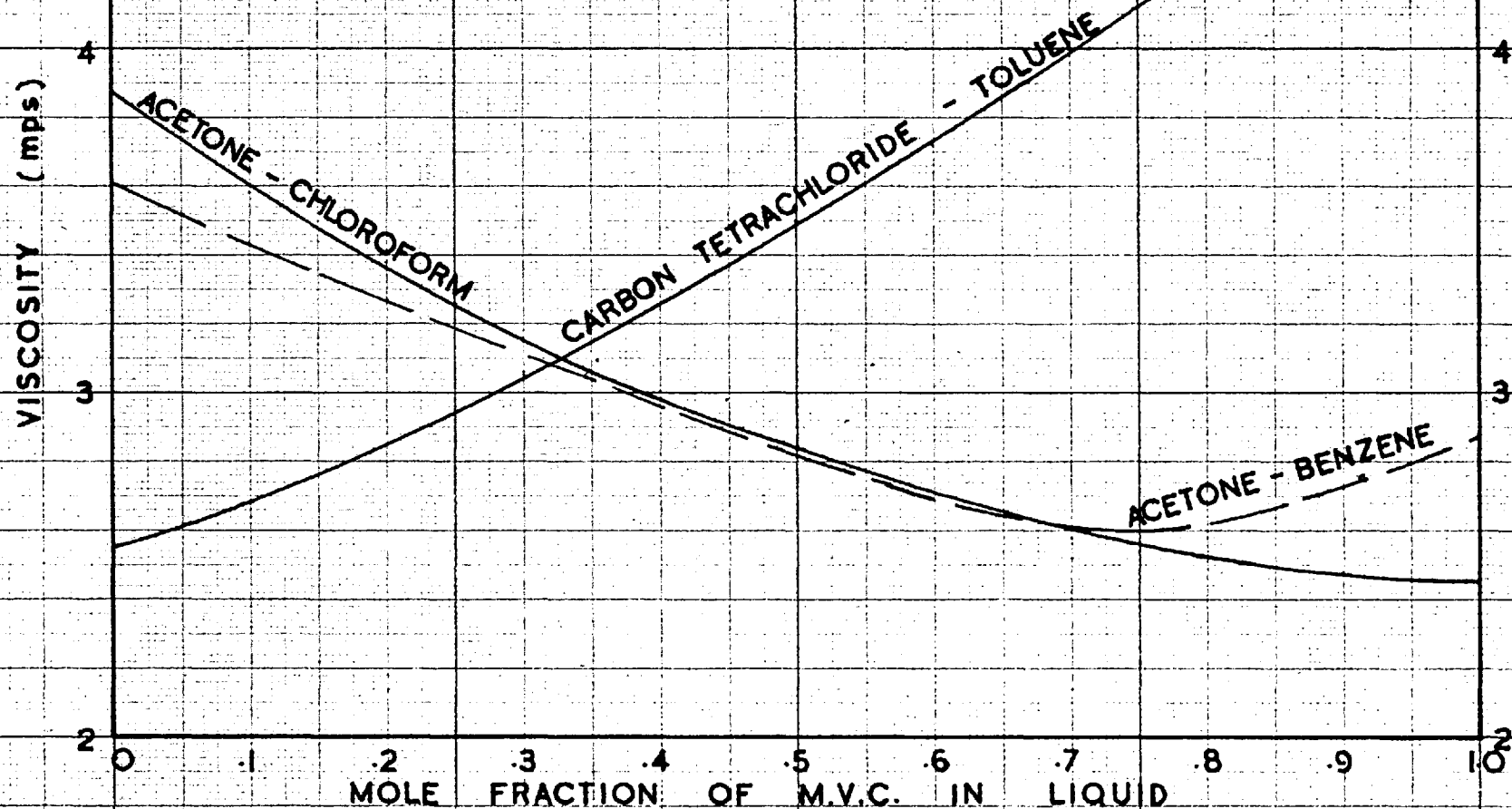


GRAPH 4.3

LIQUID DENSITY OF THE BINARY MIXTURES
AT BOILING POINT

GRAPH 4.4

LIQUID VISCOSITY OF THE BINARY MIXTURES
AT BOILING POINT



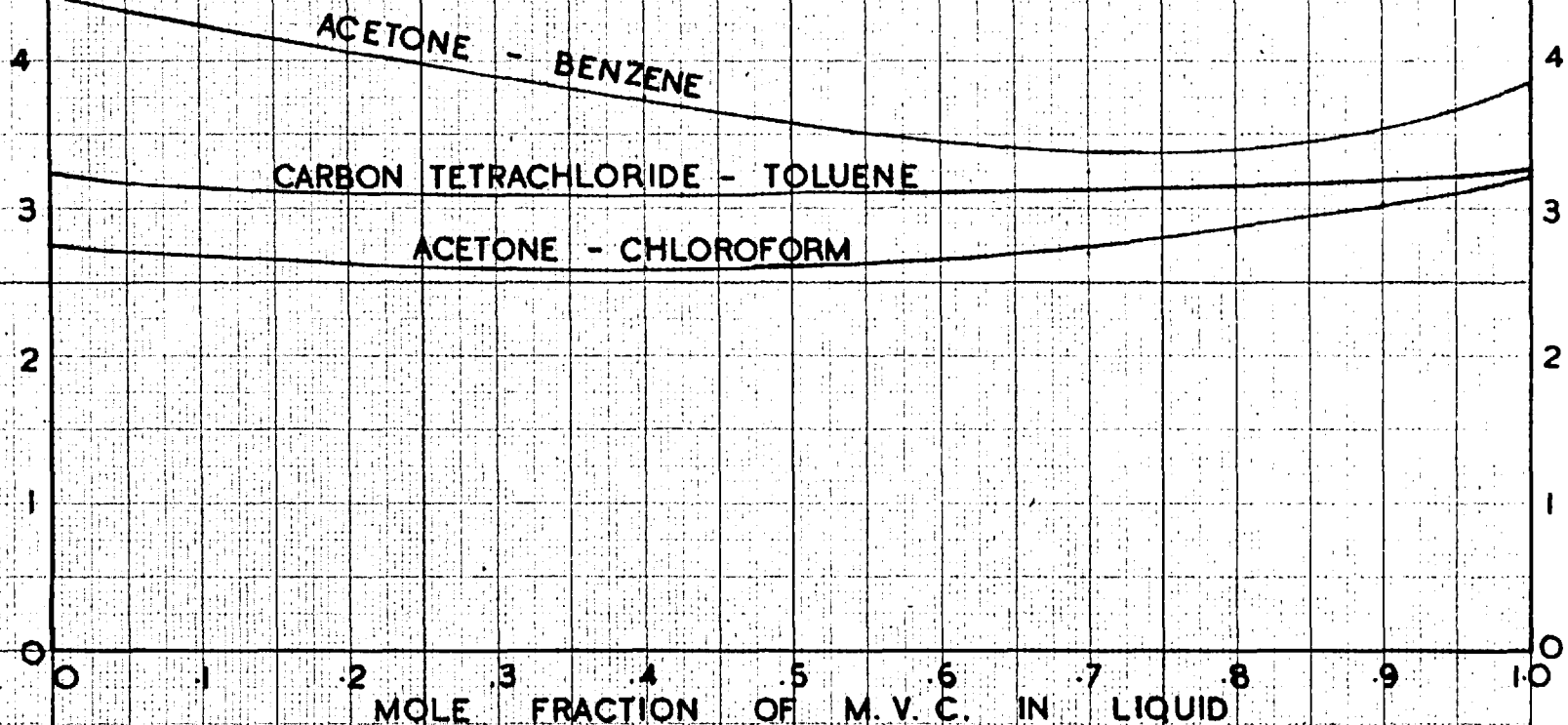
GRAPH 4.5

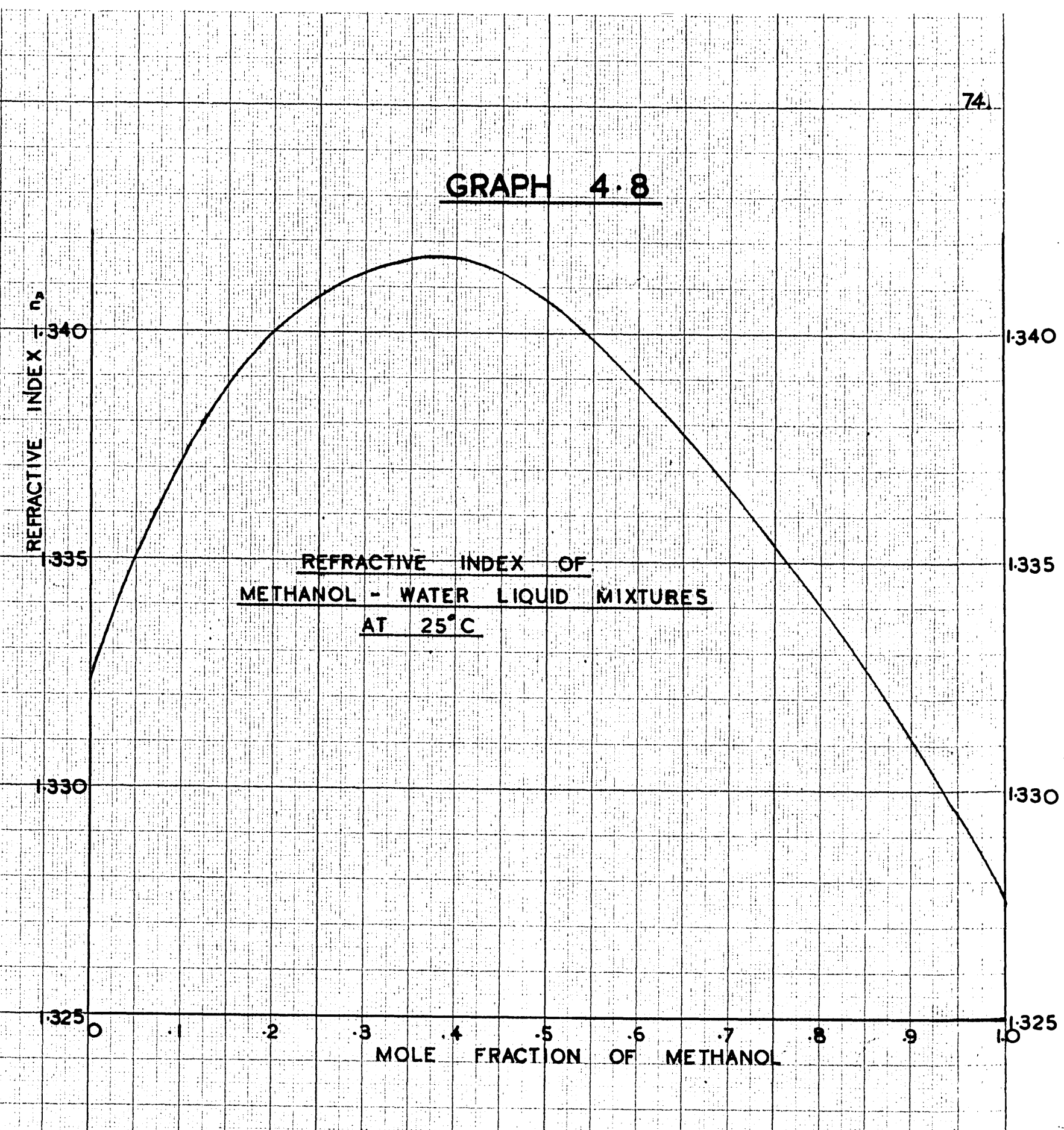
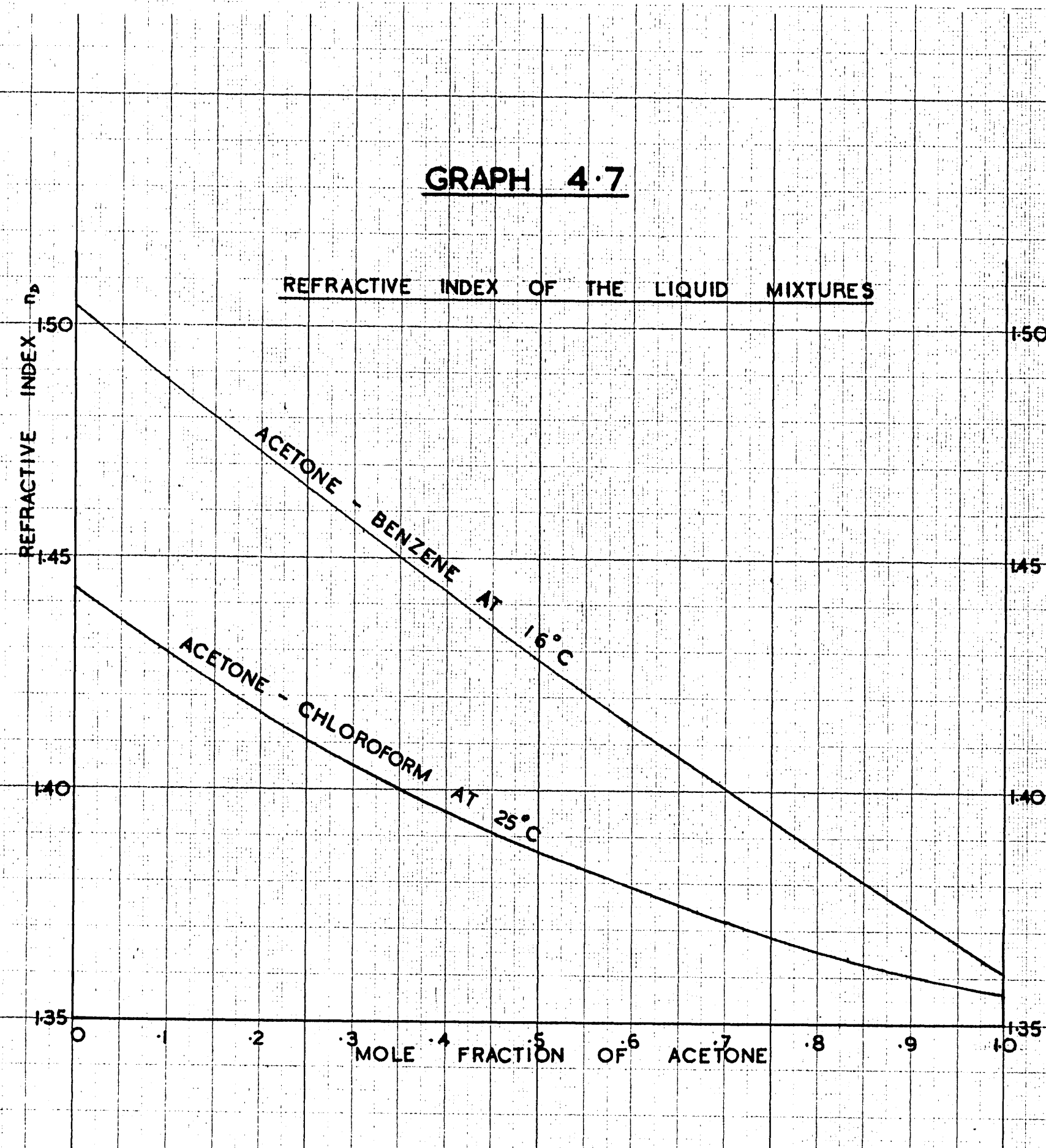
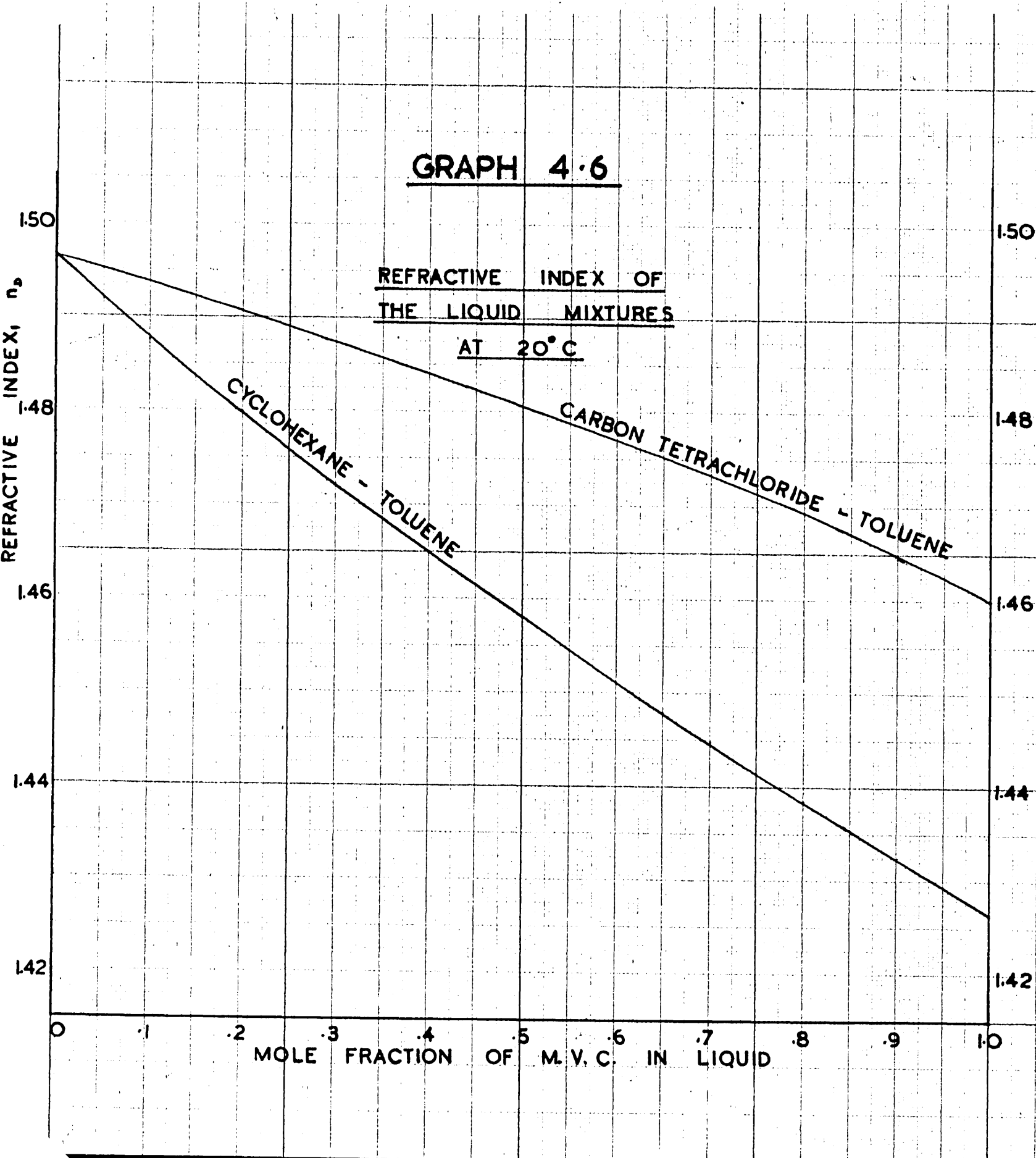
KINEMATIC VISCOSITY OF
THE LIQUID MIXTURES AT
BOILING POINT

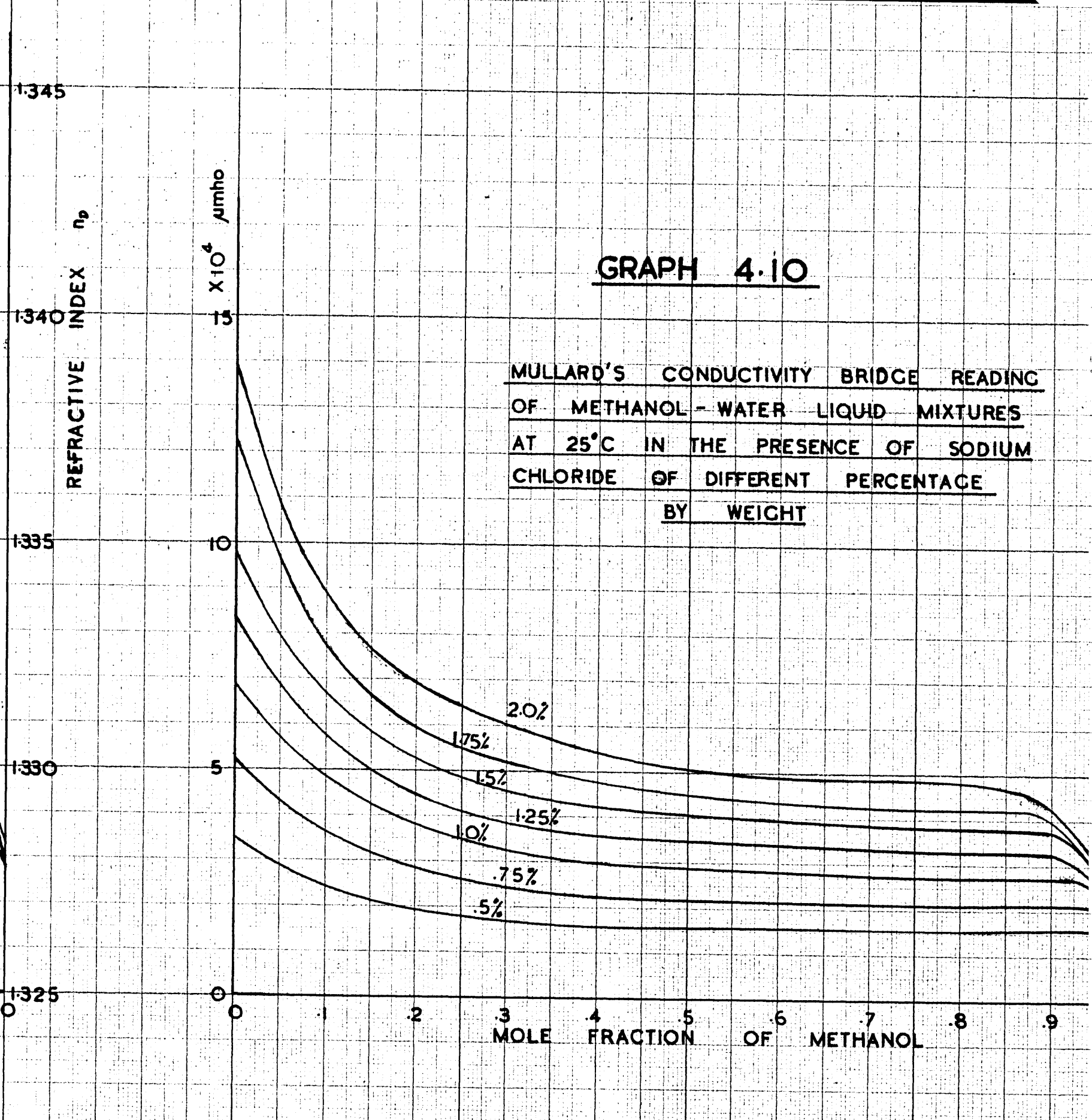
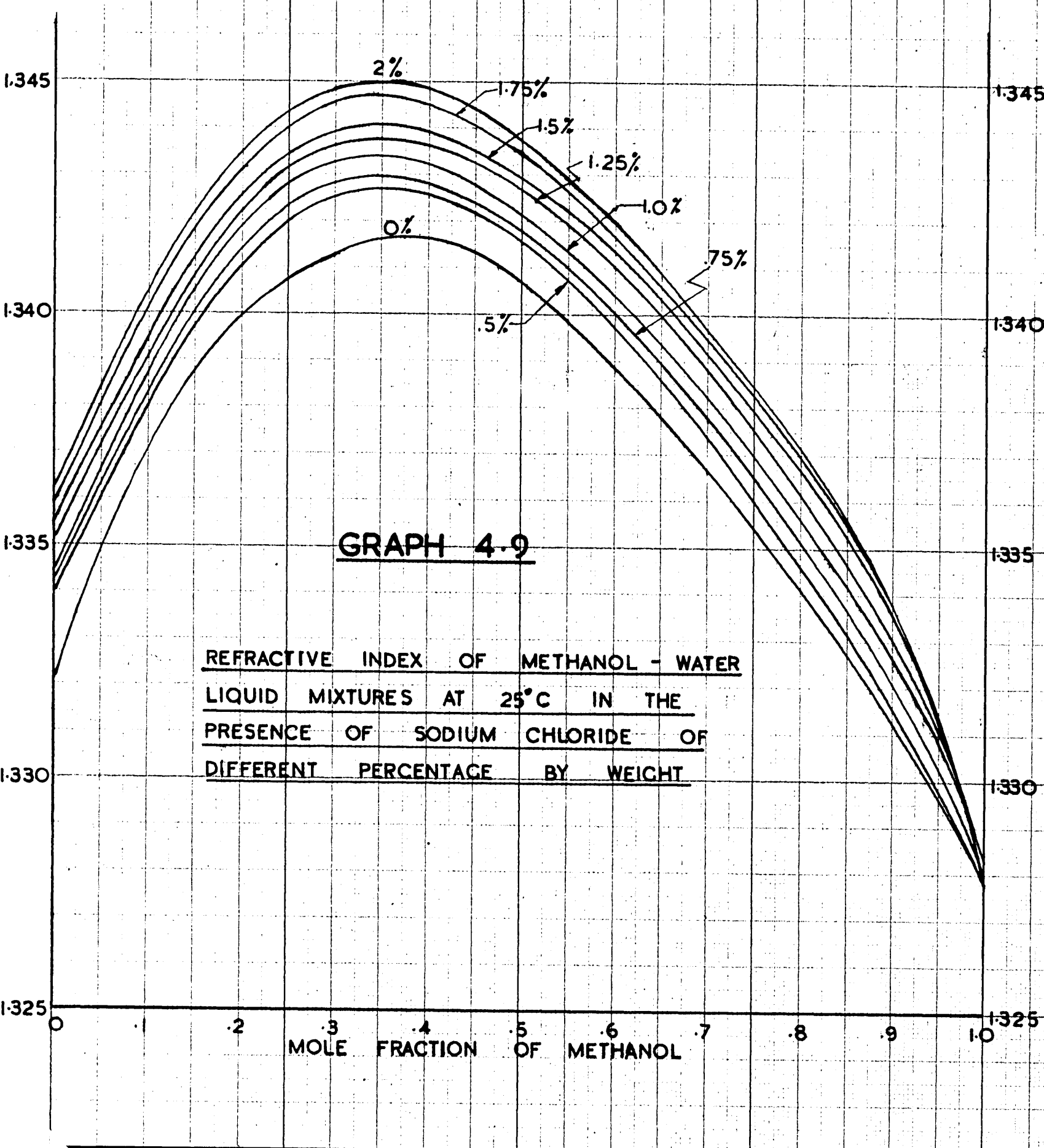
KINEMATIC VISCOSITY
 η [millistokes]

η [millistokes]

73.







GRAPH 4-11

CALIBRATION CURVE FOR CAPILLARY FLOWMETER

LIQUID HEAD h

15

10

5

0

13

15

20

25

30

VOLUME

FLOW

RATE

Q

cc/min.

$\gamma = 1.0 \text{ c.s.}$

$\gamma = 0.8 \text{ c.s.}$

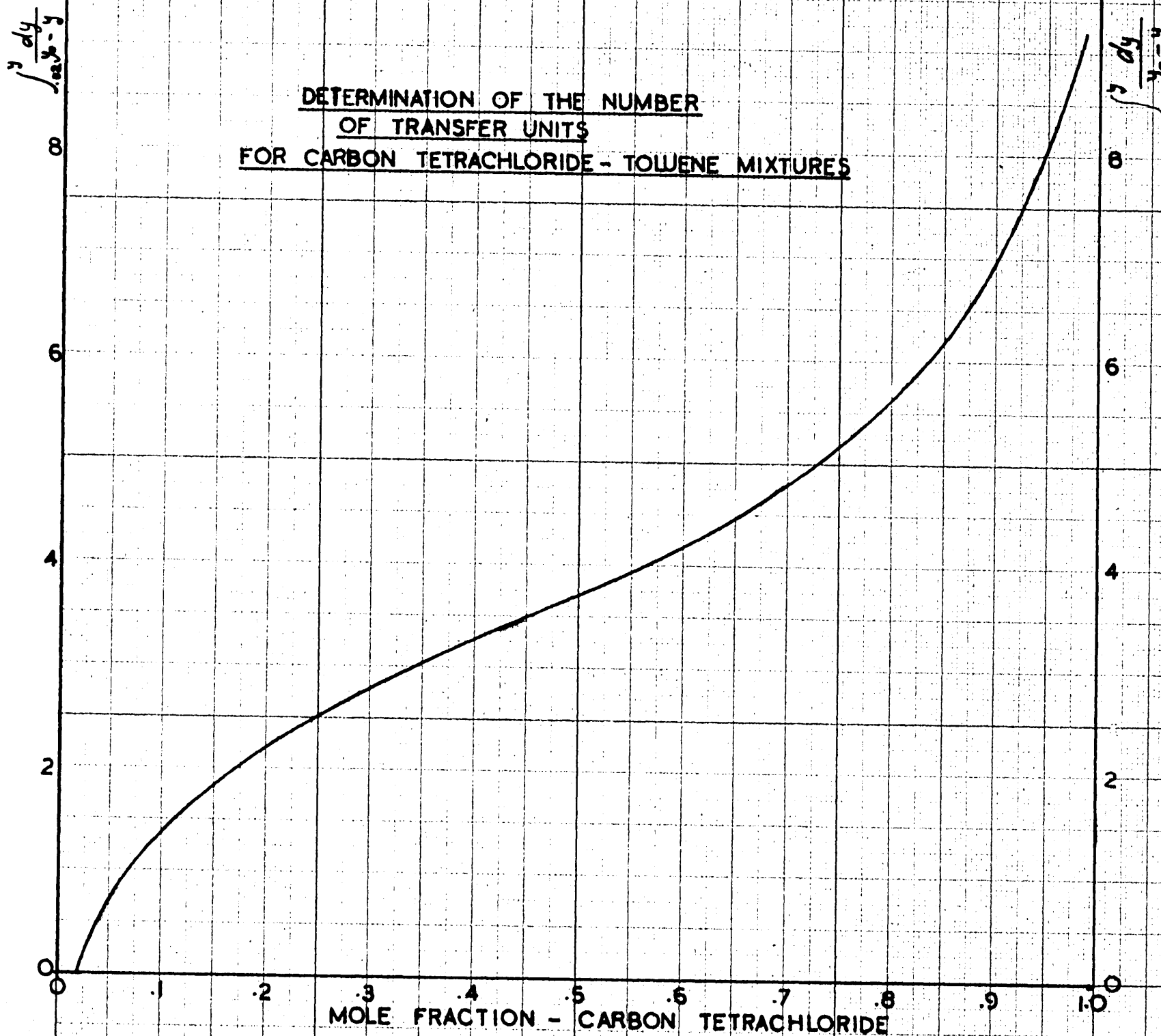
$\gamma = 0.6 \text{ c.s.}$

$\gamma = 0.4 \text{ c.s.}$

$\gamma = 0.2 \text{ c.s.}$

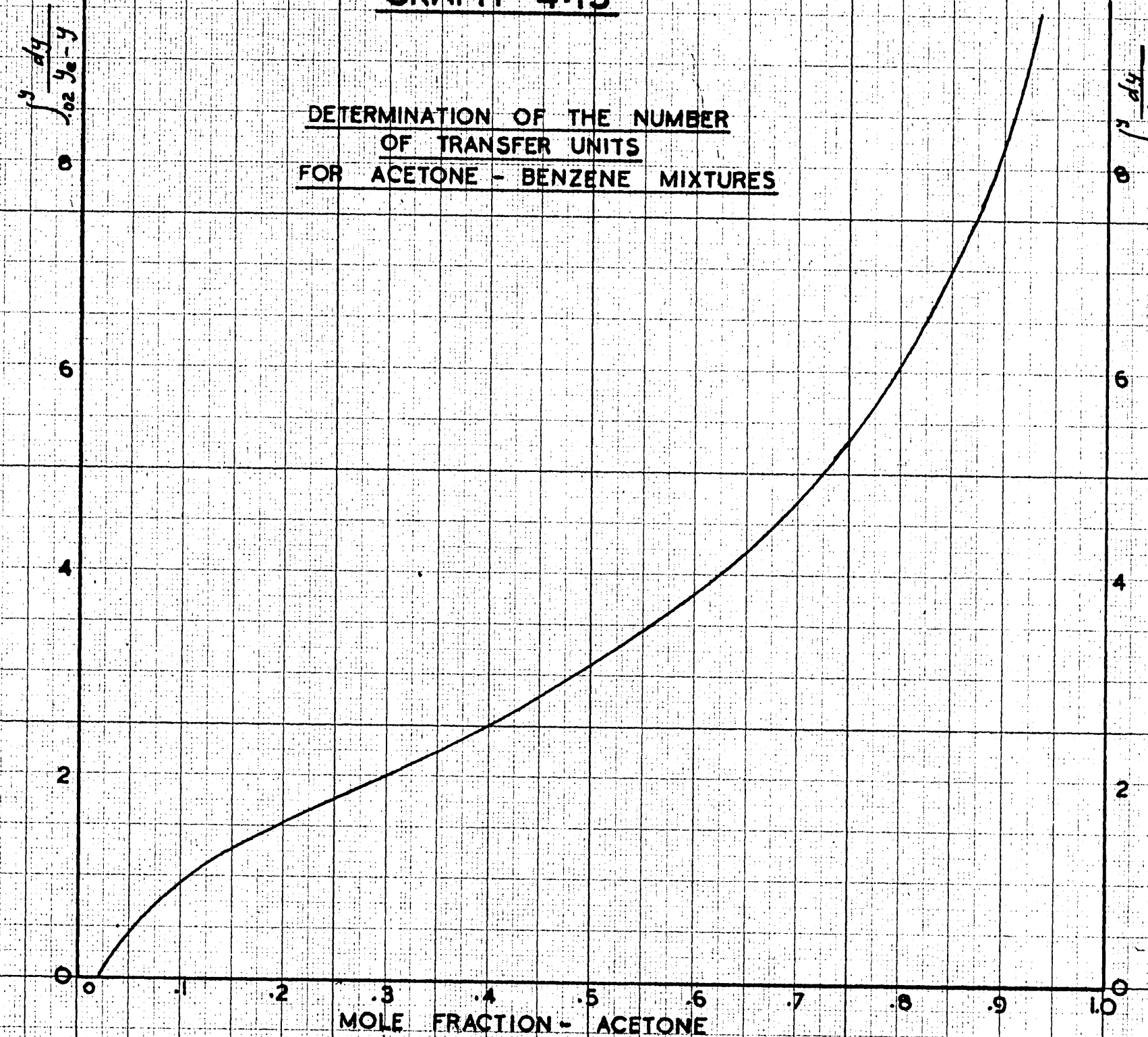
GRAPH 4.12

DETERMINATION OF THE NUMBER
OF TRANSFER UNITS
FOR CARBON TETRACHLORIDE - TOLUENE MIXTURES



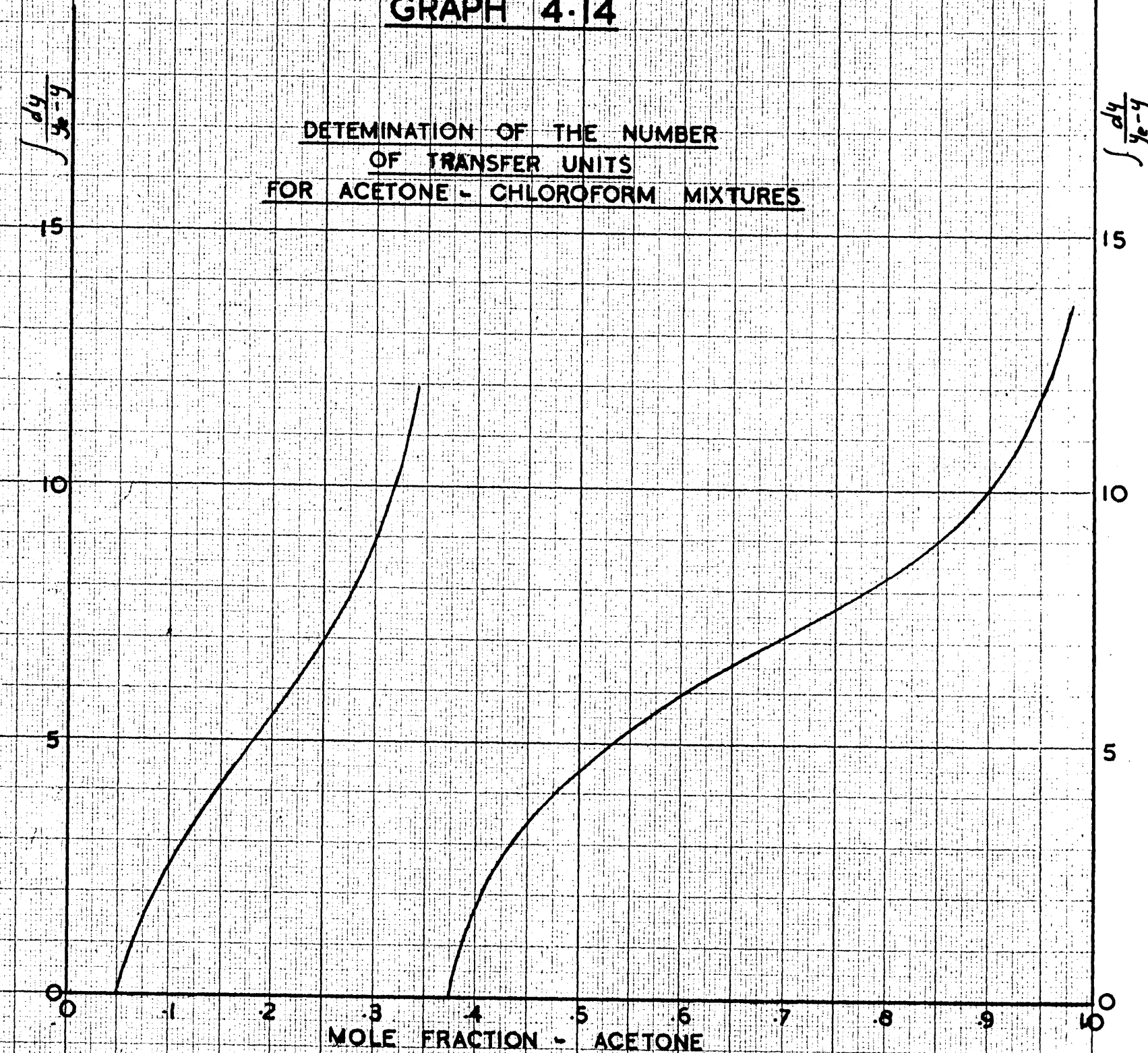
GRAPH 4.13

DETERMINATION OF THE NUMBER
OF TRANSFER UNITS
FOR ACETONE - BENZENE MIXTURES



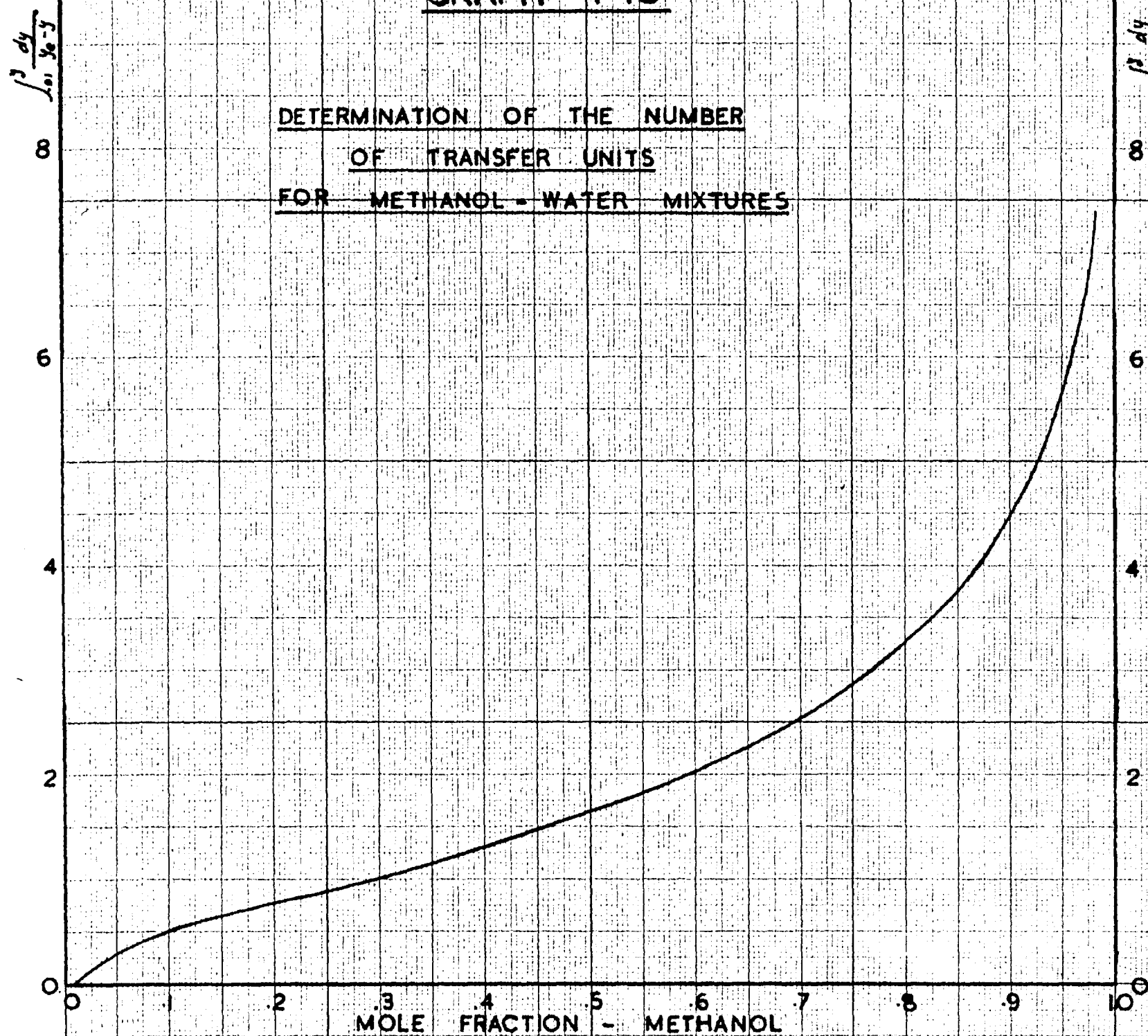
GRAPH 4.14

DETERMINATION OF THE NUMBER
OF TRANSFER UNITS
FOR ACETONE - CHLOROFORM MIXTURES



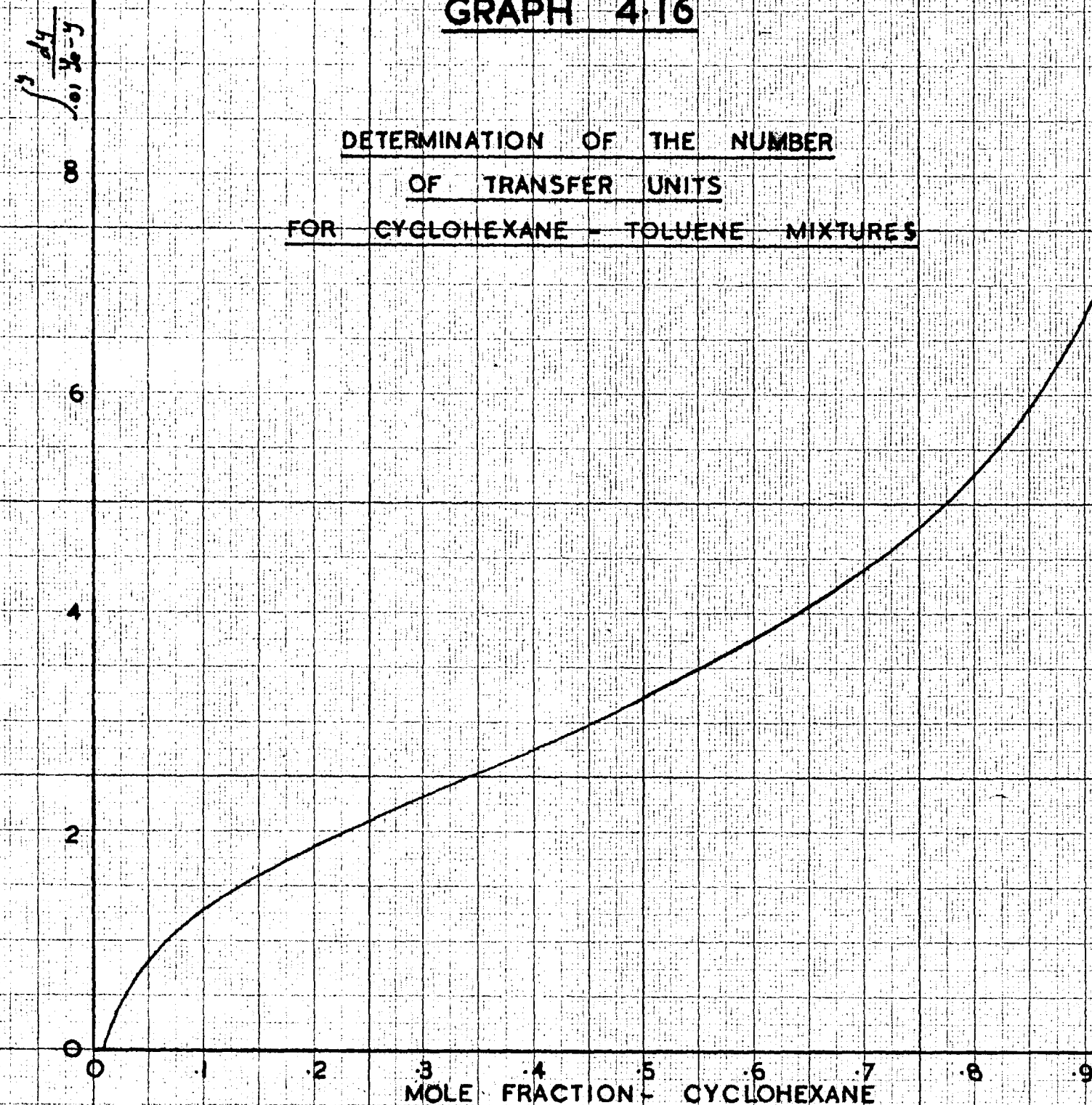
GRAPH 4-15

DETERMINATION OF THE NUMBER
OF TRANSFER UNITS
FOR METHANOL - WATER MIXTURES



GRAPH 4-16

DETERMINATION OF THE NUMBER
OF TRANSFER UNITS
FOR CYCLOHEXANE - TOLUENE MIXTURES



GRAPH 4.17

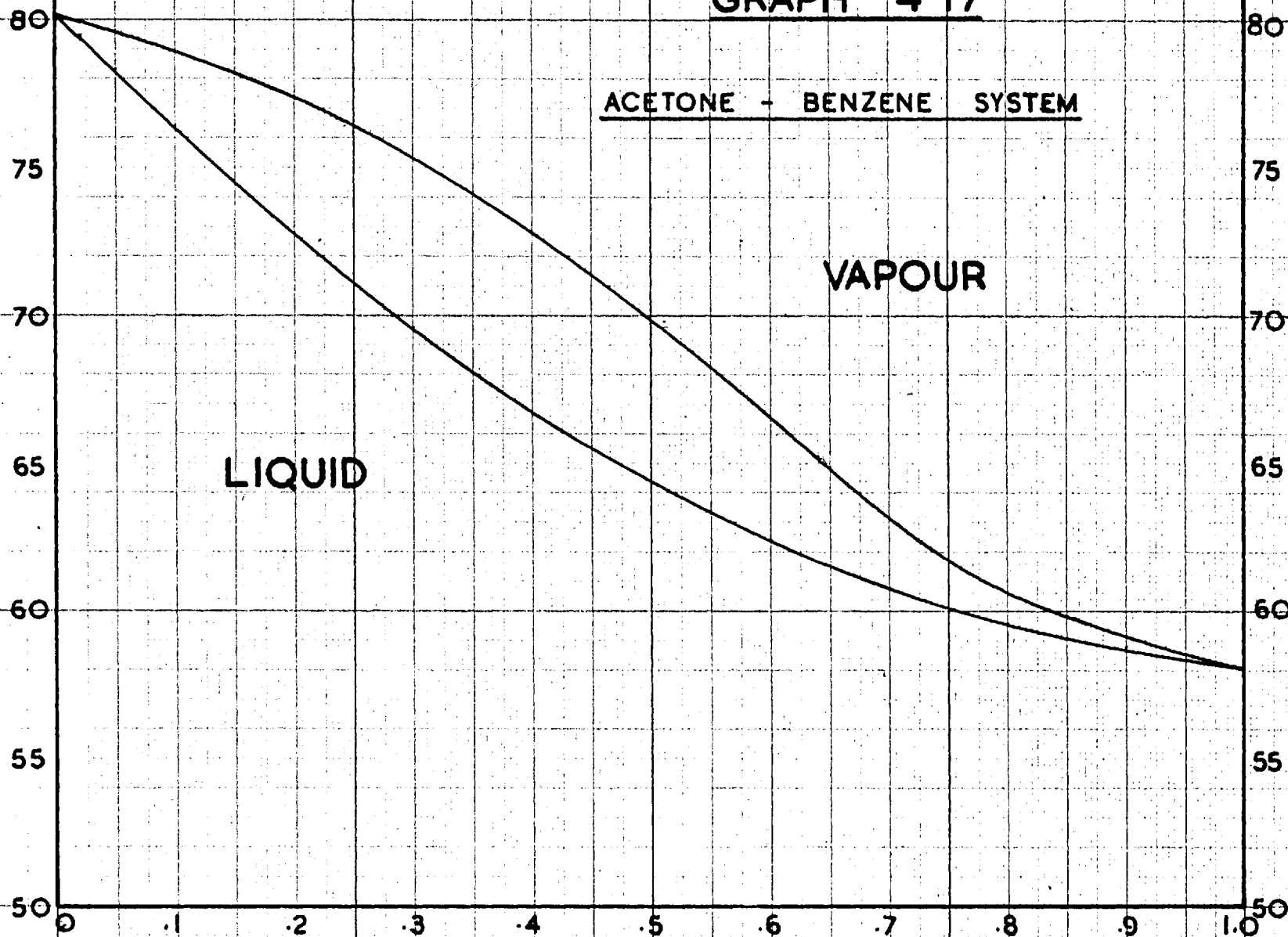
ACETONE - BENZENE SYSTEM

BOILING POINT TEMPERATURE (°C)

VAPOUR

LIQUID

MOLE FRACTION OF ACETONE



BOILING POINT TEMPERATURE (°C)

65

60

55

50

0

.1

.2

.3

.4

.5

.6

.7

.8

.9

1.0

MOLE FRACTION OF ACETONE

LIQUID

VAPOUR

GRAPH 4.18

ACETONE - CHLOROFORM SYSTEM

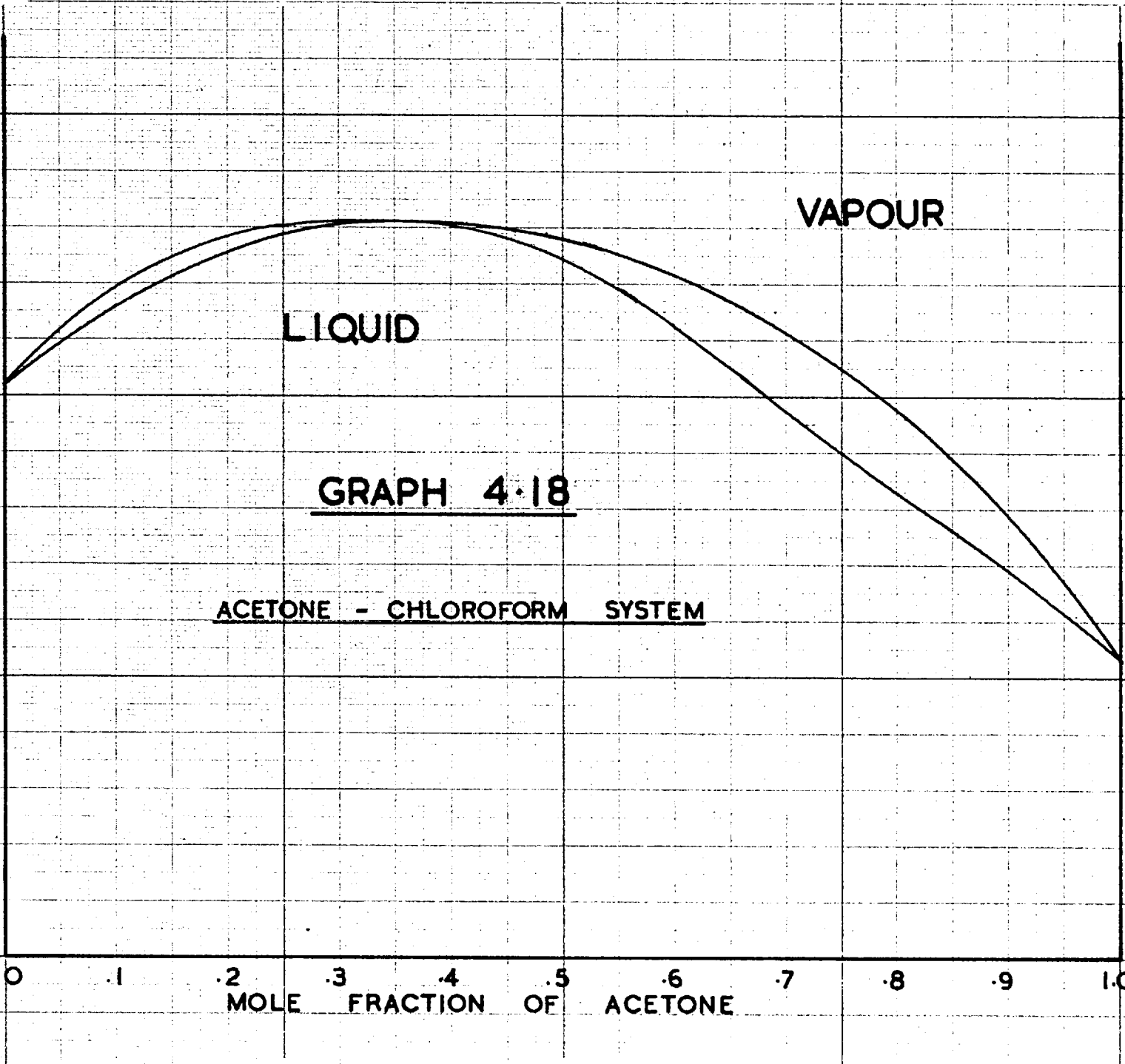
65

60

55

50

80.



PART III

RESULTS AND DISCUSSION

CHAPTER 5.

THE SHORT PACKED COLUMN.

Three binary systems have been used in the packed column described in Chapter 2 : namely carbon tetrachloride-toluene, acetone-benzene and acetone-chloroform. The results obtained show that the separation of these mixtures, as measured by the number of transfer units (N_{OG}) (see Graph 5.1, 5.2, 5.3), depends upon the concentration of the mixture in the column: for a short column the arithmetic mean of the terminal concentrations is used. A distinct maximum in the number of transfer units is obtained for each of the first two systems at a mean concentration of high mole fraction of the more volatile component (m.v.c.); and in the acetone-chloroform system, which has an azeotropic, two maxima are noted. These findings confirm those found by many workers, including Smith [1], Sawistwoski [2] and Qureshi [11]. These workers mostly used tall columns and plotted their results as number of transfer units against the log mean values of concentrations and some physical properties. The maximum values of the performance of the column for all the binary systems they used were, under the same conditions, found to occur at the same log mean

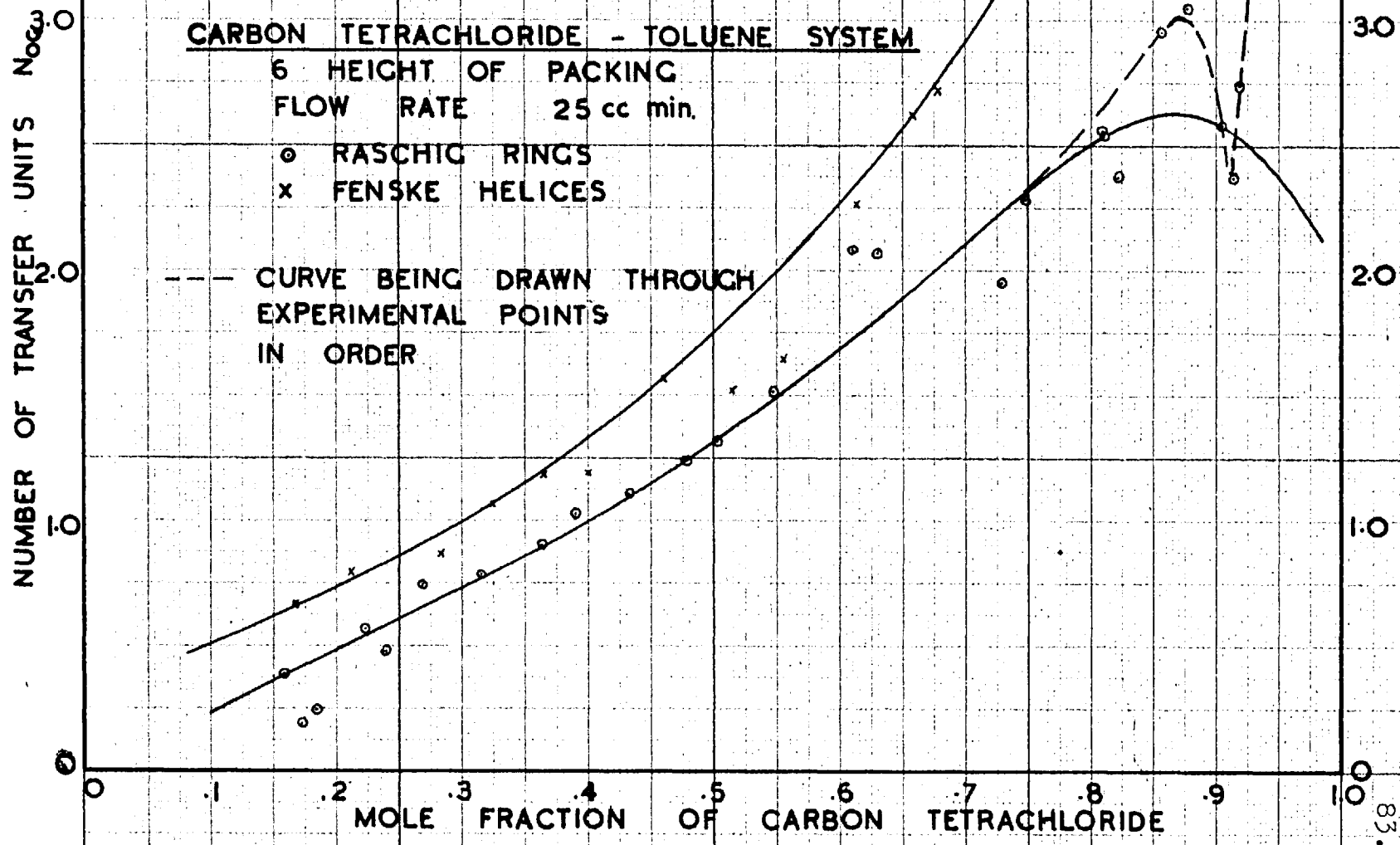
GRAPH 5.1

CARBON TETRACHLORIDE - TOLUENE SYSTEM

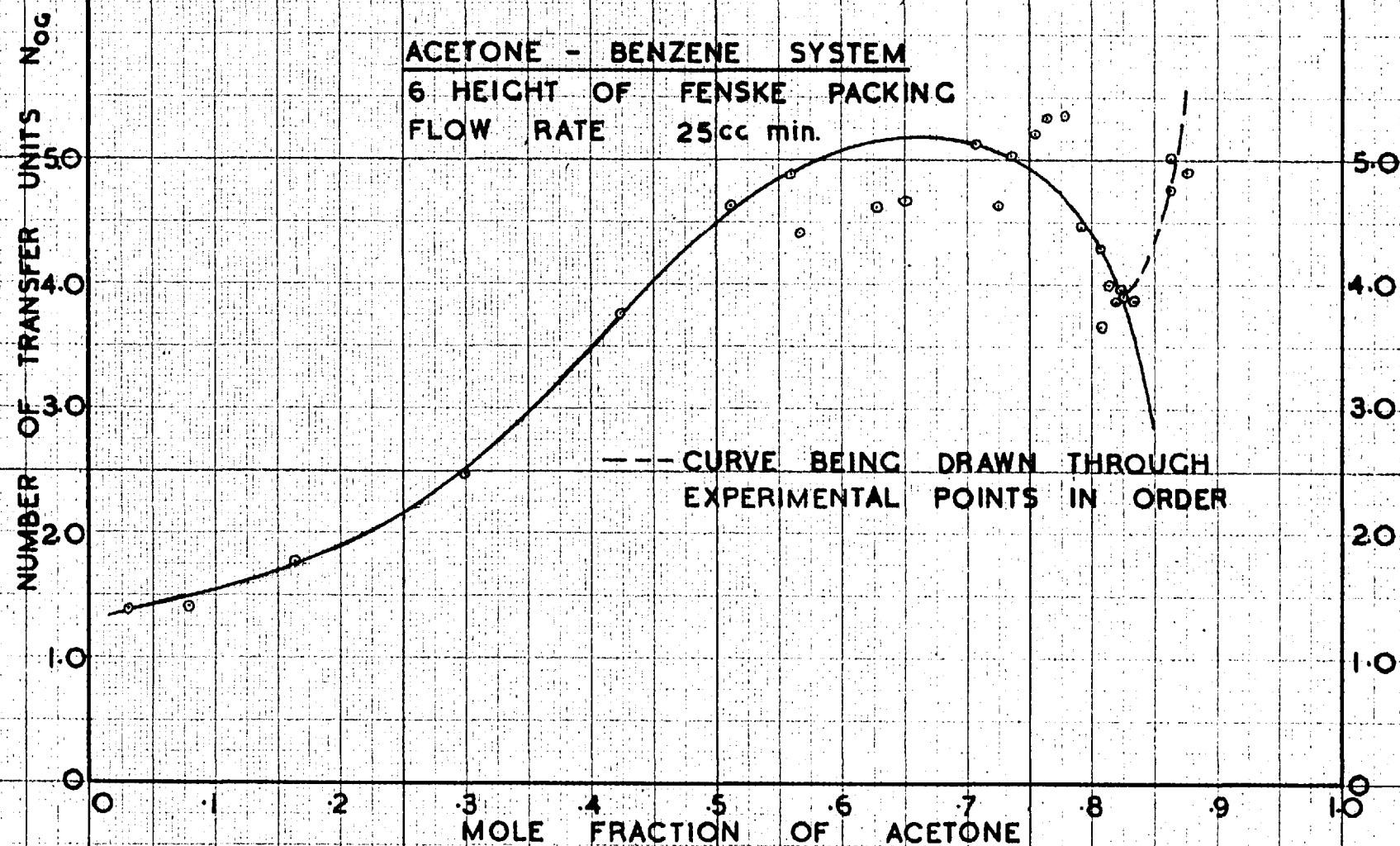
6 HEIGHT OF PACKING
FLOW RATE 25 cc min.

○ RASCHIG RINGS
x FENSKE HELICES

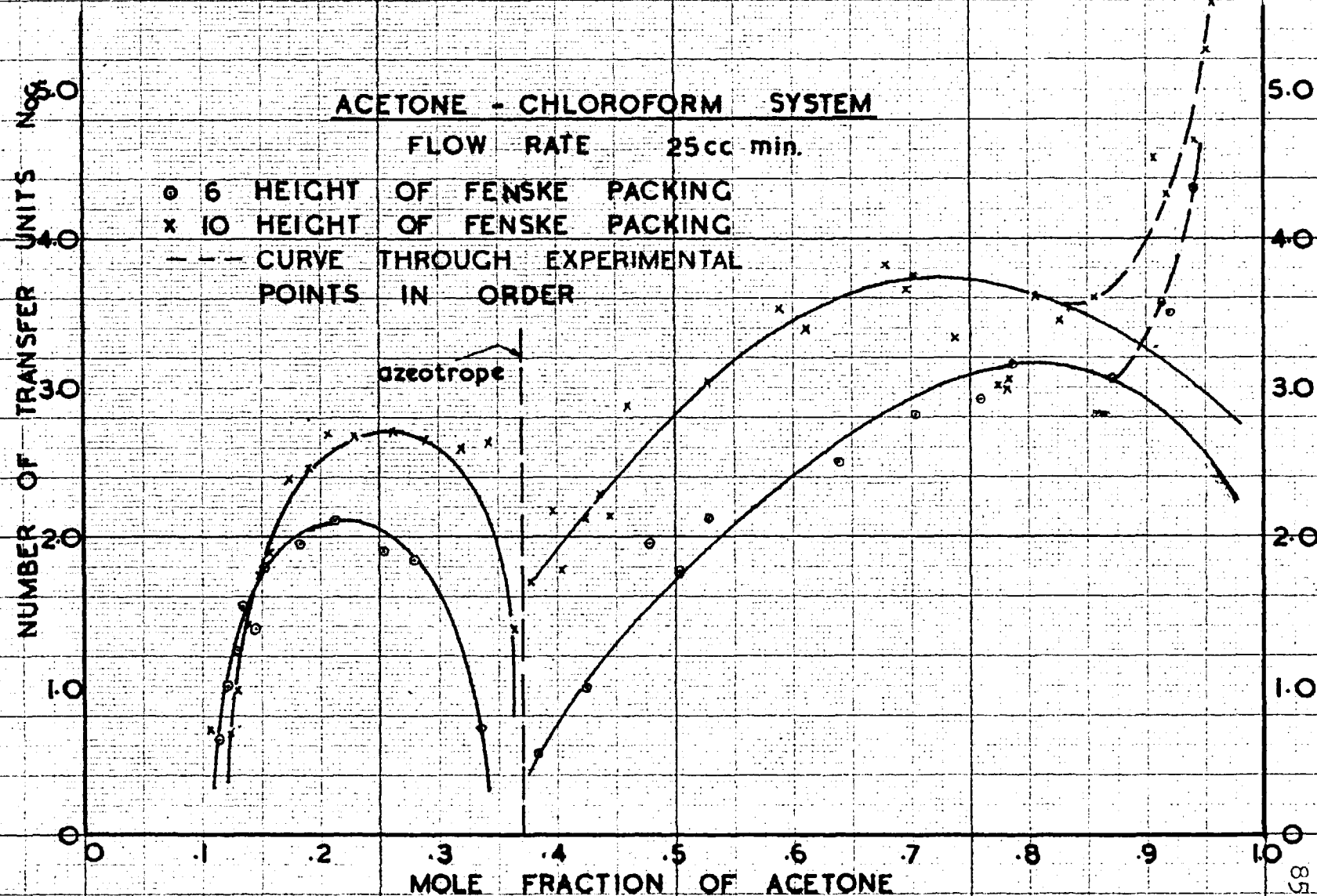
--- CURVE BEING DRAWN THROUGH
EXPERIMENTAL POINTS
IN ORDER



GRAPH 5.2



GRAPH 5.3



concentration as is found here and to be independent of the flow rate. The separation measured for the carbon tetrachloride-toluene system (Graph 5.1) agrees with that obtained by Sawistowski [2], the slight difference in the position of the maximum being caused by the use of log mean instead of arithmetic mean concentrations.

The distillation of mixtures which are dilute has especially been studied, using the packed column, for all the binary systems used in this work: also the separation of mixtures close to the azeotropic composition has been measured. In the carbon tetrachloride-toluene and acetone-benzene systems, the results show a rapid decrease in the number of transfer units (N_{OG}) or in the efficiency as the mixture becomes dilute and appears to indicate that the efficiency is zero for a pure component: this is inconsistent with mass transfer theory.

For the acetone-chloroform system, which has an azeotrope, two distinct maxima have been found, and also a minimum in the number of transfer units occurs at the azeotropic composition. Qureshi [11] who distilled this system in a wetted-wall column also noticed the two maximum values of the efficiency of this system. Besides the maxima found in all three binary systems used, the N_{OG}

risers sharply as the mole fraction of the m.v.c. tends to one (Graphs 5.1, 5.2, 5.3). This increase in efficiency has also been noted by Stanislas and Smith [57] when distilling an acetone-water system in a plate column. Values measured for the efficiency are not accurate when there is only a little of one component present and it is uncertain whether or not this result may be due to experimental error.

Many reasons have been suggested for the dependence of N_{OG} (or $H_{OG} = \frac{\text{Height of column}}{N_{OG}}$) on concentration. They are as follows.

A. HYDRODYNAMIC EFFECT.

Sawistowski [2] studied an identical distillation separation in columns which brought the liquid and vapour into contact in several different ways. He found that the shape of the curve of N_{OG} versus concentration and the position of the maximum, but not its magnitude, stayed the same. Since maxima were found by Qureshi [11] using a wetted-wall column and have also been noted for distillation in plate columns [10], it is concluded that the nature of the phase flow does not affect the form of the relationships between N_{OG} and the composition of mixture undergoing separation.

B. PHYSICAL PROPERTIES.

The relationship between the overall (H_{OG}) and the film coefficients of mass transfer (H_G and H_L in vapour and liquid phases respectively) is

$$H_{OG} = H_G + m \frac{G}{L} H_L \quad (5.1)$$

The dependence of H_{OG} on concentration has often been ascribed to the dependence of the film coefficients of mass transfer on values of the physical properties of the mixture, which themselves vary with composition, and to the fact that m , the mean slope of the equilibrium curve, also varies with composition. Several relationships between the film coefficients of mass transfer and the physical properties of the mixture and flow rates of the phases have been suggested. These are expressed in terms of dimensionless groups; for instance by Pratt [58], for vapour phase resistance, and by Chari and Storrow [59] for liquid phase resistance. The magnitude of constants required to correlate these expressions with experimental results was determined from experimental data, mostly using wetted-wall columns. Typical results found by these methods, for a particular arrangement of packing and at a given flow rate, with those properties depending upon concentration as the

only variables, are

$$K_G \propto D_G^{0.67} \cdot \gamma_G^{-0.42} \quad (5.2)$$

$$\text{and } K_L \propto D_L^{0.48} \cdot \gamma_L^{-0.15} \quad (5.3)$$

where K_G and K_L are mass transfer coefficients in vapour and liquid phase respectively,

D_G and D_L are diffusion coefficients in vapour and liquid phase respectively,

γ_G and γ_L are kinematic viscosities in vapour and liquid phase respectively.

Smith and Sawistowski [18] calculated the possible variation of K_G over the whole range of composition and found it to be less than 5%: thus, for practical purposes, K_G can be assumed constant. K_L , however, varies by about 25% over the full composition range for the cyclohexane-toluene system.

Accounting for this Smith and Sawistowski [18] obtained the equation

$$\int_{y_B}^{y_T} \left(\frac{A}{K_G} + \frac{m}{1.98 - 0.54 x} \right) \frac{dy}{y_e - y} = \frac{aAZ}{G} \quad (5.4)$$

for the cyclohexane-toluene system and found that $\frac{aAZ}{G}$ still varies considerably in value with composition. Evidently the effect of a change in value of physical properties on the transfer unit, as predicted by these correlations, based essentially on diffusion theory, is insufficient to account for the measured variation.

There is the possibility that the interfacial area or interfacial resistance, if present, is determined by the value of some physical property, for instance surface tension. This will be discussed in more detail later, but it can be seen qualitatively that the action of any physical property cannot be the major factor determining the shape of the N_{OG} versus composition curve; for the value of a physical property (e.g. surface tension) at the azeotropic composition, which appears to act in distillation as a pure component, has no unique value, and yet there is a unique value of N_{OG} at this composition. For these reasons it is suggested that variation in physical properties is not the main factor affecting the dependence of N_{OG} on composition.

C. SLOPE OF THE EQUILIBRIUM CURVE, m .

Although physical properties do not have a maximum or minimum value either at the azeotropic composition or for pure components, equilibrium properties, such as temperature difference or concentration difference between the phases, do have a unique value at these compositions. An equilibrium property is $\bar{m}$, the mean slope of the equilibrium curve. Shilling, Beyer and Watson [10] noticed that the efficiency of separation in bubble-plate column depended on composition

and suggested that this was due to the change in slope of the equilibrium curve with composition. Equation (5.1) shows that H_{OG} is dependent on the vapour-liquid equilibrium relations, owing to $\bar{m}$, and will not be constant throughout the packing if $\bar{m}$ varies. The film coefficients should be independent of composition, except in so far as they depend upon the physical properties of the mixtures, which themselves depend upon composition [58, 59]. As has been pointed out, this is relatively a small effect and therefore, at total reflux ($G=L$), the H_{OG} versus $\bar{m}$ curve should be almost a straight line with slope equal to H_L and intercept on the ordinate axis equal to H_G . Smith and Sawistowski [1, 2, 18] plotted this curve from experimental data for binary mixtures and obtained curves concave downwards. A curve of this type implies negative film transfer units over a part of the composition range. Thus, a negative slope at low values of m means a negative value of H_L , and a negative intercept on the ordinate axis means negative values of H_G . This is anomalous, for negative film units indicate that material is being transferred against its own concentration gradient.

D. MALDISTRIBUTION EFFECT.

For maximum efficiency in the transfer of mass, both the liquid and the vapour should be uniformly distributed over the column cross-section, and this uniformity should persist throughout the packed height. In practice, however, the distribution alters considerably as the liquid proceeds down the packing, even though there is a uniform distribution at the top. Mullin [60] has studied the effect of liquid maldistribution on the performance of a packed distillation column and has developed an equation relating the operating group, $\frac{G}{L}$, (G and L being the molal rate of flow of vapour and liquid per unit cross-sectional area of column) and the degree of maldistribution, M. For the purpose of calculation, Mullin considered a type of maldistribution where a tower was divided, theoretically, into two vertical sections, each taking half the total gas flow but unequal flows of liquid. By the degree of maldistribution is meant the ratio of the flows of liquid into these two vertical sections. The equation developed is

$$n\left(\frac{G}{L}\right)_{\text{mean}} = \frac{(M+1)^2}{4M} \cdot n\left(\frac{G}{L}\right)_{\text{overall}} \quad (5.5)$$

It may be seen from this equation that an increase in the

liquid distribution would decrease the separation of the mixture, although not directly the performance of the column, since it increases the value of the operating gas to liquid ratio (G/L), that is, it decreases the slope of the operating line (L/G), resulting in a closer approach of this line to the equilibrium line at one end of the tower. Thus, the measured separation would be less and the apparent performance of the column smaller.

Zuiderweg [61] showed how maldistribution can affect the performance of a distillation column, packed with 1/4 inch glass spheres, and operated at total reflux. Serious maldistribution of liquid was obtained by inserting a glass rod through the entire length of the column. Results obtained at constant boil-up rate showed a decrease in overall efficiency by about 35% to 40%.

In our experiments, the liquid rate and other column conditions have been kept constant and hence the degree of maldistribution may be considered to remain almost constant. The conclusions of Mullin can only be used as an explanation of our results in the region where the mole fraction of m.v.c. is low, as already explained, but cannot explain the decrease in performance at high mole fraction of m.v.c. or at azeotropes. Maldistribution

would be serious in a tall packed column, but for such a short column as that used in this work (6" and 10" height of packing) the degree of maldistribution will be small and only a slight reduction in performance will occur. Mullin showed that the percentage reduction in performance is small for values of M less than 2, which applies to our short columns. The reduction in performance caused by the liquid maldistribution also depends on the number of transfer units exhibited by the column. Thus Morris [62] and Mullin [60] showed that maldistribution has only a slight effect on the performance of columns equivalent to 5 transfer units or less (even though $M > 2$). The short columns used in this work have fewer transfer units than 5. In addition, it is unlikely that maldistribution will occur in the pear column (v.i.) unless there is channelling down the walls. Hence it is believed that maldistribution cannot explain satisfactorily the results obtained in this work.

E. INTERFACE RESISTANCE.

In solvent extraction the effect of surface active agents on the rate of mass transfer has been studied [30, 32, 33]. The agent causes an interface resistance

but simultaneously increases the surface area of contact, giving a net increase in the rate of transfer in some cases. Similar behaviour has been noted in absorption [36]. In neither case does there appear to be a large resistance in the absence of surface-active agents.

That there may be a resistance at the interface in distillation has been suggested in the past by many workers [2, 14, 63], but there is no experimental evidence for this. The small size of an interfacial resistance possible can certainly not account for the magnitude of the change in H_{OG} with composition.

F. SURFACE TENSION AND WETTING.

Zuiderweg and Harmens [12] have studied the effect of changes in surface tension on the formation of interfaces in different types of equipment. They found that liquid films are stabilised when the surface tension of the reflux increases down the column; otherwise the liquid stream breaks up into narrow rivulets or droplets. They concluded that in columns with a supported interfacial area, (for instance, wetted wall columns and packed columns), for positive changes in surface tension (that is, the surface tension of reflux increases down the column) there is no influence of the driving force on the performance. This is because

wetting of the supporting surface is almost complete. This also applies to negative systems (i.e., the surface tension of reflux decreases down the column) at low driving forces in columns with supported interfacial area. But in plate columns there will be an effect of driving force on performance in positive systems, since the dispersion of vapour into liquid frequently leads to the formation of foam, the life-time of which may increase considerably because of the reinforcement of weak spots (i.e. the weak spots become preferentially saturated with the component with the highest surface tension). These weak spots occur between adjacent bubbles and the preferential saturation of these spots with heavy component causes the surface tension to rise locally. Liquid is, therefore, drawn in between the bubbles and coalescence of adjacent bubbles is prevented. Thus this increases the interfacial area and hence the performance. For neutral or negative systems, however, the stabilising factor tending to produce prolonged bubble life is absent, and there is no effect on the performance.

Sawistowski [28] has put forward another explanation for the effect of surface tension on performance. In addition to the effect on the magnitude of the

interfacial area, as a result of the presence of a surface tension gradient, surface tension may also influence mixing and surface renewal in the liquid phase for positive systems. The local concentration change at the interface, owing to the transfer of mass, introduces unbalanced surface tension forces. Since in a positive system the bulk liquid has a lower surface tension than the liquid at the interface, the liquid at the interface will spread across the surface and cause convection currents. These currents will reinforce the normal surface-renewal mechanism and thus increase the mass-transfer coefficient. This phenomenon has been observed in liquid-liquid extraction [64, 65, 66] and is confirmed indirectly by the results of Grassmann and Anderes [67] on transfer from single bubbles: they found that a higher mass-transfer coefficient is obtained if mass transfer takes place in the direction of increasing surface tension at the interface. In negative systems, however, any liquid brought to the surface from the interior will have a higher surface tension and the resulting forces will tend to prevent this liquid from spreading over the surface and renewing it. Hence the rate of mass transfer will be decreased.

The binary mixtures used in this work are positive systems but the difference in surface tension between the pure components for carbon tetrachloride-toluene and cyclohexane-toluene systems is comparatively small (less than 4 dynes/cm.), though some systems show large differences in surface tension. Since in the carbon tetrachloride-toluene and cyclohexane-toluene systems it is still found that the performance changes markedly with composition, gradients in surface tension cannot be the main factor determining the distillation process.

Many other explanations have been put forward for the dependence of H_{OG} on composition, but none appears to account for the magnitude of the observed change. Nor is there any explanation of the fact that two maxima in the relation are obtained when an azeotropic mixture is distilled, the azeotropic compound acting like a pure component in this sense. Evidently the phenomenon is principally dependent upon some equilibrium property, governed by the equilibrium curve (e.g. $y_e - y$) and not upon some physical property (e.g.

surface tension). This leads to the idea, described in the next chapter, that thermal distillation, which depends upon a temperature difference between the phases, may be causing the maximum.

CHAPTER 6.THERMAL DISTILLATION : RESULTS AND DISCUSSION.A. PRELIMINARY DISCUSSION.

As has been pointed out, the results obtained indicate that the performance of a column appears to depend upon the equilibrium rather than the physical properties of a mixture. For this reason it is thought that thermal distillation, which is governed by the temperature difference between the phases (an equilibrium property) may be the prime cause of the dependence of N_{OG} on composition. By thermal distillation is meant condensation or evaporation in the bulk phases or at the interface, this being caused by local excess or shortage of heat over and above that required to keep the phases at their boiling or dew points. Distillation by diffusional process is referred to as contact distillation.

Thermal distillation was first considered by Kirschbaum. He assumed that heat and mass transfer between the phases were equivalent throughout the column; and, essentially, by his method there is no distinction between temperature and composition as a measure of separation. According to Kirschbaum the performance, E , of a column, expressed in terms of temperature driving force [18,68] is

$$E = \frac{\frac{dy}{dz}}{t_G - t_L} = \frac{h a}{\gamma G} \quad (6.1)$$

where y = bulk vapour composition of the more volatile component; mole fraction

z = height of packing,

t = temperature,

h = heat transfer coefficient,

a = interfacial area per unit volume of column,

G = molal rate of flow of vapour per unit cross-sectional area of column,

γ = molal latent heat.

Subscripts G and L refer to the vapour and liquid phase respectively.

Performance expressed on this basis also has a definite maximum value when plotted against composition [18], and as with the H_{OG} this cannot be explained by considering the change in the magnitude of those physical properties affecting individual heat transfer coefficients [18]. Kirschbaum has recently abandoned this approach [69].

Smith [28], however, has drawn attention to the fact that the rate of heat transfer and the rate of mass transfer are not necessarily equivalent. As has been described in the Introduction, depending upon the ratio of the heat to the mass transfer coefficient, the vapour phase can superheat or undergo partial condensation and the liquid phase can subcool or partially boil. If there is

condensation or boiling in the bulk phases this will affect the transfer of components in addition to that given by contact transfer, that is, by diffusion. Such thermal distillation will depend upon temperature difference between the phases, which is a maximum towards the centre of the composition range.

In this thesis we are not principally concerned with predicting or postulating mechanisms to explain why the heat or mass transfer coefficients through the boundary layers close to the interface should have the value indicated by these experiments; nor are we concerned with the mechanism by which vapour condenses or liquid boils in the bulk phases, nor with the efficiency with which this latter process takes place. Speculation on these subjects has been made before [28] and it is hoped to investigate some of these problems later. Instead we are going to show by making heat and material balances that taking into account thermal distillation partly explains the experimental results obtained, and that values of flow rate and temperature measured along the column are consistent with the theory developed.

A qualitative explanation of the action of thermal distillation on the separation of a binary mixture by a

continuous countercurrent fractionating column operating in a steady state is as follows.

We assume the rate of transfer of mass by contact distillation is determined by the composition difference between the bulk phases. In the vapour less volatile component (l.v.c.) moves to the interface carrying its latent heat. By some means this is transferred to the more volatile component (m.v.c.) which carries it back to the bulk vapour. In a simple process where the molal latent heat is constant there would be no net energy transfer owing to this source, and in other cases the net energy transfer is small, although it will be taken into account here.

In addition to this energy transfer there is heat transfer between the phases owing to and determined by the temperature difference which always exists between the phases at the same point in the column. Most of this latter heat transfer accounts for the change in temperature of the phases as they pass through the column. It is the discrepancy between this heat transfer and the change in heat content of the phases which partly causes thermal distillation. The heat transfer coefficient between phases will depend upon the state of turbulence of the moving phases as does the mass transfer coefficient but the rate of heat transfer is not related in

any direct way to the rate of mass transfer [18] which additionally depend upon temperature and composition difference. The rate of heat transfer will not necessarily correspond exactly with that needed to keep the phases just at their condensation or boiling points. For instance, if in equivalent terms the heat transfer was much less than the mass transfer, the condensation point would fall more rapidly than the temperature of the vapour as it passes through the column, and the vapour would thereby superheat: likewise the liquid may subcool (line SS on Fig.6.1). In the opposite case where the mass transfer is in equivalent terms less than the heat transfer, the temperature of the vapour would remain at the condensation point and additional transfer of heat would cause condensation of vapour: likewise the liquid may boil. Condensation or boiling gives a product of different composition than the bulk phase it came from, and if this product physically joins the other phase, a transfer of mass has been effected in addition to the normal contact transfer.

Further, it is not necessary and indeed it is unlikely that the rate of heat transfer in each phase is equal, since the ratio of the heat to mass conductivity although about equal in the vapour, is high in a liquid.

A net transfer of mass must then take place at the interface, condensation being more likely [18], the latent heat so liberated making up the difference in heat transfer rate. Condensation or evaporation at the interface will effect mass transfer in the same way as it does in the bulk phases.

Mass transfer by thermal distillation under control of temperature difference is additional to that by contact distillation under control of a concentration gradient and this extra transfer of mass may partly account for the discrepancy between the theory of contact transfer processes and the experimental results obtained for distillation in this work and previous work [1,2,10,11]. The experimental work carried out in the pear column was devised to investigate and to measure those quantities required to study thermal distillation: namely the temperatures of the bulk phases and the net change in the molal flow rate along the column.

Another result of condensation or evaporation was pointed out by Hartnett and Eckert [70] who studied mass transfer in a laminar boundary layer with constant fluid properties. They found that mass added or removed by condensation or evaporation destabilises the boundary layer.

The destabilisation of the boundary layer causes breaking up of the interface and a wave motion or turbulence is set up. The rate of transfer of mass is then governed by both molecular and eddy diffusion.. There is an increase in interfacial area and this additionally increases the rate of mass transfer. Again Schulte-Vieting [71] assumed vapour side control and pointed out that condensation and evaporation depends on the value $\Delta y = y - y_e$, that is, whether it is negative or positive. The amount depends on the value of Δy , and in general they increase as Δy increases.

B. EXPERIMENTAL RESULTS.

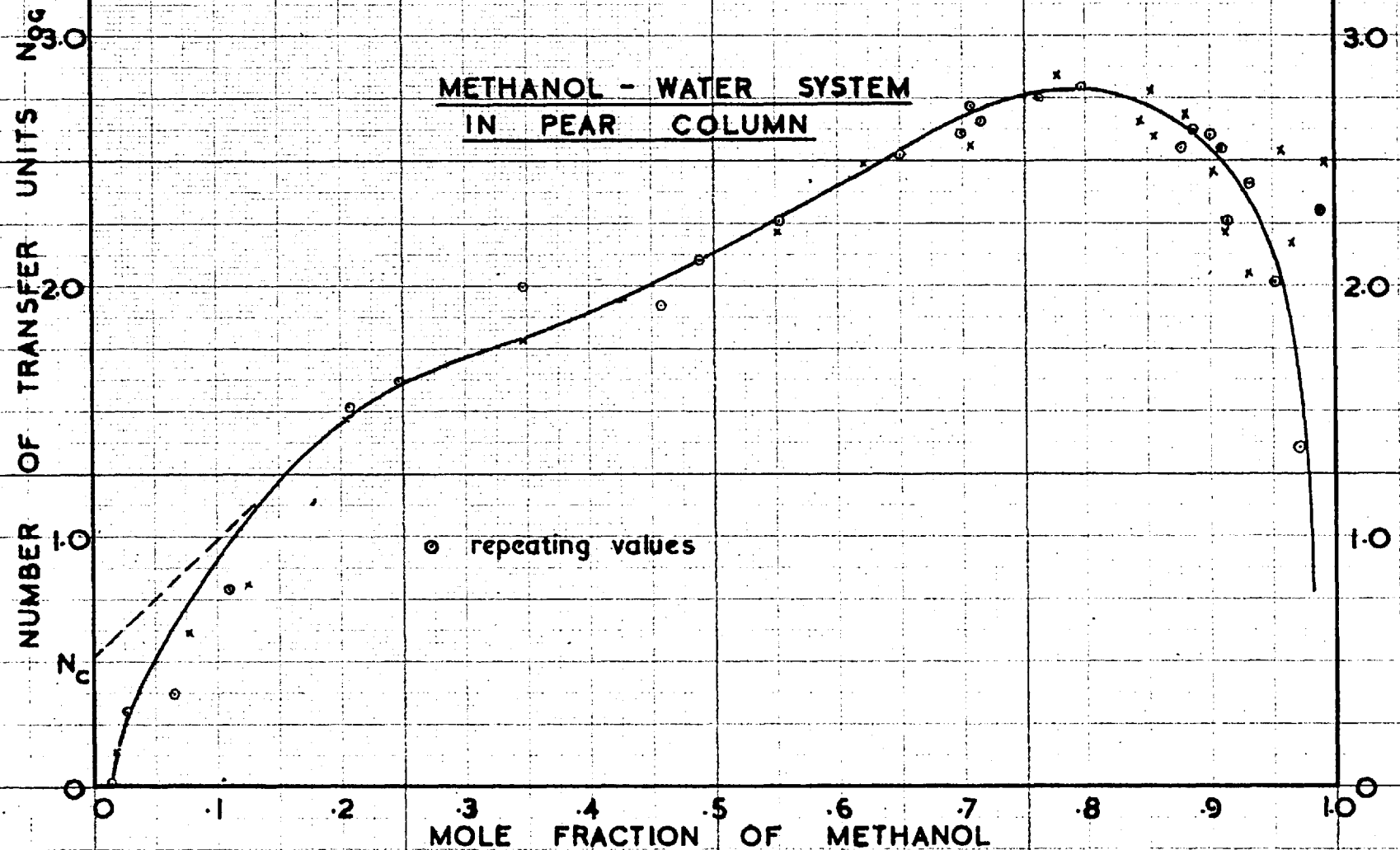
In the pear column, the N_{OG} versus composition curves obtained (Graph 6.1, 6.2) for the binary systems, methanol-water and cyclohexane-toluene, have a maximum value and the same shape as was found with the short packed column. This provides further evidence that the method by which phases flow through a column, different again in the pear column, does not affect the shape but only the magnitude of the relationship between N_{OG} and composition.

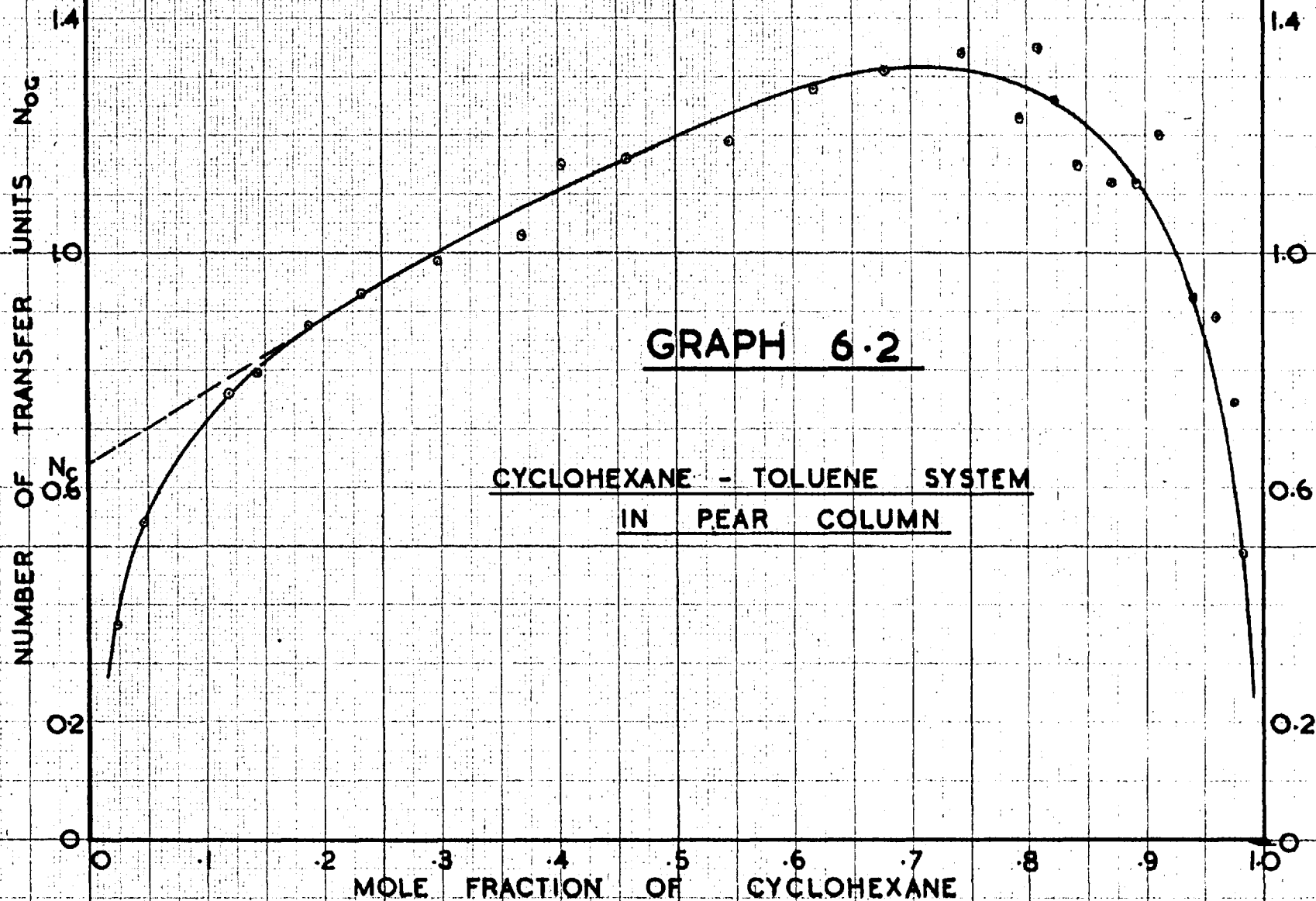
The temperatures of the vapour and liquid phases measured for both systems show that the liquid phase stays

at its boiling point while vapour phase appears to be slightly subcooled. In the methanol-water system, the greatest degree of subcooling measured is 4.6°C . (Graph 6.3) as compared with a maximum of 18°C . of the temperature difference between the condensation and boiling points. In the cyclohexane-toluene system, the greatest value of subcooling is 0.8°C . (Graph 6.4) as compared with the maximum temperature difference of 6.4°C . between the two phases.

The degree of subcooling (Graphs 6.3, 6.4) varies with composition: it is greater near the centre where the temperature difference between the phases is wide and is small when the mixture is dilute. Although it is conceivable that a vapour could be slightly subcooled, it is more likely that the measured temperature is less than the condensation point owing to cooler liquid entrained in the vapour phase. When the composition of the still is near 0.5 m.f., the number of theoretical stages will be greater, and the difference in temperature of the two phases will be larger. Then a small quantity of entrainment, say less than 1% of liquid flow rate (v.i.), will have a greater influence on the measured vapour temperature than would be the case if the temperature difference between the phases is small.

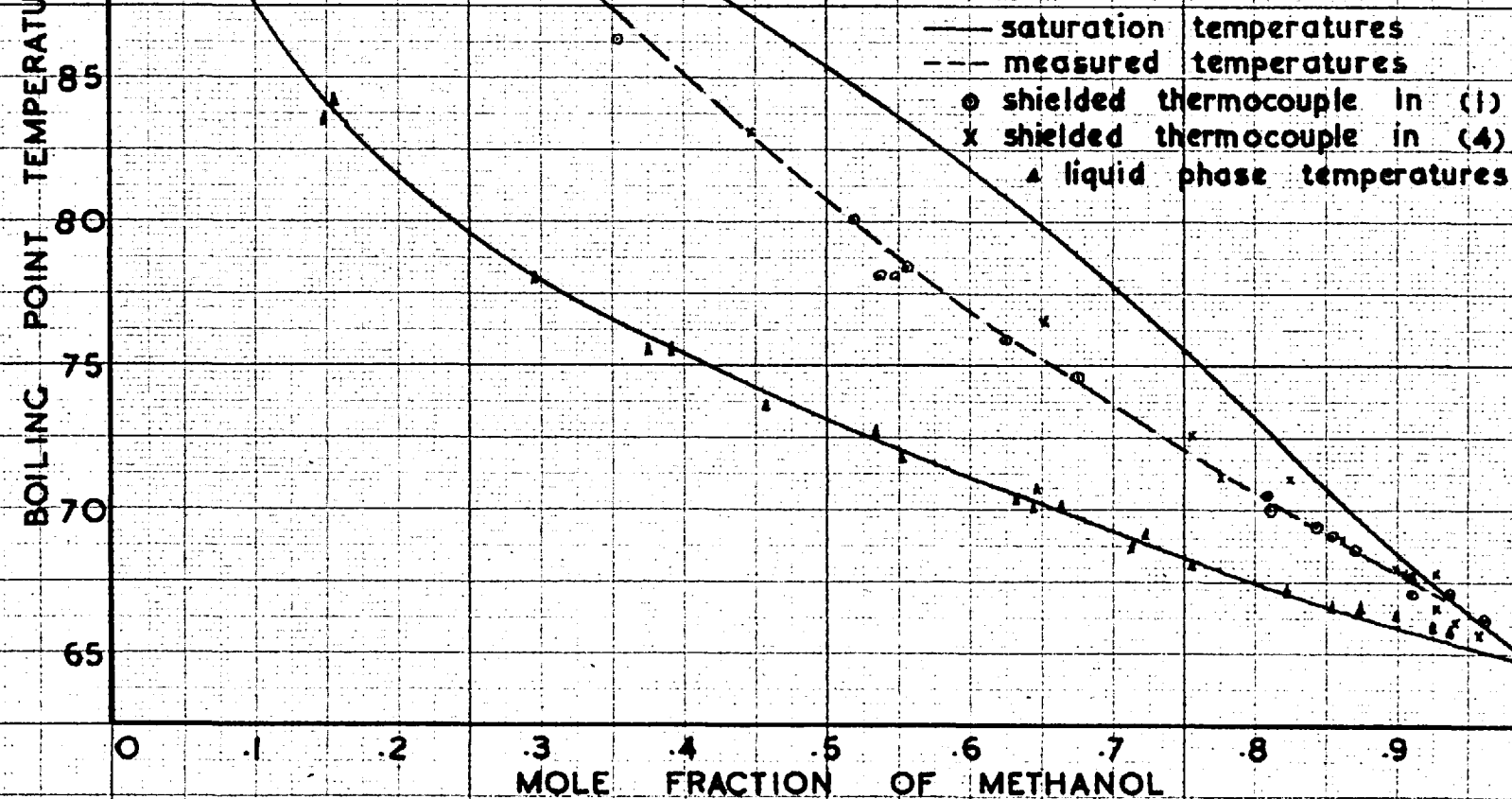
GRAPH 6-1





GRAPH 6.3

THE MEASURED TEMPERATURES OF
THE BULK PHASES OF METHANOL-
WATER SYSTEM



GRAPH 6.4

THE MEASURED TEMPERATURES OF THE
BULK PHASES OF CYCLOHEXANE - TOLUENE SYSTEM

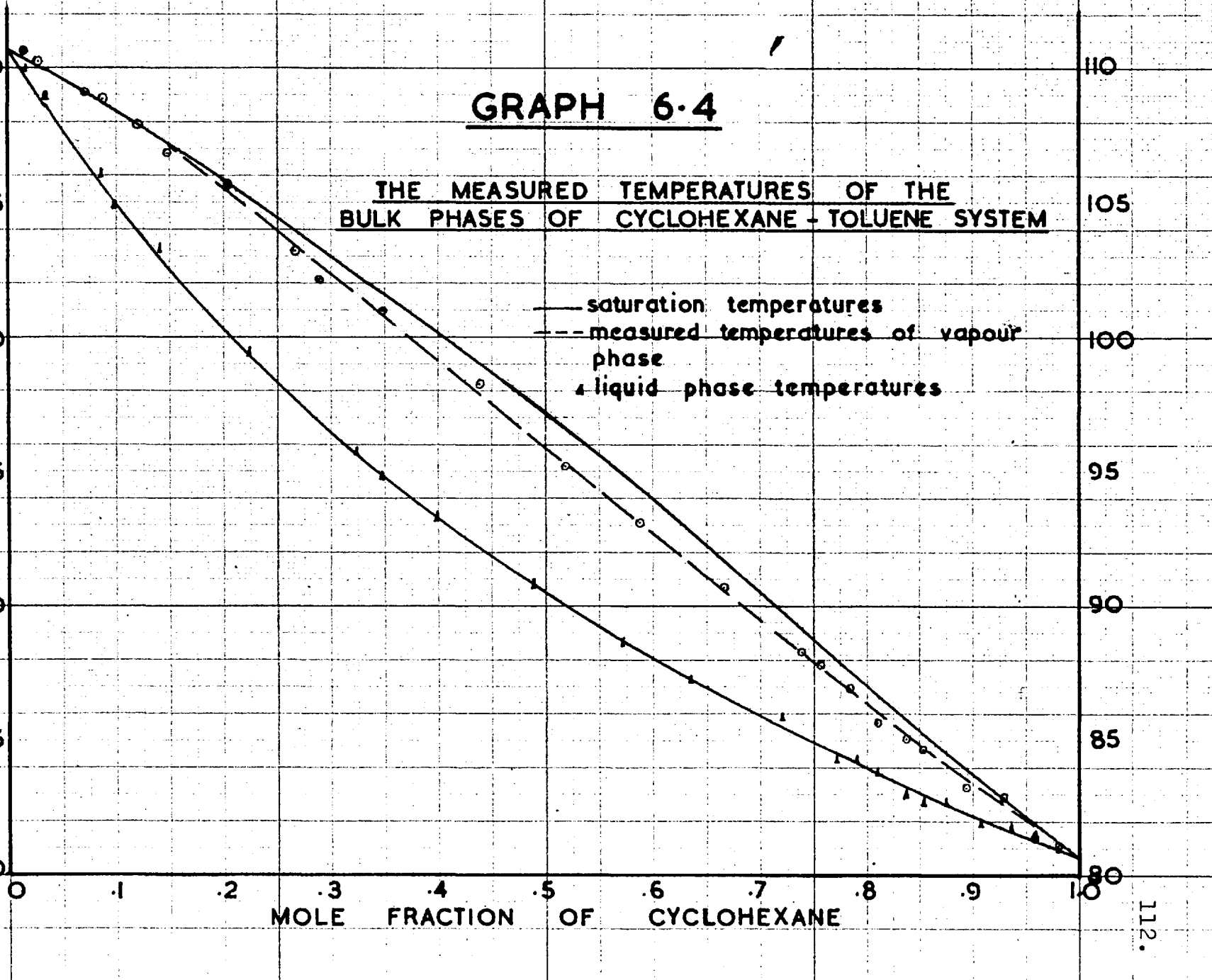
BOILING POINT TEMPERATURE °C

110
105
100
95
90
85
80

110
105
100
95
90
85
80

— saturation temperatures
--- measured temperatures of vapour phase
▲ liquid phase temperatures

MOLE FRACTION OF CYCLOHEXANE



Experimental errors, which must be taken into account, are caused by the fact that it is the liquid composition being measured and not that of the vapour phase, and the thermocouple may not measure the temperature of the vapour at the position where it has the same composition as that of the liquid phase (at total reflux). Also vaporisation of m.v.c. may occur during sampling, a small inaccuracy in composition being important in this connection. That the degree of subcooling in the methanol-water system is greater than that in the cyclohexane-toluene system may be explained by these errors and by the fact that the rate of change of temperature with composition of the former system is greater than that of the latter system: thus a slight inaccuracy in determining the composition of the vapour phase will have a greater effect on the accuracy of measurement in the methanol-water system.

Measurement of the flow rate by introducing concentrated salt solution from the top of the reflux condenser at a constant flow rate showed that there is a change in the flow rate along the length of the column and that the flow rate at the bottom of the column is less than that at the top; that is, there appears to be a net evaporation. The extent of the change in liquid flow rate

varies with composition and there is a maximum (Graph 6.5). The experimental measurement indicates that there is much more evaporation than is specified by heat balance, equation (6.15) (v.i.). It should be noted that the excess evaporation measured over that given by equation (6.15) cannot be due to entrainment in vapour phase, since entrainment will not change the salt concentration in liquid phase.

That the column was adiabatic has been checked as described in Chapter 4. Thus the discrepancy must be due to errors in sampling and in determining the percentage of salt in the samples. The error is calculated as follows.

Let L is the weight of liquid flow rate per unit time,
 C is the percentage by weight of salt in liquid,
 e is the error in salt concentration determination.
 subscripts B and T refer to the bottom and top
 samples.

By a material balance on sodium chloride,

$$\frac{L_T}{L_B} = \frac{C_B}{C_T} \quad (6.2)$$

With an error of e , the measured ratio would be a maximum of

$$\frac{L_T'}{L_B'} = \frac{C_B}{C_T} \left(\frac{1 + e}{1 - e} \right) \quad (6.3)$$

PERCENTAGE OF CHANGE IN FLOW RATE (%)

GRAPH 6.5

METHANOL - WATER SYSTEM
(EXPERIMENTAL RESULTS)

including entrainment

calculated by change in latent heat

0

5

10

.1

.2

.3

.4

.5

.6

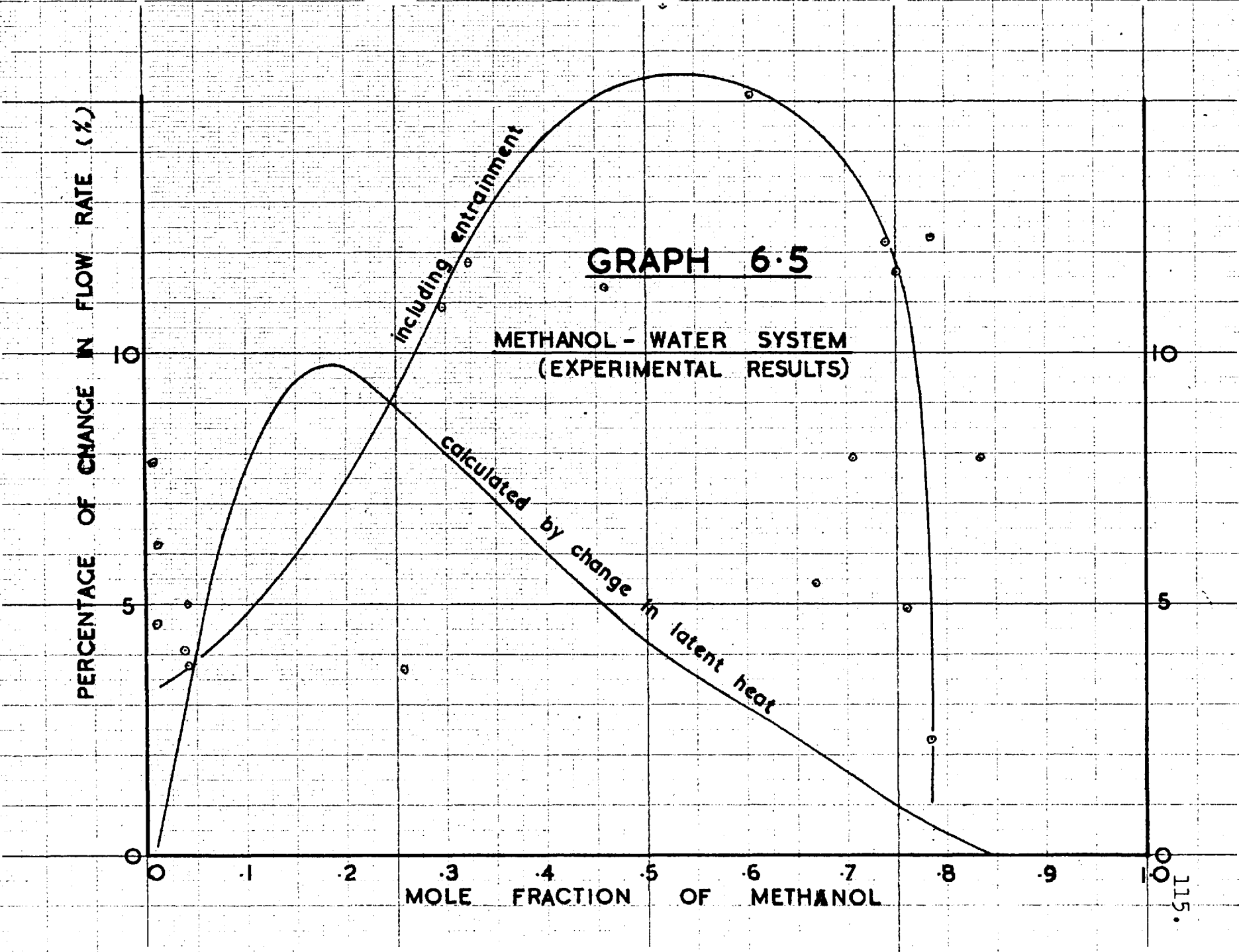
.7

.8

.9

1.0

MOLE FRACTION OF METHANOL



Thus the maximum error in flow rate is

$$\begin{aligned} \left(\frac{L_T}{L_B} - \frac{L_T'}{L_B'} \right) \frac{C_T}{C_B} &= 1 - \frac{1+e}{1-e} \\ &= \frac{-2e}{1-e} \end{aligned} \quad (6.4)$$

From equation (6.4), it is seen that the error in flow rate will be about twice that in the measurement of salt concentration. It is estimated that an error of 3 to 5% is possible in measuring salt concentration owing to vaporisation of the m.v.c. during the sampling and weighing periods and to the difficulty of determining the end point in titration. Thus, the error in flow rate could be 10% and the method is not reliable. It is concluded that there is no change in flow rate along the column, other than the change caused by variation in latent heat. It is also not likely that there is net evaporation anywhere in the column, since a heat balance requires that evaporation in one place in the column must be balanced by condensation in another place; and there is no evidence of an uneven change in flow rate occurring along the column.

That entrainment of liquid occurs in the vapour phase along the column has been confirmed by the detection of traces of sodium chloride in the reflux condensate

after concentrated salt solution had been introduced through a side arm near the top of the column. The amount of sodium chloride present in the reflux condensate cannot be detected with sufficient accuracy. But the conductivity method detects a concentration as low as 0.005% by weight. The reflux condensate contains an amount of salt which can just be detected by conductivity. The maximum value of the entrainment which is picked up by the vapour and carried through and out of the column is then as follows.

Let E be the amount of entrainment per unit time,

L be the weight of liquid flow at the bottom of the column per unit time, and

C_T and C_B be the salt concentration in weight per cent at the top and bottom of the column respectively.

For an experiment in which pure water was used (to exclude changes in flow rate owing to changes in latent heat), a material balance on salt gives :

$$\frac{E}{L} = \frac{C_T}{C_B - C_T} \quad (6.5)$$

With C_T as 0.005%, E/L is 0.34%, that is, less than 1%.

We should however note that there may be entrainment at one level in the column which is removed at another higher level, this process being continuous along the column. Such back

mixing would give a lower efficiency of separation, and if the entrainment depended upon composition may partly account for the dependence of performance upon composition.

The fact that there is a net evaporation, even if owing only to change in the molal latent heat, may as pointed out by Hartnett and Eckert [70] contribute to the variation of the performance with concentration. Suppose the number of transfer units depends upon composition, as shown in Fig.6.3. The rate of mass transfer depends not only upon the interfacial area but also upon the driving force $(y_e - y)$. At the extreme ends a,f, of the curve in Fig.6.3, the concentration driving force $(y_e - y)$ for the transfer of mass is small and the net evaporation is small, since the difference in the latent heat of vaporisation is slight and has little effect on the interfacial area. The efficiency of separation is therefore very low. When the mixture is more concentrated, the concentration driving force increases, and at the same time so also does the temperature driving force. The transfer of mass by net evaporation is larger and hence the interfacial area is larger. These increases favour the rate of mass transfer which will therefore increase rapidly, as shown by the portion 'ab' on the curve. When the mole fraction of m.v.c.

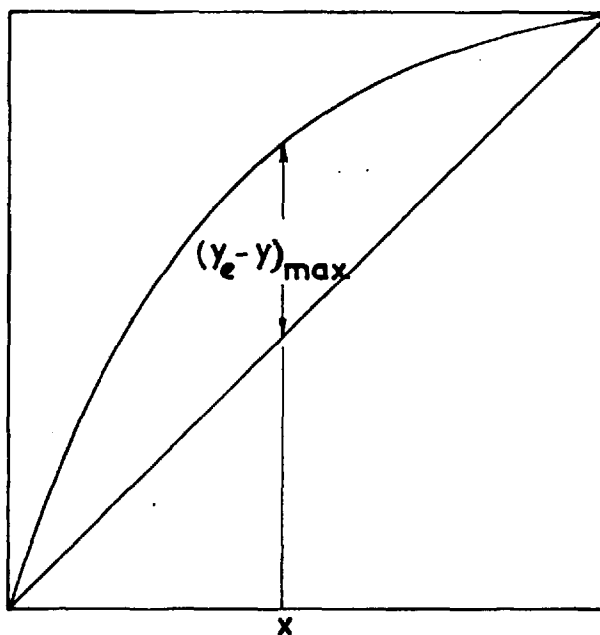


FIG. 6.2. EQUILIBRIUM CURVE

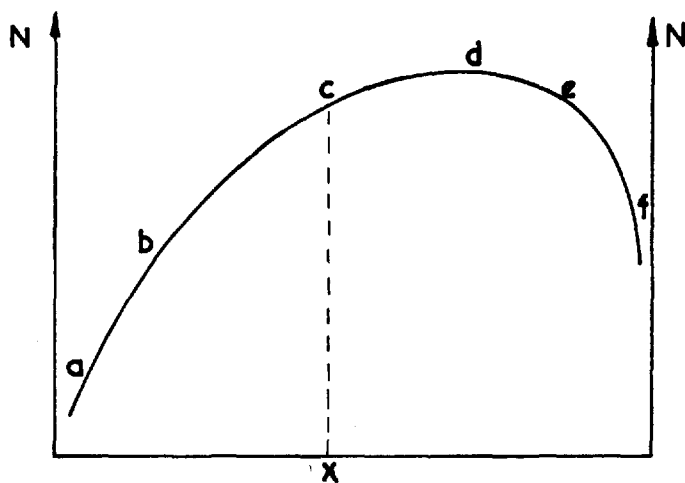


FIG. 6.3. NUMBER OF TRANSFER
UNIT - CONCENTRATION RELATION-
SHIPS FOR BINARY MIXTURE IN
THE ABSENCE OF AZEOTROPE

increases further and passes the composition at which the concentration driving force ($y_e - y$) starts to decrease, it is found experimentally that the efficiency still increases (as shown by 'bc'). At this point it could be stated that the effect of the increase in interfacial area, caused by the net evaporation is greater than the effect of the concentration driving force. At point 'd' the efficiency of separation decreases as the mole fraction of m.v.c. increases; this is caused not only by the gradual decrease in ($y_e - y$) but also by the gradual decrease in the net evaporation. In fact, though a net evaporation would increase the thickness of boundary layer on the gas side and hence increase the diffusional resistance, as reported by Schulte-Vieting [71], yet there is sometimes measured (Chapter 5,6) an increase in performance. This may be explained as follows. There is an increase in concentration driving force at low concentration of m.v.c. in liquid but the interfacial area increases at a much higher rate than does the diffusional resistance, caused by net evaporation. At higher mole fraction, however, the increased diffusional resistance becomes more important and beyond certain concentrations outbalances the advantages of the increase of the interfacial area. By this argument it is possible

to account partly for the change in performance of a column with composition. But it assumes that there is a significant amount of evaporation of the liquid at the interface, and as shown before there is no evidence for this.

C. QUANTITATIVE DISCUSSION.

It has been discussed how thermal transfer can arise in a distillation process. In this section we shall show by heat and material balances how thermal distillation affects the performance of equipment. The equations are derived in detail in Appendix I.

We make the following assumptions for the simplest case in order to derive a relationship for the performance.

- (1) The bulk phases are at their saturation temperatures.
This has been shown to be so by experiment.
- (2) The rate of contact mass transfer is determined by a simple mole fraction, overall composition difference between the phases, as is usual.
- (3) The rate of heat transfer is determined by the overall temperature difference between the phases, the heat transfer coefficient being constant.
- (4) The products of thermal distillation generated in either phase immediately join the opposite stream.

- (5) Account is taken of the fact that the latent heat and total heat of the phases may vary with composition, and lead to non-constant molal overflow.
- (6) In the first case, we assume the products of thermal distillation are in equilibrium with the bulk phase from which they have been generated, and that there is no entrainment of the liquid in bulk vapour.

Using these assumptions heat and material balances lead to the following expression for the ratio of the number of transfer units with (N_T) and without (N_c) thermal distillation (notation in Appendix III); that is, to the ratio of number which are measured experimentally to the number which are predicted by contact mass transfer theory. The latter number, as has been discussed, should be almost constant.

$$\frac{N_T}{N_c} = \frac{1 + \gamma \frac{U'}{K'} \frac{(t_v - t_L)}{(y_c - y)}}{1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) - \gamma(H^* - H)] \frac{dV}{dy}} \quad (6.6)$$

All values, except the ratio of overall heat to mass transfer coefficient (U'/K'), on the right hand side of equation (6.6) are known or can be measured. Thus, by assuming values for U'/K' , a series of N_T/N_c versus composition curves can be calculated. ($\frac{1}{V} \cdot \frac{dV}{dy}$ is calculated

as shown in Appendix I). Such a calculation has been done for both the methanol-water and the cyclohexane-toluene systems (Tables 6.6, 6.14; Graphs 6.16, 6.17). As may be seen, the relationship shows a maximum, which occur at that mole fraction where the concentration and temperature driving forces $[(y_e - y) \text{ and } (t_v - t_L)]$ are greatest, near the centre of concentration range. All values of N_T/N_c are greater than unity except for dilute mixtures when U'/K' is small (less than 10). As is expected, N_T/N_c , at all values of U'/K' , is equal to 1 for pure components.

The experimental values of N_T/N_c are plotted on the same graphs (6.16, 6.17). The number of transfer units given by pure contact mass transfer (N_c) is estimated approximately as the intercept of the tangent on the ordinate drawn from the experimental curve at that point where N_T decreases suddenly and rapidly. On comparing the experimental and theoretical curves two distinct differences are observed:

- (1) The experimental value of the maximum in N_T/N_c occurs at a higher composition.
- (2) The experimental values appear to become zero instead of approaching unity at both ends of the curve, especially at pure l.v.c.

two
 Apart from these important differences between the theory and experiment, it appears that this simplest case of thermal distillation is sufficient to give the maximum in performance. We now consider other factors in order to account for these two discrepancies, such as variation of physical properties, interfacial resistance or maldistribution etc., with concentration. Furthermore, entrainment of liquid by the vapour phase would affect the separation since it would be equivalent to back mixing along the column. Also, the assumption that the products of thermal distillation in either phase are in equilibrium with the bulk phase from which they have been generated, that is, $y^* = y_e$ and $x^* = x_e$, is unlikely to be true, although, as will be shown this can easily be allowed for. Several modifications of equation (6.6) have been made. They are as follows.

(i) Entrainment.

By assuming that there is entrainment (E) in the bulk vapour we obtain (Appendix I)

$$\frac{N_T}{N_C} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'(t_v - t_c)}{K'(y_e - y)} + \frac{\{(y^* - x) - \gamma(H^* - L)\}}{K'(y_e - y)} \frac{dL}{dz}}{1 - \gamma \frac{dR}{dz}} \quad (6.7)$$

which may be rewritten (see Appendix I)

$$\frac{N_T}{N_C} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'(t_v - t_c)}{K'(y_e - y)} - \frac{\{(y^* - x) - \gamma(H^* - L)\}}{(y_e - y)} \frac{z}{N_C} \frac{1}{V} \frac{dV}{dy} \frac{dy}{dz} + \frac{\{(y^* - x) - \gamma(H^* - L)\}}{K'(y_e - y)} \frac{dE}{dz}}{1 - \gamma \frac{dR}{dy}} \quad (6.8)$$

Alternatively,

$$\frac{N_T}{N_c} = \frac{1 + \gamma \frac{U'(t_v - t_c)}{K'(y_c - y)} + \frac{\gamma E}{K'(y_c - y)} \frac{dh}{dz} + \frac{[(y^* - x) - \gamma(H^* - h)]}{K'(y_c - y)} \frac{dE}{dz}}{(1 + \frac{E}{V}) - \gamma \frac{dh}{dy} - [(y^* - x) - \gamma(H^* - h)] \frac{1}{V} \frac{dV}{dy}} \quad (6.9)$$

The variation in liquid composition along the height of column has been measured and is shown in Graphs 6.12, 6.13. From this values of dy/dZ (for equation 6.8) can be measured and average values are shown in Graphs 6.14, 6.15. Still assuming that y^* and x^* are equilibrium values these equations are plotted in Graph 6.18 in four ways:

(a) Without entrainment

$$\text{i.e. } \frac{N_T}{N_c} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'(t_v - t_c)}{K'(y_c - y)} - \frac{[(y^* - x) - \gamma(H^* - h)]}{(y_c - y)} \cdot \frac{Z}{N_c} \cdot \frac{1}{V} \frac{dV}{dy} \frac{dy}{dz}}{1 - \gamma \frac{dh}{dy}} \quad (6.10)$$

(b) With a constant value of entrainment throughout the concentration range. It is assumed that the entrainment is 1% of liquid flow rate at the top of the column and increases linearly along the height of the column.

Equation (6.8) becomes (see Appendix I)

$$\frac{N_T}{N_c} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'(t_v - t_c)}{K'(y_c - y)} - \frac{[(y^* - x) - \gamma(H^* - h)]}{(y_c - y)} \cdot \frac{Z}{N_c} \cdot \frac{1}{V} \frac{dV}{dy} \frac{dy}{dz} + \frac{[(y^* - x) - \gamma(H^* - h)]}{(y_c - y)} \cdot \frac{0.01}{N_c}}{1 - \gamma \frac{dh}{dy}} \quad (6.11)$$

(c) By equation (6.9) assuming a 1% constant entrainment along the column,

$$\frac{N_r}{N_c} = \frac{1 + \gamma \frac{U'(t_v - t_c)}{K'(y_e - y)} + \frac{\gamma}{(y_e - y)} \frac{0.01z}{N_c} \frac{dh}{dy} \frac{dy}{dz} + \frac{[(y^* - x) - \gamma(H^* - K)]}{(y_e - y)} \frac{0.01z}{N_c}}{(1.011 - \gamma \frac{dK}{dy}) - \frac{1}{V} [(y^* - x) - \gamma(H^* - K)] \frac{dV}{dy}} \quad (6.12)$$

(d) By taking evaporation and entrainment to vary with concentration in the way determined by experiment:

$$\frac{N_r}{N_c} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'(t_v - t_c)}{K'(y_e - y)} + \frac{[(y^* - x) - \gamma(H^* - K)]}{K'(y_e - y)} \frac{dV}{dy} \frac{dy}{dz} + \frac{[(y^* - x) - \gamma(H^* - K)]}{K'(y_e - y)} \frac{dE}{dy} \frac{dy}{dz}}{1 - \gamma \frac{dK}{dy}} \quad (6.13)$$

This ignores the error which arises in the determination of salt concentration (see section B of this Chapter), and it is taken that the entrainment is the excess of the experimental value of the change in liquid flow rate over that given by theoretical calculation of evaporation by the difference in latent heats of vaporisation. The entrainment calculated in this way varies with concentration (Graph 6.10). Using equation (6.7) and the relationship shown in Graphs 6.10 and 6.14 for values of dV/dy , dE/dy and dy/dz respectively, and determining N_c as before, the values of N_r/N_c at different values of U'/K' are calculated and shown in Graph 6.19.

Graph (6.18) shows that at the same value of U'/K' ($=1000$) there is only a ^{very} slight difference between the curve given by taking into account a constant entrainment and that given by equation (6.6) where entrainment is assumed to be zero. The theoretical curves are not shifted to the right. But Graph 6.19, derived by considering variations in evaporation and entrainment with composition, puts the maximum value of N_T/N_C at a higher composition and gives better agreement between the theoretical and experimental values.

- (ii) The products of thermal distillation are not in equilibrium with the bulk phase from which they have been generated.

That the evaporated liquid and condensed vapour from the bulk phases are the equilibrium values of the compositions of the bulk phases is not likely. These values will determine $\gamma \left(\frac{y^* - x^*}{H^* - h^*} \right)$ which has a large effect on the value of N_T/N_C . We introduce a Murphree efficiency :

$$\text{Efficiency} = \frac{y^* - y}{y_e - y} \quad \text{and} \quad \text{Efficiency} = \frac{x - x^*}{x - x_e}$$

N_T/N_C has been calculated for two cases with an assumption that vapour and liquid Murphree efficiencies are the same.

(a) Without entrainment.

N_T/N_C is calculated from equation (6.6) for $U'/K' = 1000$ for three values of the Murphree efficiency, 75, 50 and 25%, and is plotted on Graph 6.20. These curves show that the Murphree efficiency has not altered the shape of the curve or the m.f. where maximum value of N_T/N_C occurs, but has simply decreased the corresponding values of N_T/N_C ; for instance, the maximum value of N_T/N_C decreases by about 50% with a 25% Murphree efficiency.

(b) With entrainment which varies with composition.

Using equation (6.7), the curve obtained at $U'/K' = 1000$ for a 75% Murphree efficiency (Graph 6.21) has the same shape as the curve for a 100% Murphree efficiency.

(iii) No contact distillation.

There remains the problem that experimental values of N_T/N_C tend towards zero at the pure components. The theoretical values of N_T/N_C at pure components, remain at unity, though the variation of entrainment with concentration does shift the position of maximum towards a higher value of liquid m.f. The only theoretical way to make the performance zero at pure components is to assume contact distillation takes very little part in the process,

and hence that the transfer of mass is by thermal distillation only. With this assumption, the following equation (see Appendix I) is obtained :

$$N_T = \frac{z}{L} \cdot \frac{\gamma U' \frac{t_v - t_L}{y_e - y} + \frac{[(y^* - x) - \gamma(H^* - h)]}{y_e - y} \frac{dL}{dz}}{1 - \gamma \frac{dh}{dy}} \quad (6.14)$$

Two series of calculations have been made for several values of U' over the whole concentration range :

- (a) Assuming that the products of thermal distillation are in equilibrium with the bulk phase from which they have been generated;
- (b) that the Murphree efficiency is 75%.

The calculated results are shown in Graph 6.22. In these cases, the number of transfer units is zero at pure l.v.c. and the maximum occurs at a high value of y , but the graph still does not coincide with the experimental curve.

Even including the effect of an entrainment which varies with composition and excluding contact distillation it is still not possible to predict precisely the experimental results obtained during the course of the work. There may be several reasons for this: for instance, there may be a variation in wetting along the column or in

GRAPH 6.6

$\frac{d \ln V}{dy}$ - COMPOSITION RELATIONSHIPS

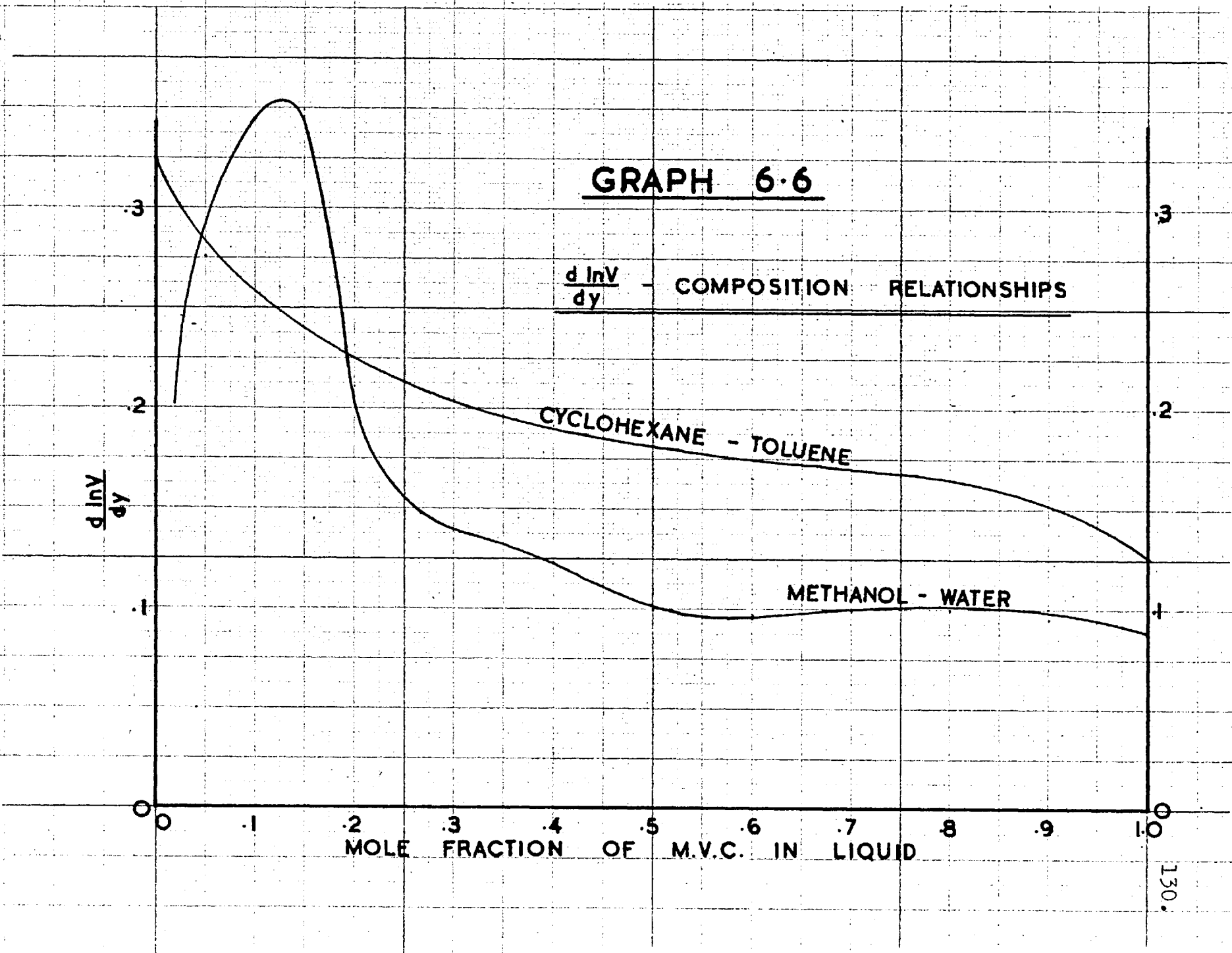
$\frac{d \ln V}{dy}$

CYCLOHEXANE - TOLUENE

METHANOL - WATER

MOLE FRACTION OF M.V.C. IN LIQUID

130.



GRAPH 6.7

$\frac{\lambda}{T_b}$ - COMPOSITION RELATIONSHIPS

cal / gm. mol. / °K

$\frac{\lambda}{T_b}$

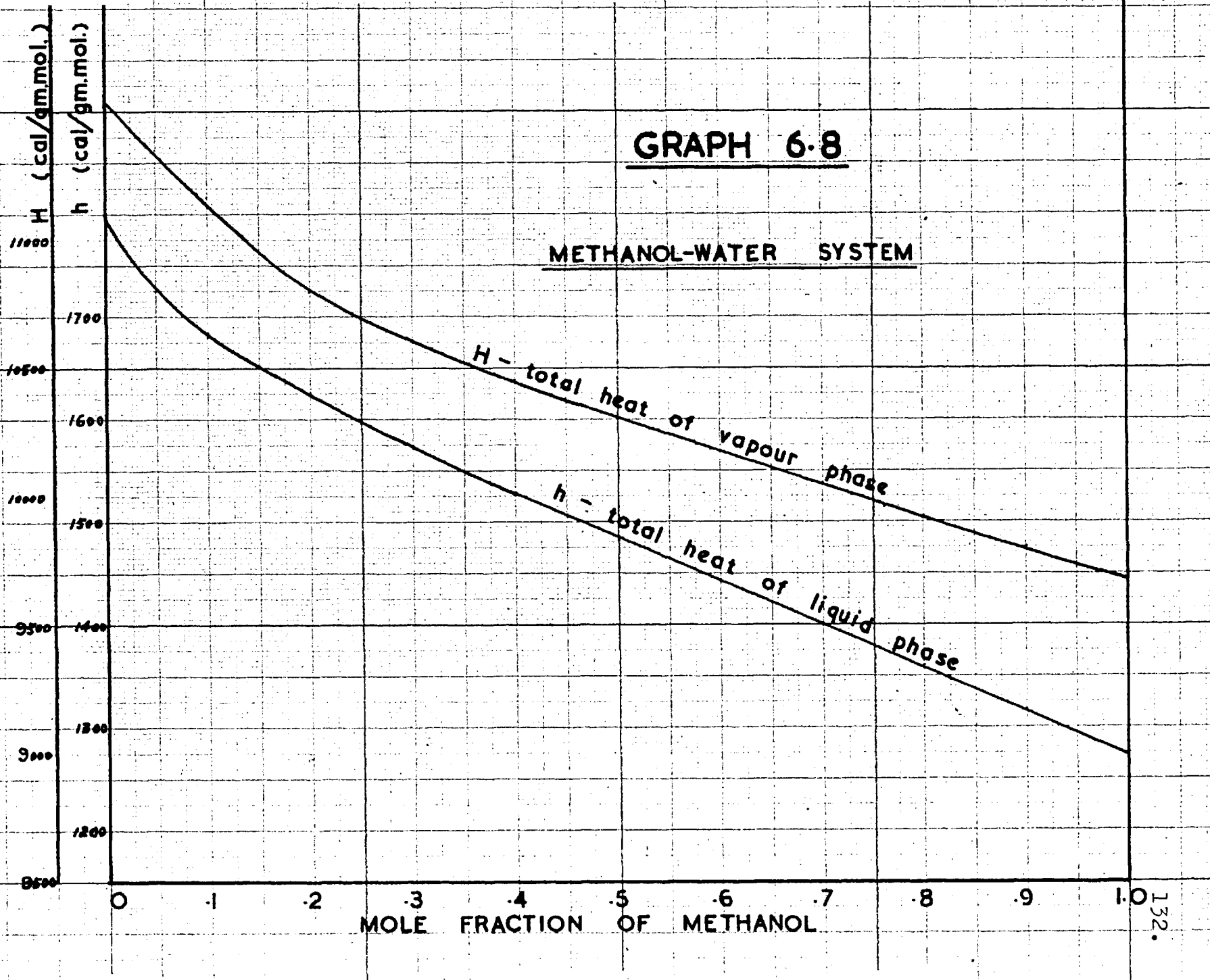
METHANOL - WATER

CYCLOHEXANE - TOLUENE

MOLE FRACTION OF M.V.C. IN LIQUID

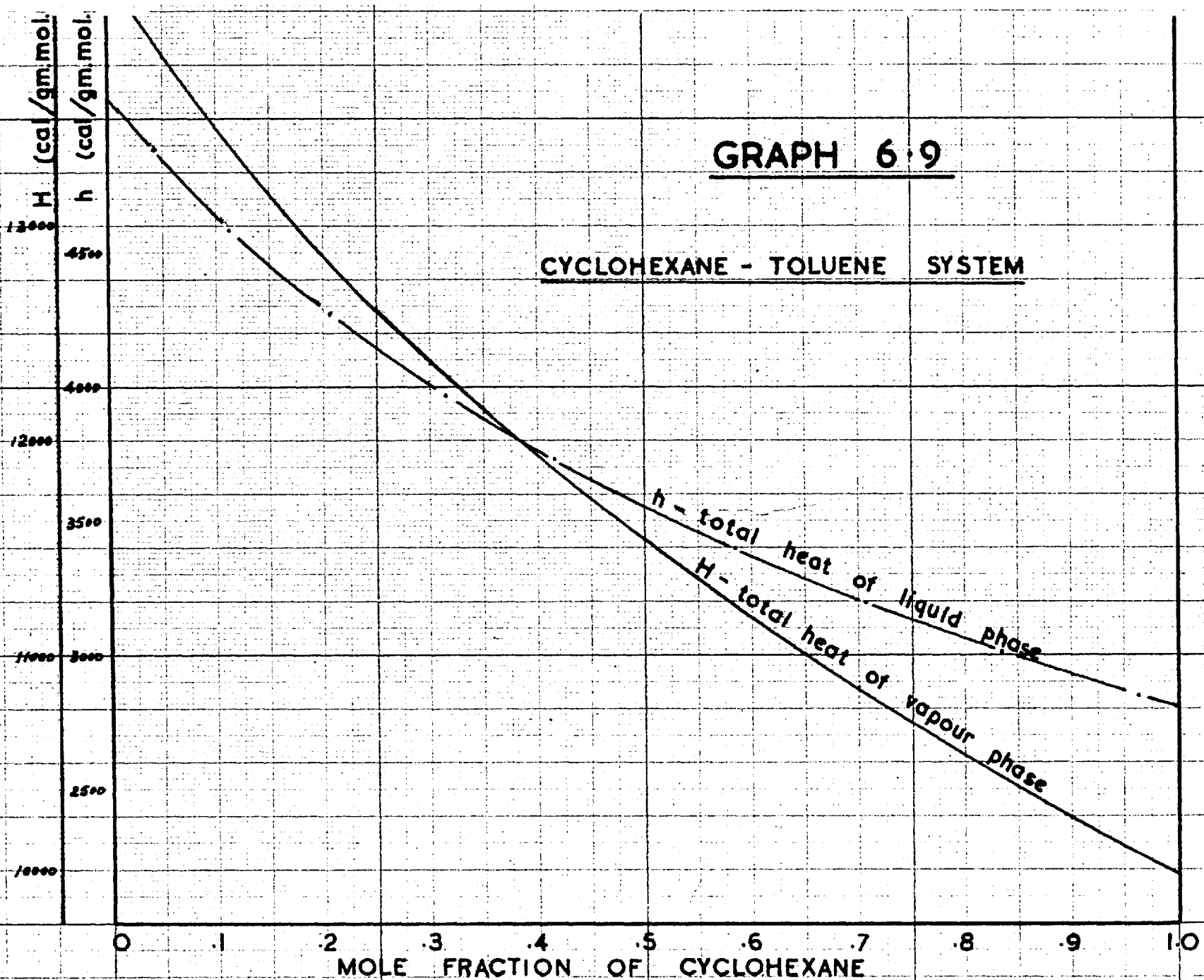
GRAPH 6.8

METHANOL-WATER SYSTEM



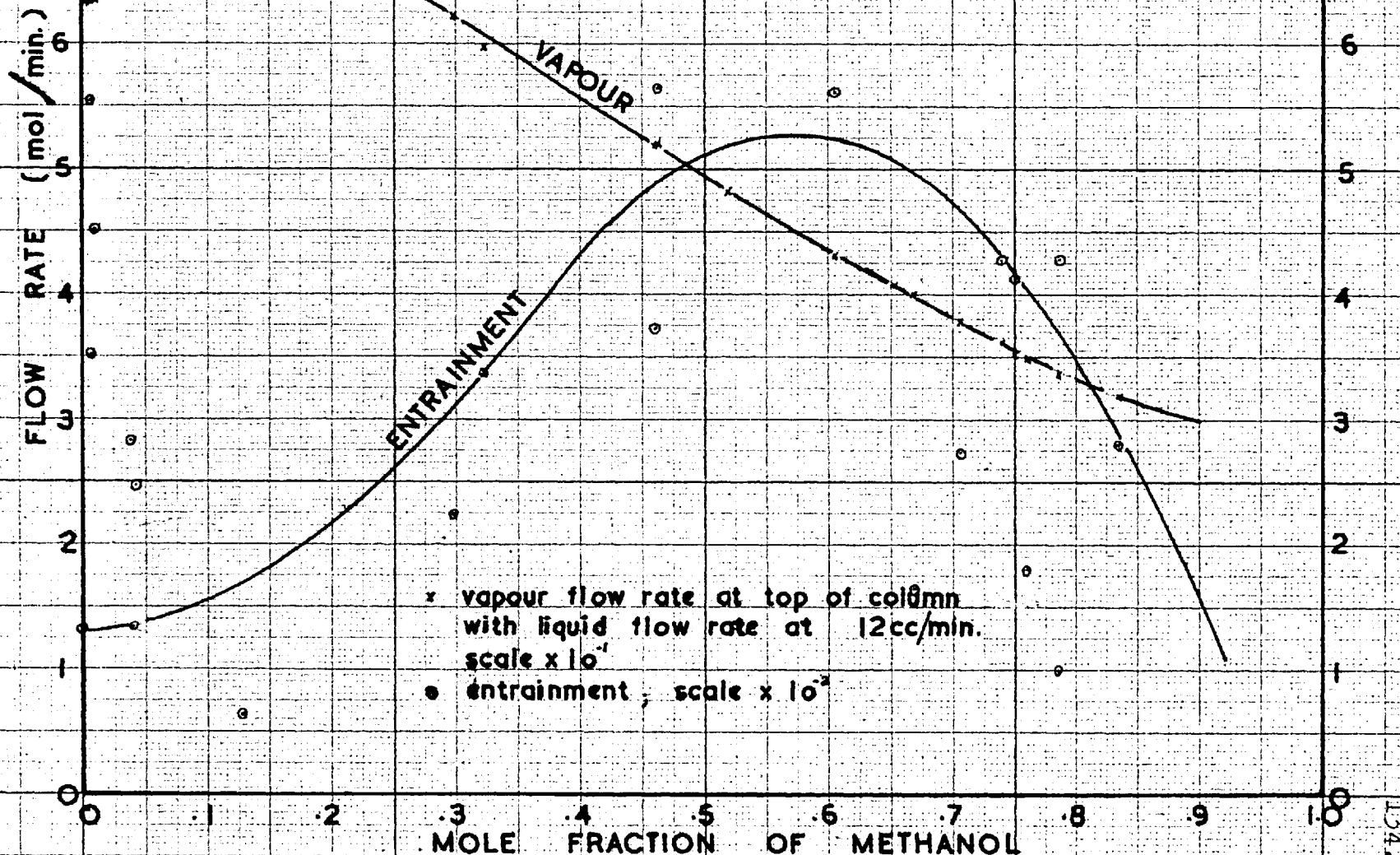
GRAPH 6.9

CYCLOHEXANE - TOLUENE SYSTEM



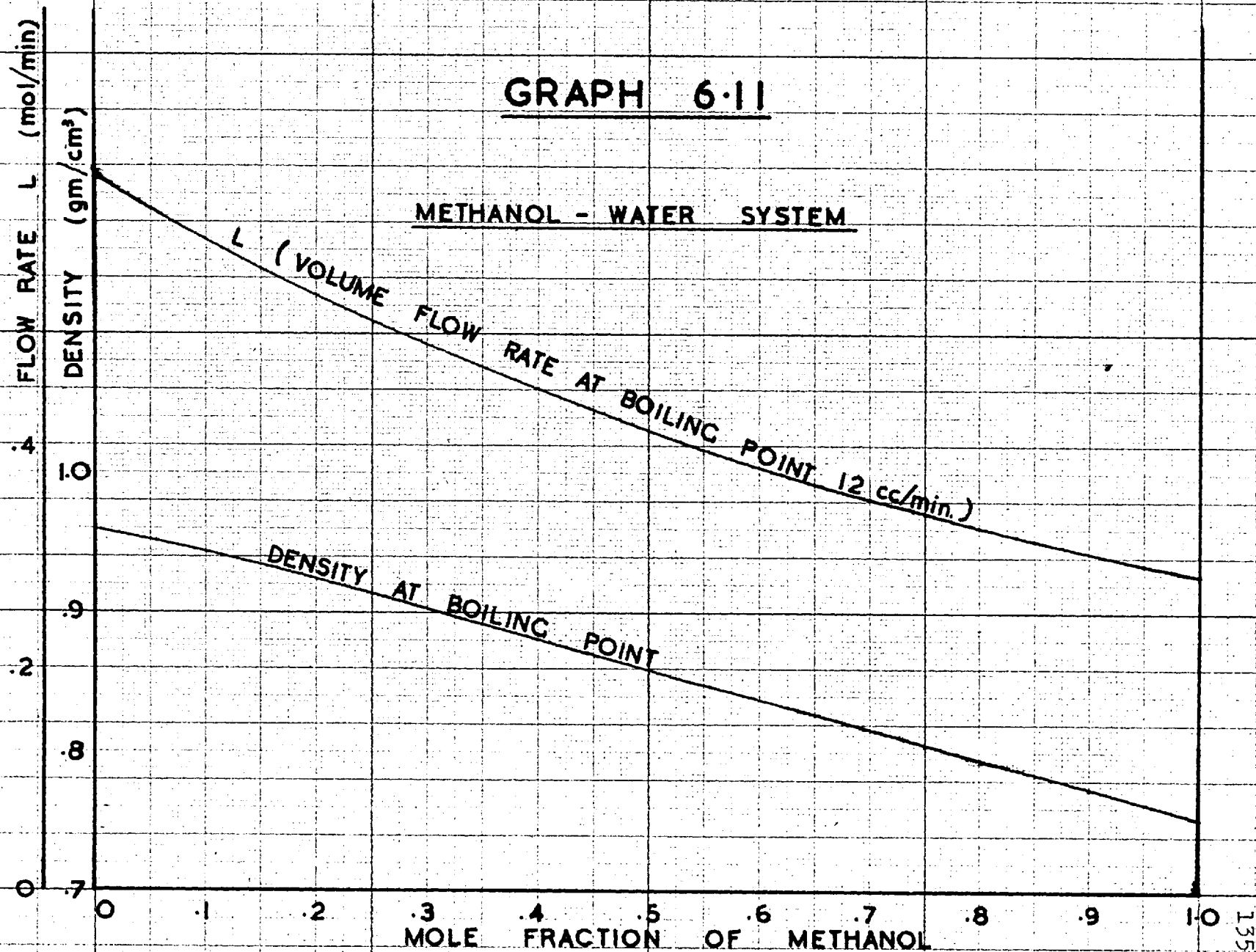
GRAPH 6.10

METHANOL - WATER SYSTEM



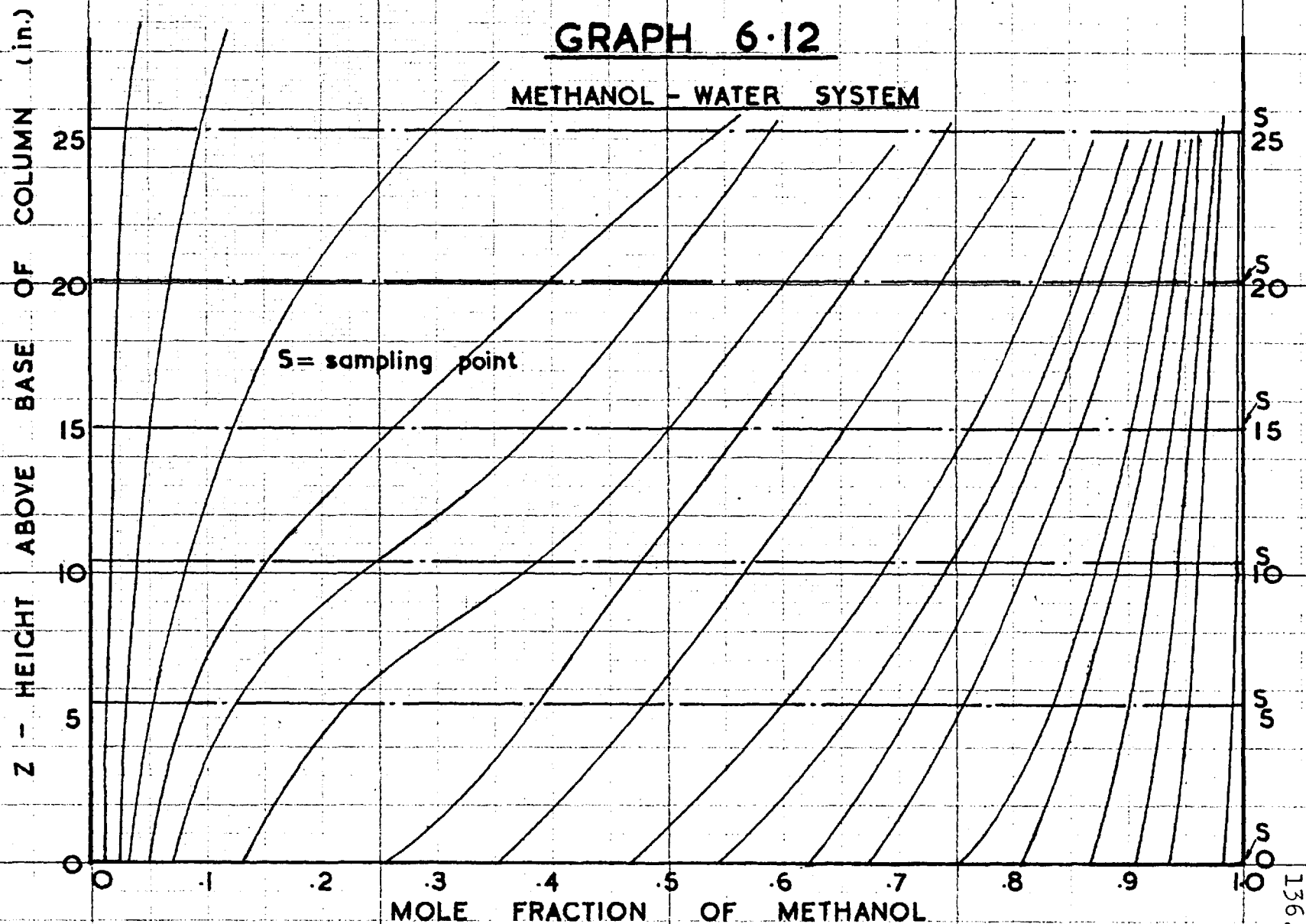
GRAPH 6.11

METHANOL - WATER SYSTEM



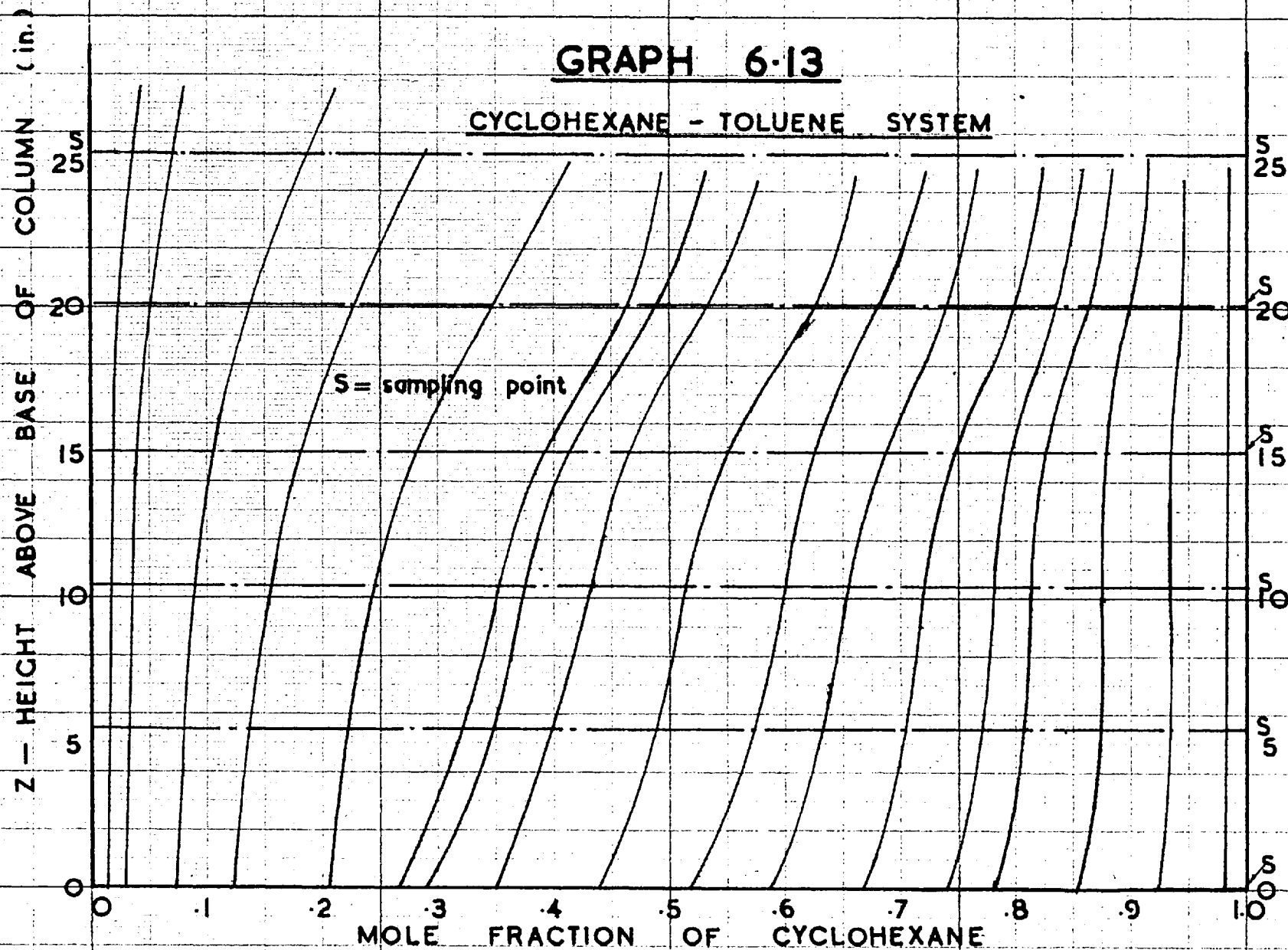
GRAPH 6.12

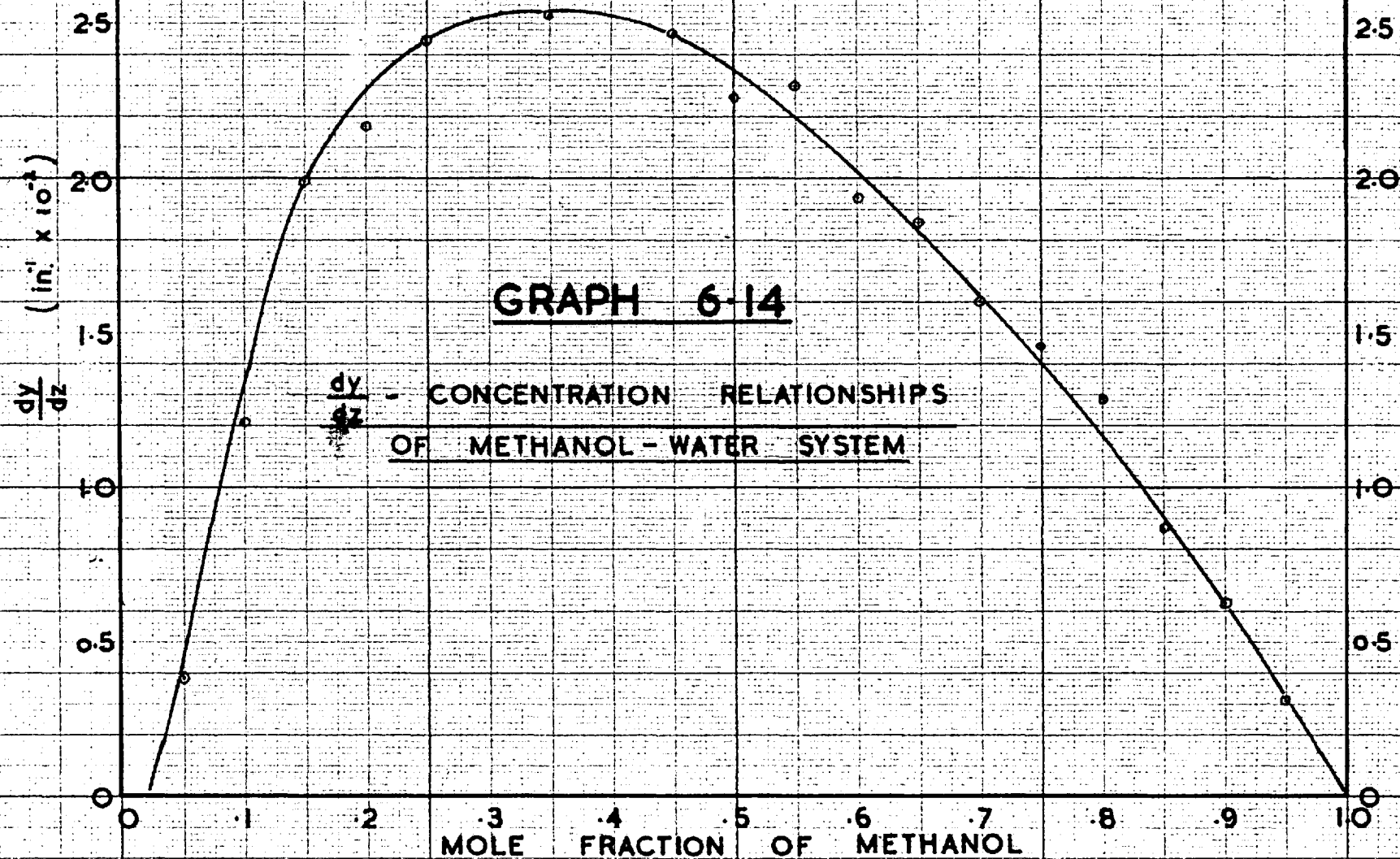
METHANOL - WATER SYSTEM



GRAPH 6-13

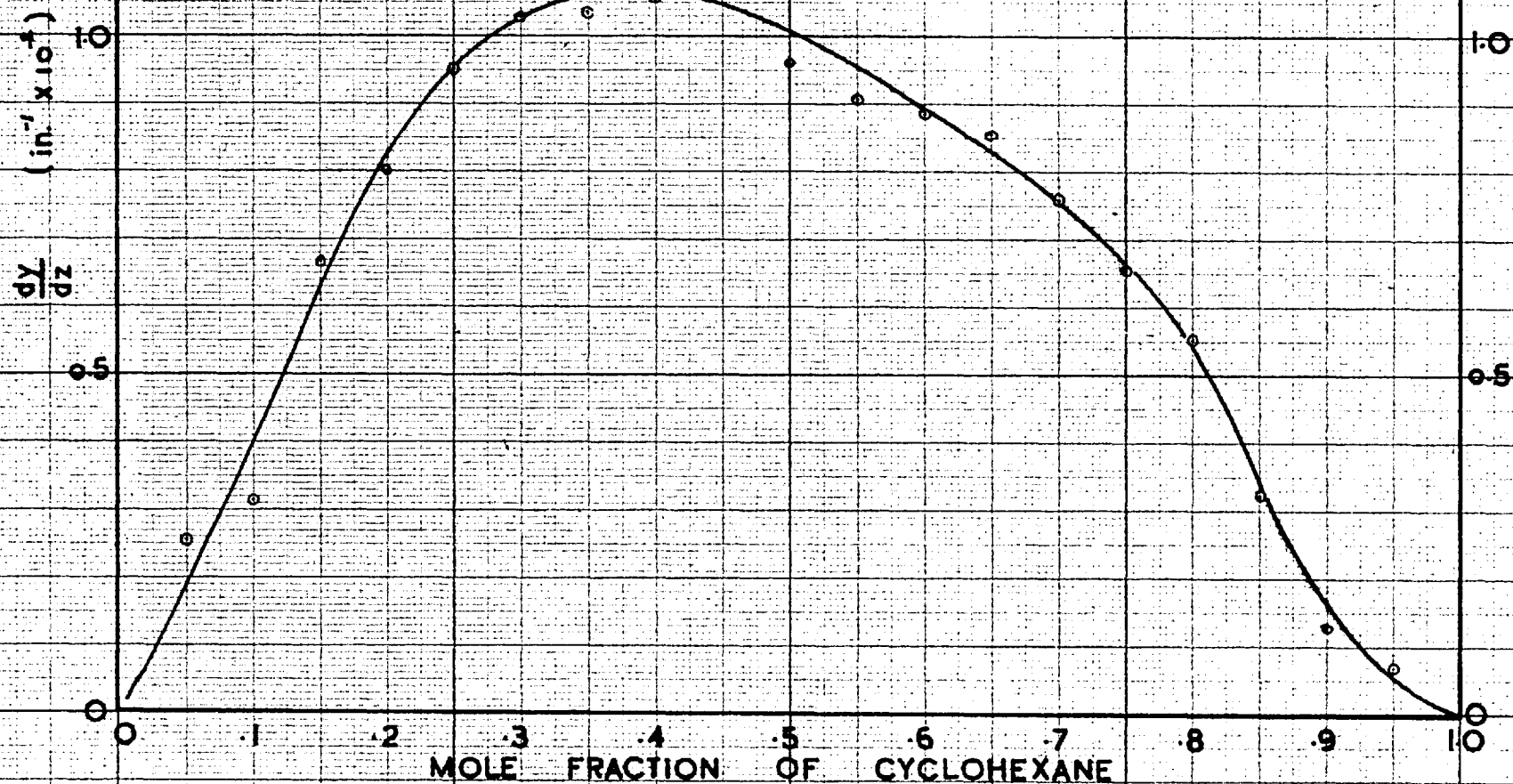
CYCLOHEXANE - TOLUENE SYSTEM

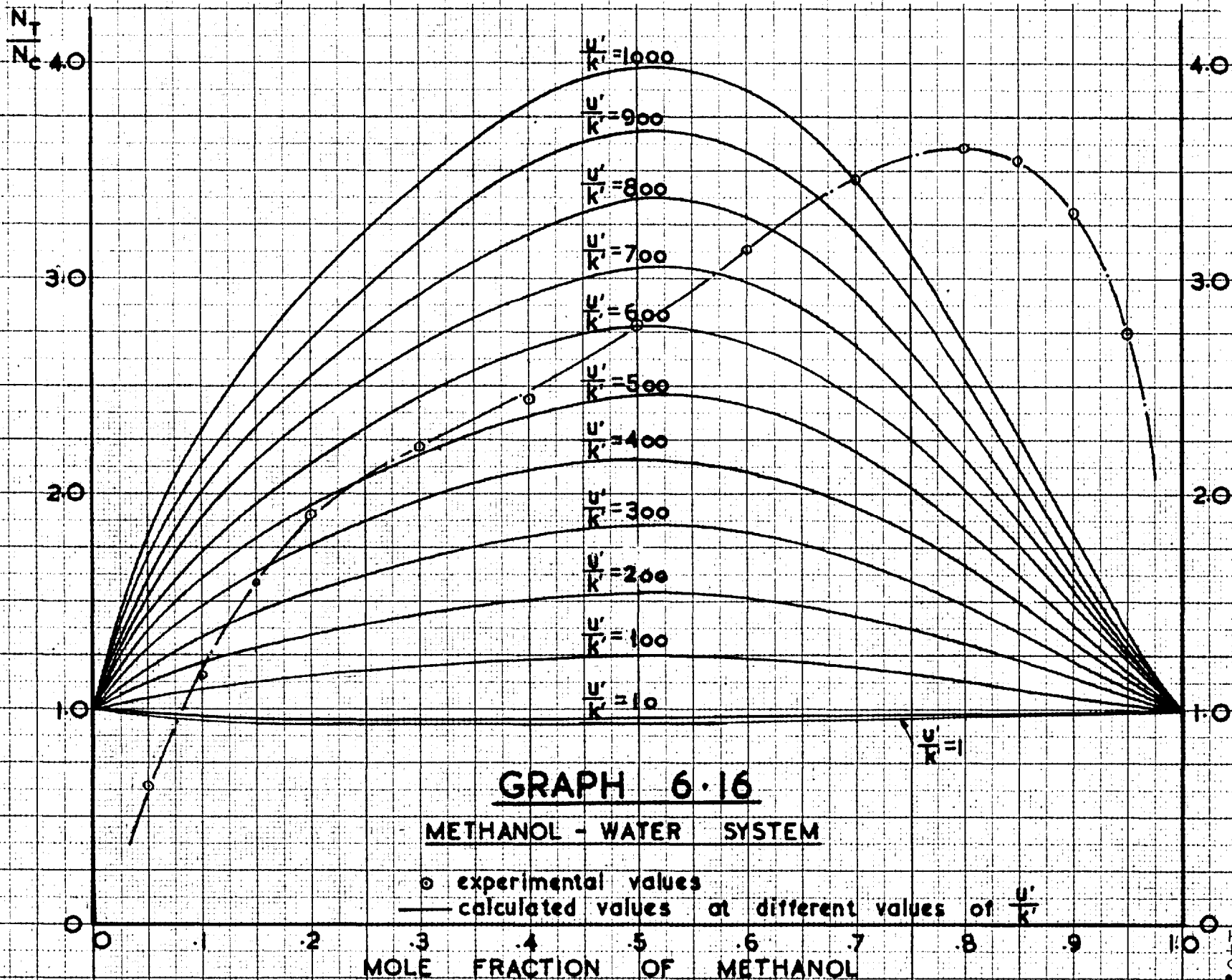




GRAPH 6.15

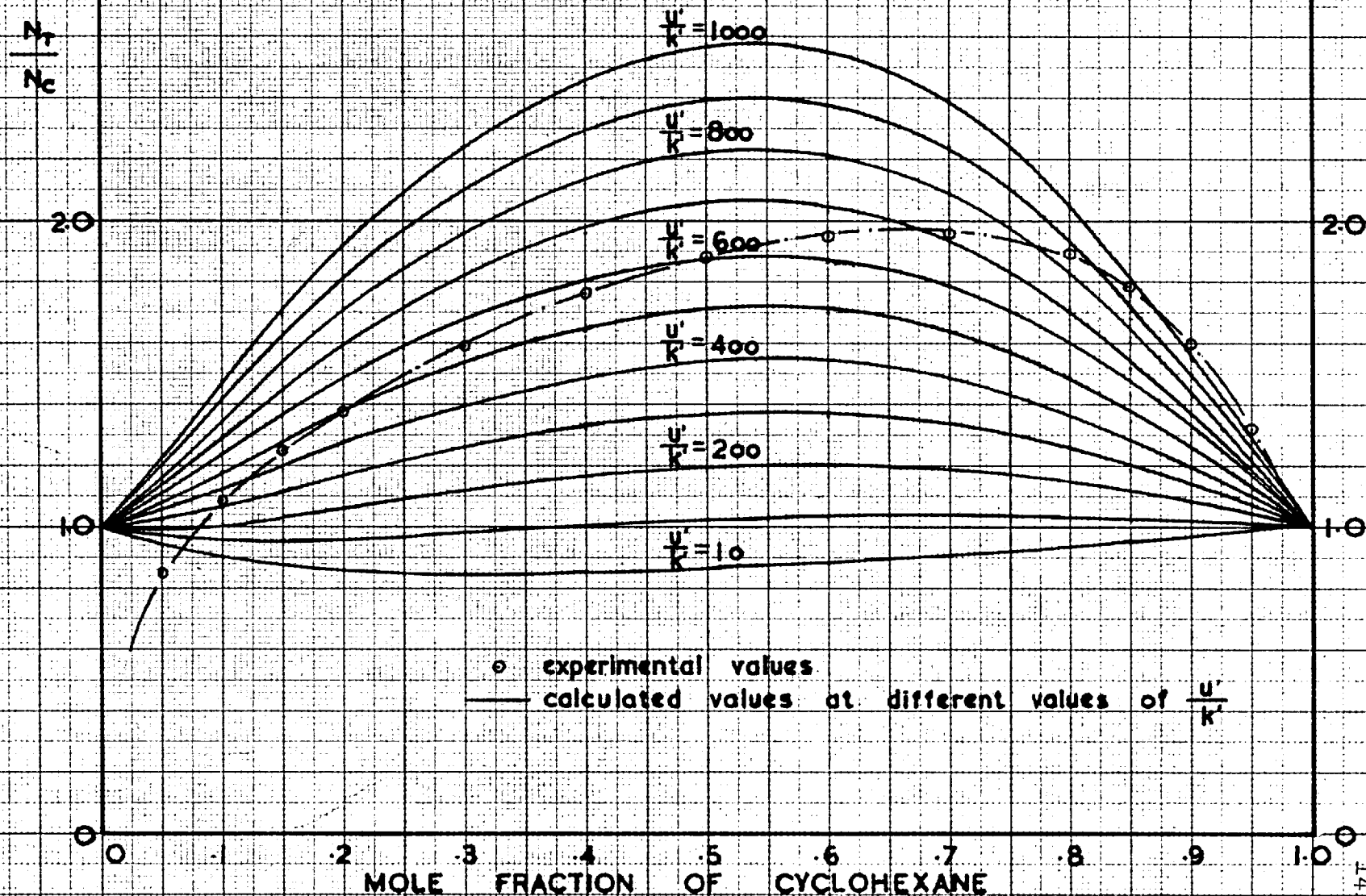
$\frac{dy}{dz}$ - CONCENTRATION RELATIONSHIPS
OF CYCLOHEXANE - TOLUENE SYSTEM

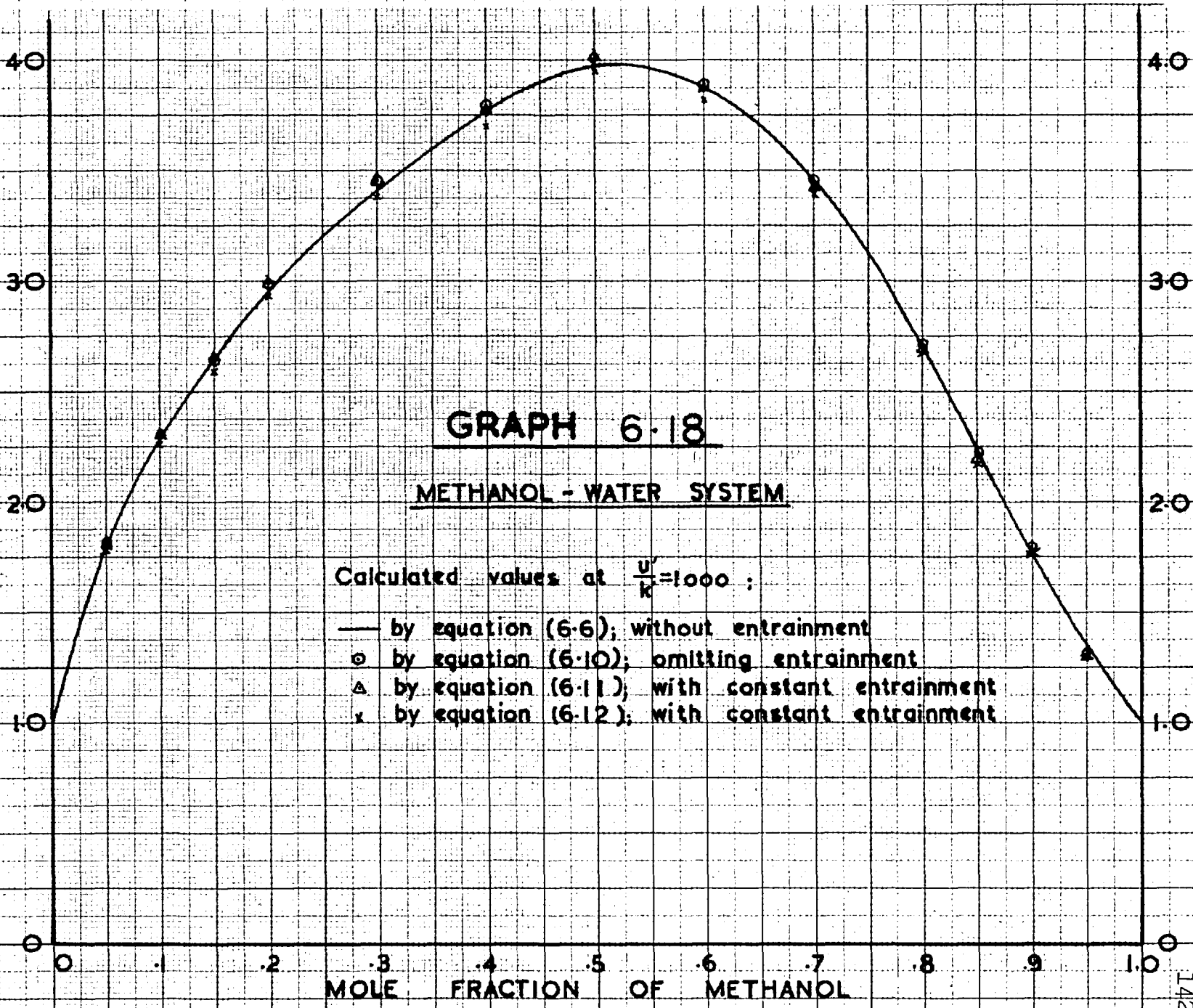


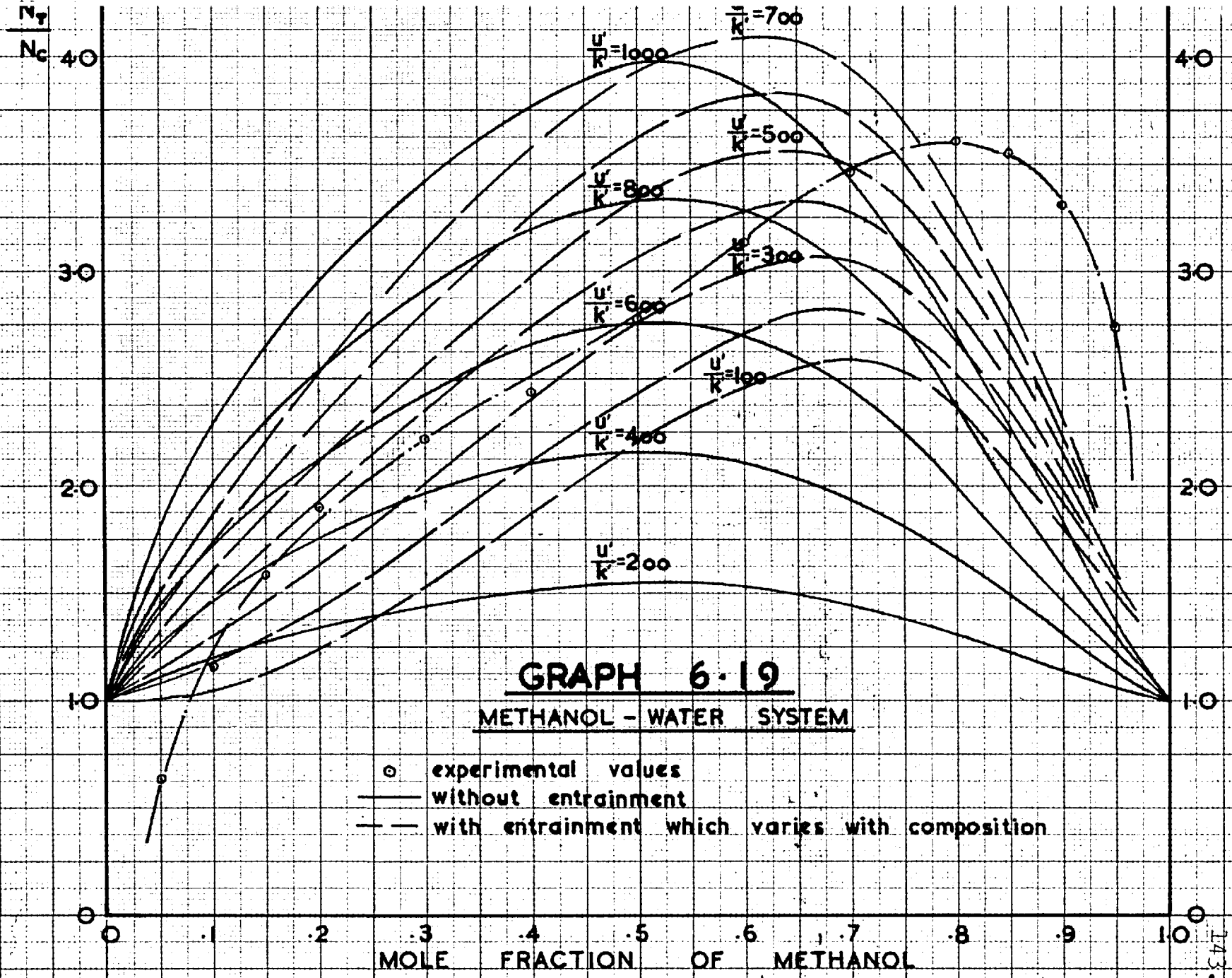


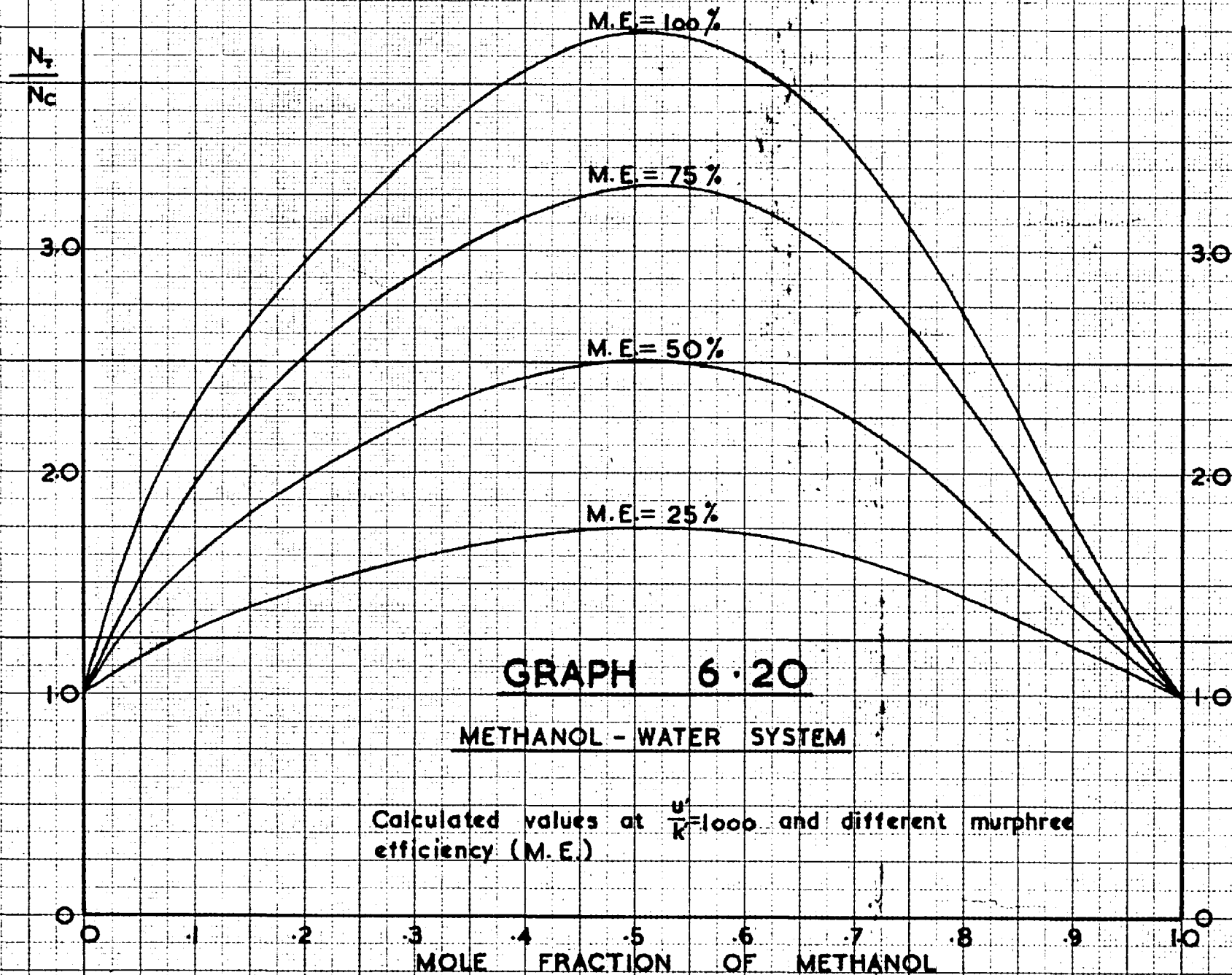
GRAPH 6.17

CYCLOHEXANE - TOLUENE SYSTEM









GRAPH 6-21

METHANOL - WATER SYSTEM

Calculated values, including entrainment, at
different Murphree efficiency (M.E.) and $\frac{V}{K}=1000$

$\frac{N_T}{N_C}$

5.0

4.0

3.0

2.0

1.0

5.0

4.0

3.0

2.0

1.0

M.E.=100% →

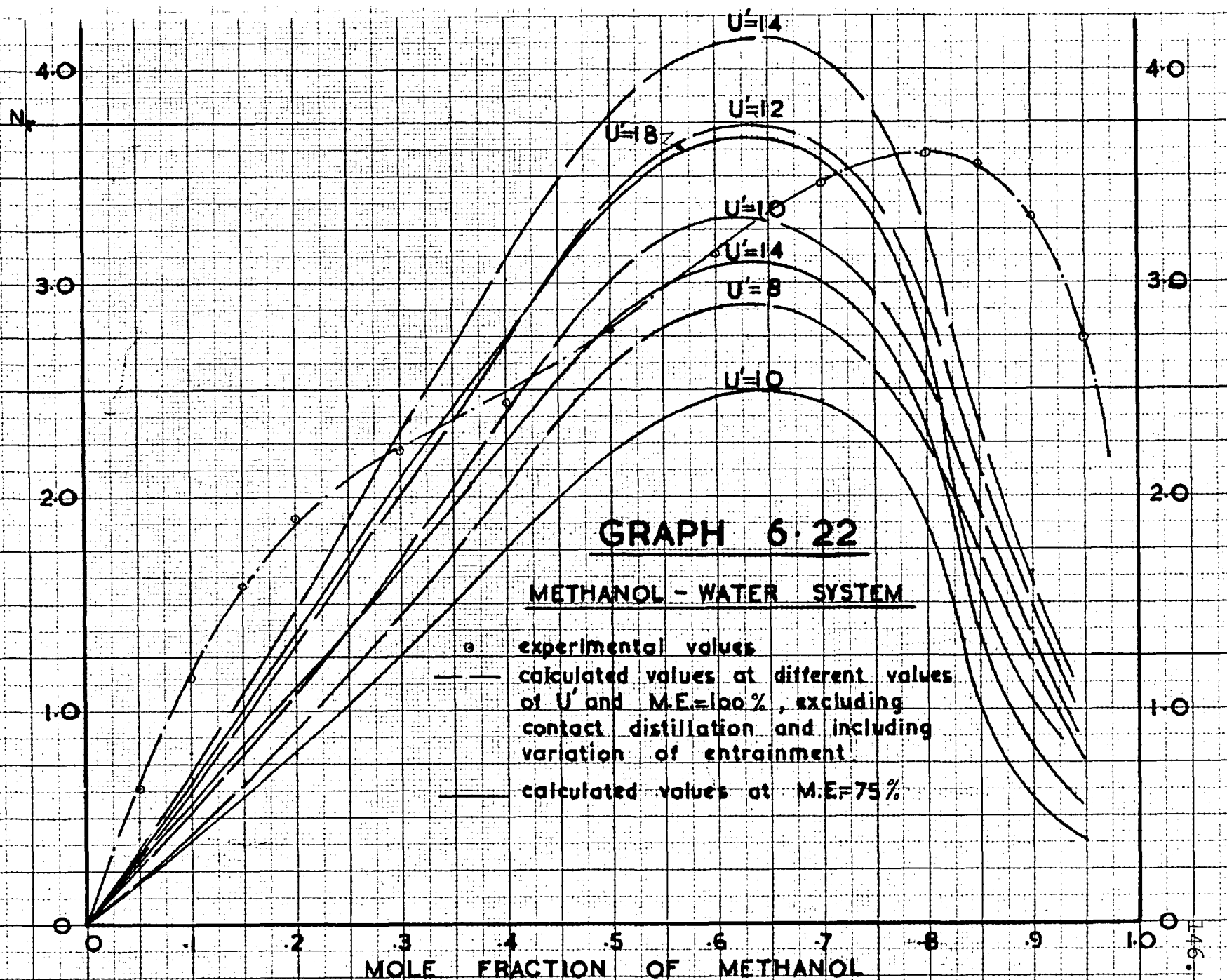
M.E.=75%

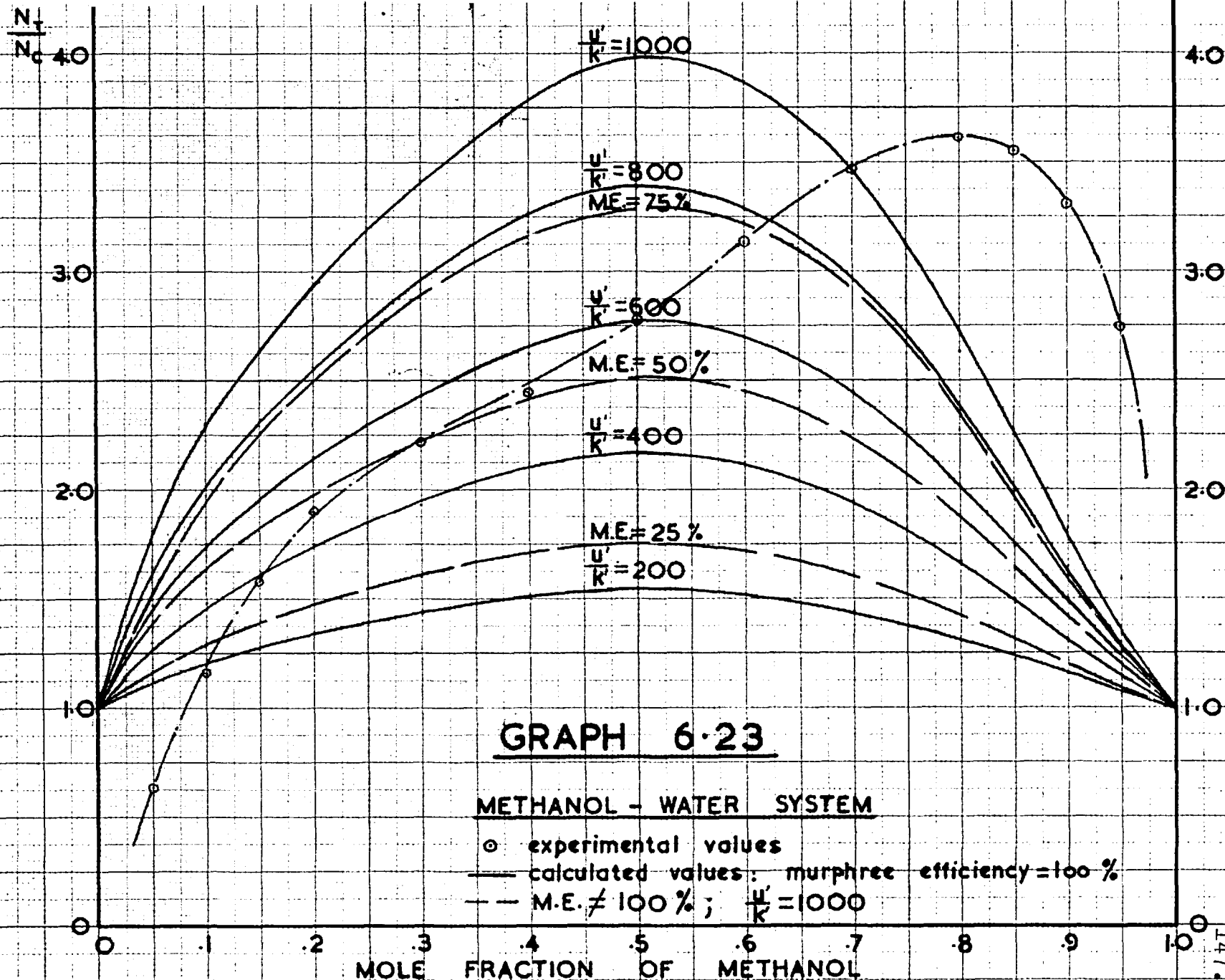
M.E.=50%

M.E.=25%

MOLE FRACTION OF METHANOL

145.





surface renewal, as has been discussed earlier. In addition the theory of thermal distillation may be further modified, as follows :

(1) U'/K' and the Murphree efficiency of thermal condensation and evaporation may not be constant, but may vary with composition. At low compositions, U'/K' or the Murphree efficiency may be low, but may increase as the liquid mole fraction increases (Graph 6.23). U'/K' may increase continuously even after passing the maximum value of N_T/N_C , but it is likely that the Murphree efficiency will increase only up to a medium liquid concentration after which it decreases. If such variations did occur the actual maximum of N_T/N_C would shift and occur at a higher value of the liquid composition than is found for a curve of constant U'/K' or Murphree efficiency, even without considering the variation of entrainment with composition.

(2) An asymmetry of the curve of N_T/N_C versus concentration over the whole range of concentration has been achieved (Graph 6.19) by assuming that the amount of entrainment varies with concentration. The value dE/dy , used in equation (6.13), is the slope of the curve of entrainment versus concentration (Graph 6.10) and it varies from positive to negative depending on the liquid composition,

that is, whether it is on the left or on the right of the maximum of the entrainment versus concentration curve. As has been pointed out, these experiments are not accurate, and we place no reliance on this explanation. If E did vary with composition, the performance of a distillation column could only be predicted from equation (6.13).

Entrainment is more likely in plate columns when the equivalent of equation (6.13) should be used. Entrainment would certainly have an unfavourable effect on the efficiency of separation, since effectively back mixing occurs along the height of the column. In a good design of a packed column entrainment should be small.

The amount of entrainment and the change in flow rate in this experiment has been obtained by introducing a small volume of concentrated solution of sodium chloride. This affects the actual composition of liquid and its heat content, and the above calculations are only approximate. In the absence of salt, the liquid composition will be higher, the curves of E vs. y and V vs. y will be shifted to the right, and it is possible that the maximum value of N_T/N_C occurs at a still higher liquid mole fraction than is shown.

(3) That the value of N_T/N_c for pure components is not unity but zero has been explained by assuming that the contact transfer of mass by ordinary diffusional mass transfer theory, with concentration difference as the driving force, is small in amount. This does not accord with measurements made in other mass transfer processes, for instance, absorption.

(We note that, since the number of transfer units

$$N = \int \frac{dy}{y_e - y} \quad , \quad (6.15)$$

both the numerator and denominator approach zero as y approaches 0 or 1, and are equal to zero for pure components. The number of transfer units is then indeterminate). When there is no contact distillation, the same arguments used previously to explain the differences between the theoretical and experimental results can be applied again; that is, U' and the Murphree efficiency vary with concentration.

CONCLUSION.

It has been shown that even elementary consideration of thermal distillation can account for the variation of performance with composition measured with continuous countercurrent distillation equipment. It is likely that the application of similar principles would account for the same behaviour noted with plate columns.

There is no doubt that thermal distillation in one form or another can take place, and we have shown quantitatively that the results conform in essential features with experiment. But we do not claim that thermal distillation provides the entire reason for discrepancies from normal mass transfer theory. The explanations based upon the change of surface tension with composition for instance can be superposed upon the mechanism suggested here and together the two effects may form a complete explanation of the behaviour of equipment. The difference between the observed and predicted maximum values of the number of transfer units or the tendency for the performance with dilute mixtures to be small may be better explained by alternative theories, rather than by the secondary and unconfirmed explanations given here: namely that the ratio U'/K' or entrainment defers upon composition.

No attempt has been made to study in detail the way in which thermal transfer is carried out. Further work on this aspect of the process is envisaged. The problems to be studied are the mechanism and efficiency of condensation or evaporation in the bulk phases or at an interface, of entrainment whether mechanical or as a result of thermal distillation, and the application of these processes to the operation of a plate column.

APPENDIX I.THEORY OF THERMAL DISTILLATION.

A. HEAT AND MATERIAL BALANCES.

We make the following assumptions, some of which are justified by the experimental results:

- (1) the vapour is at its dew point and the liquid is at its boiling point;
- (2) the rate of contact mass transfer is determined by a simple m.f. overall composition difference between the phases;
- (3) the rate of heat transfer is determined by the overall temperature difference between the phases;
- (4) account is taken of the fact that the latent and total heat of the phases may vary with composition and lead to non-constant molal overflow;
- (5) it is always assumed that the products of thermal distillation generated in either phase immediately join the opposite stream;
- (6) in the simplest treatment given first it is considered that there is no entrainment of the liquid and that the products of thermal distillation are in equilibrium with the bulk phase from which they have been generated.

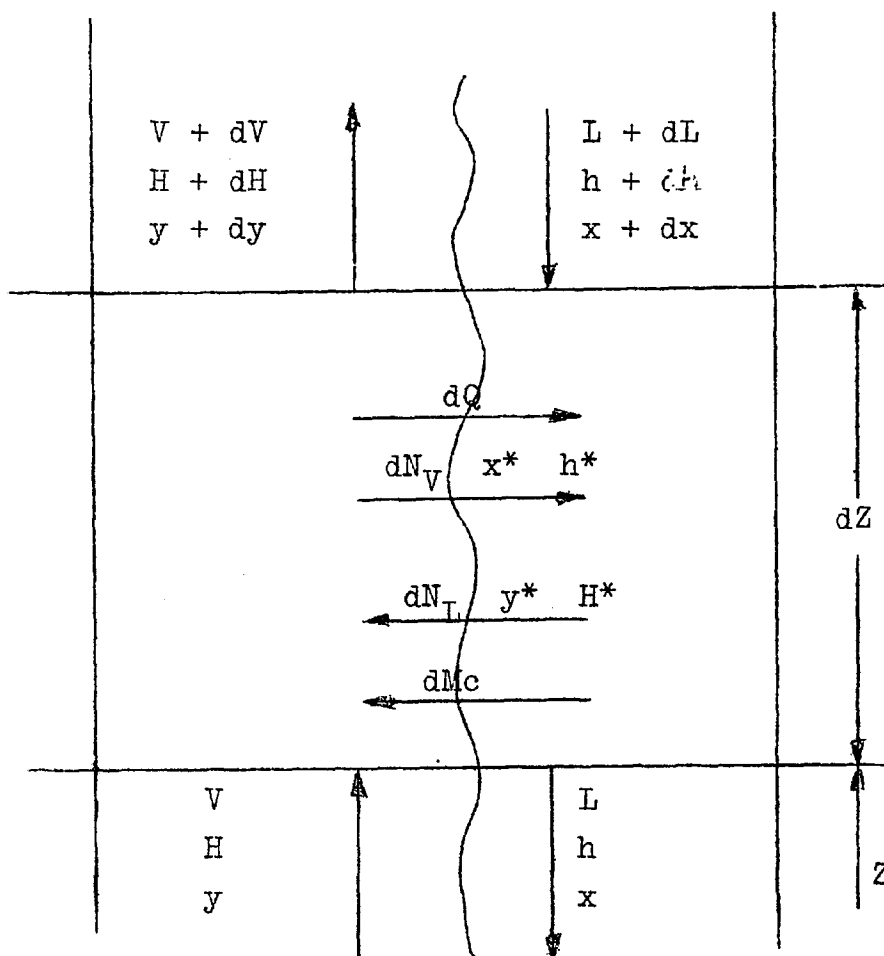


Figure I.1. MATERIAL AND HEAT TRANSFER
DIAGRAM WITHOUT ENTRAINMENT.

(Notation is given in Appendix III. Refer to Fig. I.1).

The rate of contact mass transfer is

$$dMc = K_{OG} aS (y_e - y) dZ \quad (I.1)$$

$$= K' (y_e - y) dZ \quad (I.2)$$

The rate of heat transfer between the phases is

$$dQ = U aS (t_v - t_L) dZ \quad (I.3)$$

$$= U' (t_v - t_L) dZ \quad (I.4)$$

Material balances.

The independent balances chosen are those on the whole element (liquid and vapour) and on the vapour element for total moles of both components and for moles of m.v.c.

Total moles

$$dV = dL \quad (I.5)$$

$$dV = dN_L - dN_V \quad (I.6)$$

Moles of m.v.c.

$$d(Vy) = d(Lx) \quad (I.7)$$

$$d(Vy) = y^* dN_L - x^* dN_V + dMc \quad (I.8)$$

Heat balances.

Two independent balances are obtained on the whole element and on the vapour element.

$$d(VH) = d(Lh) \quad (I.9)$$

$$d(VH) = H^* dN_L - h^* dN_V - dQ \quad (I.10)$$

B. SOLUTION - NO ENTRAINMENT.

By equation (I-5) and (I-8) dN_L is eliminated.

$$\begin{aligned} d(Vy) &= y^*(dV + dN_V) - x^*dN_V + dMc \\ &= y^*dV + (y^*-x^*) dN_V + dMc \end{aligned} \quad (I-11)$$

By equations (I-6) and (I-10) dN_L is again eliminated

$$d(VH) = H^* dV + (H^*-h^*) dN_V - dQ \quad (I-12)$$

$$\text{i.e.} \quad \frac{dN_V}{dN_V} = \frac{d(VH) - H^*dV + dQ}{H^* - h^*} \quad (I-13)$$

Substituting equation (I-13) into (I-11), then

$$d(Vy) = y^*dV + dMc + \gamma [d(VH) - H^*dV + dQ] \quad (I-14)$$

$$\text{where } \gamma = \frac{y^* - x^*}{H^* - h^*}$$

Expanding equation (I-14) and dividing by dZ ,

$$V \frac{dy}{dZ} + y \frac{dV}{dZ} = y^* \frac{dV}{dZ} + \frac{dMc}{dZ} + \gamma \left[V \frac{dH}{dZ} + H \frac{dV}{dZ} - H^* \frac{dV}{dZ} + \frac{dQ}{dZ} \right]$$

$$\text{or } V \frac{dy}{dZ} - [(y^*-y) + \gamma(H-H^*)] \frac{dV}{dZ} - \gamma V \frac{dH}{dZ} = \frac{dMc}{dZ} + \gamma \frac{dQ}{dZ} \quad (I-15)$$

Substituting equations (I-2) and (I-4) into (I-15), and expressing all differential coefficients in terms of $\frac{dy}{dZ}$,

$$\begin{aligned} V \frac{dy}{dZ} - [(y^*-y) + \gamma(H-H^*)] \frac{dV}{dy} \cdot \frac{dy}{dZ} - \gamma V \frac{dH}{dy} \cdot \frac{dy}{dZ} &= K'(y_e - y) \\ &+ \gamma U'(t_V - t_L), \end{aligned}$$

$$V \frac{dy}{dz} \left\{ 1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) + \gamma(H - H^*)] \frac{dV}{dy} \right\} = K'(y_e - y) + \gamma U'(t_v - t_c),$$

$$\text{i.e. } \frac{\frac{dy}{dz}}{y_e - y} = \frac{K' + \gamma U' \frac{t_v - t_c}{y_e - y}}{V \left\{ 1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) + \gamma(H - H^*)] \frac{dV}{dy} \right\}} \quad (\text{I.16})$$

The quantity $dy/dz/y_e - y$ is the point value of a height of a transfer unit based on overall vapour composition difference. In the absence of thermal distillation, by (I.2) and (I.8), we have

$$dMc = K' (y_e - y) dz = V dy \quad (\text{I.17})$$

$$\text{or } \frac{dy/dz}{y_e - y} = \frac{K'}{V} = \frac{K_{OG} a S}{V} \quad (\text{I.18})$$

Thus for contact distillation the point value of the height of a transfer unit is

$$\frac{K'}{V} = \frac{1}{H_c}, \quad \text{say} \quad (\text{I.19})$$

However the measured and calculated value of the height of a transfer unit, as determined when there is both thermal and contact distillation, is

$$\frac{dy/dz}{y_e - y} = \frac{1}{H_T}, \quad \text{say.}$$

Thus, (I.16) becomes

$$\frac{1}{H_T} = \frac{K' + \gamma U' \frac{t_v - t_c}{y_e - y}}{V \left\{ 1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) + \gamma(H - H^*)] \frac{dV}{dy} \right\}} \quad (\text{I.20})$$

On a rearrangement, using equation (I.19),

$$\therefore \frac{H_c}{H_T} = \frac{N_T}{N_c} = \frac{1 + \gamma \frac{U'}{K'} \frac{(t_v - t_L)}{(y_e - y)}}{1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) + \gamma(H - H^*)] \frac{dV}{dy}} \quad (\text{I.21})$$

where N_T and N_c are the number of transfer units with and without thermal distillation respectively.

In the first case we have taken y^* and x^* to be the equilibrium values of x and y respectively; that is, $y^* = y_e$ and $x^* = x_e$.

There will be reason to consider conditions where the extent of contact distillation is small or $dMc = 0$. From equation (I.15)

$$V \frac{dy}{dz} \left\{ 1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) + \gamma(H - H^*)] \frac{dV}{dy} \right\} = \gamma U' (t_v - t_L)$$

$$\frac{\frac{dy}{dz}}{y_e - y} = \frac{1}{H_T} = \frac{\gamma \frac{U'}{V} \frac{t_v - t_L}{y_e - y}}{1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) + \gamma(H - H^*)] \frac{dV}{dy}}$$

Since $N_T = \frac{Z}{H_T}$,

$$N_T = \frac{\gamma \cdot Z \cdot \frac{U'}{V} \frac{(t_v - t_L)}{y_e - y}}{1 - \gamma \frac{dH}{dy} - \frac{1}{V} [(y^* - y) + \gamma(H - H^*)] \frac{dV}{dy}} \quad (\text{I.22})$$

Value of dV/dy .

From equation (I.9) we have :

$$H \frac{dV}{dy} + V \frac{dH}{dy} = L \frac{dh}{dy} + h \frac{dL}{dy}$$

Substituting equation (I.5), then

$$H \frac{dV}{dy} + V \frac{dH}{dy} = V \frac{dh}{dy} + h \frac{dV}{dy}$$

$$\text{i.e. } (H-h) \frac{dV}{dy} = -V \frac{d(H-h)}{dy}$$

$$\therefore \frac{1}{V} \cdot \frac{dV}{dy} = - \frac{1}{(H-h)} \cdot \frac{d(H-h)}{dy} \quad (\text{I.23})$$

Equation (I.23) can be substituted for part of the last term of the denominator of equations (I.21) and (I.22).

C. SOLUTION — WITH ENTRAINMENT.

The experiments show that there is a small amount of entrainment of liquid by vapour.

Let h_E be the total heat of the entrained liquid, flow rate E , having a composition x_E (Fig.I.2).

(Notation is given in Appendix III).

Since it is liquid compositions which are measured and it ^{is} the change in liquid flow rate that is calculated, we consider the liquid element. The material and heat balances become (cf. equations (I.5) to (I.10)).

$$dL = dN_L - dN_V \quad (\text{I.24})$$

$$\text{since } dV + dE = dL \quad (\text{I.25})$$

$$d(Lx) = y^* dN_L - x^* dN_V + dMc \quad (\text{I.26})$$

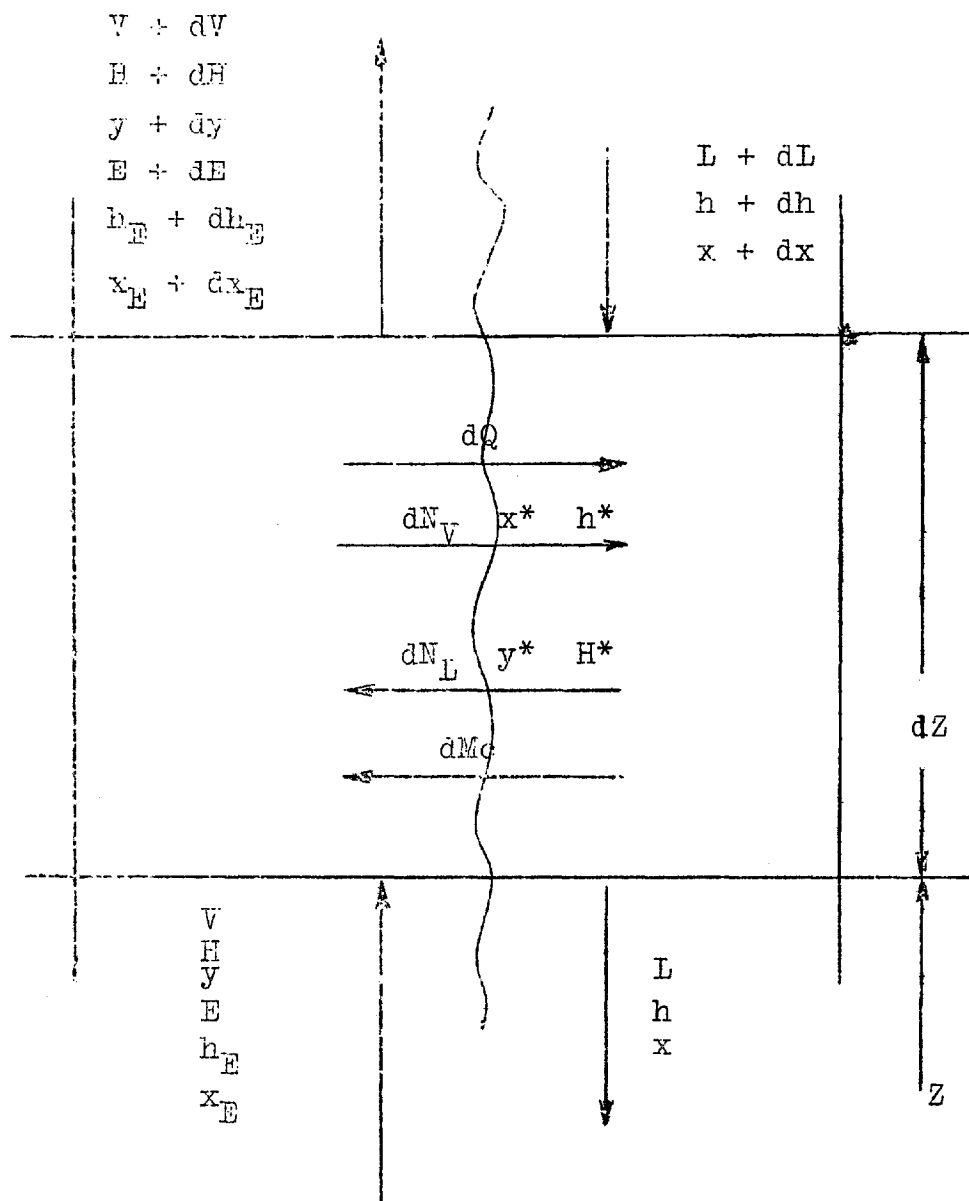


Figure I.2. MATERIAL AND HEAT TRANSFER DIAGRAM.
ENTRAINMENT BEING TAKEN INTO ACCOUNT.

$$\text{Since } d(Vy) + d(Ex_E) = d(Lx) \quad (I.27)$$

$$d(Lh) = H^* dN_L - h^* dN_V - dQ \quad (I.28)$$

$$\text{Since } d(VH) + d(Eh_E) = d(Lh) \quad (I.29)$$

By equations (I.24) and (I.26), we eliminate dN_L :

$$d(Lx) = y^*(dL + dN_V) - x^* dN_V + dMc$$

$$\text{or } d(Lx) = y^*dL + (y^* - x^*) dN_V + dMc \quad (I.30)$$

By equations (I.24) and (I.28), we eliminate

dN_L again :

$$d(Lh) = H^* dL + (H^* - h^*) dN_V - dQ$$

$$\text{or } dN_V = \frac{d(Lh) - H^*dL + dQ}{H^* - h^*} \quad (I.31)$$

Substituting for dN_V by (I.31) in (I.30)

$$d(Lx) = y^*dN_L + dMc + \gamma \{d(Lh) - h^*dL + dQ\}$$

$$\text{where } \gamma = \frac{y^* - x^*}{H^* - h^*}$$

By equations (I.2) and (I.4), as before,

$$d(Lx) = y^*dL + K'(y_e - y)dZ + \gamma \{d(Lh) - h^*dL + U'(t_V - t_L)dZ\} \quad (I.3)$$

Expanding and dividing by dZ

$$L \frac{dx}{dz} + x \frac{dL}{dz} = y^* \frac{dL}{dz} + K'(y_e - y) + \gamma \left[L \frac{dh}{dz} + h \frac{dL}{dz} - H^* \frac{dL}{dz} - U'(t_V - t_L) \right]$$

$$\text{or } L \frac{dx}{dz} \left(1 - \gamma \frac{dh}{dx}\right) = K'(y_e - y) + \gamma U'(t_v - t_L) + [(y^* - x) - \gamma(H^* - h)] \frac{dL}{dz} \quad (\text{I.33})$$

$$\text{i.e. } \frac{\frac{dx}{dz}}{y_e - y} = \frac{K' + \gamma U' \frac{t_v - t_L}{y_e - y} + \frac{[(y^* - x) - \gamma(H^* - h)]}{(y_e - y)} \frac{dL}{dz}}{L \left(1 - \gamma \frac{dh}{dx}\right)} \quad (\text{I.34})$$

For total reflux, as in our experiment, $x=y$

and

$$\frac{\frac{dy}{dz}}{y_e - y} = \frac{K' + \gamma U' \frac{t_v - t_L}{y_e - y} + \frac{[(y^* - y) - \gamma(H^* - h)]}{(y_e - y)} \frac{dL}{dz}}{L \left(1 - \gamma \frac{dh}{dy}\right)} \quad (\text{I.35})$$

Thus, as before,

$$\frac{H_c}{H_T} = \frac{N_T}{N_c} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'}{K'} \frac{(t_v - t_L)}{(y_e - y)} + \frac{[(y^* - y) - \gamma(H^* - h)]}{K'(y_e - y)} \frac{dL}{dz}}{\left(1 - \gamma \frac{dh}{dy}\right)} \quad (\text{I.36})$$

Substituting (I.25) into (I.36),

$$\frac{N_T}{N_c} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'}{K'} \frac{(t_v - t_L)}{(y_e - y)} + \frac{[(y^* - y) - \gamma(H^* - h)]}{K'(y_e - y)} \frac{dV}{dz} + \frac{[(y^* - y) - \gamma(H^* - h)]}{K'(y_e - y)} \frac{dE}{dz}}{1 - \gamma \frac{dh}{dy}} \quad (\text{I.37})$$

$$\text{or } \frac{N_T}{N_c} = \frac{V}{L} \cdot \frac{1 + \gamma \frac{U'}{K'} \frac{(t_v - t_L)}{(y_e - y)} + \frac{[(y^* - y) - \gamma(H^* - h)]}{K'(y_e - y)} \frac{dV}{dy} \frac{dy}{dz} + \frac{[(y^* - y) - \gamma(H^* - h)]}{K'(y_e - y)} \frac{dE}{dy} \frac{dy}{dz}}{1 - \gamma \frac{dh}{dy}} \quad (\text{I.38})$$

We assume as an approximation that equation (I.23) still holds in this case, that is, entrainment is small. Then

$$\begin{aligned} \frac{dV}{dz} &= \frac{dV}{dy} \cdot \frac{dy}{dz} = V \frac{dy}{dz} \cdot \frac{1}{V} \frac{dV}{dy} \\ &= V \frac{dy}{dz} \left\{ -\frac{1}{H-h} \cdot \frac{d(H-h)}{dy} \right\} \end{aligned} \quad (\text{I.39})$$

From equation (I.19)

$$\frac{1}{H_c} = \frac{K'}{V} = \frac{N_c}{Z}$$

so the third term in the numerator of equation (I.38) is

$$\begin{aligned} \frac{[(y^*-y)-\gamma(H^*-h)]}{K'(y_e-y)} \cdot \frac{dv}{dz} &= \frac{[(y^*-y)-\gamma(H^*-h)]}{K'(y_e-y)} V \frac{dy}{dz} \left[-\frac{1}{H-h} \frac{d(H-h)}{dy} \right] \\ &= \frac{[(y^*-y)-\gamma(H^*-h)]}{(y_e-y)} \frac{Z}{N_c} \frac{dy}{dz} \left[-\frac{1}{H-h} \frac{d(H-h)}{dy} \right] \end{aligned} \quad (I.40)$$

If liquid entrainment is 1% of liquid flow rate relative and is constant throughout the height of the column,

$$\frac{dE}{dZ} = \frac{E}{Z} = \frac{1\%L}{Z} \approx \frac{1\%V}{Z} \approx 0.01 \frac{K'}{N_c} \quad (I.41)$$

Substituting equations (I.40) and (I.41) into (I.38),

$$\frac{N_r}{N_c} = \frac{V}{L} \frac{1 + \gamma \frac{U'(t_v-t_c)}{K'(y_e-y)} + \frac{[(y^*-y)-\gamma(H^*-h)] Z}{(y_e-y) N_c} \frac{dy}{dz} \left[-\frac{1}{H-h} \frac{d(H-h)}{dy} \right] + \frac{[(y^*-y)-\gamma(H^*-h)] 0.01}{(y_e-y) N_c}}{1 - \gamma \frac{dh}{dy}} \quad (I.42)$$

Alternatively, re -consider equation (I.32)

$$L dx - x dL = y^* dL + K'(y_e-y) dz + \gamma [L dh + h dL - H^* dL + U'(t_v-t_c) dz] \quad (I.32')$$

Substituting $L=V + E$ and $dL = dV + dE$ into equation (I.32'), collecting common terms and dividing by dZ ,

$$V \frac{dx}{dz} - [(y^* - x) - \gamma(H^* - h)] \frac{dV}{dz} - \gamma V \frac{dh}{dz} + E \frac{dx}{dz} \\ = K'(y_e - y) + \gamma U'(t_v - t_L) + \gamma E \frac{dh}{dz} + [(y^* - x) - \gamma(H^* - h)] \frac{dE}{dz}$$

$$\text{or } V \frac{dx}{dz} \left\{ \left(1 + \frac{E}{V}\right) - \gamma \frac{dh}{dx} - [(y^* - x) - \gamma(H^* - h)] \frac{1}{V} \cdot \frac{dV}{dx} \right\} \\ = K'(y_e - y) + \gamma U'(t_v - t_L) + \gamma E \frac{dh}{dz} + [(y^* - x) - \gamma(H^* - h)] \frac{dE}{dz}$$

$$\text{thus } \frac{\frac{dx}{dz}}{y_e - y} = \frac{K' + \gamma U' \frac{t_v - t_L}{y_e - y} + \frac{\gamma E}{y_e - y} \frac{dh}{dz} + \frac{[(y^* - y) - \gamma(H^* - h)]}{(y_e - y)} \frac{dE}{dz}}{V \left\{ \left(1 + \frac{E}{V}\right) - \gamma \frac{dh}{dy} - [(y^* - y) - \gamma(H^* - h)] \frac{1}{V} \frac{dV}{dy} \right\}}$$

$$\therefore \frac{N_T}{N_c} = \frac{1 + \gamma \frac{U'(t_v - t_L)}{K'(y_e - y)} + \frac{\gamma E}{K'(y_e - y)} \frac{dh}{dz} + \frac{[(y^* - y) - \gamma(H^* - h)]}{K'(y_e - y)} \frac{dE}{dz}}{\left(1 + \frac{E}{V}\right) - \gamma \frac{dh}{dy} - [(y^* - y) - \gamma(H^* - h)] \frac{1}{V} \frac{dV}{dy}} \quad (\text{I.43})$$

If the total entrainment is still 1% of liquid flow rate, the third term in the numerator in equation (I.43) will be

$$\frac{\gamma E}{K'(y_e - y)} \cdot \frac{dh}{dz} = \frac{\gamma}{y_e - y} \cdot \frac{1\%L}{K'} \cdot \frac{dh}{dy} \cdot \frac{dy}{dz} \\ \approx \frac{\gamma}{y_e - y} \cdot \frac{1\%V}{K'} \cdot \frac{dh}{dy} \cdot \frac{dy}{dz} \\ \approx \frac{\gamma}{y_e - y} \cdot \frac{0.01Z}{N_c} \cdot \frac{dh}{dy} \cdot \frac{dy}{dz}$$

Also, in the denominator,

$$1 + \frac{E}{V} = 1 + \frac{1\%L}{L - 1\%L} = 1.011$$

Substituting into (I.43),

$$\frac{N_T}{N_c} = \frac{1 + \gamma \frac{U'(t_v - t_L)}{K'(y_e - y)} + \frac{\gamma}{(y_e - y)} \cdot \frac{0.01 z}{N_c} \cdot \frac{dh}{dy} \cdot \frac{dz}{dz} + \frac{[(y^* - y) - \gamma(H^* - h)] \cdot 0.01}{(y_e - y)} \cdot \frac{1}{N_c}}{(1.011 - \gamma \frac{dh}{dy}) - [(y^* - y) - \gamma(H^* - h)] \cdot \frac{1}{V} \cdot \frac{dV}{dy}} \quad (I.44)$$

No contact distillation.

With both entrainment and evaporation, if contact distillation does not take place, i.e. $dM_c = 0$, then from equation (I.32), we have

$$L \frac{dx}{dz} + x \frac{dL}{dz} = y^* \frac{dL}{dz} + \gamma U'(t_v - t_L) + \gamma \left[L \frac{dh}{dz} + h \frac{dL}{dz} - H^* \frac{dL}{dz} \right]$$

$$L \frac{dx}{dz} (1 - \gamma \frac{dh}{dx}) = \gamma U'(t_v - t_L) + [(y^* - x) - \gamma(H^* - h)] \cdot \frac{dL}{dz}$$

$$\text{or} \quad \frac{dx}{dz} = \frac{\gamma U'(t_v - t_L) + [(y^* - x) - \gamma(H^* - h)] \frac{dL}{dz}}{L (1 - \gamma \frac{dh}{dx})}$$

Dividing by $(y_e - y)$, for total reflux,

$$\therefore \frac{\frac{dx}{dz}}{y_e - y} = \frac{\gamma U' \frac{t_v - t_L}{y_e - y} + \frac{[(y^* - y) - \gamma(H^* - h)]}{(y_e - y)} \cdot \frac{dL}{dz}}{L (1 - \gamma \frac{dh}{dy})}$$

$$\therefore N_T = \frac{z}{L} \cdot \frac{\gamma U' \frac{t_v - t_L}{y_e - y} + \frac{[(y^* - y) - \gamma(H^* - h)]}{(y_e - y)} \cdot \frac{dL}{dz}}{1 - \gamma \frac{dh}{dy}} \quad (I.45)$$

$\frac{dL}{dz}$ is dealt with as before, by substituting equation (I.25),

hence

$$N_T = \frac{z}{L} \cdot \frac{\gamma U' \frac{t_v - t_L}{y_e - y} + \frac{[(y^* - y) - \gamma(H^* - h)]}{(y_e - y)} \cdot \frac{dV}{dy} \cdot \frac{dy}{dz} + \frac{[(y^* - y) - \gamma(H^* - h)]}{(y_e - y)} \cdot \frac{dE}{dy} \cdot \frac{dy}{dz}}{1 - \gamma \frac{dh}{dy}} \quad (I.46)$$

D. CALCULATION OF CHANGE IN FLOW RATE.

Let L'_T and L'_B be the weight of liquid flow per unit time at the column top and bottom respectively,

M_T and M_B be the molecular weight of mixture at the column top and bottom respectively,

W_B and W_T be the weight per cent of salt in liquid.

A material balance on salt (Fig.I.3) gives

$$\frac{L'_T}{L'_B} = \frac{W_B}{W_T} \quad (I.47)$$

The molal flow rate is $L_T = \frac{L'_T}{M_T}$

$$\text{and} \quad L_B = \frac{L'_B}{M_B}$$

$$\text{so that} \quad \frac{L_T}{L_B} = \frac{W_B}{W_T} \cdot \frac{M_B}{M_T} \quad (I.48)$$

Consider Fig. I.4. With entrainment, under adiabatic conditions, a heat balance in the column gives

$$V_T H_T + L_B h_B = V_B H_B + L_T h_T \quad (I.49)$$

At total reflux,

$$V_T = L_T \quad \text{and} \quad V_B = L_B$$

$$\text{so} \quad L_T (H_T - h_T) = L_B (H_B - h_B)$$

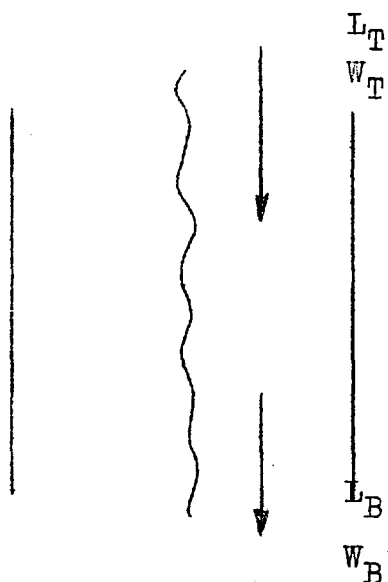


Fig. I.3

where L_T , L_B are molal flow rate of liquid per unit time at top and bottom respectively
 W_T , W_B are weight percent of salt at top and bottom.

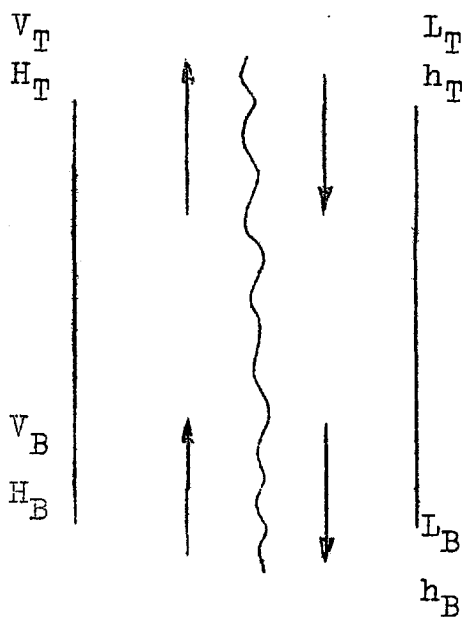


Fig. I.4

where H and h are total heat of vapour and liquid respectively.

V is molal vapour flow rate per unit time.

L is molal liquid flow rate per unit time.

$$\text{or} \quad \frac{L_T}{L_B} = \frac{H_B - h_B}{H_T - h_T} = \frac{\lambda_B}{\lambda_T} \quad (\text{I.50})$$

where λ is the molar latent heat of vaporisation.

If entrainment is present in the bulk vapour, equation (I.48) will give the measured change in flow rate, including both net evaporation and entrainment. The change in flow rate owing to change in total heat (net evaporation) is given by equation (I.50). Thus, the ratio of entrainment to liquid flow rate at column bottom will then be given by the difference of equations (I.48) and (I.50):

$$\frac{E}{L_B} = \left(\frac{W_B}{W_T} \cdot \frac{M_B}{M_T} - \frac{\lambda_B}{\lambda_T} \right) \quad (\text{I.51})$$

E. NUMBER OF TRANSFER UNITS TO SEPARATE DILUTE MIXTURES.

Consider Fig.I.5, for dilute mixtures in the still, the top and bottom compositions, x_T and x_B of the column are close, and the portion 'ab' on the equilibrium curve can be considered to be a straight line, which can be expressed as:

$$y_e = mx + c \quad (\text{I.52})$$

At total reflux,

$$y_e = my + c \quad (\text{I.53})$$

Thus, the overall number of transfer units will be:

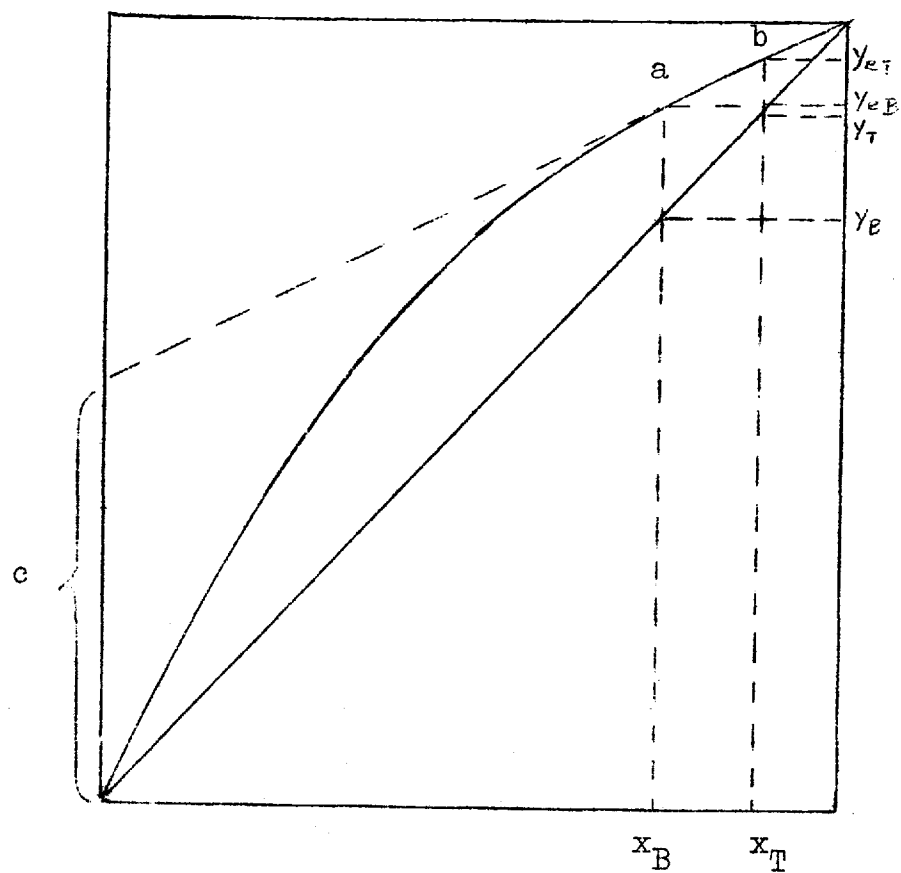


Fig. I.5

$$\begin{aligned}
 N_{OG} &= \int_{y_B}^{y_T} \frac{dy}{y_{eT} - y} \\
 &= \int_{y_B}^{y_T} \frac{dy}{(m-1)y + c} \\
 &= \frac{1}{m-1} \ln \frac{(m-1)y_T + c}{(m-1)y_B + c} \\
 &= \frac{1}{(m-1)} \ln \frac{y_{eT} - y_T}{y_{eB} - y_B}
 \end{aligned}$$

where y_{eT} , y_{eB} are the equilibrium values of the top and bottom compositions respectively.

$$\text{But } m = \frac{y_{eT} - y_{eB}}{y_T - y_B}$$

$$\text{Thus } N_{OG} = \frac{y_T - y_B}{(y_{eT} - y_T) - (y_{eB} - y_B)} \ln \frac{y_{eT} - y_T}{y_{eB} - y_B} \quad (\text{I } 54)$$

$$\text{or } N_{OG} = \frac{y_T - y_B}{\Delta y_m} \quad (\text{I } 55)$$

$$\text{where } \Delta y_m = \frac{(y_{eT} - y_T) - (y_{eB} - y_B)}{\ln \frac{y_{eT} - y_T}{y_{eB} - y_B}}, \text{ the log mean value.}$$

APPENDIX II.
TABLE 4.1.

REFRACTIVE INDEX - COMPOSITION RELATIONSHIPS FOR
METHANOL-WATER SYSTEM AT 25°C. [54].

Weight % of Methanol	$^{25}n_D$	Mole % of Methanol.
0.	1.33232	0.
21.7	1.33823	13.32
37.66	1.34090	25.52
52.22	1.34163	37.4
54.23	1.34163	40.19
64.35	1.34065	50.59
68.2	1.33984	55.01
80.44	1.33663	69.99
86.27	1.33505	78.09
92.48	1.33193	87.47
100	1.32773	100

Max. $^{25}n_D = 1.34170$ at 53% by weight of methanol,
i.e. at 39.01 mole % of methanol.

TABLE 4.2.

172.

CALIBRATION OF CONDUCTIVITY-COMPOSITION AND REFRACTIVE INDEX - COMPOSITION CURVES AT DIFFERENT SALT PER CENT FOR METHANOL-WATER SYSTEM.

mole fraction of m.v.c.	Reading	0.5%		0.75%		1.0%	
		25_{n_D}	Conductivity bridge reading $\times 10^4 \mu mho$	25_{n_D}	Conductivity bridge reading $\times 10^4 \mu mho$	25_{n_D}	Conductivity bridge reading $\times 10^4 \mu mho$
0		1.3340	3.54	1.3342	5.35	1.3342	6.98
0.05		1.3360	2.86	1.3363	4.23	1.3367	5.6
0.1		1.3380	2.38	1.3381	3.58	1.3382	4.74
0.15		1.33985	2.1	1.3400	3.2	1.3403	4.22
0.2		1.3411	1.88	1.3414	2.82	1.3419	3.72
0.25		1.3419	1.82	1.34225	2.6	1.3428	3.43
0.3		1.3428	1.69	1.3428	2.49	1.3433	3.23
0.35		1.3428	1.62	1.34285	2.39	1.3434	3.17
0.4		1.3425	1.59	1.34275	2.27	1.3430	2.97
0.45		1.3420	1.57	1.3428	2.24	1.3428	2.97
0.5		1.3413	1.55	1.3418	2.21	1.3421	2.93
0.55		1.34075	1.55	1.3410	2.21	1.3412	2.69
0.6		1.3398	1.51	1.33995	2.16	1.3404	2.79
0.65		1.33865	1.52	1.3390	2.14	1.3392	2.71
0.7		1.3373	1.52	1.3377	2.14	1.3380	2.69
0.75		1.3360	1.51	1.3364	2.14	1.33695	2.73
0.8		1.3346	1.54	1.3360	2.15	1.3358	2.73
0.85		1.3332	1.55	1.3337	2.17	1.3330	1.64
0.9		1.3315	1.59	1.3320	2.21	1.3324	2.72
0.95		1.3300	1.63	1.3305	2.24	1.3310	2.77
1.0		1.3279	0.69	1.3280	1.07	1.3278	0.92

TABLE 4.2 (continued)

173.

% of NaCl by weight Reading mole fraction of m.v.c.	1.25%		1.5%		1.75%		2.0%	
	$^{25}n_D$	Conductivity bridge reading $\times 10^4 \mu mho$	$^{25}n_D$	Conductivity bridge reading $\times 10^4 \mu mho$	$^{25}n_D$	Conductivity bridge reading $\times 10^4 \mu mho$	$^{25}n_D$	Conductivity bridge reading $\times 10^4 \mu mho$
0	1.3351	8.4	1.3358	9.8	1.3360	12.4	1.3362	13.9
0.05	1.3361	6.6	1.3374	8.0	1.3380	10.2	1.3384	11.2
0.1	1.3390	5.7	1.3392	6.4	1.3399	7.5	1.3404	8.8
0.15	1.3407	4.94	1.3411	5.8	1.3418	6.6	1.3421	7.6
0.2	1.3422	4.34	1.3428	5.1	1.3431	5.9	1.3436	6.9
0.25	1.3431	4.08	1.3432	4.73	1.34395	5.6	1.3442	6.4
0.3	1.3435	3.97	1.3439	4.52	1.3445	5.2	1.3448	6.1
0.35	1.3438	3.82	1.3439	4.42	1.3444	5.1	1.3448	5.5
0.4	1.3436	3.63	1.34375	4.33	1.3443	4.85	1.3448	5.5
0.45	1.3430	3.52	1.3435	4.18	1.3440	4.72	1.3444	5.1
0.5	1.3426	3.41	1.3429	4.09	1.3432	4.42	1.3437	4.99
0.55	1.34195	3.39	1.3420	3.97	1.3424	4.40	1.3426	4.88
0.6	1.3410	3.33	1.3413	3.88	1.3416	4.34	1.3418	4.82
0.65	1.3397	3.29	1.3401	3.86	1.3404	4.28	1.34085	4.72
0.7	1.3385	3.23	1.3391	3.83	1.3394	4.29	1.3397	4.72
0.75	1.3375	3.14	1.3375	3.83	1.3381	4.29	1.3384	4.72
0.8	1.3360	3.21	1.3366	3.63	1.3368	4.22	1.3371	4.77
0.85	1.3334	3.29	1.3350	3.74	1.3354	4.22	1.3359	4.72
0.9	1.3330	3.29	1.3335	3.74	1.3338	4.2	1.3338	4.5
0.95	1.3310	2.61	1.3313	2.98	1.3312	2.82	1.3312	2.98
1.0	1.3277	0.729	1.3280	0.88	1.3283	1.12	1.3282	1.11

TABLE 4.3.

CALIBRATED REFRACTIVE INDEX - COMPOSITION RELATION
FOR CYCLOHEXANE-TOLUENE SYSTEM AT 20°C.

Mole % of Cyclohexane in liquid.	$^{20}_{n_D}$
0	1.4964
0.1	1.48755
0.2	1.4798
0.3	1.4722
0.4	1.4650
0.5	1.4580
0.6	1.4511
0.7	1.4448
0.8	1.4385
0.9	1.4325
1.0	1.4265

TABLE 5.1

175.

CARBON TETRACHLORIDE-TOLUENE SYSTEM.

6" HEIGHT OF PACKINGS.

CONSTANT FLOW RATE : 21.5 cc/min.

RASCHIG RINGS.					FENSKE HELICES				
Run	x_T	x_B	x_{av}	N_{OG}	Run	x_T	x_B	x_{av}	N_{OG}
1	.995	.894	.945	4.13	1	.968	.668	.818	3.92
2	.993	.873	.933	3.67	2	.965	.665	.815	3.8
3	.981	.855	.918	2.74	3	.963	.645	.804	3.89
4	.973	.852	.913	2.37	4	.95	.656	.803	3.43
5	.973	.835	.904	2.58	5	.948	.637	.793	3.51
6	.97	.783	.877	3.05	6	.94	.615	.778	3.45
7	.958	.753	.856	2.96	7	.925	.593	.759	3.29
8	.922	.722	.822	2.37	8	.919	.559	.739	3.35
9	.923	.697	.810	2.54	9	.91	.533	.722	3.3
10	.922	.696	.809	2.56	10	.859	.499	.679	2.72
11	.874	.62	.747	2.28	11	.843	.47	.657	2.62
12	.843	.614	.729	1.95	12	.8	.435	.613	2.27
13	.792	.466	.629	2.07	13	.713	.397	.655	1.65
14	.783	.437	.61	2.08	14	.67	.357	.514	1.52
15	.696	.397	.547	1.52	15	.632	.287	.46	1.62
16	.643	.36	.502	1.32	16	.531	.268	.4	1.19
17	.613	.34	.477	1.24	17	.491	.237	.364	1.18
18	.559	.305	.432	1.11	18	.43	.217	.324	1.07
19	.503	.276	.39	1.03	19	.364	.201	.283	0.87
20	.456	.267	.362	0.9	20	.268	.154	.211	0.79
21	.392	.237	.315	0.78	21	.213	.126	.166	0.67
22	.335	.201	.268	0.74					
23	.278	.197	.238	0.48					
24	.268	.174	.221	0.57					
25	.2	.166	.183	0.24					
26	.185	.158	.172	0.19					
27	.163	.155	.159	0.385					

TABLE 5.2.

ACETONE-BENZENE SYSTEM.

6" HEIGHT OF PACKING: FENSKE HELICES.

CONSTANT FLOW RATE ; 25 cc/min.

Run	x_T	x_B	x_{av}	N_{OG}	Run	x_T	x_B	x_{av}	N_{OG}
1	.958	.791	.875	4.9	14	.905	.565	.735	5.03
2	.95	.776	.863	5.01	15	.893	.556	.725	4.63
3	.92	.746	.833	3.87	16	.898	.514	.706	5.12
4	.917	.733	.825	3.9	17	.867	.433	.65	4.67
5	.916	.727	.822	3.96	18	.844	.412	.628	4.62
6	.913	.725	.819	3.86	19	.823	.308	.566	4.41
7	.913	.712	.813	3.99	20	.841	.278	.560	4.89
8	.904	.71	.807	3.66	21	.812	.21	.511	4.63
9	.916	.695	.806	4.3	22	.714	.132	.423	3.76
10	.914	.671	.793	4.46	23	.52	.078	.299	2.49
11	.926	.629	.778	5.35	24	.322	.04	.153	1.77
12	.921	.606	.764	5.33	25	.139	.017	.078	1.42
13	.915	.592	.754	5.2	26	.056	.003	.03	1.4

TABLE 5.3.

ACETONE-CHLOROFORM SYSTEM.

FENSKE HELICES PACKING.

CONSTANT FLOW RATE : 25 cc/min.

HEIGHT OF PACKING = 6".

Run	x_T	x_B	x_{av}	N_{OG}	Run	x_T	x_B	x_{av}	N_{OG}
1	.978	.893	.94	4.34	14	.33	.336	.333	.71
2	.969	.87	.92	3.51	15	.257	.301	.279	1.84
3	.966	.859	.913	3.57	16	.225	.283	.254	1.91
4	.937	.805	.871	2.86	17	.174	.249	.212	2.11
5	.899	.675	.787	3.16	18	.148	.219	.184	1.96
6	.871	.647	.759	2.92	19	.122	.182	.152	1.79
7	.82	.586	.703	2.81	20	.122	.167	.145	1.38
8	.741	.536	.639	2.5	21	.109	.157	.133	1.53
9	.592	.462	.527	2.12	22	.109	.148	.129	1.24
10	.554	.453	.504	1.78	23	.105	.135	.12	0.99
11	.527	.431	.479	1.96	24	.103	.122	.113	0.64
12	.439	.41	.425	0.99					
13	.388	.38	.384	0.55					

TABLE 5.3 (continued)

HEIGHT OF PACKING = 10"

Run	x_T	x_B	x_{av}	N_{OG}	Run	x_T	x_B	x_{av}	N_{OG}
1	.992	.922	.957	5.59	26	.359	.367	.363	1.38
2	.99	.912	.951	5.26	27	.335	.348	.341	2.64
3	.985	.896	.941	4.67	28	.308	.337	.32	2.6
4	.975	.859	.917	4.3	29	.26	.317	.289	2.65
5	.975	.839	.907	4.54	30	.225	.3	.262	2.72
6	.944	.769	.857	3.6	31	.185	.272	.229	2.58
7	.927	.724	.826	3.46	32	.16	.252	.206	2.59
8	.921	.693	.807	3.61	33	.145	.234	.19	2.47
9	.893	.674	.784	3.06	34	.13	.214	.172	2.39
10	.889	.672	.781	2.99	35	.133	.21	.172	2.19
11	.886	.664	.775	3.02	36	.124	.188	.156	1.89
12	.867	.607	.737	3.33	37	.119	.176	.148	1.75
13	.852	.554	.703	3.76	38	.116	.162	.139	1.41
14	.842	.55	.696	3.65	39	.114	.145	.13	0.98
15	.831	.527	.679	3.82	40	.112	.133	.123	0.69
16	.74	.483	.612	3.4	41	.097	.117	.107	0.71
17	.714	.465	.589	3.53					
18	.615	.436	.526	3.04					
19	.517	.403	.460	2.88					
20	.483	.406	.445	2.13					
21	.462	.395	.436	2.29					
22	.451	.394	.423	2.12					
23	.424	.386	.405	1.78					
24	.413	.379	.396	2.175					
25	.383	.37	.377	1.71					

METHANOL-WATER SYSTEM.

Measurement of the number of transfer units in the
 pear column and the temperature of the bulk phases.

I. SHIELDED THERMOCOUPLE IN VAPOUR
 PHASE AT POSITION (4).

Run	x	Temperature °C.		x_m	N_{OG}
		Vapour	Liquid		
IA.6	.9975	65.2	65.2		
5	—	65.52	65.55		
4	.995	64.88			
3	.9915	65.85	65.88		
2	.99	65.9	65.3		
1	.9855	65.72	65.88	.991	2.49
IB.6	.9855	65.28	65.92		
5	.97	65.8	66.3		
4	.957	65.7			
3	.951	66.35	66.88		
2	.937	67.08	65.8		
1	.928	66.92	66.52	.956	2.542
IC.6	.969	65.32	66.1		
5	.963	66.05	66.6		
4	.94	66.08			
3	.94	66.48	67.32		
2	.925	67.35	65.9		
1	.9035	67.32	66.8	.93	2.055
ID.6	.963	65.6	66.6		
5	.957	66.28	67.18		
4	.927	66.58			
3	.9215	66.8	68.1		
2	.895	68.2	66.35		
1	.867	68.1	67.32	.912	2.22

TABLE 6.1 (continued)
Temperature

180.

Run	x	Vapour	Liquid	x_m	N_{OG}
IE.6	.963	66.13	66.88		
5	.957	66.25	66.82		
4	.927	<u>67.82</u>			
3	.9035	<u>67.25</u>	68.6		
2	.88	68.9	66.9		
1	.848	69.02	67.9	.902	2.455
IF.6	.958	66.35	67		
5	.951	66.4	67.08		
4	.91	<u>67.72</u>			
3	.898	<u>67.6</u>	69.1		
2	.864	69.48	67.2		
1	.805	69.85	68.4	.88	2.68
IG.6	.949	66.22	67.12		
5	.94	66.5	67.08		
4	.905	<u>67.8</u>			
3	.874	<u>69.3</u>	67.68		
2	.843	69.58	67.22		
1	.774	70.55	68.6	.855	2.6
IH.6	.952	65.98	67.4		
5	.925	66.72	67.95		
4	.899	<u>68.0</u>			
3	.857	<u>69.5</u>	67.62		
2	.821	69.8	67.2		
1	.763	70.5	68.55	.851	2.78
II.6	.946	67.4	66.2		
5	.934	66.9	66.32		
4	.901	<u>68.05</u>			
3	.87	<u>69.72</u>	67.8		
2	.836	70.12	67.4		
1	.751	71.0	68.8	.844	2.66
IJ.6	.929	67.9	66.6		
5	.904	67.2	66.65		
4	.861	<u>68.95</u>			
3	.806	<u>71.22</u>	69.5		
2	.751	71.72	68.18		
1	.644	73.05	70	.778	2.84

TABLE 6.1 (continued)

181.

Temperature °C.					
Rui.	x	Vapour	Liquid	x_m	N_{OG}
IK.6	.892	69.25	67.05		
5	.843	70.32	68.02		
4	.775	<u>71.12</u>			
3	.69	<u>73.9</u>	68.68		
2	.657	74.62	69.6		
1	.546	76.72	71.35	.706	2.56
IL.6	.861	69.88	67.4		
5	.804	71.3	68.5		
4	.755	<u>72.6</u>			
3	.632	<u>75.82</u>	70.3		
2	.587	76.6	70.4		
1	.43	79.2	72.4	.62	2.49
IM.6	.812	71.88	69		
5	.73	73.82	70.2		
4	.652	<u>76.52</u>			
3	.537	<u>79.88</u>	72.6		
2	.474	80.72	73.28		
1	.353	84.22	75.7	.551	2.22
IN.6	.676	75.68	69.8		
5	.582	76.92	71.62		
4	.447	<u>83.12</u>			
3	.375	<u>88.2</u>	75.42		
2	.239	87.38	77.52		
1	.147	90.92	83.5	.347	1.78
IO.6	.534	79.75	72.75		
5	.39	86.7	76.6		
4	.266	<u>93.3</u>			
3	.147	<u>96.7</u>	86		
2	.074	87.68	88		
1	.051	98.4	95.5	.205	1.47
IP.6	.295	89.3	78.05		
5	.186	93.6	83.8		
4	.124	<u>96.68</u>			
3	.091	<u>97.82</u>	90.5		
2	.053	98.35	91.68		
1	.033	99.42	97.02	.125	0.81

TABLE 6.1 (continued)

Run	x	Temperature		x_m	N_{OG}
		Vapour	Liquid		
IQ.6	.195	95.05	82.3		
5	.116	98	89.42		
4	.072	<u>98.88</u>			
3	.058	<u>99.5</u>	92.85		
2	.032	99.3	95.22		
1	.025		99.12	.077	.62
IR.6	.028	99.73	96.95		
5	.021	100	98.65		
4	.018	<u>99.8</u>			
3	.016		99.5		
2	.012		98.55		
1	.0115		100.22	.0183	.14

II. SHIELDED THERMOCOUPLE IN VAPOUR PHASE AT
POSITION (1).

Run	x	Vapour	Liquid	x_m	N_{OG}
II.A.6	.995	65.02	65.0		
5	.989	65.28	65.25		
4	.976	65.38			
3	.981	65.42	65.38		
2	.979	65.35	65.42		
1	.98	<u>64.52</u>		.988	2.3
II.B.6	.983	66.1	65.62		
5	.982	66.35	66.0		
4	.969	66.38			
3	.967	66.58	66.55		
2	.965	66.65	65.82		
1	.96	<u>66.22</u>		.971	1.35
II.C.6	.981	66.08	66.0		
5	.977	66.2	66.0		
4	.963	66.6			
3	.959	66.88	66.8		
2	.94	67.18	66.2		
1	.936	<u>67.1</u>		.951	2.02

TABLE 6.1 (continued)

Run	x	Temperature °C.		x_m	N_{OG}
		Vapour	Liquid		
II.D.6	.98	66.4	65.78		
5	.966	66.65	66.12		
4	.946	67.02			
3	.94	67.38	66.52		
2	.928	67.5	66.12		
1	.91	<u>67.08</u>		.931	2.41
II.E.6	.965	66.65	65.9		
5	.954	67.12	66.35		
4	.934	67.22			
3	.928	67.92	66.92		
2	.899	68.22	66.4		
1	.87	<u>68.62</u>		.913	2.26
II.F.6	.966	66.45	65.9		
5	.961	66.38	65.75		
4	.936	67.1			
3	.913	67.92	66.82		
2	.892	68.28	66.42		
1	.854	<u>69.02</u>		.909	2.54
II.G.6	.965	66.78	65.8		
5	.943	67.25	66.5		
4	.914	67.4			
3	.906	68.5	67.2		
2	.874	68.78	66.68		
1	.843	<u>69.38</u>		.9	2.6
II.H.6	.957	66.7	65.65		
5	.942	67.25	66.4		
4	.911	67.45			
3	.898	68.75	67.22		
2	.854	69.02	66.6		
1	.81	<u>69.95</u>		.882	2.62
II.I.6	.956	66.8	66.15		
5	.946	66.72	66.02		
4	.909	67.52			
3	.895	68.9	67.3		
2	.854	69.1	66.85		
1	.808	<u>70.5</u>		.877	2.55

TABLE 6.1 (continued)

184.

Run	x	Temperature °C		x_m	N_{OG}
		Vapour	Liquid		
II.J.6	.934	68.2	66.35		
5	.88	68.9	67.08		
4	.859	69			
3	.795	71.3	68.28		
2	.755	71.92	68.08		
1	.675	<u>74.58</u>		.797	2.79
II.K.6	.922	68.3	66.6		
5	.87	69.4	67.38		
4	.843	69.42			
3	.775	72.05	68.78		
2	.714	72.7	68.45		
1	.625	<u>75.9</u>		.761	2.75
II.L.6	.901	68.98	66.8		
5	.852	70.1	67.68		
4	.813	70.15			
3	.722	73.1	69.22		
2	.676	74.08	69.18		
1	.556	<u>78.38</u>		.715	2.66
II.M.6	.902	69.08	66.9		
5	.843	70.05	67.8		
4	.807	70.28			
3	.722	73.48	69.4		
2	.666	74.22	69.38		
1	.543	<u>78.1</u>		.706	2.72
II.N.6	.894	69.02	66.88		
5	.836	70.4	67.82		
4	.8	70.45			
3	.701	73.78	68.8		
2	.644	74.62	69.4		
1	.536	<u>78.12</u>		.699	2.61
II.O.6	.872	69.72	67.12		
5	.804	71.02	68.25		
4	.77	71.18			
3	.664	75.02	70.18		
2	.604	75.75	70.12		
1	.469	<u>80.05</u>		.650	2.52

TABLE 6.1 (continued)

Ru.	x	Vapour	Liquid	x_m	N_{OG}
II.P.6	.817	70.9	68.15		
5	.739	72.95	69.42		
4	.684	72.92			
3	.551	79.28	71.72		
2	.481	80.35	72.22		
1	.353	<u>86.35</u>		.552	2.26
II.Q.6	.744	83.84	69.12		
5	.646	76.42	70.78		
4	.573	74.82			
3	.456	82.05	73.58		
2	.39	84.7	74.85		
1	.256	<u>92.6</u>		.457	1.92
II.R.6	.71	76.48	70.6		
5	.581	79.7	74.95		
4	.469	77.8	9		
3	.39	88.7	76.95		
2	.222	89.8	78.1		
1	.134	<u>97.2</u>		.346	2.0
II.S.6	.594	79.55	71.82		
5	.465	83.22	75		
4	.39	80.3			
3	.249	96.4	81.18		
2	.125	95.65	83		
1	.072	<u>99.38</u>		.247	1.62
II.T.6	.545	80	72.42		
5	.39	84.52	75.5		
4	.266	84.55			
3	.154	96.1	84.2		
2	.082	97.4	87		
1	.05	<u>99.02</u>		.207	1.52
II.U.6	.272				
5	.113				
4	.093				
3	.072				
2	.045				
1	.029	<u>99.85</u>		.108	0.79

TABLE 6.1 (continued)

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Temperature °C.					
Run	x	Vapour	Liquid	x_m	N_{OG}
II.V.6	.091	97.4	92.15		
5	.067	98.32	93.08		
4	.005	98.15			
3	.045	101.88	94.92		
2	.027	100.65	98.15		
1	.023	<u>99.35</u>		.065	0.345
II.W.6	.054	99.25	91.8		
5	.027	99.8	96.8		
4	.023	99.8			
3	.014		99.4		
2	.011		94.9		
1	.0105	<u>100.9</u>		.026	0.295
II.X.6	.0185	99.7	98.05		
5	.015	99.9	99.25		
4	.014				
3	.012		99.8		
2	.0105				
1	.0105			.014	0.014

TABLE 6.2 METHANOL-WATER SYSTEM.

Measurement of the salt percentages in liquid phase to determine the change in flow rate along the pear column. The rate of liquid flow at column bottom being kept constant at 12 cc/min. and the concentrated salt solution being introduced at constant rate of 0.625 cc/min.

RUN	n_D^{25}	Wt.% of Salt	x	M	H	h	$\frac{L_n}{L_B}$	$\frac{(H-h)_B}{(H-h)_n}$	x_m	$\frac{E}{L_B}$	L_B	V	E
A.6	1.33395	.6804	.839	29.78	9875	1344	1.089	1.001		.088		.319	.028
5	1.3342	.7216	.835	29.72	9880	1345	1.026	1.001					
4	1.3344	.7762	.829	29.62	9889	1348	0.96	1.001					
3	1.3343	.7445	.830	29.64	9884	1348	0.998	1.001					
2	1.3344	.7593	.827	29.6	9891	1349	0.983	1.001					
1	1.3342	.7430	.832	29.68	9884	1346	1	1	.835		.3185		
B.6	1.3350	.7802	.807	29.32	9913	1356	1.033	1.003		.03		.334	.01
5	1.3353	.7563	.795	29.14	9931	1362	1.071	1.002					
4	1.3356	.7972	.787	29.04	9940	1364	1.021	1.001					
3	1.33575	.8317	.784	29.0	9946	1366	0.979	1.001					
2	1.33595	.8160	.776	28.89	9953	1369	1.002	1.001					
1	1.3360	.8189	.774	28.86	9958	1370	1	1	.785		.333		
C.6	1.3347	.7051	.813	29.4	9908	1354	1.133	1.005		.128		.336	
5	1.33505	.7575	.803	29.28	9919	1358	1.06	1.003					
4	1.33575	.7643	.778	28.91	9951	1378	1.064	1.002					
3	1.3360	.8029	.772	28.84	9959	1371	1.015	1.001					
2	1.3360	.807	.768	28.79	9963	1372	1.011	1.001					
1	1.33615	.8182	.766	28.76	9967	1373	1	1	.786		.335		
D.6	1.33535	.7500	.793	29.12	9933	1362	1.059	1.007		.052		.349	.0428
5	1.3355	.776	.788	29.06	9939	1364	1.024	1.006					
4	1.3361	.801	.768	28.79	9964	1372	1.002	1.004					
3	1.3366	.8051	.749	28.5	9989	1380	1.008	1.004					
2	1.3370	.817	.734	28.3	10007	1386	1	1.001					
1	1.3374	.823	.719	28.1	10026	1393	1	1	.76		.347		.187

Run	n_D^{25}	Wt. % of salt	α	M	H	h	$\frac{L_n}{L_B}$	$\frac{(H-h)_B}{(H-h)_n}$	χ_m	$\frac{E}{L_B}$	L_B	V	E
E.6	1.3351	.735	.799	29.2	9925	1360	1.126	1.008		.118		.351	.0411
5	1.33595	.763	.771	28.82	9960	1371	1.098	1.004					
4	1.3363	.7960	.764	28.72	9970	1374	1.055	1.003					
3	1.3370	.843	.736	28.34	10003	1385	1.011	1.002					
2	1.3374	.827	.717	28.07	10028	1393	1.039	1					
1	1.33755	.861	.714	28.02	10032	1395	1	1	.75		.348		
F.6	1.3351	.711	.798	29.2	9927	1360	1.132	1.012		.12		.361	.0428
5	1.3360	.748	.768	28.79	9964	1372	1.091	1.009					
4	1.3368	.783	.739	28.37	10000	1385	1.06	1.005					
3	1.3374	.814	.716	28.07	10030	1394	1.029	1.003					
2	1.33805	.827	.691	27.71	10060	1403	1.029	1.001					
1	1.33835	.853	.682	27.59	10073	1407	1	1	.74		.3565		
G.6	1.33495	.599	.797	29.18	9929	1360	1.089	1.016		.073		.378	.0272
5	1.3357	.604	.768	28.79	9964	1372	1.092	1.014					
4	1.3363	.653	.749	28.51	9989	1380	1.02	1.012					
3	1.3374	.675	.707	27.94	10041	1397	1.008	1.009					
2	1.3381	.670	.678	27.53	10079	1407	1.028	1.005					
1	1.33935	.706	.630	26.87	10139	1428	1	1	.706		.372		
H.6	1.33535	.634	.784	29.0	9945	1366	1.064	1.022		.042		.40	.0164
5	1.3360	.624	.759	28.66	9975	1376	1.091	1.02					
4	1.3370	.662	.722	28.14	10023	1391	1.048	1.016					
3	1.3384	.719	.669	27.4	10090	1413	0.991	1.011					
2	1.33905	.719	.644	27.03	10121	1423	1.005	1.008					
1	1.3408	.754	.564	25.92	10223	1455	1	1	.669		.392		
I.6	1.3354	.545	.776	28.9	9953	1369	1.21	1.023		.187		.407	.0744
5	1.3363	.601	.746	28.48	9992	1381	1.114	1.020					
4	1.3374	.625	.704	27.9	10045	1399	1.092	1.016					
3	1.3390	.687	.643	27.02	10121	1423	1.027	1.01					
2	1.3400	.698	.601	26.45	10176	1440	1.033	1.007					
1	1.3411	.743	.545	25.64	10248	1464	1.0	1	.654		.398		

Run	n_D^{25}	Wt % of salt	α	M	H	h	$\frac{L_n}{L_B}$	$\frac{(H-h)_B}{(H-h)_n}$	χ_m	$\frac{E}{L_B}$	L_B	V	E
J.6	1.33615	.658	.755	28.5	9980	1377	1.161	1.027		.134		.431	.0562
5	1.3376	.683	.700	27.84	10049	1400	1.146	1.022					
4	1.3383	.727	.674	27.47	10085	1410	1.09	1.018					
3	1.34015	.819	.601	26.46	10176	1440	1.004	1.012					
2	1.3411	.816	.551	25.75	10239	1460	1.039	1.006					
1	1.3422	.883	.478	24.71	10332	1494	1	1	.605		.42		
K.6	1.33695	.625	.722	28.13	10022	1391	1.179	1.039		.14		.481	.0649
5	1.3381	.675	.679	27.53	10078	1408	1.115	1.034					
4	1.33915	.720	.639	26.98	10128	1425	1.067	1.031					
3	1.3416	.815	.52	25.3	10280	1475	1.005	1.029					
2	1.3423	.823	.457	24.41	10360	1502	1.031	1.012					
1	1.3432	.899	.359	23.03	10510	1545	1	1	.519		.464		
L.6	1.3373	.659	.709	27.97	10038	1397	1.163	1.049		.114		.52	.0565
5	1.33915	.741	.640	26.99	10127	1424	1.072	1.041					
4	1.3404	.744	.583	26.2	10199	1448	1.102	1.035					
3	1.3428	.860	.416	23.84	10420	1520	1.047	1.018					
2	1.34315	.879	.372	23.21	10490	1539	1.051	1.012					
1	1.3431	.977	.281	21.97	10643	1583	1	1	.463		.496		
M.6	1.3375	.549	.696	27.78	10056	1402	1.123	1.048		.075		.519	.0372
5	1.3392	.607	.631	26.88	10138	1428	1.049	1.041					
4	1.34045	.624	.575	26.07	10210	1451	1.052	1.035					
3	1.34235	.691	.446	24.25	10378	1507	1.022	1.021					
2	1.3428	.719	.386	23.41	10468	1533	1.019	1.016					
1	1.34275	.779	.284	21.99	10639	1581	1	1	.46		.495		
N.6	1.33975	.55	.605	26.5	10171	1439	1.128	1.068		.06		.598	.0336
5	1.3418	.625	.491	24.89	10316	1488	1.058	1.056					
4	1.3429	.67	.372	23.22	10490	1529	1.058	1.042					
3	1.3427	.740	.284	21.98	10640	1557	1.011	1.027					
2	1.34095	.799	.178	20.49	10868	1638	1.002	1.011					
1	1.3400	.820	.146	20.04	10978	1655	1	1	.323		.56		

Run	n_D^{25}	Wt. % of Salt	x	M	H	h	$\frac{L_n}{L_B}$	$\frac{(H-h)_E}{(H-h)_n}$	x_m	$\frac{E}{L_B}$	L_B	V	E
O.6	1.3398	.641	.607	26.52	10169	1438	1.119	1.08		.039		.621	.0224
5	1.3420	.721	.481	24.77	10328	1491	1.063	1.067					
4	1.3429	.779	.378	23.3	10481	1537	1.047	1.054					
3	1.34235	.850	.235	21.29	10729	1607	1.051	1.032					
2	1.3403	.950	.147	20.04	10973	1654	1.001	1.011					
1	1.33925	.969	.117	19.63	11093	1672	1	1	.298		.575		
P.6	1.3404	.635	.576	26.09	10207	1450	1.047	1.088		.41		.644	—
5	1.3425	.703	.429	24.02	10402	1515	1.028	1.073					
4	1.3428	.759	.296	22.15	10616	1575	1.031	1.054					
3	1.3417	.815	.209	20.93	10782	1622	1.018	1.041					
2	1.3390	.859	.116	19.62	11097	1672	1.029	1.011					
1	1.33785	.905	.084	19.18	11222	1694	1	1	.256		.593		
Q.6	1.3415	.647	.514	25.21	10288	1477	1.11	1.098		.012		.68	.0063
5	1.3429	.734	.364	23.09	10503	1543	1.069	1.079					
4	1.34195	.799	.222	21.11	10755	1615	1.075	1.059					
3	1.3401	.893	.144	20.0	10988	1656	1.014	1.039					
2	1.3375	.939	.074	19.03	11260	1701	1.012	1.011					
1	1.33595	.98	.035	18.48	11407	1741	1	1	.178		.62		
R.6	1.3380	.801	.094	19.31	11190	1686	1.06	1.022		.038		.646	.0246
5	1.3370	.840	.066	18.93	11290	1708	1.032	1.013					
4	1.33635	.88	.048	18.68	11357	1726	1.0	1.009					
3	1.3360	.887	.039	18.52	11392	1736	1.0	1.006					
2	1.33525	.917	.02	18.28	11460	1761	.981	1.002					
1	1.3350	.903	.014	18.19	11483	1771	1	1	.042		.632		
S.6	1.3384	.828	.102	18.43	11154	1681	1.048	1.027		.021		.650	.0133
5	1.3366	.851	.055	18.77	11333	1719	1.056	1.011					
4	1.3361	.851	.042	18.58	11379	1732	1.067	1.008					
3	1.3356	.902	.028	18.4	11430	1750	1.018	1.004					
2	1.3349	.901	.011	18.14	11493	1775	1.031	1.001					
1	1.3349	.929	.01	18.13	11498	1776	1	1	.04		.634		.061

Run	n_D^{25}	wt. % of Salt	x	M	H	h	$\frac{L_n}{L_B}$	$\frac{(H-h)_B}{(H-h)_n}$	x_m	$\frac{E}{L_B}$	L_B	V	E
T.6	1.3380	.896	.089	19.24	11207	1690	1.051	1.022		.029		.647	.0283
5	1.3370	.905	.063	18.89	11300	1711	1.041	1.014					
4	1.3362	.935	.042	18.57	11380	1733	1.044	1.008					
3	1.3358	.979	.031	18.42	11420	1745	1.006	1.005					
2	1.3351	.994	.015	18.20	11482	1770	1.003	1.001					
1	1.3350	.999	.012	18.15	11490	1774	1	1	.038		.633		
U.6	1.3350	.908	.014	18.19	11483	1771	1.072	1.001		.071		.635	.045
5	1.3350	.939	.013	18.18	11487	1772	1.038	1.001					
4	1.33495	.941	.012	18.15	11490	1774	1.037	1.001					
3	1.3349	.983	.01	18.13	11498	1776	0.995	1.001					
2	1.3350	.99	.0125	18.17	11489	1774	0.985	1.001					
1	1.33485	.978	.009	18.11	11500	1780	1	1	.011		.634		
V.6	1.3350	.955	.014	18.18	11486	1772	1.088	1.001		.087		.637	.0554
5	1.3350	.973	.013	18.17	11489	1772	1.069	1.001					
4	1.33495	.999	.011	18.14	11494	1775	1.041	1.001					
3	1.3349	1.023	.008	18.11	11505	1781	1.018	1.001					
2	1.33485	1.044	.006	18.07	11513	1784	1.002	1.0					
1	1.33485	1.046	.0055	18.07	11513	1784	1	1	.009		.637		
W.6	1.3348	.937	.0095	18.12	11500	1778	1.056	1.001		.055		.636	.035
5	1.3348	.963	.009	18.11	11502	1780	1.03	1.001					
4	1.3348	.965	.009	18.11	11502	1780	1.024	1.001					
3	—	—	—	—	—	—	—	—					
2	1.3348	.987	.0075	18.10	11508	1782	1.003	1.001					
1	1.33475	.994	.0065	18.08	11510	1784	1		.008		.635		

TABLE 6.3.CYCLOHEXANE-TOLUENE SYSTEM.

Measurements of the number of transfer units in the pear column and the temperatures of the bulk phases with the shielded thermocouple at position (6).

Run	x	Temperature °C.		x_m	N_{OG}
		Vapour	Liquid		
A.6	.985	81.05	80.9		
5	.985	81	81.2		
4	.985	81	-		
3	.984	81	-		
2	.98	80.98	80.98		
1	.98	<u>81.1</u>		.982	.489
B.6	.981	80.35	80.2		
5	.975	80.35	80.55		
4	.975	80.5	-		
3	.975	80.45	-		
2	.974	80.4	80.2		
1	.972	<u>80.95</u>		.976	.744
C.6	.971	80.85	80.6		
5	.971	80.9	81.1		
4	.963	81.2	-		
3	.957	81.2	-		
2	.957	80.75	81.35		
1	.953	<u>81.35</u>		.96	.89
D.6	.954	81.85	81.75		
5	.944	82	81.98		
4	.936	81.9	-		
3	.935	82.35	-		
2	.935	82.4	81.68		
1	.928	<u>82.95</u>		.941	.925
E.6	.941	81.9	81.5		
5	.924	82.2	81.45		
4	.915	82.0	-		
3	.905	82.6	-		
2	.906	82.8	81.85		
1	.893	<u>83.25</u>		.911	1.2

TABLE 6.3 (continued)

Temperature °C.

Run	x	Vapour	Liquid	x_m	N_{OG}
F.6	.916	82.78	82.05		
5	.903	82.78	82.3		
4	.880	82.98			
3	.876	83.75			
2	.875	84	82.7		
1	.854	<u>84.65</u>		.891	1.12
G.6	.905	82.9	82.22		
5	.886	83.25	82.6		
4	.866	82.8			
3	.856	84			
2	.854	84.35	82.7		
1	.837	<u>85.05</u>		.87	1.12
H.6	.891	83.1	81.85		
5	.875	83	81.9		
4	.850	83			
3	.836	84.25			
2	.836	84.8	83		
1	.810	<u>85.65</u>		.842	1.15
I.6	.881	83.55	82.55		
5	.867	83.45	82.62		
4	.826	83.78			
3	.811	85.28			
2	.810	85.85	83.8		
1	.783	<u>86.95</u>		.823	1.26
J.6	.874	83.9	82.85		
5	.851	84.4	82.7		
4	.811	84.0			
3	.795	86.2			
2	.790	86.5	84.3		
1	.756	<u>87.8</u>		.809	1.35
K.6	.855	83.9	82.55		
5	.835	84.1	82.7		
4	.793	84.15			
3	.781	85.8			
2	.771	87.2	84.28		
1	.739	<u>88.3</u>		.793	1.23

TABLE 6.3 (continued)

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Run	x	Temperature °C.		x_m	N_{OG}
		Vapour	Liquid		
L.6	.821	85.35	83.98		
5	.803	85.48	83.8		
4	.746	85.3			
3	.720	87.6			
2	.720	89.18	85.9		
1	.666	<u>90.7</u>		.743	1.34
M.6	.772	86.0	84.25		
5	.743	86.0	83.9		
4	.686	86.3			
3	.655	89.3			
2	.635	90.55	87.22		
1	.588	<u>93.1</u>		.677	1.31
N.6	.726	87.18	85.38		
5	.699	87.3	85.2		
4	.621	87.65			
3	.604	91.6			
2	.571	93.55	88.6		
1	.518	<u>95.18</u>		.617	1.28
O.6	.664	88.9	86.8		
5	.637	88.45	85.6		
4	.547	90.0			
3	.517	94.38			
2	.489	96.0	90.8		
1	.439	<u>98.25</u>		.545	1.19
P.6	.591	92.7	91.3		
5	.533	94.3	90.7		
4	.464	91.7			
3	.435	97.3			
2	.389	98.8	93.3		
1	.348	<u>100.95</u>		.458	1.16
Q.6	.540	92.0	88.45		
5	.498	93.5	89.5		
4	.410	93.7			
3	.356	98.6			
2	.347	99.9	94.8		
1	.289	102.15		.402	1.15

TABLE 6.3 (continued)

195.

Run	x	Vapour	Liquid	x_m	N_{OG}
R.6	.497	93.1	89.45		
5	.489	93.2	89.3		
4	.387	94.55			
3	.352	99.55			
2	.303	101.0	95.75		
1	.266	103.2		.369	1.03
S.6	.418	98.9	97.8		
5	.353	100.55	96.8		
4	.280	98.0			
3	.266	102.8			
2	.223	104.2	99.4		
1	.203	<u>105.38</u>		.297	0.985
T.6	.341	98.0	94.55		
5	.329	98.3	92.8		
4	.229	100.0			
3	.198	104.55			
2	.183	105.45	101.55		
1	.148	<u>106.8</u>		.231	0.93
U.6	.286	102.8	102.1		
5	.234	104.1	101.4		
4	.174	102.9			
3	.171	106.15			
2	.140	107.1	103.32		
1	.120	<u>107.9</u>		.187	0.875
V.6	.224	104.48	104.05		
5	.173	105.6	103.4		
4	.124	104.5			
3	.123	106.8			
2	.098	107.75	104.9		
1	.087	<u>108.85</u>		.144	0.795
W.6	.186	106	105.45		
5	.137	107.2	105.2		
4	.105	105.65			
3	.103	108.15			
2	.086	108.6	106.1		
1	.070	109.1		.119	0.76
X.6	.070	108.8	108.8		
5	.050	108.8	108.6		
4	.039	109.5			
3	.037	109.7			
2	.035	109.8	109.1		
1	.029	110.25		.046	0.54

TABLE 6.4.

CALCULATED VALUES OF dy/dz AT DIFFERENT COMPOSITIONS
FROM EXPERIMENTAL RESULTS IN THE PEAR COLUMN.

A. METHANOL-WATER		B. CYCLOHEXANE-TOLUENE.	
x	dy/dz	x	dy/dz
.05	.00385	.05	.00265
.1	.0121	.1	.00315
.15	.0199	.15	.0067
.2	.0217	.2	.00800
.25	.0245	.25	.00952
.3	.0266	.3	.0103
.35	.02528	.35	.01035
.4	.02655	.4	.01055
.45	.02465	.45	.01082
.5	.0226	.5	.0096
.55	.023	.55	.00908
.6	.01938	.6	.00885
.65	.0186	.65	.00855
.7	.016	.7	.00758
.75	.0146	.75	.00653
.8	.01282	.8	.00552
.85	.00865	.85	.00322
.9	.00628	.9	.00128
.95	.00315	.95	.0007

TABLE 6.5. METHANOL - WATER SYSTEM.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
y _e	0	.27	.42	.515	.58	.665	.729	.78	.824	.87	.915	.937	.957	.98	1.0
x _e	0	.007	.015	.023	.033	.058	.092	.14	.22	.352	.545	.657	.758	.881	1.0
y _e -y	0	.22	.32	.365	.38	.365	.328	.28	.224	.17	.115	.087	.058	.03	0
t _v	100	98.7	97.3	95.9	94.5	91.4	88.3	85.0	80.8	76.5	72.4	70.2	68.1	66.0	64.5
t _L	100	92.2	87.6	84.3	81.7	78	75.3	73.1	71.2	69.3	67.6	66.8	66.0	65.1	64.5
t _v -t _L	0	6.55	9.68	11.6	12.8	13.4	12.9	11.9	9.6	7.22	4.8	3.42	2.1	.9	0
H	11533	11349	11160	10961	10797	10608	10447	10304	10177	10049	9923	9861	9799	9737	9677
h	1797	1724	1682	1653	1627	1574	1528	1483	1440	1400	1359	1339	1318	1296	1271
H _e	11533	10663	10416	10286	10201	10095	10014	9949	9895	9838	9781	9752	9727	9700	9677
h _e	1797	1782	1769	1756	1743	1715	1687	1658	1616	1548	1464	1413	1376	1325	1271
H-h	9736	9625	9478	9308	9170	9034	8919	8821	8737	8649	8564	8522	8481	8441	8406
H _e -h _e	9736	8881	8647	8530	8458	8380	8327	8291	8281	8290	8317	8339	8351	8375	8406
H-H _e	0	686	744	675	596	513	433	355	282	211	142	109	72	37	0
H _e -h	9736	8939	8734	8633	8574	8521	8486	8466	8455	8438	8422	8413	8409	8404	8406
$\frac{1}{H-h} \times 10^4$	1.027	1.039	1.055	1.074	1.091	1.107	1.121	1.134	1.145	1.16	1.168	1.173	1.18	1.185	1.19
$-\frac{dH}{dy}$	4355	3645	3880	3780	2477	1745	1534	1307	1265	1265	1265	1265	1246	1246	1246
$-\frac{dh}{dy}$	1960	985	615	536	542	498	436	435	421	408	411	409	439	449	518
$-\frac{d(H-h)}{dy}$	2395	2660	3265	3244	1935	1247	1098	872	844	857	854	856	807	797	728
$-\frac{1}{H-h} \cdot \frac{d(H-h)}{dy} \cdot 246$	.246	.276	.344	.349	.211	.138	.123	.099	.097	.099	.0997	.1005	.095	.094	.087
y _e -x _e	0	.263	.405	.492	.547	.607	.636	.64	.604	.518	.37	.28	.2	.099	0

TABLE 6.5. (continued)

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
$\gamma \times 10^4$	0	.296	.469	.577	.647	.725	.764	.772	.73	.625	.445	.336	.24	.118	0
$\frac{t_v - t_h}{\gamma_e - \gamma}$	0	29.8	30.2	31.7	33.7	36.8	39.5	42.5	42.8	42.5	41.7	39.3	36.2	30	0
$\gamma(H - H_e) \times 10^2$	0	2.03	3.48	3.9	3.86	3.72	3.31	2.74	2.06	1.32	.632	.366	.172	.044	0
$\gamma(H_e - h)$	0	.265	.409	.498	.555	.618	.649	.654	.618	.527	.375	.283	.202	.099	0
$-\gamma \frac{dH}{dy} \times 10$	0	1.08	1.78	2.18	1.60	1.26	1.17	1.01	.923	.791	.563	.425	.299	.147	0
$-\gamma \frac{dh}{dy} \times 10$	0	.292	.288	.31	.351	.361	.333	.336	.307	.255	.183	.137	.105	.053	0
$\gamma \frac{t_v - t_h}{\gamma_e - \gamma} \times 10^4$	0	8.82	14.18	18.3	21.8	26.65	30.18	32.82	31.28	26.55	18.58	13.21	8.67	3.55	0
$1 - \gamma \frac{dH}{dy}$	1	1.11	1.18	1.22	1.16	1.13	1.12	1.10	1.09	1.08	1.06	1.04	1.03	1.01	1
$1 - \gamma \frac{dh}{dy}$	1	1.029	1.029	1.031	1.035	1.036	1.034	1.034	1.031	1.026	1.018	1.014	1.011	1.005	1
$[(\gamma_e - \gamma) - \gamma(H_e - H)]$	0	.240	.355	.404	.419	.402	.361	.307	.245	.183	.121	.091	.06	.030	0
$\frac{1}{\gamma} [(\gamma_e - \gamma) - \gamma(H_e - H)] \frac{dV}{dy}$	0	.066	.122	.141	.088	.056	.044	.030	.024	.018	.012	.009	.006	.003	0
$-(\gamma_e - \gamma) - \gamma(H_e - h)$	0	.045	.089	.133	.175	.253	.321	.375	.394	.357	.26	.196	.144	.069	0
$-\frac{[(\gamma_e - \gamma) - \gamma(H_e - h)]}{\gamma_e - \gamma} \frac{dV}{dz}$	0	.202	.277	.365	.46	.692	.977	1.336	1.758	2.1	2.26	2.252	2.475	2.313	0
$\frac{dV}{dz}$	0	.0042	.0127	.0198	.0229	.0252	.0251	.0235	.0201	.0163	.0117	.0091	.0061	.0032	0
$\frac{dV}{dy}$	.54	.418	.18	-.015	-.243	-.605	-.645	-.618	-.555	-.5	-.413	-.37	-.325	---	---
$\frac{dE}{dy}$	.01	.0235	.0395	.063	.079	.109	.118	.051	-.015	-.088	-.138	-.185	-.194	---	---
$\frac{dV}{dz} \times 10^2$	0	.1756	.228	-.297	-.556	-1.52	-1.62	-1.45	-1.12	-.815	-.483	-.336	-.198	---	---
$\frac{dE}{dz} \times 10^2$	0	.0098	.05	.125	.181	.274	.296	.12	-.030	-.144	-.161	-.168	-.118	---	---
$\frac{dL}{dz} \times 10^2$	0	.185	.278	-.172	-.375	-1.25	-1.32	-1.33	-1.15	-.959	-.644	-.504	-.316	---	---
$-\frac{[(\gamma_e - \gamma) - \gamma(H_e - h)]}{(\gamma_e - \gamma)} \frac{dL}{dz} \times 10^2$	0	.037	.077	-.063	-.17	-.86	-1.29	-1.78	-2.01	-2.01	-1.45	-1.13	-.77	---	---

TABLE 6.5. (continued)

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
V	.63	.654	.669	.674	.667	.622	.555	.494	.433	.38	.334	.315	.3	---	---
$\frac{V}{L}$	---	.98	.978	.975	.969	.953	.928	.907	.892	.89	.907	.925	.949	---	---

If $N_c = .77$

$\frac{1}{K'}$	52.1	50.25	49.1	48.75	49.25	52.85	59.3	66.5	75.85	86.5	98.4	104.2	109.5	---	---
$\frac{(y_e - y) - r(h - h)}{K'(y_e - y)} \frac{dl}{dz}$	0	.0188	.0378	-.031	-.085	-.456	-.767	-1.18	-1.52	-1.73	-1.43	-1.17	-.84	---	---

From experiment:

N_T	--	.49	.89	1.22	1.46	1.71	1.88	2.14	2.41	2.67	2.78	2.73	2.55	2.11	---
$\frac{N_T}{N_c}$	--	.636	1.157	1.585	1.898	2.22	2.44	2.78	3.13	3.47	3.61	3.55	3.31	2.74	---

TABLE 6.6. METHANOL - WATER SYSTEM.

$\frac{N_T}{N_C}$ being calculated from equation (6.6) at different values of $\frac{U'}{K'}$.

$\frac{U'}{K'}$	$\frac{N_T}{N_C}$	x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
1	1		1	.961	.948	.93	.935	.937	.934	.938	.939	.945	.96	.969	.978	.988	1
10	1		1	.969	.961	.944	.953	.958	.959	.965	.966	.966	.974	.981	.984	.991	1
100	1		1	1.04	1.08	1.1	1.14	1.18	1.21	1.24	1.23	1.19	1.14	1.09	1.06	1.02	1
200	1		1	1.13	1.22	1.27	1.34	1.43	1.5	1.55	1.52	1.44	1.31	1.22	1.15	1.06	1
300	1		1	1.21	1.35	1.44	1.54	1.62	1.78	1.85	1.82	1.69	1.49	1.35	1.23	1.09	1
400	1		1	1.3	1.48	1.61	1.75	1.93	2.12	2.16	2.11	1.94	1.67	1.48	1.31	1.13	1
500	1		1	1.38	1.62	1.78	1.95	2.18	2.34	2.47	2.41	2.20	1.85	1.61	1.4	1.16	1
600	1		1	1.47	1.75	1.95	2.15	2.43	2.62	2.78	2.7	2.45	2.27	1.73	1.48	1.2	1
700	1		1	1.55	1.89	2.12	2.36	2.68	2.9	3.08	2.98	2.69	2.20	1.86	1.57	1.23	1
800	1		1	1.64	2.02	2.29	2.56	2.92	3.18	3.39	3.28	3.14	2.38	1.99	1.65	1.27	1
900	1		1	1.72	2.16	2.46	2.76	3.17	3.46	3.69	3.58	3.2	2.56	2.12	1.72	1.30	1
1000	1		1	1.81	2.29	2.63	2.97	3.42	3.75	3.99	3.87	3.45	2.71	2.25	1.82	1.34	1

TABLE 6,7. METHANOL - WATER SYSTEM.

$\frac{N_T}{N_c}$ being calculated from equation (6.13) at different values of $\frac{U'}{K'}$.

$\frac{U'}{K'}$	$\frac{N_T}{N_c}$	x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
100	1	1.02	1.05	1.15	1.22	1.59	1.86	2.20	2.46	2.6	2.33	2.11	1.81	---	---		
200	1	1.10	1.18	1.32	1.42	1.84	2.13	2.49	2.72	2.84	2.5	2.23	1.9	---	---		
300	1	1.19	1.32	1.5	1.63	2.02	2.4	2.78	2.99	3.07	2.66	2.34	1.97	---	---		
400	1	1.27	1.45	1.67	1.83	2.32	2.72	3.07	3.27	3.3	2.82	2.47	2.06	---	---		
500	1	1.36	1.58	1.84	2.04	2.57	2.94	3.36	3.54	3.53	2.99	2.58	2.14	---	---		
600	1	1.44	1.72	2.01	2.24	2.81	3.21	3.64	3.82	3.76	3.16	2.71	2.22	---	---		
700	1	1.52	1.86	2.18	2.44	3.06	3.48	3.93	4.08	3.99	3.32	2.83	2.31	---	---		
800	1	1.61	1.99	2.36	2.65	3.31	3.75	4.22	4.35	4.21	3.49	2.95	2.38	---	---		
900	1	1.69	2.12	2.53	2.86	3.55	4.02	4.5	4.62	4.45	3.66	3.07	2.45	---	---		
1000	1	1.69	2.26	2.7	3.06	3.79	4.3	4.8	4.89	4.68	3.82	3.19	2.55	---	---		

TABLE 6.8. METHANOL - WATER SYSTEM.

($y^*=y_e$ and $x^*=x_e$)

A. $\frac{N_T}{N_C}$ being calculated from equation (6.10) at $\frac{U'}{K'}=1000$ and $N_C=.77$.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
$Z \frac{dy}{dz}$	---	.106	.322	.501	.579	.638	.635	.595	.509	.413	.296	.231	.154	.081	---
$Z \frac{dy}{dz} \cdot \frac{1}{V} \cdot \frac{dV}{dy}$	---	.029	.111	.175	.122	.088	.078	.059	.049	.041	.03	.023	.015	.008	---
$-\frac{[(y_e-y)-r(H_e-h)]}{(y_e-y)} \cdot \frac{z}{N_C} \cdot \frac{1}{V} \cdot \frac{dV}{dz}$	---	.008	.04	.083	.073	.079	.099	.102	.112	.112	.087	.068	.047	.023	---
$\frac{N_T}{N_C}$	---	1.82	2.31	2.66	2.99	3.46	3.79	4.04	3.89	3.46	2.72	2.22	1.80	1.32	---

B. $\frac{N_T}{N_C}$ being calculated from equation (6.11) at $\frac{U'}{K'}=1000$ and $N_C=.77$.

$-\frac{[(y_e-y)-r(H_e-h)]}{y_e-y} \cdot \frac{0.01}{N_C}$	---	.003	.004	.005	.006	.009	.013	.017	.023	.027	.029	.029	.032	.030	---
$\frac{N_T}{N_C}$	---	1.82	2.31	2.66	2.99	3.46	3.78	4.03	3.87	3.43	2.69	2.19	1.77	1.3	---

C. $\frac{N_T}{N_C}$ being calculated from equation (6.12) at $\frac{U'}{K'}=1000$ and $N_C=.77$.

$\frac{N_T}{N_C}$	---	1.78	2.26	2.59	2.93	3.38	3.7	3.95	3.82	3.4	2.68	2.17	1.77	1.3	---
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TABLE 6.9. METHANOL - WATER SYSTEM.

$\frac{N_T}{N_C}$ being calculated from theoretical equations with Murphree efficiency=75%.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
y*	0	.215	.34	.428	.485	.578	.646	.71	.768	.828	.886	.915	.944	.973	1.0
x*	0	.018	.036	.055	.075	.119	.169	.23	.315	.439	.609	.705	.794	.898	1.0
y*-x*	0	.197	.304	.373	.410	.459	.477	.48	.453	.389	.277	.21	.15	.074	0
y*-y	0	.165	.24	.278	.285	.278	.246	.21	.168	.128	.086	.062	.044	.023	0
H*	11533	10770	10542	10402	10322	10205	10118	10038	9963	9890	9817	9780	9747	9708	9677
h*	1797	1765	1738	1719	1700	1671	1643	1610	1566	1510	1437	1398	1362	1319	1271
H*-h*	9736	9005	8804	8683	8622	8534	8475	8428	8397	8380	8380	8382	8385	8389	8406
H*-h	9736	9046	8860	8749	8695	8631	8590	8555	8523	8490	8458	8441	8429	8412	8406
$\gamma \times 10^4$	0	.183	.345	.43	.476	.538	.563	.57	.54	.464	.331	.251	.179	.089	0
H-H*	0	579	618	559	475	403	329	266	214	159	106	81	52	29	0
$\gamma(H-H^*) \times 10^2$	0	1.06	2.13	2.4	2.26	2.17	1.85	1.52	1.15	.737	.351	.203	.093	.026	0
$\gamma(H^*-h)$	0	.166	.306	.376	.414	.465	.484	.488	.46	.394	.28	.212	.151	.075	0
$\gamma \frac{t_s - t_r}{y_s - y} \times 10^4$	0	5.46	10.44	13.61	16.03	19.79	22.25	23.8	23.12	19.7	13.8	9.86	6.48	2.66	0
$-\gamma \frac{dh}{dy} \times 10$	0	.668	1.34	1.62	1.18	.939	.864	.745	.682	.586	.419	.317	.223	.110	0
$-\gamma \frac{dh}{dy} \times 10$	0	.18	.212	.23	.258	.268	.246	.248	.228	.189	.136	.103	.079	---	---

TABLE 6.9. (continued)

A. Only evaporation being considered and $\frac{U'}{K'}=1000$.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
$[(y^*-y)-\gamma(H^*-H)]$	0	.186	.284	.349	.388	.437	.459	.465	.442	.381	.274	.208	.149	.074	0
$\frac{1}{V}[(y^*-y)-\gamma(H^*-H)]\frac{dV}{dy}$	0	.052	.098	.122	.082	.060	.056	.046	.043	.038	.027	.021	.014	.007	0
$\frac{N_T}{N_c}$	1	1.52	1.97	2.27	2.51	2.88	3.13	3.29	3.23	2.91	2.35	1.96	1.63	1.26	1

B. Both evaporation and entrainment being considered and $\frac{U'}{K'}=1000$.
(Entrainment is assumed to vary with composition.)

$-(y^*-y)-\gamma(H^*-h)$	0	.001	.066	.098	.129	.187	.238	.278	.292	.266	.194	.15	.107	.052	0
$-\frac{(y^*-y)-\gamma(H^*-h)}{y_e-y}\frac{dL}{dz}$	0	$.6 \times 10^{-5}$	$.6 \times 10^{-3}$	$.3 \times 10^{-3}$	-.001	-.006	-.010	-.013	-.015	-.015	-.011	$-.87 \times 10^{-2}$	$-.59 \times 10^{-2}$	---	---
$-\frac{(y^*-y)-\gamma(H^*-h)}{K'(y_e-y)}\frac{dL}{dz}$	0	.003	.0295	-.015	-.064	-.338	-.569	-.878	-1.13	-1.29	-1.06	-.91	-.65	---	---
$\frac{N_T}{N_c}$	1	1.49	1.923	2.26	2.52	3.08	3.43	3.77	3.87	3.73	3.08	2.65	2.16	---	---

C. Both evaporation and entrainment being considered in the absence of contact distillation. N_T being calculated at different values of U' .

$U'=10; N_T$	0	.222	.419	.589	.801	1.32	1.75	2.22	2.47	2.45	1.88	.854	.607	---	---
$U'=14; N_T$	0	.311	.597	.831	1.09	1.72	2.25	2.8	3.07	3.01	2.31	1.22	.83	---	---
$U'=18; N_T$	0	.40	.773	1.07	1.39	2.12	2.74	3.35	3.68	3.57	2.73	1.54	1.05	---	---

TABLE 6.10. METHANOL - WATER SYSTEM.

$\frac{N_T}{N_C}$ being calculated from theoretical equations with Murphree efficiency=50%.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
y*	0	.16	.26	.333	.39	.483	.564	.64	.712	.785	.858	.894	.929	.965	1.0
x*	0	.029	.058	.087	.117	.179	.246	.32	.41	.526	.673	.754	.829	.916	1.0
y*-x*	0	.131	.203	.246	.274	.304	.318	.32	.302	.259	.185	.14	.1	.05	0
y*-y	0	.11	.16	.183	.19	.183	.164	.14	.112	.085	.058	.044	.029	.015	0
H*	11533	10927	10682	10553	10461	10326	10222	10127	10036	9943	9851	9808	9762	9718	9677
h*	1797	1749	1716	1691	1671	1638	1602	1568	1523	1471	1411	1379	1348	1310	1271
H*-h*	9736	9176	8976	8862	8790	8688	8620	8559	8513	8472	8440	8429	8414	8408	8406
H*-h	9736	9203	9000	8900	8834	8752	8694	8644	8596	8543	8492	8469	8444	8422	8406
$\gamma \times 10^4$	0	.143	.226	.278	.312	.35	.369	.374	.355	.306	.219	.166	.119	.059	0
H-H*	0	422	478	408	336	282	225	177	141	106	72	53	37	19	0
$\gamma(H-H^*) \times 10^2$	0	.605	1.08	1.13	1.05	.986	.829	.662	.5	.324	.158	.088	.044	.011	0
$\gamma(H^*-h)$	0	.1317	.204	.248	.276	.306	.321	.324	.305	.262	.186	.141	.101	.05	0
$\gamma \frac{t_y - t_x}{y_c - y} \times 10^4$	0	4.27	6.82	8.8	10.49	12.86	14.59	15.89	15.21	12.98	9.15	6.53	4.3	1.77	0
$-\gamma \frac{dh}{dy} \times 10$	0	.523	.876	1.049	.771	.61	.566	.489	.449	.387	.277	.21	.148	.073	0
$-\gamma \frac{dh}{dy} \times 10$	0	.141	.139	.149	.169	.174	.161	.163	.149	.125	.09	.057	.052	.027	0

TABLE 6.10. (continued)

A. Only evaporation being considered and $\frac{U'}{K'}=1000$.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
$[(y^*-y)-r(H^*-h)]$	0	.116	.171	.194	.201	.192	.172	.147	.117	.088	.059	.052	.033	.016	0
$\frac{1}{V}[(y^*-y)-r(H^*-h)]\frac{dV}{dy}$	0	.032	.059	.068	.423	.027	.021	.014	.011	.009	.006	.005	.003	.002	0
$\frac{N_T}{N_c}$	1	1.4	1.63	1.81	1.98	2.21	2.38	2.51	2.44	2.23	1.87	1.63	1.41	1.17	1

B. Both evaporation and entrainment being considered and $\frac{U'}{K'}=1000$.

(Entrainment is assumed to vary with composition.)

$-(y^*-y)-r(H^*-h)$	0	.022	.044	.065	.086	.123	.157	.184	.193	.177	.128	.097	.071	.035	0
$\frac{[(y^*-y)-r(H^*-h)]dL}{(y_e-y)}dz$	0	.0183	.0378	-.030	-.084	-.421	-.634	-.876	-.987	-.998	-.716	-.562	-.387	--	--
$\frac{[(y^*-y)-r(H^*-h)]dL}{K'(y_e-y)}dz$	0	.0092	.0186	-.015	-.042	-.222	-.376	-.583	-.749	-.863	-.704	-.586	-.424	--	--
$\frac{N_T}{N_c}$	1	1.37	1.60	1.82	1.99	2.35	2.59	2.83	2.87	2.78	2.36	2.06	1.75	--	--

TABLE 6.11. METHANOL - WATER SYSTEM.

$\frac{N_T}{N_C}$ being calculated from theoretical equations with Murphree efficiency=25%.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
y*	0	.105	.18	.241	.295	.391	.482	.57	.656	.743	.829	.872	.915	.958	1.0
x*	0	.039	.079	.118	.158	.24	.323	.41	.505	.613	.736	.802	.865	.933	1.0
y*-x*	0	.066	.101	.123	.137	.152	.159	.16	.151	.13	.093	.07	.05	.025	0
y*-y	0	.055	.08	.091	.095	.091	.082	.07	.056	.043	.029	.022	.015	.008	0
H*	11533	11139	10859	10716	10619	10459	10326	10215	10104	9998	9890	9834	9781	9728	9677
h*	1797	1735	1698	1671	1648	1605	1562	1523	1481	1435	1385	1358	1333	1303	1271
H*-h*	9736	9404	9161	9045	8971	8854	8764	8692	8623	8563	8505	8476	8448	8425	8406
H*-h	9736	9415	9177	9063	8992	8885	8798	8732	8664	8598	8531	8495	8463	8432	8406
$Y \times 10^4$	0	.07	.110	.136	.152	.171	.181	.184	.175	.151	.109	.083	.059	.029	0
H-H*	0	210	301	245	178	149	121	89	73	51	33	27	18	9	0
$\gamma(H-H^*) \times 10^2$	0	.147	.332	.333	.271	.256	.219	.166	.128	.077	.036	.022	.011	.003	0
$\gamma(H^*-h)$	0	.066	.101	.123	.137	.152	.159	.161	.152	.13	.093	.071	.05	.025	0
$\gamma \frac{t_v - t_h}{y_e - y} \times 10^4$	0	2.08	3.34	4.31	5.14	6.31	7.16	7.82	7.5	6.43	4.53	3.25	2.14	.88	0
$-\gamma \frac{dH}{dy} \times 10$	0	.255	.428	.514	.377	.299	.278	.241	.222	.191	.137	.104	.074	.037	0
$-\gamma \frac{dh}{dy} \times 10$	0	.069	.068	.073	.082	.085	.079	.08	.075	.062	.045	.034	.026	.013	0

TABLE 6.11. (continued)

A. Only evaporation being considered and $\frac{U'}{K'}=1000$.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
$(y^*-y)-\gamma(H^*-H)$	0	.057	.083	.095	.098	.094	.084	.072	.057	.044	.029	.022	.015	.008	0
$\frac{1}{\gamma}[(y^*-y)-\gamma(H^*-H)]\frac{d\gamma}{dy}$	0	.016	.029	.033	.021	.013	.010	.007	.006	.004	.003	.002	.0012	.0007	0
$\frac{N_T}{N_c}$	1	1.2	1.32	1.41	1.49	1.60	1.69	1.75	1.72	1.62	1.44	1.31	1.21	1.08	1

B. Both evaporation and entrainment being considered and $\frac{U'}{K'}=1000$.

(Entrainment is assumed to vary with composition.)

$-(y^*-y)-\gamma(H^*-h)$	0	.011	.021	.032	.042	.061	.077	.091	.096	.087	.064	.049	.035	.017	0
$-\frac{[(y^*-y)-\gamma(H^*-h)]}{(y_e-y)}\frac{dL}{dz} \times 10^3$	0	.0092	.0182	-.016	-.041	-.209	-.311	-.432	-.49	-.489	-.358	-.281	-.191	--	--
$-\frac{[(y^*-y)-\gamma(H^*-h)]}{K'(y_e-y)}\frac{dL}{dz}$	0	.0046	.009	-.007	-.020	-.110	-.184	-.287	-.372	-.423	-.352	-.293	-.209	--	--
$\frac{N_T}{N_c}$	1	1.17	1.29	1.39	1.48	1.65	1.75	1.87	1.88	1.83	1.63	1.49	1.35	--	--

TABLE 6.12. METHANOL - WATER SYSTEM.

$$(y^*=y_e \text{ and } x^*=x_e)$$

Number of transfer units in the absence of contact distillation at different values of U' . Both evaporation and entrainment being considered.

	x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
	L	.639	.611	.583	.557	.533	.488	.448	.412	.381	.352	.326	.315	.303	.292	.282
	$\frac{Z}{L}$	39.6	41.5	43.4	45.3	47.5	51.9	56.5	61.4	66.5	72	77.6	80.4	83.5	86.6	89.7
$U'=8;$	N_T	0	.269	.446	.67	.79	1.50	2.03	2.62	2.91	2.9	2.24	1.73	1.21	--	--
$U'=10;$	N_T	0	.34	.566	.83	1.08	1.77	2.36	3.01	3.31	3.32	2.52	1.94	1.35	--	--
$U'=12;$	N_T	0	.411	.686	.991	1.28	2.04	2.69	3.39	3.72	3.65	2.81	2.15	1.49	--	--
$U'=14;$	N_T	0	.481	.806	1.15	1.48	2.31	3.02	3.79	4.12	4.08	3.09	2.36	1.64	--	--

TABLE 6.13. CYCLOHEXANE - TOLUENE SYSTEM.

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
y _e	0	.132	.236	.324	.399	.523	.621	.701	.773	.835	.894	.922	.949	.975	1.0
x _e	0	.016	.036	.058	.082	.136	.201	.28	.377	.499	.642	.725	.81	.901	1.0
y _e -x _e	0	.116	.2	.266	.317	.387	.42	.421	.396	.336	.252	.197	.139	.074	0
y _e -y	0	.082	.136	.174	.199	.223	.221	.201	.173	.135	.094	.072	.049	.025	0
t _v	110.6	109.5	108.3	107.1	105.8	103.1	100.1	97.05	93.8	90.6	87.2	85.5	83.8	82.2	80.7
t _L	110.6	107.6	104.9	102.5	100.1	96.4	93.3	90.5	88.1	85.9	84	83.1	82.2	81.5	80.7
t _v -t _L	0	1.85	3.4	4.62	5.68	6.65	6.82	6.55	5.75	4.63	3.18	2.33	1.53	.77	0
H	14155	13780	13453	13147	12865	12355	11923	11530	11171	10850	10533	10380	10238	10116	9973
h	5070	4835	4638	4452	4290	3990	3758	3552	3363	3205	3058	2990	2923	2868	2808
H _e	14155	13255	12670	12280	11940	11455	11100	10835	10615	10425	10260	10180	10110	10035	9973
h _e	5070	4998	4905	4810	4710	4498	4280	4052	3812	3560	3300	3170	3048	2928	2808
H-h	9085	8945	8815	8695	8575	8365	8165	7978	7808	7645	7475	7390	7315	7248	7165
H _e -h _e	9085	8257	7765	7470	7230	6957	6820	6783	6803	6865	6960	7010	7062	7107	7165
H-H _e	0	525	783	867	925	900	823	695	556	425	273	200	128	81	0
$\frac{1}{H-h} \times 10^3$	1.10	1.119	1.134	1.151	1.167	1.196	1.224	1.253	1.281	1.309	1.339	1.353	1.368	1.380	1.39
$-\frac{dh}{dy}$	7850	7100	6450	5900	5380	4600	4100	3750	3505	3180	2970	2880	2790	2680	2420
$-\frac{dh}{dy}$	4586	4288	3910	3504	3134	2560	2182	1935	1758	1547	1340	1332	1266	1240	1160
$-\frac{d(H-h)}{dy}$	3264	2812	2540	2396	2246	2040	1918	1815	1647	1633	1630	1548	1524	1440	1260
$-\frac{1}{H-h} \frac{d(H-h)}{dy}$	.359	.315	.288	.276	.262	.244	.235	.228	.211	.214	.218	.21	.209	.199	.176
$\gamma \times 10^4$	0	.14	.258	.356	.439	.556	.616	.621	.582	.489	.362	.281	.197	.104	0

TABLE 6.13. (continued)

x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
$\frac{t_y - t_c}{y_c - y}$	0	22.55	25.0	26.58	28.55	29.82	30.88	32.6	33.25	34.3	33.8	32.38	31.22	30.8	0
$\gamma \frac{t_y - t_c}{y_c - y} \times 10^4$	0	3.17	6.44	9.46	12.51	16.59	19.02	20.25	19.37	16.78	12.23	8.77	6.15	3.21	0
$\gamma(H - H_c) \times 10^2$	0	.74	2.02	3.09	4.06	5.01	5.07	4.32	3.24	2.08	.99	.56	.252	.084	0
$-\gamma \frac{dH}{dy}$	0	.0997	.166	.21	.236	.256	.253	.233	.204	.155	.108	.081	.055	.028	0
$[(y_c - y) - \gamma(H_c - H)]$	0	.089	.156	.205	.24	.273	.272	.244	.205	.156	.104	.078	.052	.026	0
$\frac{1}{\gamma} [(y_c - y) - \gamma(H_c - H)] \frac{d\gamma}{dy}$	0	.028	.045	.057	.063	.067	.064	.056	.043	.033	.023	.016	.011	.005	0

From experiment:

N_T	--	.547	.692	.797	.882	1.018	1.123	1.201	1.245	1.251	1.208	1.142	1.022	.847	.611
If $N_c = .64$, then															
$\frac{N_T}{N_c}$	--	.855	1.08	1.25	1.38	1.59	1.76	1.88	1.95	1.96	1.89	1.79	1.6	1.32	.96

TABLE 6.14. CYCLOHEXANE - TOLUENE SYSTEM.

$\frac{N_T}{N_c}$ being calculated from equation (6.6) at different values of $\frac{U'}{K'}$.

$\frac{U'}{K'}$	$\frac{N_T}{N_c}$	x	0	.05	.1	.15	.2	.3	.4	.5	.6	.7	.8	.85	.9	.95	1.0
10	1		.935	.898	.875	.864	.855	.858	.867	.878	.906	.933	.947	.964	.981	1	
100	1		.963	.949	.949	.959	.981	1.00	1.02	1.03	1.04	1.04	1.02	1.02	1.01	1	
200	1		.992	1.01	1.03	1.07	1.12	1.16	1.20	1.19	1.19	1.15	1.10	1.01	1.04	1	
300	1		1.02	1.06	1.11	1.17	1.26	1.32	1.37	1.36	1.34	1.26	1.19	1.14	1.07	1	
400	1		1.05	1.12	1.19	1.28	1.4	1.48	1.54	1.53	1.49	1.37	1.27	1.19	1.10	1	
500	1		1.08	1.18	1.28	1.39	1.54	1.64	1.71	1.7	1.64	1.49	1.35	1.25	1.13	1	
600	1		1.11	1.24	1.36	1.49	1.68	1.80	1.88	1.86	1.79	1.6	1.43	1.31	1.17	1	
700	1		1.14	1.29	1.44	1.6	1.82	1.96	2.06	2.03	1.94	1.71	1.52	1.37	1.2	1	
800	1		1.17	1.35	1.52	1.71	1.96	2.12	2.23	2.2	2.09	1.82	1.6	1.43	1.23	1	
900	1		1.2	1.41	1.60	1.81	2.1	2.28	2.4	2.37	2.24	1.94	1.68	1.49	1.26	1	
1000	1		1.23	1.47	1.69	1.92	2.24	2.44	2.57	2.53	2.39	2.05	1.76	1.55	1.29	1	

APPENDIX III.NOMENCLATURE.

Symbols are defined in the text after the first equation in which they appear. Symbols which occur only more than once are given in the appendix.

<u>Symbol</u>	<u>Definition.</u>
C_A	Molar concentration of diffusing component A.
D	Diffusivity.
E	Molal flow rate of entrainment per unit time in bulk vapour.
G, V	Molal vapour rate of flow.
H	Molar total heat of vapour.
H^*	Molar total heat of evaporated liquid.
H_G	Height of transfer unit, gas phase.
H_{OG}	Height of transfer unit, overall gas.
H_L	Height of transfer unit, liquid phase.
H_{OL}	Height of transfer unit, overall liquid.
K_G	Overall film coefficient of mass transfer, gas phase.
K_L	Overall film coefficient of mass transfer, liquid phase.
L	Molal liquid rate of flow.
M	Molecular weight.
dM_c	Mass transfer of m.v.c. by contact process at a differential height dz .

N_g	Rate of transfer per unit time in gas phase.
N_l	Rate of transfer per unit time in liquid phase.
dN_L	Moles of liquid evaporated in bulk liquid.
dN_V	Moles of vapour condensed in bulk vapour.
N_{OG}	Overall number of transfer unit, gas phase.
N_{OL}	Overall number of transfer unit, liquid phase.
N_c	Number of transfer unit by contact distillation.
N_T	Number of transfer unit by thermal distillation.
P	Total pressure.
dQ	Heat transfer by contact process in a differential height of column, dz .
R	Gas constant.
Re	Reynolds number.
S	Cross-sectional area of column.
Sc	Schmidt number.
T	Temperature.
U	Overall heat transfer coefficient.
Z	Total height of column from base.
a	Interfacial area between vapour and liquid per unit volume of packing.
d	Representative linear dimension.
h	Molar total heat of liquid.
h^*	Molar total heat of condensed vapour.
k_g	Film coefficient of mass transfer, gas phase.
k_l	Film coefficient of mass transfer, liquid phase.

l_f	Distance in the direction of diffusion.
m	Slope of equilibrium curve - point value.
p	Partial pressure of m.v.c.
t	Temperature.
t_v, t_L	Saturation temperatures of bulk vapour and bulk liquid respectively.
x	Mole fraction of m.v.c. in liquid.
x^*	Mole fraction of m.v.c. of condensed vapour.
y	Mole fraction of m.v.c. in vapour.
y^*	Mole fraction of m.v.c. of evaporated liquid.
α	Relative volatility.
$\gamma = \frac{y^* - x^*}{H^* - h^*}$	
λ	Molar latent heat of vaporisation.
μ	Viscosity.
ρ	Density.

Subscripts.

A,B	Components A or B.
B	Column bottom.
E	Entrainment.
G,V,g	Vapour phase.
L,l	Liquid phase.
O	Overall values.
T	Column top
e	Equilibrium value.
f	Film value.
i	Interface values.
m	Log mean values.

APPENDIX IV.LITERATURE REFERENCES.

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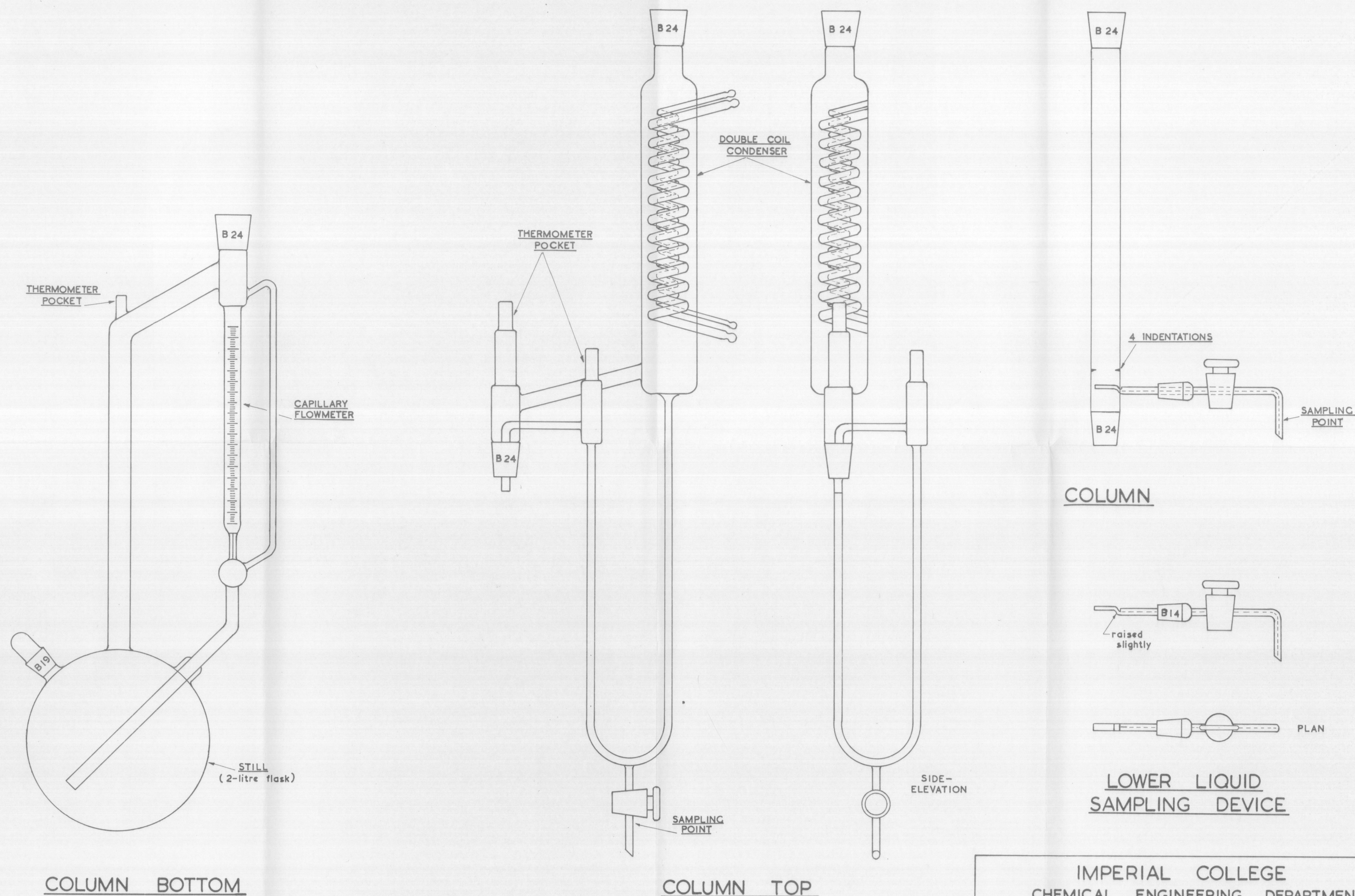
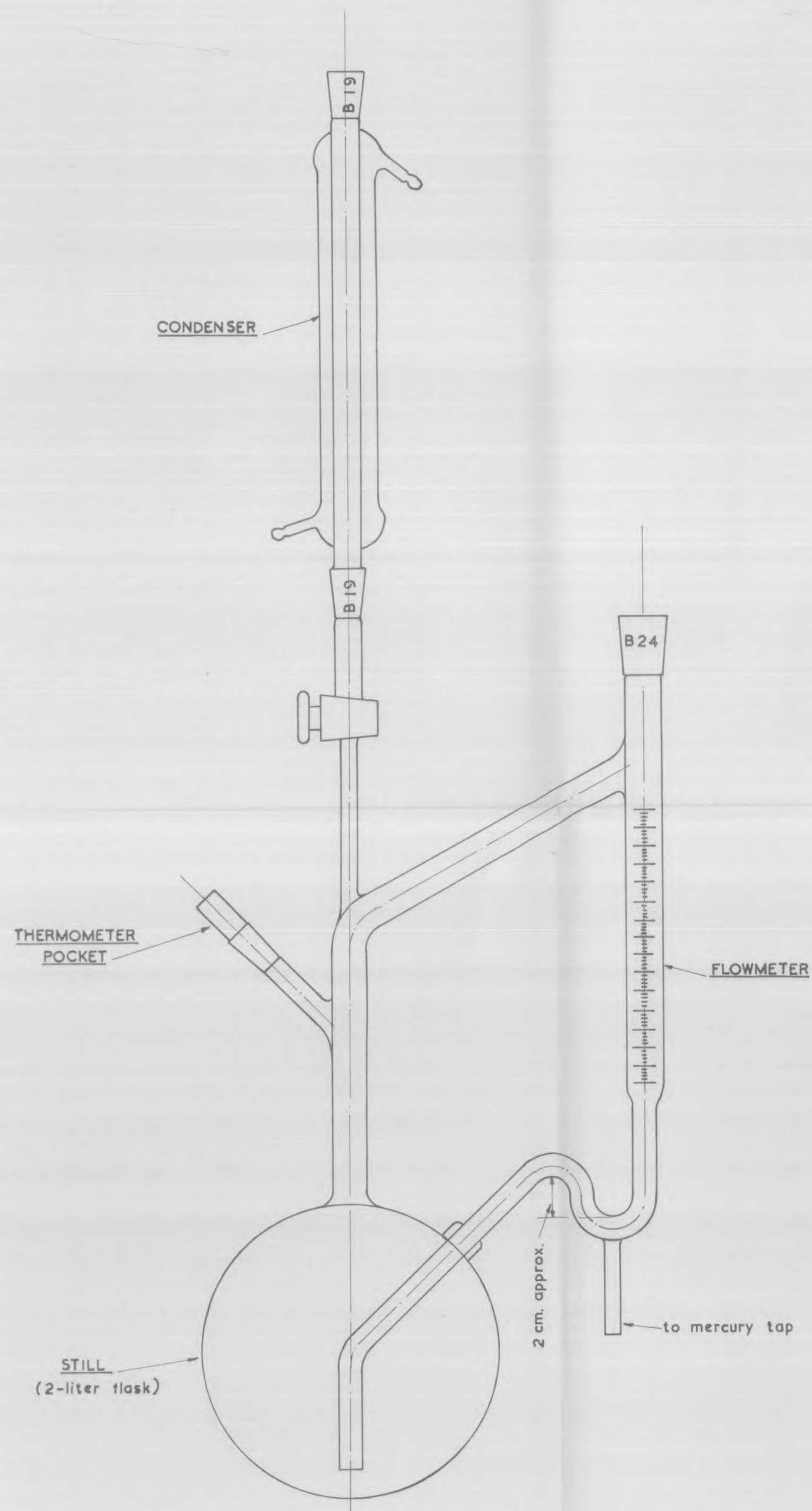


FIG. 2.1

IMPERIAL COLLEGE
CHEMICAL ENGINEERING DEPARTMENT
DISTILLATION UNIT
PACKED COLUMN

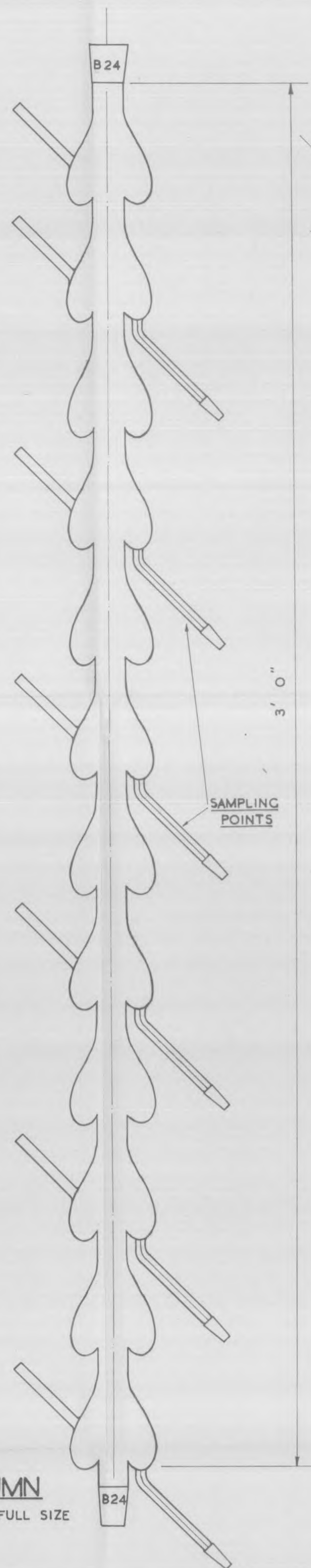
SCALE: HALF FULL SIZE DRAWN BY: S.Y.L. DATE: 26-8-1960



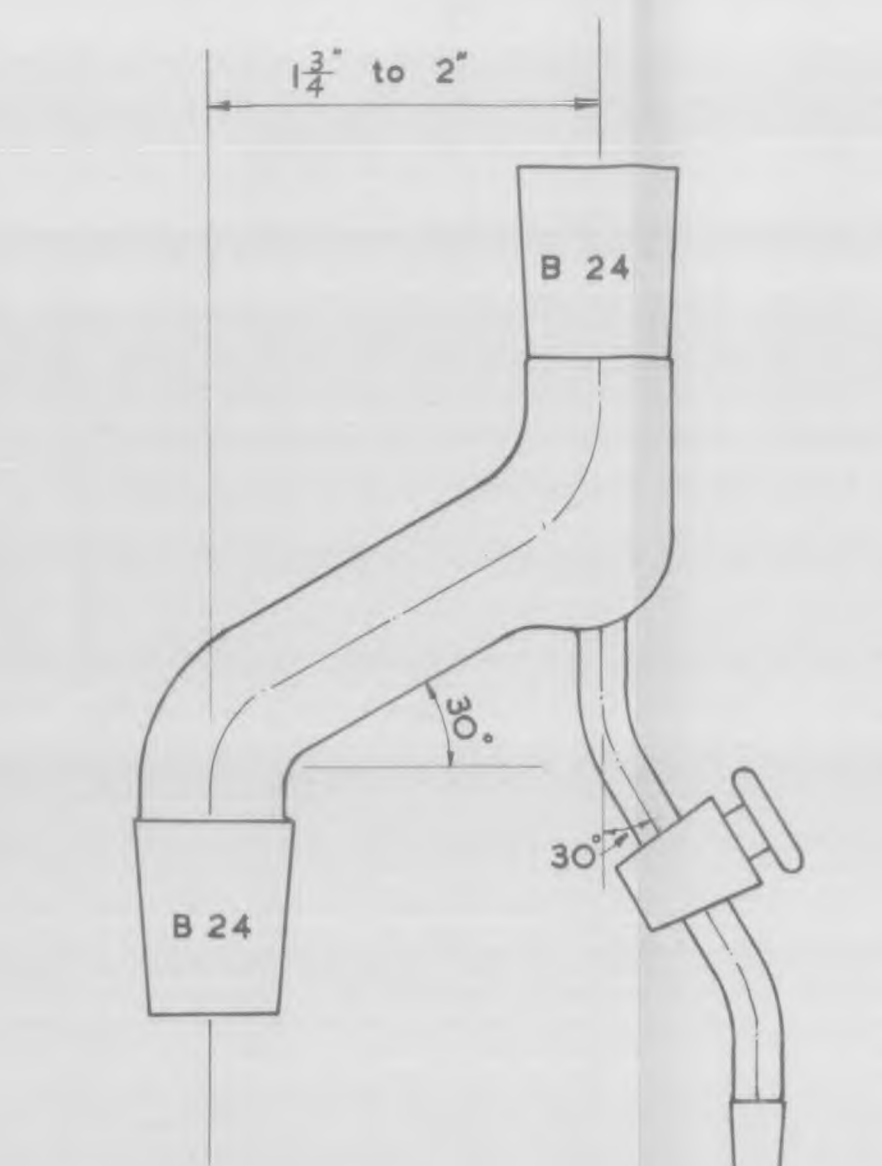
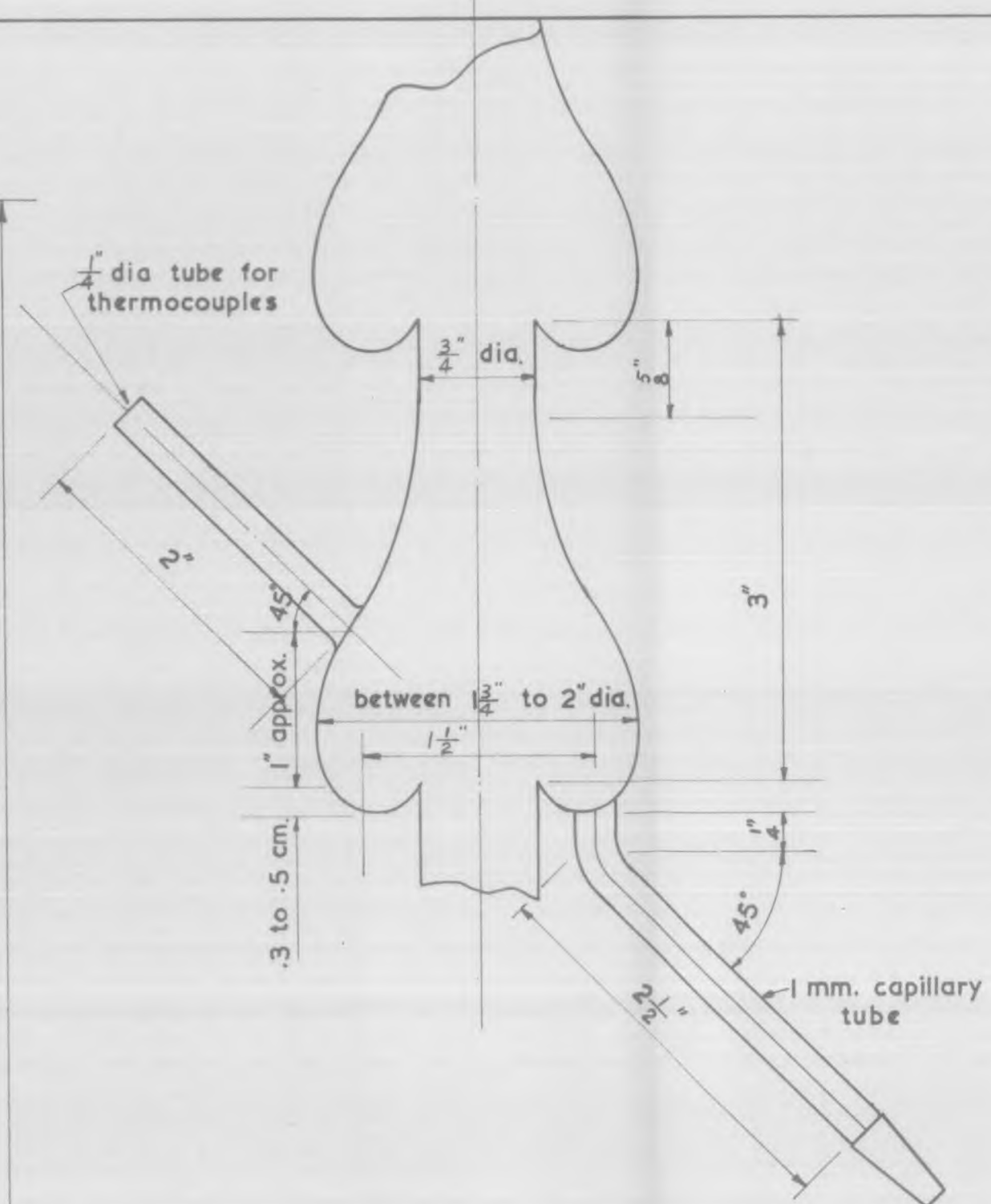
COLUMN BOTTOM

SCALE: HALF FULL SIZE

FIG. 3.1

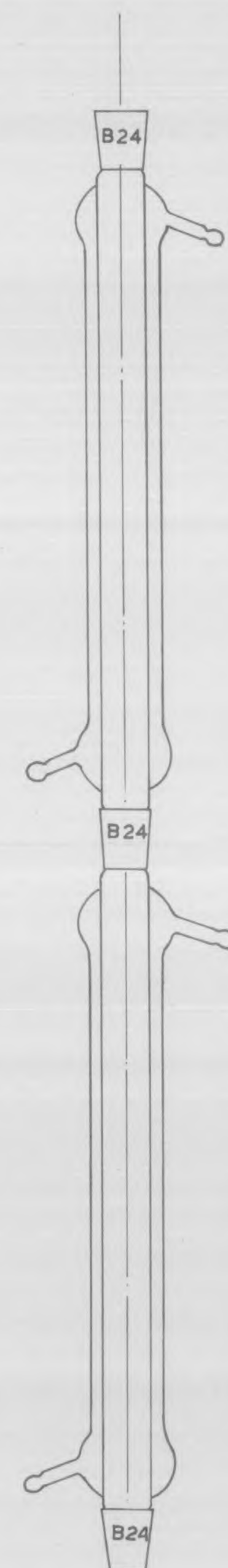


COLUMN
SCALE: HALF FULL SIZE



VAPOUR LINE ADAPTOR

SCALE: FULL SIZE



REFLUX CONDENSERS

SCALE: HALF FULL SIZE

IMPERIAL COLLEGE
CHEMICAL ENGINEERING DEPARTMENT
DISTILLATION UNIT
PEAR COLUMN

SCALE: AS SHOWN DRAWN BY: S.Y.L. DATE: 26-8-1960