

SIEVE-PLATE PERFORMANCE
IN THE FOAM AND SPRAY REGIMES
OF OPERATION

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by

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ABSTRACT

Foaming and non-foaming systems have been distilled in a sieve plate column. It has been confirmed that the plate efficiency of the foaming (surface tension positive) system is higher than that of the non-foaming (surface tension negative) system under foam inducing conditions.

At high vapour rates and low weir heights a liquid phase dispersed 'spray regime' is encountered. Under these conditions the surface tension positive system gives a lower plate performance than the negative system. It is thought that the difference is caused by the higher proportion of small drops present in the dispersion for the negative system.

For both types of system the performance in the spray regime is found to increase with decrease in the absolute value of surface tension. This is in qualitative agreement with an analogy based on spray atomisation, and consistent with the effect of surface tension on drop size.

A model is proposed for the spray regime which assumes a distribution of drop sizes projected upwards at initial velocities which are a function of drop size. The model has been used to deduce values for the drop size distribution parameters from the measured dispersion density profiles. It is found that the deduced mean drop size increases with surface tension. The

dependence is similar to that predicted by the spray analogy.

Values of plate performance in the spray regime estimated from the deduced drop size distribution with the help of the model are close to the measured values of plate performance.

To Paulette

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CHAPTER ONE

INTRODUCTION

Although distillation has been practised for centuries it has remained an art, rather than a science, until recently. Investigations into the basic phenomena that take place during distillation have been stimulated by the economic necessity of designing reliable and efficient equipment. What has resulted is a generalised approach to designing distillation equipment with a limited degree of confidence. On the other hand it is evident from the literature that the highly complex mass, momentum and heat transfer phenomena occurring in distillation are slow to yield their secrets.

Following a survey of research activity in the U.K. the Association of British Chemical Manufacturers (A.B.C.M.) and the British Chemical Plant Manufacturers Association (B.C.P.M.A.) decided to set up an ad hoc "Distillation Panel". The function of this group was to assess the state of the art, and recommend directions for further research. As a result six projects were sponsored by the A.B.C.M./B.C.P.M.A.

One of the topics investigated was concerned with the roles of heat and mass transfer in distillation. The findings to date have been reported by Sawistowski et al (1) and Bainbridge and Sawistowski (2). Particular interest was shown in the reported phenomena concerning a maximum in efficiency with

concentration (3) (4). The work of Theobald (1) indicated that the sensible heat transfer between the phases played no significant role in the separation process for the systems investigated. Consequently the concept of 'Thermal distillation' (5) cannot be evoked to explain the variation of efficiency with composition. On the other hand it has been confirmed that under certain conditions the distillation efficiency is a function of composition, decreasing at the extremes of concentration and close to an azeotropic point. It was suggested (1) that this behaviour could be explained by surfact tension effects.

In a classic paper (6) Zuideweg and Harmens demonstrated how surface tension gradients, set up by interphase mass transfer, could affect the type and magnitude of the interfacial area generated in distillation. They characterised the surface tension properties of the distilling systems in terms of the rate of change of the surface tension with composition ($d\sigma/dx$). Their work concluded that foaming systems ($d\sigma/dx < 0$) could be expected to give greater interfacial area than non-foaming systems ($d\sigma/dx > 0$), and as a consequence would give higher efficiencies. As a corollary it was added that the variation of efficiency with composition could be explained by the variation of local surface tension differences ($\Delta\sigma$) with composition. Thus foaming systems would be expected to show a maximum in the efficiency/composition relationship and non-foaming systems would not. These findings

were substantiated by Catchpole (7) who stated that when foaming and non-foaming systems were distilled in the same column the plate efficiency tended to be higher for the foaming system.

In contrast Bainbridge and Sawistowski (2) showed that under certain conditions the reverse behaviour prevailed. Not only did the non-foaming system present a maximum in the efficiency/composition relationship it also gave a higher efficiency than the foaming system. An explanation was provided by the observation that the dispersion-type on the plate was predominantly spray. Since the factors governing foam generation are the opposite of those governing spray generation it follows that in the "spray regime" the non-foaming system would be expected to give the greater interfacial area and hence the higher efficiency.

It is relevant to note that the "spray regime" is no longer an undesirable consequence of increasing column throughput beyond its design specifications, but is increasingly being employed as deliberate policy (8) (9) (10).

As a result it was decided to investigate the performance of foaming and non-foaming systems over a range of conditions. By judicious arrangement of flow rates and liquid hold-up on the plate it is possible to induce bubbling dispersions, droplet or "spray regime" dispersions, or any intermediate condition. This thesis presents the findings of such an investigation.

Measurements were taken in an attempt to characterise the dispersion types, and to highlight the differences between foaming and non-foaming dispersions.

It was also hoped that evidence could be gathered which might suggest an approach to modelling the spray regime.

CHAPTER TWO
LITERATURE REVIEW

2.1 Some basic concepts

The basic equations describing interphase mass transfer have been extensively covered in the standard text books. Only those relationships of direct relevance will be discussed here.

According to the two-film model of interphase mass transfer, proposed by Whitman (11), the resistance to transfer under turbulent bulk conditions is assumed to lie in two laminar films, one on each side of the interface. The interface itself is assumed to offer no resistance to mass transfer, In consequence the overall resistance to transfer is the sum of the two separate film resistances. The resistance is taken as the reciprocal of the mass transfer coefficient. For a gas-liquid system this leads to the relationship

$$\frac{1}{K_{OG}} = \frac{1}{K_G} + \frac{m}{K_L} \quad \dots (2.1)$$

where K_{OG} , K_G and K_L are the overall gas phase, the individual gas phase to the individual liquid phase mass transfer coefficients respectively, and m is the slope of the equilibrium line.

From the transfer unit theory of Chilton and Colburn (12) (13) combined with equation 2.1 it follows that

$$\frac{1}{N_{OG}} = \frac{1}{N_G} + \frac{mG/L}{N_L} \quad \dots (2.2)$$

in this case N_{OG} , N_G and N_L are the number of overall and individual transfer units, and N_{OG} can be expressed by

$$N_{OG} = K_{OG} \cdot S/G \quad \dots (2.2a)$$

An alternative approach to the expression of mass transfer performance in plate columns is the use of the efficiency concept. In general the efficiency is a comparison of actual separation with the ideal separation obtainable if certain equilibrium criteria are satisfied.

The most commonly applied definitions, due to Murphree (14), preceded the transfer unit theory by a number of years.

The Murphree point efficiency is given by

$$E_P = \frac{y_n' - y_{n+1}'}{y_n'^* - y_{n+1}'} = 1 - \exp(-K_{OG} S/G) \quad \dots (2.3)$$

where y_n' and y_{n+1}' are local vapour compositions above and below the tray, and $y_n'^*$ is the vapour composition in equilibrium with x_n' , the local liquid composition, which is assumed to have no vertical concentration gradient. Murphree obtained a result of this type by attempting to relate the change in composition at a locality with the basic mass transfer concepts of mass transfer coefficient, K_{OG} , and interfacial area S .

In the light of the transfer unit theory equation 2.3 can be written

$$E_P = 1 - \exp(-N_{OG}) \quad \dots (2.4)$$

The Murphree efficiency taken across a whole plate is

$$E_{MV} = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}} \quad \dots (2.5)$$

where the y_n^* is in equilibrium with x_n , the exit liquid composition. The Murphree 'vapour-phase plate efficiency', and the analogous liquid-phase efficiency, are related to the more basic 'point efficiency' by mixing criteria.

Recent mixing models have been based on the eddy diffusion concept. An example is provided by Gerster et al (15) who relate E_P , E_{MV} and the modified Peclet number P_e

$$E_{MV} = \frac{1 - \exp(-R - P_e)}{(R + P_e)(1 + R + P_e)} + \frac{\exp(R) - 1}{R(1 + \frac{R}{R + P_e})} \quad \dots (2.6)$$

$$\text{where } R = \frac{P_e}{2} \left((P_e + \frac{4AE}{P_e})^{\frac{1}{2}} - 1 \right) \quad \dots (2.7)$$

$$\text{and } A = mG/L \quad \dots (2.8)$$

By combining the foregoing equations the A.I.Ch.E. research committee (16) have presented a general design method which is shown to provide a useful means of predicting column performance. On the other hand it should be pointed out that the A.B.C.M./B.C.P.M.A. projects were partly motivated by the knowledge that the A.I.Ch.E. design method has serious limitations.

The advantage of using the addition of resistances approach is that it is possible to conduct separate experiments on gas-phase controlled and liquid-phase controlled systems and obtain correlations for N_G and N_L . For a distilling system where both phase resistances may be significant, N_{OG} is found using equation 2.2 which combines the N_G and N_L values derived from the correlations.

As a general rule it is necessary to correct the N_G and N_L values for the variation in gas and liquid physical properties. From the analogy between heat, mass and momentum transfer Chilton and Colburn (17) have obtained the following relationship for mass transfer

$$(j_D)_G = \frac{K_G \cdot S}{G} (Sc_G)^{\frac{2}{3}} = f(Re_G) \quad \dots (2.9)$$

where j_D is the so-called j-factor for mass transfer.

It follows from equation 2.9 that

$$\frac{(N_G)_1}{(N_G)_2} = \left(\frac{Sc_2}{Sc_1} \right)^{\frac{2}{3}} \quad \dots (2.10)$$

Gerster, Colburn, Bonnet and Carmody (18) used this relationship to correlate their gas phase performance data. However for the liquid phase they employed the relation

$$\frac{(N_L)_1}{(N_L)_2} = \left(\frac{D_{L1}}{D_{L2}} \right)^{\frac{1}{2}} \quad \dots (2.11)$$

which is in agreement with the penetration theory of Higbie (19).

This approach has received subsequent support, although the exponent in equation 2.10 is often given as $\frac{1}{2}$, as in the A.I.Ch.E. design method.(15)(16).

2.2 Previous Work

Before discussing the findings of the A.B.C.M./B.C.P.M.A. project 3 it is appropriate to mention the work of Zuideweg and Harmens (6). In a series of experiments in different types of distillation equipment they demonstrated the influence of the surface tension gradient on the interfacial area formed, and the associated mass transfer performance. They classified systems in terms of the magnitude and direction of the surface forces set up during mass transfer. Systems in which the surface tension decreases with increasing composition are denoted as "positive" whilst systems with increasing surface tension are termed "negative".

The combined effect of mass transfer and surface tension gradient is illustrated in Fig.2.1. In this example the system is positive (b). Illustration (c) represents a liquid lamella at bulk composition x_b , and interface composition x_i , with a "thin spot" which, owing to its low capacity, has rapidly approached equilibrium with the bulk vapour y_b . It is seen that the local difference in surface tension, represented by $\sigma^* - \sigma_i$, is influenced by the slope of the surface tension curve, and by $x_i - x^*$, which approaches $x_b - x^*$ for a gas phase controlled system. The nett effect in this example is that liquid is drawn into the "thin spot" and tends to stabilise

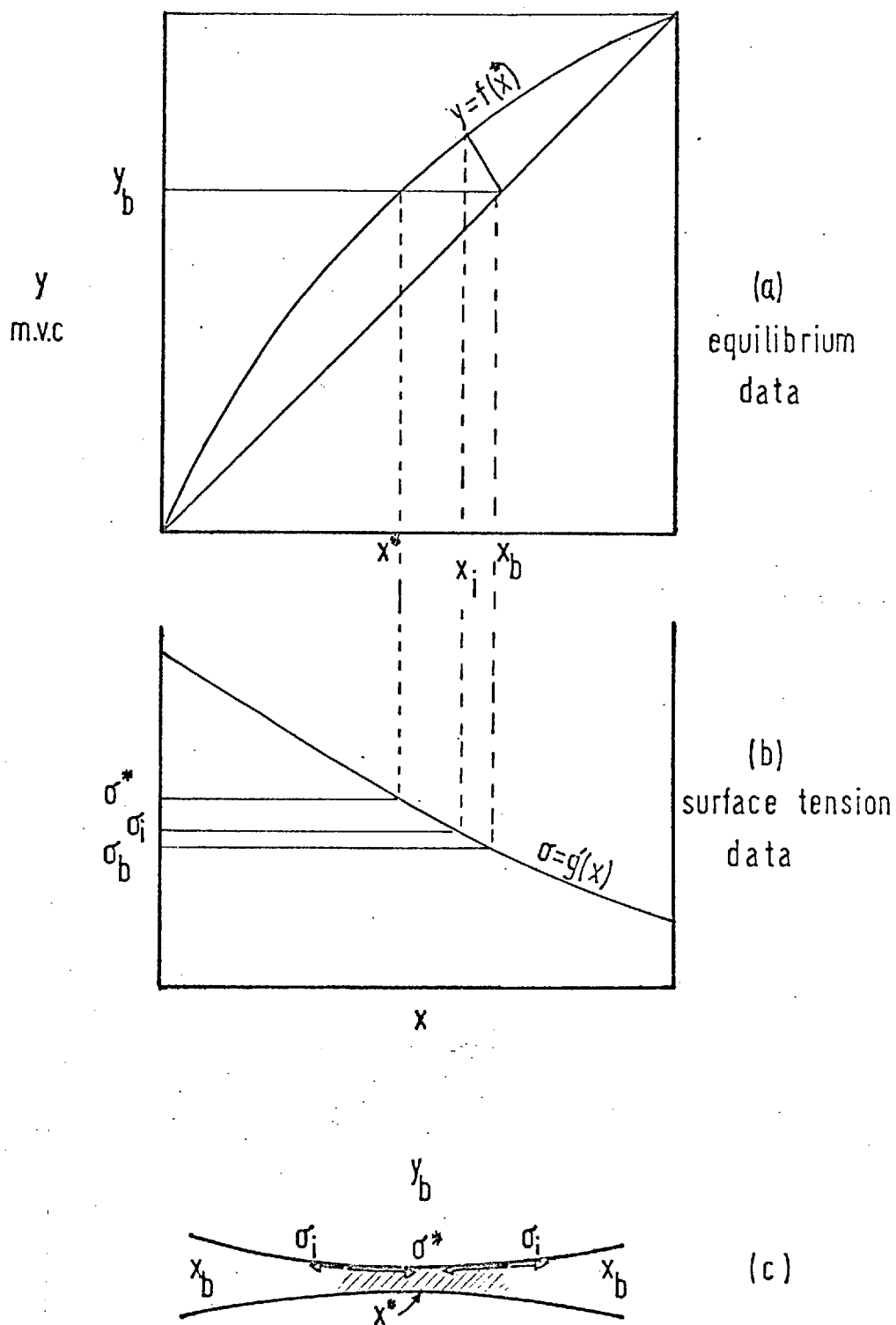


FIG 2.1 ILLUSTRATION OF FILM STABILISATION
(POSITIVE SYSTEM)

the film. For a "negative" system σ^* will be less than σ_i and the thinning process will be augmented. The classification of Zuiderweg and Harmens can be summarised as follows:

positive systems, $d\sigma/dx < 0$, relative volatility α is not small;
 negative systems, $d\sigma/dx > 0$, relative volatility α is not small;
 neutral systems, $d\sigma/dx \approx 0$, and/or $\alpha \rightarrow 1$.

An added condition for promoting neutral behaviour, is when the driving force is reduced as a result of working in the "pinch condition".

From their work on Oldershaw sieve plate columns it was concluded that the positive systems promote foam, and the negative and neutral systems promote spray. It is also shown that positive systems tend to give considerably higher plate efficiencies than neutral systems which in turn tend to give slightly higher efficiencies than negative systems. In addition the influence of the driving force on the surface tension effects will be marked for the positive system and insignificant for the neutral and negative systems. In other words the positive systems can be expected to show higher efficiencies than the negative systems, all other things being equal. The effect of composition on performance will be considerable for the positive systems and absent for the negative.

Zuiderweg and Harmens add that the differences between positive and negative systems would be expected to diminish with an increase in the size of the column, owing to the greater

tendency for spray formation. Results from a 0.45 m. diameter tray, for positive, weakly positive and weakly negative systems all showing similar performance, are offered as evidence to this claim. It will be shown later that these results are open to re-interpretation.

Additional comment is made by Catchpole (7) who reviewed a number of investigations and concluded that when a positive and negative system are distilled in the same column the plate efficiency is higher for the positive system.

The initial terms of reference of the A.B.C.M./B.C.P.M.A. project 3 were to study the roles of heat and mass transfer in distillation. Attention was focused on the observed variation of distillation performance with composition. Whilst this variation was confirmed, it was shown by Theobald (1) that the role of heat transfer, or 'thermal distillation', is insignificant.

Furthermore the generalised predictions of Zuiderweg and Harmens and Catchpole fail to apply. Thus, under certain conditions, the negative systems tended to show a higher efficiency than the positive systems, and both types exhibited a variation in performance with composition. An explanation is offered by Bainbridge and Sawistowski (2) who noted that the dispersion-type in these investigations was predominantly spray. In this regime it is suggested that the surface tension properties of the negative system will assist droplet formation and those of the positive

system will discourage droplet formation. Since, in the spray regime, the mass transfer surface is predominantly the surface of droplets, the negative system will be expected to give a higher surface area than the positive system. This implies that the criteria of Zuiderweg and Harmens fails to cover behaviour in the spray regime.

A further property of the spray regime noted by Bainbridge and Sawistowski, is that plate performance decreases with increase in absolute surface tension, σ . This has since been confirmed by de Goederen (20). In both cases the performance, N_G , was found to be inversely proportional to σ .

In the light of these observations it is necessary to re-examine the results quoted by Zuiderweg and Harmens for the 0.45 m. column, in which positive and negative systems gave very similar values for plate efficiency. At the compositions employed, the negative system, benzene/toluene, had an absolute surface tension of 20 dyn/cm, whilst the two positive systems, n-heptane/toluene, and n-heptane/methylcyclohexane had values of 15 and 13.5 dyn/cm respectively. This is important since spray formation was observed to be predominant. A comparison of systems would only be justified if the absolute value of surface tension were the same in all cases. In addition, if the performance is expressed in terms of N_G , and then corrected by equation 2.10, the negative system having a Schmidt number of 0.86 (compared with 0.67 for the

two positive systems) shows up to even greater advantage. It is recognised that expressing the performance in terms of N_G is only valid if the plate is well mixed, so that E_p is very close to E_{MV} , and if the systems can be assumed to be gas-phase controlled. However there is at least a prima facie case for stating that these results of Zuideweg and Harmens are compatible with the findings of Bainbridge and de Goederen.

From the foregoing discussion it would appear that the criteria of Zuideweg and Harmens can only be applied when the operating conditions are such as to encourage foaming in a positive system and spraying in a negative system. Such conditions are typical of Oldershaw columns in the normal operating range. When larger scale equipment is used frothing tends to be suppressed and spray-type dispersions are encountered, as observed by Zuideweg and noted by the E.F.C.E. Working Party on Distillation (21).

As a consequence it was decided to compare the performance of positive and negative systems under foam-inducing and spray-inducing conditions. Apart from the surface tension properties of the systems, the major influences on the type of dispersion generated in a given column are judged to be the liquid hold-up and the vapour rate. For a column operating at total reflux the variables become weir height and vapour rate.

It is of interest to note that liquid holdup and gas rate are major parameters in the work of Wong (22), and Shakhov, Noskov

and Romankov(23) who studied the transition from foam to spray, and in the work of de Goederen(20) in his study of mass transfer in the foam and spray regimes.

It remains to discuss briefly the dispersion regimes on sieve plates with particular reference to the properties and occurrence of the spray regime.

2.3 Dispersion Regimes on Sieve Plates

2.3.1 Hydrodynamic considerations

It is generally agreed that several types of dispersion, or regime have been observed on sieve plates. The factors governing the type of dispersion generated are the plate geometry, the fluid flow rates and the fluid physical properties.

The regimes are roughly characterised by the nature, gas or liquid, and the volume fraction of the dispersed phase. As an illustration Figure 2.2 shows plots of volume fraction of liquid, often termed mean dispersion density $\bar{F}_L$, against gas velocity.

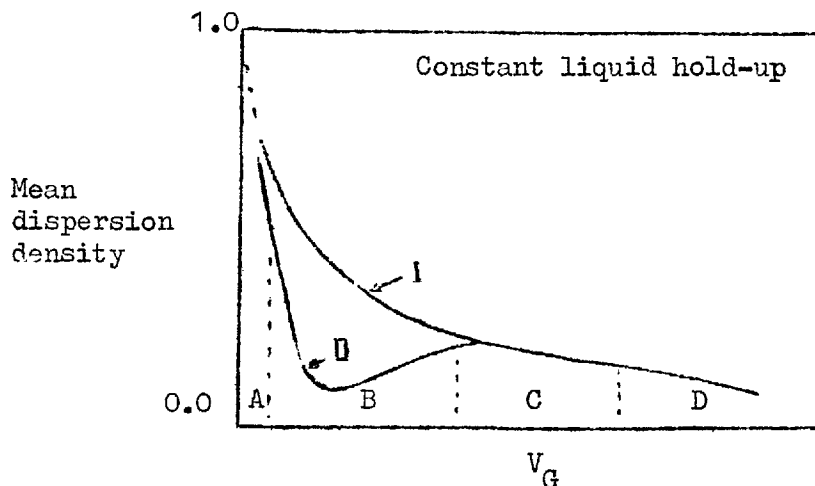


Figure 2.2

The Effect of vapour rate on Mean Dispersion Density

Curve I is taken from Williams et al (24) and can also be derived from de Goederen's (20) data on hold-up and dispersion height. Curve II is taken from Wong (22) and Calderbank and Moo-Young (25), and can again be derived from de Goederen's data; it is also consistent with the observations of Sargent (A.B.C.M./B.C.P.M.A. Final report on Distillation, 1964, page 132).

At low gas rates, liquid tends to weep through some of the perforations and gas bubbles through others, this is region A on Figure 2.2, and corresponds to the 'weeping range' of Prince (26) and the condition of non-uniform behaviour of Solomakha and Matvozov (27). In this region the bubbles rise under the effects of buoyancy alone.

It is important to note that for a continuously bubbling orifice the bubble frequency appears to be constant at approximately 20 per sec at all except the lowest flow rates. This was originally observed by Calderbank (28) over the range of flow rates from 20 to 250 cc/sec., and have recently been confirmed by Porter et al (29).

As the gas rate is gradually increased the weep rate decreases, and the majority of the perforations bubble. In this region, B, there is a decrease in the mean dispersion density and a gradual tendency for buoyancy rise to be replaced by forced bubbling. As seen in Fig. 2.2 this region is characterised by

two types of behaviour. Under certain circumstances the dispersion is comprised of large cellular bubbles of low surface area per unit volume and high gas hold-up (25). This regime is known as the cellular foam regime, and is represented by curve II. According to Schroter (30) this is an "almost rigid" structure. Factors encouraging this type of dispersion are those favouring the build-up of a mass of rigid bubbles. These include the presence of foam baffles, high overflow weir, and the influence of side walls; this latter factor being a significant difference between small scale and large scale equipment, as observed by Zuideweg (6). Added to these external influences there is the important effect of surface tension forces. A decrease in the absolute value of the surface tension encourages cellular foam build-up as can be seen from the results of de Goederen. Similar results have been reported by Schroter, who added a surface active agent to his air-water system and obtained unmanageable quantities of cellular foam. However in this case the change in surface elasticity may have been a contributing factor. In addition the surface tension properties discussed under Zuideweg's work (section 2.2) will be highly significant. A 'positive' system will tend to encourage the formation of stable bubbles and favour cellular foam build up. It has also been pointed out by Andrew (31) that the stabilising effect of surface tension forces in two-component mixtures is at a maximum when $\theta(x)(d\sigma/dx)^2$ is a maximum, here σ is the surface tension, x is the composition and the function $\theta(x)$ is

adequately approximated by x .

As the gas rate is increased the added turbulence causes bubble coalescence and breakdown of the rigid structure of the cellular foam. The resultant highly agitated dispersion is termed mobile foam, or froth. The transition is heralded by the onset of liquid circulation, manifesting itself in the downward movement of small bubbles near the walls, according to Porter et al (29). Moreover Schroter claims that the transition occurs at a superficial velocity of approximately 1m/s independent of plate geometry, and liquid bulk viscosity and surface tension. However, as he acknowledges, de Goederen's results at low liquid hold-up indicate a transition from cellular to non-cellular dispersion at a superficial velocity of 0.5 m/s.

Before considering the consequence of further increasing the gas rate it is necessary to return to Figure 2.2 and curve I in region B. It has been seen that under certain conditions region B presents a rigid cellular foam which gradually gives way to a highly agitated froth, region C. However, it is not uncommon to find that forces supporting cellular foam formation are absent or at least outweighed by other factors. This is represented by curve I on Figure 2.2, and is essentially a mobile foam regime, characterised by bubble coalescence and non-cellular behaviour. Low weir height, the absence of foam baffle and large column diameter are some of the external factors. Added to this

Calderbank (25) has shown that low orifice Reynolds numbers favour cellular foam formation. This suggests that at equal hole velocities small orifice diameters favour curve II and large orifice diameters favour curve I. It follows that bubble cap plates with their high effective orifice diameters, and coarse bubbling action, will not favour cellular foam formation, and will tend to show curve I behaviour in region B, as seen in the Michigan report (24).

Liquid physical properties such as surface tension and surface tension gradients will be important factors. In particular it is known that surface tension negative systems will not sustain cellular foam even at low orifice Reynolds number in small equipment such as Oldershaw columns (6). Such systems would inevitably give curve I behaviour.

At higher gas rates all system types will exhibit non-cellular behaviour, this is the froth, or mobile foam regime, C in Figure 2.2. This regime is characterised by a high degree of turbulence and liquid circulation. There is also an increasing tendency for the upper zone of the dispersion to be present as spray droplets. It is a fair generalisation to say that region C is the most ubiquitous in industrial sieve plate operation (6).

As the gas velocity is raised there is an increasing tendency for the liquid to replace the gas as the dispersed phase, as observed by Foss and Gerster (32) and Garner (33). When the majority of the liquid in the dispersion is in the form of droplets

the regime is termed the Spray Regime, region D in Fig. 2.2.

The importance of any of the regions A to D in the velocity spectrum of a given sieve plate is governed by the 'basic factors' of plate geometry, fluid flow rates and fluid physical properties. As already pointed out curve II, cellular foam regime, does not always prevail in region B. In most cases it appears to be difficult to pinpoint the transition from one regime to another. The transition from cellular foam to froth is not clearly defined according to Wong (22), and the results of de Goederen (20) confirm this. There appears to be some disagreement in the literature as to whether the transition between froth and spray is gradual or can be pinpointed. Wong claims that the transition occurs at a point, although he measured the transition from spray to foam in a system without liquid flow. This is in partial agreement with Shakhov, Noskov and Romankov (23) who observed critical gas velocities at transition, and also found a hysteresis effect. Shakhov et al characterised the transition by measurement of entrainment and pressure drop, claiming that a distinct jump in the entrainment rate and the pressure drop occurred at the onset of the spray regime.

In contrast, the performance data of de Goederen and of Yoshida et al (8), which cover the transition range from froth to spray, show no noticeable break in the curves, which would be expected from Shakhov's results. Moreover the pressure drop curves

of Yoshida are smooth, unlike those of Shakhov.

2.3.1.i The Spray Regime: Hydraulic behaviour

The spray regime is favoured by high gas rate and low liquid hold-up. In the Michigan report (24) the dispersion at low submergences is described as "like an inverted rainstorm". Yoshida et al (8) draw the analogy with a fluidised bed. This analogy is carried a stage further by Andrew (34) in a recent investigation into the hydrodynamic and performance characteristics of large diameter absorption columns at high superficial gas velocities. Considering the dispersion as a 'fluidised bed of fluid' Andrew assumes the drop size to be approximately equal to the maximum stable drop size consistent with the given gas rate and the liquid properties. Having an estimate of the particle size it is possible to estimate the bed (or dispersion) density from the terminal velocity and the prevailing gas velocity using the Richardson and Zaki (35) equation. Mean dispersion densities of 0.25 and less are estimated for the spray regime, which agrees with figures of 0.25 to 0.1 obtainable from de Goederen's results.

According to de Goederen the droplets from his lowest surface tension system were the smallest and appeared to be thrown up highest above the plate. It has also been shown (36) (37) that the smallest drops reach the highest levels. Another characteristic of the spray regime is that pressure drop and the liquid mixing parameter, the Peclet number, become independent of the weir height (8).

Active interest in the spray regime has had two major motivations. Firstly, it is an inevitable consequence of trying to increase the throughput of a column that a more spray-like dispersion is formed, this will have undesirable results if entrainment becomes heavy. Secondly it is believed that liquid dispersed systems offer a more intensive mass transfer zone. This is the philosophy behind recent investigations into the behaviour of a slot tray operating in the drop regime (9). It is claimed by the authors, Dytneriskii, Planovski and Matvozov, that such plates are already operating successfully in industry. Spray regime operation is also deliberate policy in the high capacity trays investigated by Manning (10) and Yoshida et al (8). It is also known that sieve trays used for vacuum distillation are often operating in the spray regime (33).

2.3.1.ii Nonhomogeneity of Dispersions

It is important to realise that whilst Figure 2.2 is a useful hook on which to hang the concept of different regimes of operation of sieve plates, it conceals an important property of the dispersions. As shown by Macmillan (38) and others the gas-liquid dispersions are generally non-homogeneous. This means that the local value of dispersion density varies with height above the plate floor. It will also vary with time, but it has been shown by Bernard (39) that time averaged values are reproducible. This non-homogeneity suggests that different types of contacting may take

place in a given dispersion; as substantiated by visual observation.

According to Garner and Porter (33) three zones can be distinguished at different levels in the dispersion above a sieve tray:

- (1) At the bottom of the dispersion, a clear liquid layer in which bubbles form and rise at high velocity.
- (2) Above this, a mass of bubbles (cellular or mobile).
- (3) Above this, a spray of liquid droplets of various sizes.

The overall significance of each zone will be different in the various regimes. In the cellular foam regime zone (3) will be negligible, in the mobile foam regime all zones may be significant, and in the spray regime zone (3) will be predominant, (2) will be absent and (1) negligible or absent.

2.3.2 Mass transfer on sieve plates

A rigorous approach to description of the mass transfer processes occurring on sieve plates would have to take into account not only the different regimes but also the separate mass transfer zones present. Some effort has been made along these lines, but this usually demands the simplifying assumption that one or more of the zones make a negligible contribution to mass transfer.

Considering the case of bubbly dispersions Garner and Porter (33) have assumed that the spray zone is insignificant. They use the solution of the unsteady state diffusion equation for a rigid sphere with constant surface concentration.

$$\frac{C_1 - C_2}{C_1 - C^*} = E_p = 1 - \frac{6}{\pi^2} \sum_{n=1}^{\infty} \left\{ \frac{1}{n^2} \exp \left[- (D_E n^2 \pi^2 t / r^2) \right] \right\} \quad \dots (2.12)$$

where C_1 , C_2 , and C^* are the initial, final and interface concentrations in the gas phase. The diffusivity term, D_E , is an effective diffusivity, which incorporates both molecular and eddy diffusion. Assuming gas phase control equation 2.4 can be written

$$N_G = -\ln(1 - E_p) \quad \dots (2.4a)$$

Equations 2.4a and 2.12 can be used to investigate the relationship between N_G and t for given values of D_E and r . Garner and Porter have done this for $r = 0.5$ cms. and $D_E = 0.1, 0.2, 0.4 \text{ cm}^2/\text{s}$; and their results are shown in Figure 2.3.

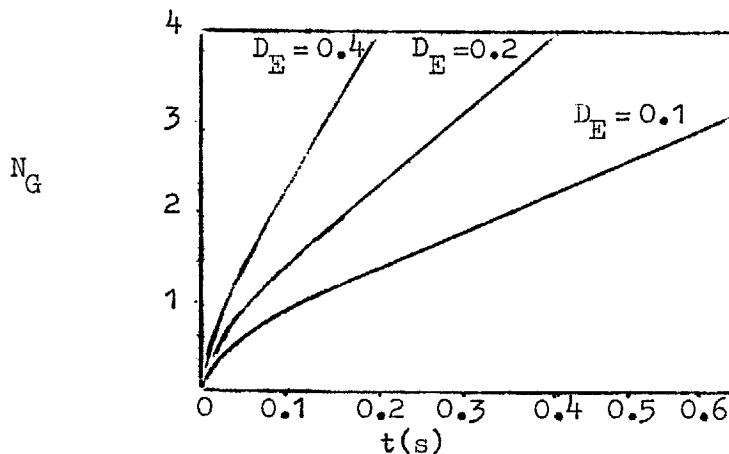


Figure 2.3

(taken from ref (33))

As a simplification they also point out that the series in equation 2.12 is closely approximated by the first term if $D_E \pi^2 t / r^2$ is greater than 0.5. Thus equations 2.4a and 2.12 give

$$N_G = 0.502 + (D_E \pi^2 t / r^2) \quad \dots (2.13)$$

They plotted data from the Delaware Report (15) and obtained a correlation of this form except at the highest gas rates.

Taking this approach a stage further Porter, Wong and Davies (29) have analysed mass transfer in the cellular foam and in the froth (mobile foam). Equation 2.13 is taken to hold in the cellular foam and two models tested for the froth. The first froth model is based on the observations of Calderbank and Moo-Young (25) that the froth comprises small bubbles of uniform size in a completely mixed cloud. The second model is based on two bubble sizes, large and small, with uniform residence time. These models give the relationships:

$$(I) \text{ Calderbank model, } N_G = 0.48 + \ln(D_E \pi^2 t / r^2) \quad \dots (2.14)$$

$$(II) \text{ Porter model, } N_G = \ln(1/L_b) + 0.5 + (D_E \pi^2 t / r^2)_{L_b} \quad (2.15)$$

where L_b is the fraction of gas in the large bubbles. When L_b is unity equations 2.13 and 2.15 are identical. It is shown that the Calderbank model is not invalid but that the Porter model is probably a much better representation of the physical conditions.

Values of L_b indicate that the major portion of the gas passes through as large bubbles of formation size, this means that the intercept on a plot of N_G versus t is nearly 0.5. In other words mass transfer in the cellular foam and in the froth should be correlated by similar relationships; this is important since few correlations specify the type of dispersion they cover.

Porter et al also determine D_E from the slopes of the N_G versus t plots. The ratio of effective to molecular diffusivity, D_E/D_M is shown to increase almost linearly with orifice Reynolds number, the values for the cellular foam being only slightly lower than those for the froth. This correlation is questionable on two counts. Firstly, since D_E is a characteristic of the bubble rise zone it is surprising it can be correlated by the orifice Reynolds number. Secondly, considering the rigidity of the cellular foam bubbles and the more chaotic nature of the froth it would be expected that the ratio D_E/D_M for the froth would be significantly greater than that for the cellular foam.

The original model of Garner and Porter indicates a high rate of mass transfer at low residence time (Fig. 2.3); dN_G/dt is very high as $t \rightarrow 0$. This is as expected from the assumption of unsteady state mass transfer to a rigid sphere with constant surface concentration. No special consideration is given to the bubble formation period, which covers the bubble growth period up to detachment. However as pointed out by Thorogood (40) it is to be

expected that during bubble formation the kinetic energy of the vapour decreases rapidly, and considerable amounts of energy will be dissipated as turbulence. Under these conditions it is to be expected that D_E will be much larger than in the bubbly dispersion. Thorogood solved the equation for mass transfer into a growing bubble for the condition of constant upward velocity, U , of the top of the bubble. The efficiency at bubble formation is then

$$E_B \approx 2K_G' / (2K_G' + U) \quad \dots (2.16)$$

where K_G' is defined by Thorogood as the mass transfer coefficient during formation. He obtained values for E_B from equation 2.4a with N_G taken as $(N_G)_{t=0}$, which is the intercept value obtained from plotting N_G versus t .

In general the plate performance can be written

$$N_G = (N_G)_0 + (N_G)_{r,t}$$

where $(N_G)_0$ is the contribution due to the bubble growth period and $(N_G)_{r,t}$ is due to the bubble rise period, and is time dependent. Thus the intercept on a plot of N_G versus t is approximately given by

$$(N_G)_{t=0} = (N_G)_0 + (N_G)_{r,t=0}$$

If it is assumed that $(N_G)_{r,t=0}$ is approximated by the constant 0.5 in equation 2.13, the extra contribution due to bubble growth is very small. As a result it would appear that the original assumption of constant D_E throughout the bubble life time is a reasonable approximation. Macmillan(38) has demonstrated the

diminishing significance of the bubble formation period in his work, as the gas rate is increased.

The foregoing analysis of bubbly dispersions suggests the following:

- (1) It is probable that under most conditions of bubbly dispersions both cellular and mobile foam mass transfer can be correlated in a similar manner.
- (2) The bubble formation zone has high mass transfer, the significance of which diminishes as the gas rate increases or the overall residence time increases.

As this analysis is based on bubbly dispersions it will not be relevant when spray formation predominates, as at the highest velocity condition plotted in Fig. 2 of the paper of Garner and Porter. Spray regime mass transfer is discussed in section 2.3.2.ii.

2.3.2.i Mass transfer correlations for bubbly dispersions

There are numerous correlations for mass transfer on bubble cap and sieve plates. They often prove difficult to compare since they may be presented in terms of efficiency, the number of transfer units, or mass transfer coefficients. The latter require either dispersion height or surface area data to aid comparison.

Another distinction is the theoretical basis of the correlations. Most of the Russian work is based on dimensional analysis, which allows for the interaction of parameters. In

contrast the A.I.Ch.E. project results are correlated by 'best-fitting' them to chosen parameters. It is not clear from their work if they tested for interactions between parameters. Bernard (39) found that his results showed considerable interactions and that no usable pattern emerged from a statistical analysis.

Bearing in mind the conclusions of Thorogood (40) on the bubble formation zone, and the approach of Garner and Porter (33), Fig. 2.3, it would appear that some allowance should be made for the residual $(N_G)_{t=0}$; this cannot be covered by the dimensional analysis approach.

As a general rule performance, expressed either as efficiency or the number of transfer units, is found to decrease with gas rate.

The work of Gerster et al (15) correlates the gas phase performance by

$$N_G \cdot Sc^{\frac{1}{2}} = 0.776 + 0.116W - 0.290F_b + 0.0217L_W \dots (2.17)$$

where W is the weir height, F_b is the F-factor based on the bubbling area and L_W is the liquid flow per unit of column width.

An alternative approach is given by Foldes (41) who demonstrates the analogy between hydrodynamic and mass transfer processes on perforated plates. His results are correlated by

$$E_{MV} = 0.33 Re_G^{-0.2} Ga_G^{0.1} Ga_L^{0.04} \left[\frac{d_c}{W} \right]^{0.1} \dots (2.18)$$

where W is the weir height and d_c is the column diameter. Ga is the

Galileo number $= \frac{W^3 \rho^2}{\mu^2 g} = Re^2 (Wg/V^2)$. The linear dimension used in Re is W .

Chu (42) used the Chilton and Colburn (17) analogy to calculate the gas phase mass transfer coefficient. This relation may be expressed as

$$K_G = 0.023 V R_e^{-0.2} Sc^{-0.67} \quad \dots (2.19)$$

or

$$Sh = 0.023 Re^{0.8} Sc^{0.33} \quad \dots (2.19a)$$

where Re is based on the bubble diameter and velocity.

For the condition

Interfacial Area (S) $\neq f(Re)$

$$N_G \propto Re^{-0.2} Sc^{-0.67} \quad \dots (2.20)$$

Rodianov and Radikovskii (43) have recently measured $K_G S$ and S on a sieve plate. They correlate the gas phase mass transfer coefficient by

$$Sh = 0.017 Re_G^{0.85} Sc^{0.5} \quad \dots (2.21)$$

where $Re_G = \rho V d / (\mu \bar{F}_G)$, $\bar{F}_G$ being the fraction of gas in the dispersion. This can be shown to be equivalent to

$$K_G = 0.017 V_o Re^{-0.15} Sc^{-0.50} \quad \dots (2.22)$$

where V_o is the actual gas velocity ($V_o = V/\bar{F}_G$). Equations 2.22 and 2.19 can be seen to be similar.

De Goederen (20) correlates his gas phase data for

cellular foam conditions with

$$Sh = 0.67 Re^{0.3} F_{OD}^{-0.45} \quad \dots (2.23)$$

where F_{OD} = diffusional Fourier number = $\frac{Dh(1-\bar{F}_L)}{d^2 V}$

D = diffusivity, h = height dispersion, $\bar{F}_L$ = fraction liquid in the dispersion, d = bubble diameter, V = superficial gas rate.

Equation 2.23 can be re-arranged to give

$$Sh = 0.67 Re^{0.75} Sc^{0.45} \left[\frac{h(1-\bar{F}_L)}{d} \right]^{-0.45} \quad \dots (2.24)$$

or

$$K_G = 0.67 V Re^{-0.25} Sc^{-0.55} \left[\frac{h(1-\bar{F}_L)}{d} \right]^{-0.45} \quad \dots (2.25)$$

Assuming $(1-\bar{F}_L)$ can be adequately accommodated in the Reynolds number as in equation (2.21), equation (2.24) becomes

$$Sh = 0.67 Re^{0.75} Sc^{0.45} \left[\frac{h}{d} \right]^{-0.45} \quad \dots (2.26)$$

Dytnerskii et al (44) have produced what they claim to be an almost universal correlation, starting from the theoretical expression

$$Sh = 0.46 Pe^{0.5} \left(\frac{\mu_G}{\mu_G + \mu_L} \right)^{0.5} \quad \dots (2.27)$$

for mass transfer from a spherical bubble at $Re < 1$. It is then assumed that an equation of the same type will hold for bubbly dispersions, if some allowance is made for the increased mass transfer rate due to the introduction of turbulence. This hydrodynamic factor Γ_h , is taken as $\Delta P / \rho_L dg$, where ΔP is the

pressure drop caused by the presence of the dispersion.

The correlation is

$$Sh = A' P_e^{0.5} \left[\frac{\mu_G}{\mu_G + \mu_L} \right]^{0.5} \quad \dots (2.27a)$$

Assuming $\Delta P \approx h(1 - \bar{F}_G) \rho_L g$

and noting $Pe \approx Re \times Sc$, equation 2.27a can be re-arranged to

$$Sh = A' Re^{0.5} Sc^{0.5} \left[\frac{h}{d} \right] \left[\frac{\mu_G}{\mu_G + \mu_L} \right]^{0.5} \quad \dots (2.28)$$

Dytneriskii et al have shown that their correlation equation (2.27) holds good for numerous systems on a variety of plate types. This is less surprising when equation 2.28 is compared with the equations of de Goederen (2.26), Rodionov (2.21) and Chu (2.19a).

This analysis shows that a number of seemingly different correlations can be reduced to a common form

$$Sh = \alpha Re^\beta Sc^\gamma \left[\frac{\delta}{h} \right] \quad \dots (2.29)$$

where

$$\alpha = \text{a constant}$$

$$\beta = 0.5 \text{ to } 0.85$$

$$\gamma = 0.33 \text{ to } 0.50$$

$$\delta = 0.0 \text{ to } 1.0$$

It is significant that in none of the correlations for the bubbly dispersions is there a term which includes the surface tension.

2.3.2.ii Mass transfer in the Spray Regime

There is only a limited amount of mass transfer data available for the spray regime. However there are certain points which emerge from a comparison of these data. It will be shown that the results are in qualitative agreement with theory.

Yoshida et al (8) found that the plate performance, expressed as E_{MV} or N_G was virtually independent of weir height in the spray regime. This accords with similar observations on the pressure drop and the Peclet number. Yoshida's results show that the performance gradually increases with vapour rate in the spray regime. This applies to absorption of ammonia from aqueous solution as well as distillation of the methanol-water system.

An increase in performance with gas rate was also observed by de Goederen (20) in his vaporisation studies. In addition he confirmed the findings of Bainbridge (2), that the performance was dependent on the absolute value of surface tension. They both show that N_G is inversely proportional to the surface tension, σ . This dependence on the absolute value of surface tension distinguishes the spray regime from the bubbly dispersions. De Goederen correlates his results by

$$N_G Sc^{\frac{2}{3}} = 1.62 F_H^{1.5} / \sigma \quad \dots (2.30)$$

where F_H is the hole F factor = $V_H \rho_G^{0.5}$

Andrew (34) has used the fluidised bed analogy for his analysis of the hydrodynamic and mass transfer performance in the spray regime. As his work is unpublished the fluidised bed analogy is developed from a description of his model.

The approaches due to de Goederen and Andrew are analysed to assess the possible influence of gas velocity and absolute surface tension on the performance in the spray regime.

I. Spray Analogy

A constant liquid hold up Q_L is assumed which is dispersed into uniform drops of diameter d . The total interfacial area, S , is given by

$$S = 6Q_L/d \quad \dots (2.31)$$

It is assumed that the mass transfer from a single drop can be expressed by a correlation of the type

$$Sh = \text{const.} Re^{0.5} Sc^{0.33} \quad \dots (2.32)$$

where Re is based on the drop diameter and the gas velocity V .

The performance, in terms of the number of transfer units in the gas phase, can then be shown to be

$$N_G.Sc^{2/3} \propto (\mu_G/\rho_G)^{0.5} V^{0.5} d^{-1.5} \quad \dots (2.33)$$

where $N_G.Sc^{2/3}$ is termed $(N_G)_{Sc1}$ and, as indicated by equation 2.10, is used for comparative purposes. In order to develop this equation it has been assumed that the gas velocity is the superficial velocity in the column.

De Goederen (20) on the other hand assumed that the presence of the liquid will alter the cross-sectional area available for gas flow. For constant liquid hold up

$$N_G \cdot Sc^{\frac{2}{3}} \propto A_o^{-1.0} \cdot (\mu_G / \rho_G)^{0.5} \cdot V_o^{-0.5} \cdot d^{-1.5} \quad \dots (2.34)$$

where A_o is the actual cross-sectional area available for flow, and V_o is the actual gas velocity. He then refers to Mugele (45) for a relationship between drop size, surface tension and velocity for atomisers, and uses

$$d = \text{const.} \cdot \sigma / (\rho_G V_o^2) \quad \dots (2.35)$$

which is obtained from a balance between the inertia forces and the surface tension forces; i.e. Weber number is a constant.

From equations 2.34 and 2.35 de Goederen obtains

$$N_G \cdot Sc^{\frac{2}{3}} \propto A_o^{-1.0} \cdot \rho_G \cdot V_o^{2.5} \cdot \sigma^{-1.5} \quad \dots (2.36)$$

for the condition of constant gas viscosity that prevailed in his investigations.

He then assumes that A_o will depend on the gas hold-up and both will be dependent on the surface tension and the gas load. It follows that N_G will no longer be proportional to $V_o^{2.5} \cdot \sigma^{-1.5}$. This fact is then used by de Goederen to explain the difference between equations 2.30 and 2.36. These equations become identical if

$$A_o \propto V_o^{0.4} \sigma^{-0.2} \quad \dots (2.37)$$

De Goederen's application of equation 2.35 is questionable since Mugele points out that its validity is limited to the case of a droplet acted on only by the drag of the continuous medium. Mugele discards this equation in favour of a dimensional analysis approach. He summarises the factors governing the mean and maximum drop size from sprays with the relationship

$$\frac{\bar{x}}{L'} = A \left[\frac{L' \rho_L V_R}{\mu_L} \right]^B \left[\frac{\mu_G V_R}{\sigma} \right]^C \quad \dots (2.38)$$

where $\bar{x}$ = the mean diameter (Sauter) d or, the maximum diameter $d_{p \max}$.

L' = a characteristic linear dimension of the equipment

V_R = the relative phase velocities at formation.

A, B and C are constants taken from the literature for straight pressure or tangential atomisers. The values are 5.0, -0.35, -0.20 respectively for the mean diameter, and 57, -0.48, -0.18 for the maximum diameter. In an example which is particularly relevant to this work Mugele tested equation 2.38 against the results taken from a grid tray. He obtained close estimates for d and $d_{p \max}$ taking the characteristic linear dimension as twice the slot width.

From equation 2.38 the dependence of the mean drop size on fluid properties and gas velocity can be written as

$$d \propto (\rho_L / \mu_L)^B \cdot (\mu_G)^C \cdot V^{(B+C)} \sigma^{-C} \quad \dots (2.39)$$

In order to isolate the effect of velocity and surface tension it

is convenient to define a reference drop size d^0 such that

$$* \quad d^0 \propto (\rho_L^0 / \mu_L^0)^B \cdot (\mu_G^0)^C \cdot (V^0)^{(B+C)} (\sigma^0)^{-C} \quad \dots (2.40)$$

Combining equations 2.39 and 2.40

$$d' = d \cdot f' \propto V^{(B+C)} \cdot \sigma^{-C} \quad \dots (2.41)$$

$$\text{where } f' = \left(\frac{\rho_L \cdot \mu_L^0}{\rho_L^0 \cdot \mu_L} \right)^{-B} \left(\frac{\mu_G}{\mu_G^0} \right)^{-C}$$

and d' is the drop size corrected for the effects of gas and liquid viscosities and for the effect of liquid density. For the values of B and C given by Mugele for the mean drop size, equation 2.41 can be written

$$d' \propto V^{-0.55} \sigma^{0.20} \quad \dots (2.42)$$

A similar analysis combining equations 2.33 and 2.39 gives the plate performance corrected for the effects of viscosity and density, N_G' . Thus

$$N_G' \propto V^{0.35} \sigma^{-0.30} \quad \dots (2.43)$$

$$\text{where } N_G' = (N_G) Sc^{1.0} g' \quad \dots (2.44)$$

$$\text{and } g' = \left(\frac{\rho_L \rho_G^0 \mu_G \mu_L^0}{\rho_L^0 \rho_G \mu_G^0 \mu_L} \right)^{-0.50} \left(\frac{\mu_G}{\mu_G^0} \right)^{-0.30}$$

Assuming the applicability of equation 2.38 to sieve plate operation in the spray regime, it would appear from equation 2.43 that the

* (where the symbols superscripted 0 are the chosen reference values of the various parameters.)

plate performance increases gradually with gas velocity and decreases gradually with surface tension. The dependence on velocity and surface tension is significantly different from the empirical equation of de Goederen (equation 2.30).

The results of Dombrowski et al (46) (47) for atomisers indicate that the exponent C in equation 2.38 can be in the range -0.16 to -0.33, which gives the relationship

$$N_G' \propto V^n \sigma^m \quad \dots (2.45)$$

where $n = 0.25$ to 0.5

$m = -0.25$ to -0.5

The correction factor g' in equation 2.44 will of course differ slightly if the exponent C in equation 2.38 is altered.

II Fluidised bed model

In his model Andrew (34) assumed that the drop size in the dispersion was reasonably approximated by the maximum drop size possible. However it would seem more correct to take the mean drop size given by Mugele's equation. Combining equations 2.31 and 2.41 for f' equal to unity (i.e. $d' = d$)

$$S \propto V^{0.55} \sigma^{-0.2} \quad \dots (2.46)$$

Chu et al (48) have satisfactorily correlated fluidised bed mass transfer data by plotting the j-factor j_D against a modified Reynolds number. For turbulent conditions they give

$$j_D = \frac{K_G \cdot Sc^{\frac{2}{3}}}{V} = 1.77 \left(\frac{V d \rho}{\mu (1-e)} \right)^{-0.44} \quad \dots (2.47)$$

where e = the voidage.

$$\text{For the velocities considered } e \propto V^{0.42} \text{ ref (49)} \quad \dots (2.48)$$

The number of transfer units $(N_G)_{Sc1}$ is found by

$$(N_G)_{Sc1} = j_D \cdot S \quad \dots (2.49)$$

Combining equations 2.46, 2.47, 2.48 and 2.49 gives

$$(N_G)_{Sc1} \propto V^{0.54} \sigma^{-0.29} \quad \dots (2.50)$$

Comparison of equations 2.50 and 2.45 indicates that the two models predict similar relationships between the performance, the gas velocity and the absolute surface tension.

CHAPTER THREE

APPARATUS

3.1 Introduction

The distillation unit consisted of the reboiler, the column and the condenser system. A line drawing of the unit is shown in Fig. 3.1, and a photograph in Fig. 3.2. At the back of the thesis is a one-eighth scale drawing of the column, showing details of the test plate (DRG.No.1).

Under normal operation vapour, generated in the reboiler, passed into the column via a cyclone-separator and a three-way valve. The separator had been introduced by Bainbridge (50) to relieve the effects of foaming in the reboiler. After flowing through the two-plate column, the vapour was piped by two externally heated vapour lines to the condenser system. This consisted of two water-cooled glass condensers, placed one on top of the other. Reflux emerged at a temperature just below its boiling point, and passed back into the column through a rotameter and a reheater. This electrically heated exchanger was of the shell-and-tube type, with liquid paraffin circulating through the shell, and reflux through the tubes. By adjusting the heat load to the reheater the reflux was returned to the column within 1°C of its boiling point. Liquid, leaving the base of the column, passed through another rotameter and into the reboiler.

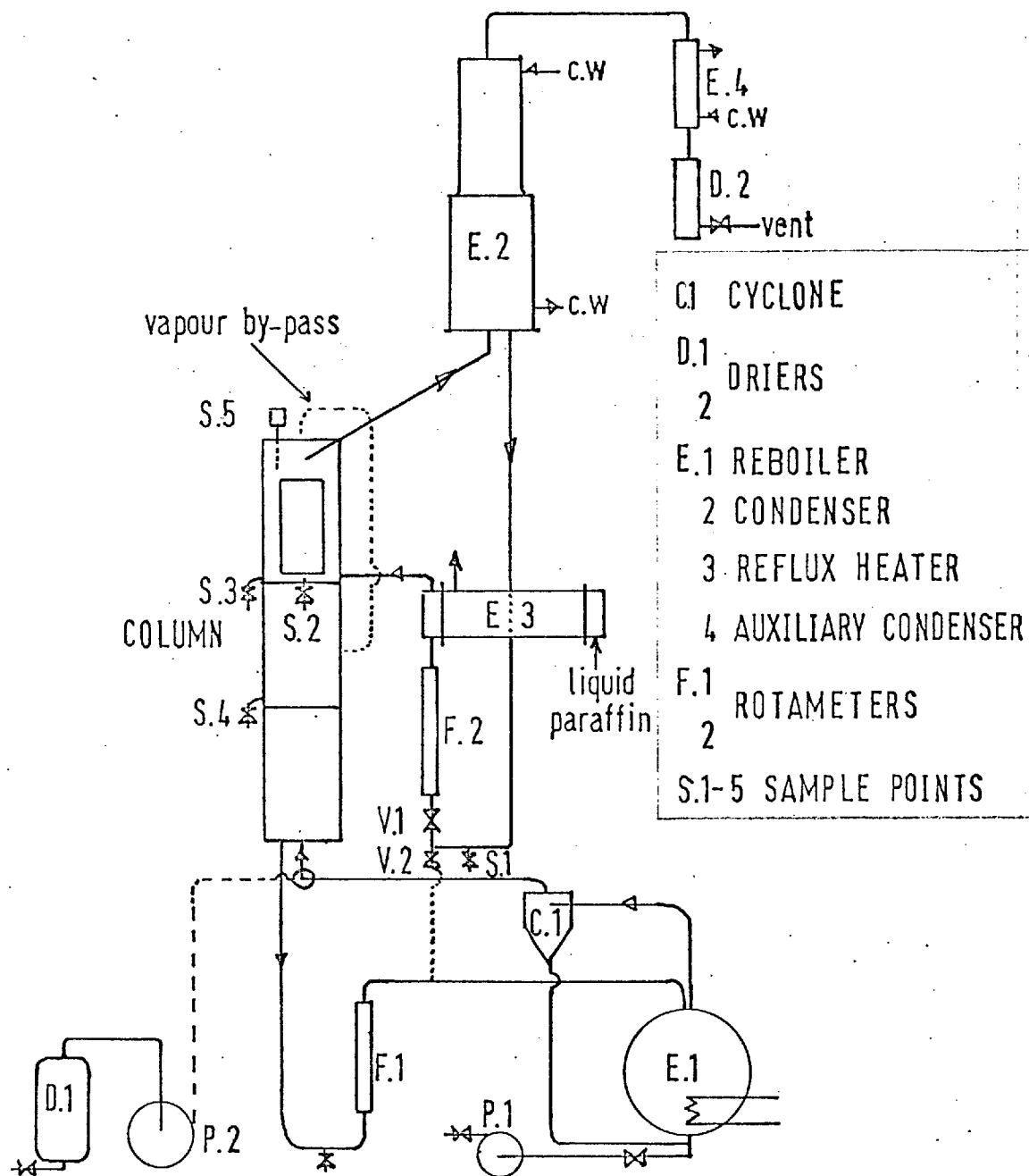


FIG 3.1 LINE DIAGRAM OF DISTILLATION EQUIPMENT

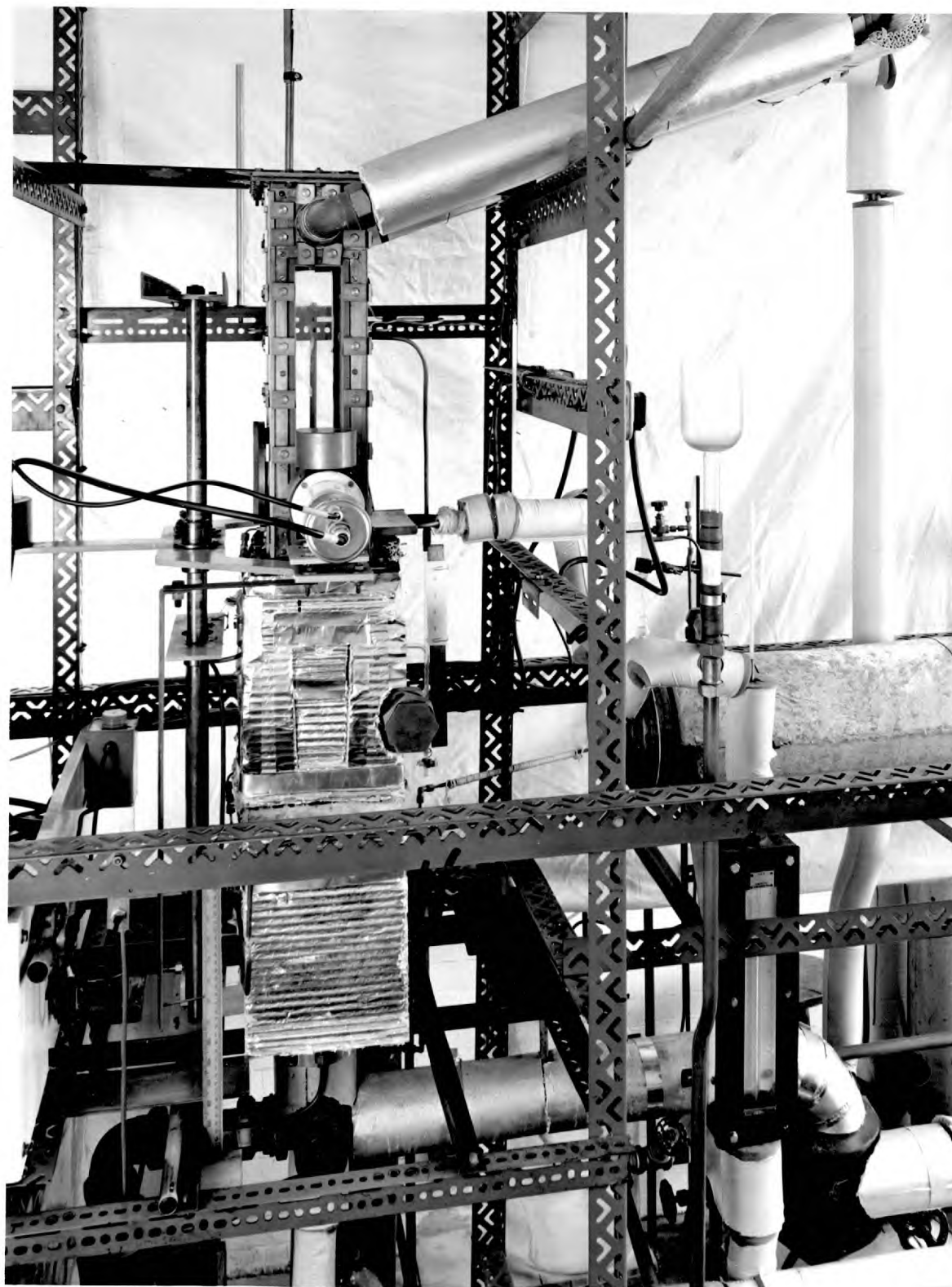


FIG 3.2 DISTILLATION COLUMN

In the design of the unit, particular attention was paid to the requirements of controlling and observing the type of dispersion generated on the test plate. To this end, provision was made for varying the column vapour velocity over a wide range, for altering the plate hold-up, and for taking both photographic and dispersion density measurements of the dispersion from the plate floor upwards. It was decided that the plate geometry should be similar to that used in the previous work (1).

3.2 The Column

The column consisted of three sections with two plates between them, the upper plate (number 1, see DRG.No.1) being the test plate and the lower (number 2) acting as a vapour flow distributor. Both plates had $148 \times \frac{1}{8}$ in. holes on $\frac{3}{8}$ in. pitch giving approximately 10 percent free area (based on bubbling area). The plate geometry is summarised in Table 3.1,

Table 3.1

Plate Geometry	
Hole size	$\frac{1}{8}$ in.
No. of holes	148
Pitch	$\frac{3}{8}$ in. triangular
Free area (i)	10 percent (of bubbling area)
Free area (ii)	7 percent (of superficial area)
Weir heights	$\frac{3}{8}$, $\frac{3}{4}$, 3 ins.
Weir length	7 ins.
Plate length	$2\frac{3}{4}$ ins.

The three sections, numbered 1, 2 and 3, from the top downwards, were of rectangular cross section ($7\frac{1}{2} \times 4\frac{1}{4}$ in. internally) and constructed from angle brass, as shown on DRG.No.1. Side walls of either $\frac{1}{4}$ in. brass plate or $\frac{1}{4}$ in. borosilicate glass were clamped into the framework made by the angle sections. Details are shown in Fig. 3.3.

This mode of construction gave a very flexible design, and speeded up column modifications, when required. Windows could be replaced by brass walls, or vice versa, and the brass walls could be removed and branches or pressure points added without undue delay.

3.2.1 Column sections 2 and 3

The column base plate was made from $\frac{1}{4}$ in. brass plate, $10\frac{1}{2} \times 7\frac{1}{4}$ in., and had vapour inlet, liquid outlet and liquid drain branches. The centrally placed vapour inlet consisted of a 2 in. copper pipe passed up through a specially machined Instantor B.S.P. compression fitting brazed into the plate. It protruded $4\frac{1}{2}$ in. into the column and was capped by an inverted cone section. Even distribution of vapour was aided by the drilling of four 1 in. dia. holes at the top of the pipe, and four $\frac{1}{2}$ in. holes in the cone.

The liquid outlet, a 1 in. dia. copper pipe, 12 ins. long, protruded $1\frac{3}{4}$ ins. into the column through another Instantor fitting. It served as a stand pipe, creating a liquid seal for the downcomer.

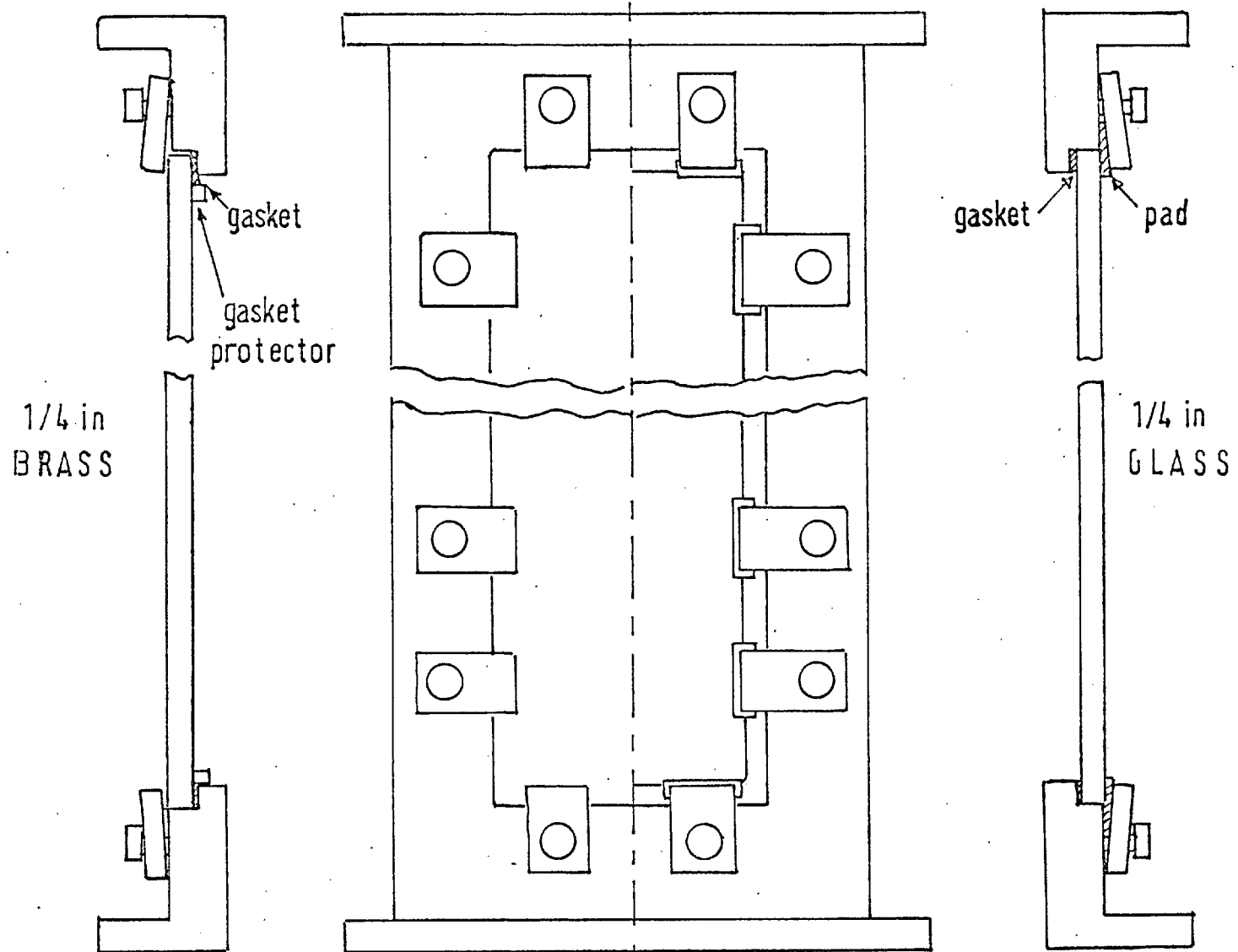


FIG 3.3 DETAILS OF BRASS SIDE PLATE AND GLASS WINDOW FITTING

from the plate above, as well as a free-running downcomer, which prevented entrained vapour being carried through the base rotameter.

A $\frac{1}{4}$ in. 'pet' cock brazed into the base plate provided a drain point. During column operation it was also possible to return samples and sample flushings through this point.

Column section 3, the bottom section, was placed on top of the base plate. The side walls of this section were brass plate, except during the commissioning of the unit when one window was used.

Sieve plate No. 2 was placed in between sections 3 and 2. It was constructed from $\frac{1}{4}$ in. brass plate, and had a downcomer soldered into it. This downcomer, made of 16 gauge copper sheet, was of rectangular cross section and brazed along one edge. The plate geometry was similar to that of the test plate (Table 3.1). Although the inlet and outlet weirs were changeable, they were kept constant, at $\frac{3}{4}$ and $1\frac{1}{2}$ in. respectively, throughout the experiments.

A liquid sample point was sited before the inlet weir, below the test plate downcomer. The samples could be drawn out through a $\frac{1}{16}$ in. drilled hole which ran vertically downwards from the point of sampling, and then horizontally out, through the edge of the plate.

The two narrower sides of section 2 were windowed, whilst the others were brass plate. Into one of these brass side plates were brazed a pressure tapping, and a 2 in. B.S.P. elbow, which was

either blanked off or connected to the vapour by-pass system (section 4.3).

3.2.2 Column Test Section

Details of the test plate can be seen in DRG.No.1. The inverted box construction was necessary because the plate floor had to be raised to the level of the bottom of the window. It was constructed from lengths of angle brass ($1\frac{1}{2} \times 1\frac{1}{2}$ in.) machined and then brazed together, with a recessed ledge on the inside to support the perforated plate, which was soldered in. The perforated plate included a downcomer of similar construction to that used in plate 2. A mid-plate sample point was provided with a 16 gauge pipe carrying the sample down the inside of the box and out through the flange.

The inlet weir was $\frac{3}{4}$ in. high and secured to the plate at three points by 2 BA screws, $\frac{1}{4}$ in. long. Three outlet weirs were used $\frac{3}{8}$, $\frac{3}{4}$ and 3 in. high. They were secured at three points by extended wing nuts, used for ease of removal. Asbestos paste was used to seal the weirs at the sides.

Section 1, which was placed over the test plate was 21 ins. long and had $1\frac{1}{2}$ in. B.S.P. vapour outlet branches either side of the section at the top of the narrower sides. The sides are numbered 1, 2, 3 and 4 as shown in DRG.No.1. Sides 1 and 3 were windowed with borosilicate glass plates, which had a $\frac{1}{4}$ in. rebate machined round their edges. This rebating had the

advantage of decreasing the dead space on the plate, it also meant that the gaskets between the glass and the brass were partially protected from the solvents used in the column.

Side 2, on the downstream side of the plate, was made from $\frac{1}{4}$ in. brass and had a hand hole ($3\frac{1}{2}$ in. I.D.) sited to give easy access to the outlet weir. A sampling cup, positioned above the outlet downcomer was supported by the sample outlet pipe, which passed through a $\frac{1}{4}$ in. Simplifix fitting brazed into the side plate.

The liquid inlet side, 4, was of recessed box construction, aimed to reduce the column superficial area to that of a plate with an inlet downcomer. A liquid inlet manifold packed with small Raschig rings, to promote even flow of liquid over the inlet weir, was clamped to side 4. A pressure tapping on side 4 was connected via a manometer to the pressure tapping in section 2, and measured the overall plate pressure drop.

The top cover of the column was a $\frac{1}{4}$ in. brass plate with a 3 in. dia. window. Removal of this window, and replacement by a specially modified 2 in. B.S.P. elbow, allowed a vapour by-pass to be connected between section 2 and the top of section 1. The use of this by-pass is described in Chapter 4.

The column sections, plates and end covers were bolted together, with $\frac{1}{4}$ in. B.S.F. m.s. bolts through the flanges. Care

was taken to ensure that the test plate was horizontally aligned. All metal to metal, and glass to metal surfaces were separated by gaskets cut from $\frac{1}{8}$ in. thick Lioncelle material, which is a cork base impregnated with neoprene. This material was chosen on consideration of compressibility, resistance to attack by solvents, and cost. It should be pointed out that some trouble was encountered due to swelling and subsequent failure of the gaskets, particularly at the windows and side plates of sections 2 and 3. This was partially cured by soldering a rim of brass strip ($\frac{1}{4} \times \frac{1}{4}$ in.) on the inside of the side plates (see Fig. 3.3), which prevented swelling and reduced attack of the gaskets.

3.3 Ancillary Equipment

3.3.1 Reboiler

This piece of equipment has been fully described by Bainbridge (50). It was 10 in. I.D. by 25 in. long, having a working capacity of 5 gallons. Heat was provided by three 4.6 KW Santon immersion heaters, wired to give infinitely variable input. Two of these three-phase heaters were switched directly, whilst the third had separate switching on each phase with an 8 Amp variac control on one of them.

The reboiler had branches for vapour outlet, liquid return, liquid charging and a pressure relief valve. A liquid drain line, from the bottom, ran through a Tuscan (P.T.F.E.) pump.

3.3.2 Condensers

The two water-cooled condensers, placed one on top of the other, gave a total cooling surface of 30 ft^2 . Details of the units have been given by Bainbridge (50). The main alterations for this work were to the bottom cover. It was decided that the liquid outlet should be of larger diameter, so that it acted as a free-running downcomer. This was done to avoid gas entrainment in the liquid stream which would have interfered with the rotameter readings. To this end the liquid outlet pipe was $1\frac{1}{2}$ in. dia. for the first 18 ins., before reducing down to $\frac{3}{4}$ in. A liquid drain cock was added to the condenser base plate.

The system was vented to the atmosphere through a small auxilliary condenser and a drying column, filled with calcium chloride. The latter prevented moist air from entering the system at shut down.

3.3.3 Reflux heater

This was a 4-pass shell-and-tube exchanger with reflux flowing through the tubes and liquid paraffin heating medium pumped through the shell. Mechanical details have been given by Bainbridge. Heat input was provided by two 2 KW immersion heaters. The paraffin was circulated through the exchanger by a Stuart-Turner No. 25 centrifugal pump.

In the previous work the reflux heater had been mounted vertically, but this was difficult to drain. For this reason it was mounted horizontally and a drain cock added to the bottom of the end-box.

The heating medium was altered from glycerol to liquid paraffin, since the former is corrosive to copper, the material of construction.

3.3.4 Blower

When altering solvent systems it was imperative to clean and dry-out the unit. To assist this a blower was linked into the column base through the 3-way valve. The air was dried by pulling it through a 6 in. dia. column, 18 in. long, packed with calcium chloride pellets.

3.4 Equipment used for data collection

This section deals with the parts of the apparatus and those pieces of equipment associated with the collection of data and other information. The following topics are described:

- (1) Sampling system
- (2) Analysis
- (3) Measurement of dispersion density
- (4) Photographic technique.

3.4.1 Sampling system

The system used is shown schematically in Fig. 3.4. Four liquid sampling points were provided, provision was also made for

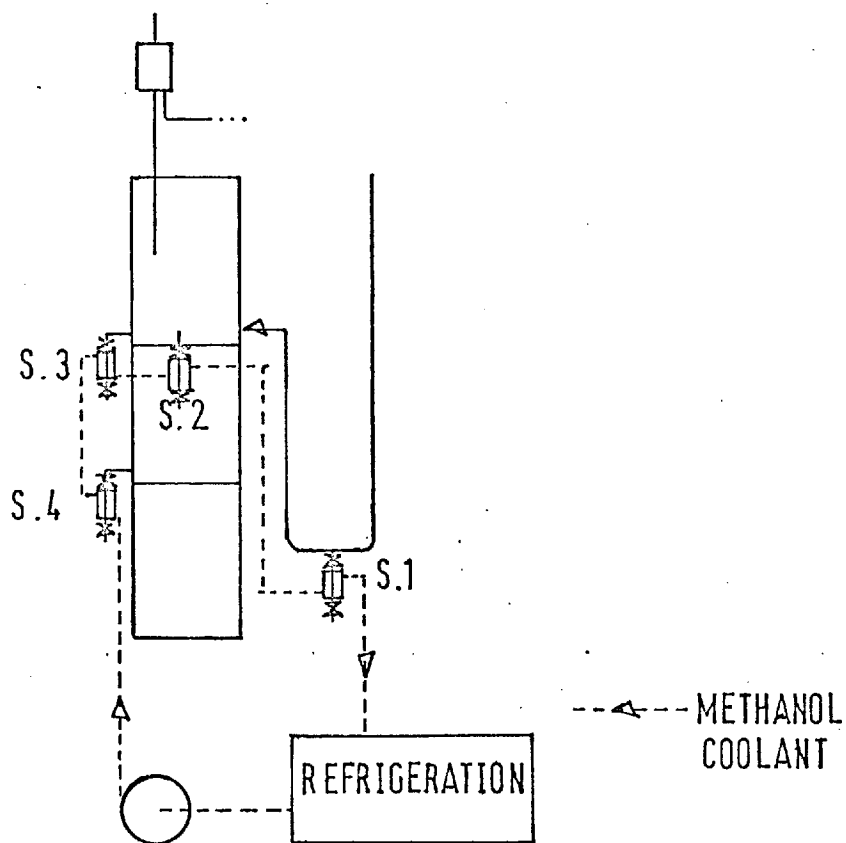


FIG 3.4 SAMPLING SYSTEM (schematic)

sampling the vapour from the test plate. Liquid sampling point 1 was sited in the reflux line, and placed, for convenience, below the reflux rotameter. The mid-plate sample, 2, was not used during the main runs, since it was found during commissioning to be unreliable. The off-plate sample, 3, was used in all the runs and considered to be more reliable than point 4. This latter sample point gave samples of unduly low composition on occasion, probably as a result of the downcomer liquid being contaminated with liquid from plate 2 due to backmixing. Bearing in mind the work of Sheldon and Ellis (51) on the errors introduced by sampling below the downcomer in small columns, it was decided to concentrate on samples S1 and S3 (subsequently labelled x_0 and x_1).

The samples were drawn into a cooling section, where they were held for a number of minutes before running into sample bottles. In all cases great care was taken in flushing through the sample point and ensuring that the samples were 'fresh'.

Refrigerated methanol was circulated through the sample coolers from an A.J. and E. Hall type TV2 Refrigeration unit.

3.4.2 Analysis

Initially two methods of analysis were contemplated, refractometry and gas chromatography. However, it was found that the chromatographic method was as accurate as the refractometer and more convenient, so the bulk of the analyses were carried out on

the chromatograph. Occasional checks were made with the refractometer.

The chromatograph unit chosen was the Perkin Elmer Model 452, with hot wire detector. The separation column was a 2 metre x $\frac{1}{4}$ in. o.d. stainless column packed with DE-301, carbowax 1500, and situated in a constant temperature oven, providing regulation to less than 0.1°C over the range 50 to 225°C . Other features incorporated were pressure regulation of the carrier gas, in this case hydrogen, and a separately heated injection block. In addition a gas sampling valve, Loenco type 208-6V, was situated in the oven, the original intention being to sample the distillation unit continuously, vaporise the sample streams and inject the vapours automatically into the column. Considerable difficulties were encountered in obtaining representative samples under these conditions, the particular problem being to ensure complete vaporisation of the sample. Consequently it was decided to resort to the well-tried method of manual sampling and injection.

The output from the chromatograph was quantified by an electronic integrator, Perkin Elmer Model D2. This highly sensitive piece of equipment was vital to the degree of accuracy demanded in this work. It had a maximum counting rate of 72,000 counts per minute, automatic range change and valley sensor. The latter facility removed much of the subjectivity from the use of

this type of instrument. The valley sensor actuated the printer, a Kienzle digital type, when the input to the integrator had a rate of change greater than a preset tolerance. Thus, automatic print out occurred at the beginning of each peak, and manual print out was necessary after the last peak. A visual record of the peaks was provided by a Hitachi-Perkin-Elmer Model 159 Flatbed recorder.

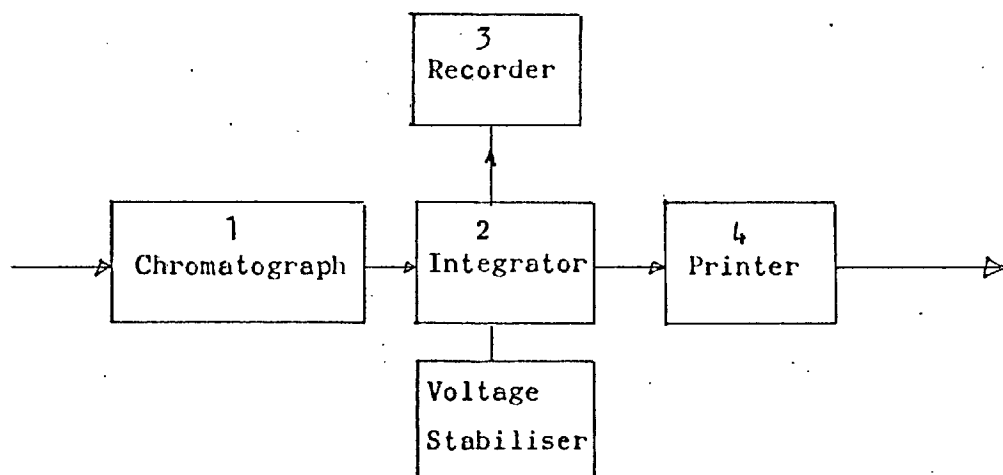
Details of the instrument settings are given in section 4.1, and the system is shown schematically in Fig. 3.5a.

3.4.3 Measurement of dispersion density

The measurement of dispersion density by means of gamma ray absorption has been carried out successfully by a number of workers (30) (38) (39). It was decided to use this approach since it provides information on the vertical distribution of the liquid phase, as well as measuring the total liquid hold up.

1. Traverse system

The gamma ray source and the detector were mounted on the support shown in DRG.No.2. It consisted of a three-armed beam running up and down two $1\frac{1}{2}$ in. dia. steel shafts, mounted vertically and secured top and bottom. The third arm carried a counterweight. An hydraulic ram attached to the base plate and acting through the centre of gravity of the three-armed beam was used to adjust its level. The axis of the source/detector system was carefully aligned to the lateral centre line of the test plate.



1. PERKIN ELMER TYPE 452

2. P.E. TYPE D2

3. HITACHI P E . TYPE 159

4. KIENZLE DIGITAL PRINTER

FIG 3.5.a GAS CHROMATOGRAPH AND ANCILLARIES

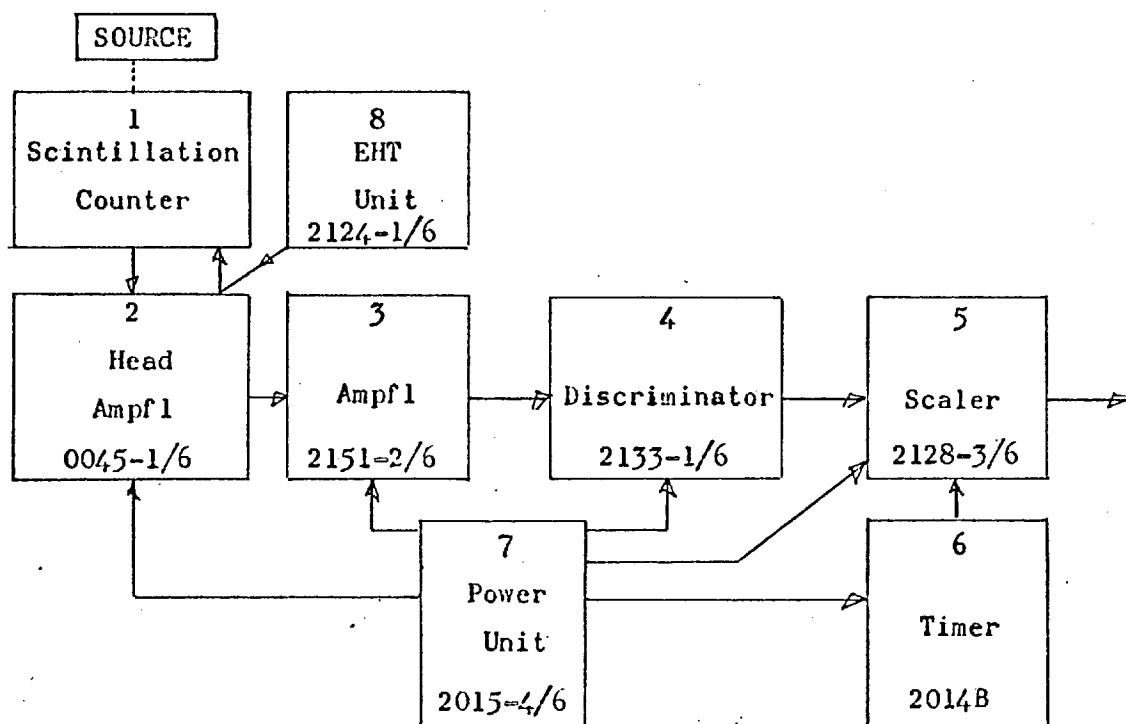


FIG 3.5.b COUNTING EQUIPMENT (HARWELL 2000)

ii. Source and Detector

The source was a one millicurie Caesium 137 slug mounted in a shielding castle, consisting of a lead-filled brass cylinder. A collimating section was designed to give a slightly convergent beam $\frac{1}{8}$ in. deep. The collimating section of the detector was geometrically compatible with the collimating section of the source, and the plate area 'seen' by the rays had a percentage free area similar to the whole plate.

The detector was a Ladgear D4133 directional scintillation counter with built in photomultiplier tube.

iii. Counting Equipment

The output from the scintillation counter photomultiplier was amplified and counted by the units shown schematically in Fig. 3.5b. The signals were boosted close to the detector by the head amplifier. Pulse shaping and the bulk of the gain was provided by the main amplifier, and pulse amplitude selection was made by the discriminator. The fixed value output pulses were counted by the scaler, and the duration of counting time was controlled by the timer.

It was decided to use the Harwell 2000 series counting equipment, which has the advantages of full electronic compatibility, versatility and modular construction.

Details of the final settings are given in section 4.2.

3.4.4 Photographic technique

Photographs of the gas/liquid dispersion were taken throughout the experimental programme. Some initial investigations were carried out with a half plate camera, but for the bulk of the work a Praktika IV, 35 mm, reflex camera was used.

The high speed events occurring demanded the use of a flash unit, and an Ernest Turner Speediflash, type HS/5, was found to be suitable. This gave a 5 microsecond flash at a power output of 5 joules.

Owing to the uneven nature of the image, with areas of high and low light intensity it was necessary to place graded filters between the light source and the image, and between the image and the camera. The position of the filters depended on the height and type of dispersion photographed.

A number of films were tried, and the bulk of the work was done on Ilford Pan F, which was developed in the normal way.

CHAPTER FOUR

EXPERIMENTAL

4.1 Introduction

This chapter covers the following topics:

- (1) Calibration of the Chromatograph
- (2) Calibration of the gamma-ray traverse system
- (3) Column erection and testing
- (4) Typical operating procedure
- (5) Experimental programme.

4.2 Calibration of the chromatograph

4.2.1 Instrument settings

The adjustable parameters were carrier gas flow, oven temperature, hot wire detector temperature, sensitivity, and sample volume. Bearing in mind the needs for maximum sensitivity, and fairly short retention time, the system was operated under the conditions given in Table 4.1. These conditions were chosen after extensive trials.

Table 4.1

Carrier gas (hydrogen) pressure	5 psig
Oven temperature	100°C
Hot wire detector setting	4
Injection block temperature	125°C
Sensitivity	1 (max)
Sample volume	1 μ litre

4.2.2 Measurement of the substance-specific correction factors

In order to convert the ratio of peak areas of a given mixture to the weight ratio, it was necessary to measure the substance-specific correction factors. According to Kaiser (52) this factor is given by

$$\frac{W_i}{W_j} = \frac{f_i \cdot A_i}{f_j \cdot A_j}$$

where W_i/W_j = the weight ratio;

A_i/A_j = the peak area ratio;

and f_i and f_j = the respective correction factors.

It is standard practice to take f_1 (benzene) equal to unity. This convention was followed, although for the system n-heptane/toluene, f_1 (toluene) was taken as unity.

No data were available on the variation of f_j with composition, and it was necessary to measure this for the three systems used.

Calibration mixtures were made up at a number of concentration levels and analysed. It was found that there was a non-linear dependence of f_j on concentration. Using the calibration technique described in Appendix I it was possible to obtain highly reproducible results.

The results are shown in Appendix I. Also included is an analysis of the effect of errors in f_j on the calculated weight fraction W .

4.3 Calibration of the gamma ray traverse system

The technique of dispersion density measurement by gamma ray absorption has been discussed by Macmillan (38) and others. The essential fact is that the log of the count rate can be taken as a linear function of the fraction of liquid present. This means that for a given liquid and vapour mixture, only two points are required for a calibration, the count rate for a vapour filled column, n_0 , and that for a liquid filled column, n . A straight line can then be drawn on semi log paper through the points ($\ln n_0$, $\bar{F}_L = 0$ and $\ln n$, $\bar{F}_L = 1$) where $\bar{F}_L$ is the volumetric fraction of liquid.

Prior to taking readings with the system it was necessary to find the optimum settings for each module in the counting equipment. The voltage to be applied to the photomultiplier was selected after obtaining a count rate/voltage profile. A count rate plateau was evident between 850 and 1050 volts, so the applied voltage was chosen at 950 volts. Pulse shaping was carried out using an oscilloscope, the final settings are given in Table 4.2.

Since the source/detector system was best calibrated in situ facilities for providing a vapour-free and a liquid-free dispersion were required. A vapour by-pass from column section 2 to the top of section 1 could be inserted, and a 1/16 in. thick brass sheet, cut to size, was placed over the bubbling area. Using the highest weir it was possible to build up a 3 in. deep pool of liquid on the plate, with the vapour taking the path of

least resistance, through the by-pass.

Table 4.2
Counting Equipment Settings

E.H.T.	950 volts
Head amplifier time constants	0.02 μ sec
Main amplifier: gain	65.5 dB
differentiating t.c.	1 μ s
integration t.c.	1 μ s
Discriminator: level	0.16 V
dead time	10 μ s

By closing V1 (Fig. 3.1) and opening V2, the reflux could be returned directly to the reboiler and vapour only passed through section 1.

Since these calibrations required some modifications to the column they were only carried out at three composition levels for each binary system. As the results in Appendix II indicate, the variation with composition was regular.

A vertical traverse was carried out from below plate level to 10 inches above plate level, with an empty column, to check for irregularities in the windows and to establish the base line for the subsequent traverses. Intermittent checks were made on the base line to detect any misalignment.

4.4 Column erection and testing

Care was taken when assembling the column to ensure that the test plate was horizontal. Attention was also paid to the levelness of the test plate inlet and outlet weirs. Both of the rotameters were carefully mounted to ensure verticality.

Before erecting, the column sections were tested under 15 psig. hydraulic pressure. Further tests were carried out on the assembled unit with nitrogen at 10 psig. However under distilling conditions it proved impossible to eliminate all the vapour leaks. Since the highly toxic benzene was used for a number of experiments it was necessary to keep a close check on the benzene level in the laboratory, using Drager tubes. The level never exceeded 10 p.p.m.

During the commissioning of the unit observations were made on the response of the column to changes. It was found that the response times were very short and that the compositions levelled out within half an hour. Sampling did not appear to affect the subsequent performance.

4.5 Typical operating procedure

Before charging the reboiler with a new system, the whole unit was blown through with dry air for a number of hours. Approximately five gallons of mixture were charged. The condenser water, the reboiler heaters and the reflux heater and pump were

switched on. When vapour reached the vapour pipe to the condensers, the pipe heaters were also switched on. The reboiler heat load was adjusted to give the desired column F factor. After approximately one hour liquid samples were taken. The procedure followed was to switch on the refrigerated methanol circuit a few minutes before sampling, then the sample points were slowly flushed and the required sample was held in the sample coolers for approximately two minutes. Sampling was carried out from the base upwards to minimise the effect of sample-taking on the subsequent samples. The chilled liquid was then run off into small 10 cc glass stoppered bottles which were partially immersed in iced water. Samples were then injected into the chromatograph, care being taken to ensure that the integrator was zeroed before injecting each sample.

Sampling was repeated every 20 to 30 minutes, as convenient, until consecutive sample compositions were within 0.1 mole percent or the calculated efficiencies were within 2%, whichever was the lesser.

Dispersion density measurements were carried out concurrently with the sampling. The source/detector level was incremented 0.05 to 0.3 ins. depending on dispersion and the position being measured. Counting was stopped after 100 seconds by the Timer and the number of counts was displayed on the Scaler. These instruments were zeroed and reset manually.

When the vertical traverse was completed the source was plugged with a lead plug and then raised as high as possible; this enabled the camera, flash unit and filters to be swung into position. When the photographic record had been made the camera was swung back and the source/detector system lowered.

The boil-up rate was then adjusted to the next level and the procedure was repeated. Samples and sample flushings were returned when the flow rate was changed.

4.6 Experimental programme

Three binary systems were studied. These were,

System A, Benzene/Cyclohexane

System B, Benzene/n-Heptane

System C, n-Heptane/Toluene.

System A forms an azeotrope at approximately 50% benzene, and is weakly surface-tension negative below and weakly positive above the azeotrope. System B is strongly surface-tension negative and C is strongly positive.

Experiments were conducted, as described in section 4.5, across the composition range for each system. At each composition measurements were made at approximate F_H factors of 7.5, 10.0, 12.5 and 15 ($(\text{lbs})^{\frac{1}{2}}/(\text{ft}.\text{sec})^{\frac{1}{2}}$). In S.I. units these become 9.15, 12.2, 14.9, 18.3.

For System A the weir heights used were 3 ins. (7.6 cm.) and $\frac{3}{4}$ in. (1.9 cm.). For System B, as above, and in addition $\frac{3}{8}$ in. (0.95 cm.). For System C, weir heights of $\frac{3}{4}$ and $\frac{3}{8}$ in. were employed.

The results from this experimental programme are tabulated in Appendix VI.

CHAPTER FIVE

PHYSICAL PROPERTIES OF THE BINARY SYSTEMS

5.1 Equilibrium data

The vapour-liquid equilibrium data for benzene/cyclohexane (System A) and benzene/n-heptane (System B) have been measured by Bainbridge (50), who also carried out extensive thermodynamic consistency tests on his results.

System C, n-heptane/toluene has been investigated by a number of workers. Kogan et al (53) list the data from eight different sources. The data of Black (54) were chosen following consistency tests.

Hala et al (55) and Ellis et al (1) recommend the use of the Redlich-Kister-Herington (56) test for isobaric data. The basis of this test is the relationship

$$\int_0^1 \log \frac{\gamma_A}{\gamma_B} \cdot dx_A = I \quad \dots (5.1)$$

where $\gamma_{A,B}$ are the activity coefficients and x is the mole fraction. If $\log(\gamma_A/\gamma_B)$ is plotted against x_A the area A_1 above, and A_2 , below the abscissa can be determined. The value of the integral I , is the sum of A_1 and A_2 , paying regard to sign; i.e. taking A_2 as a negative quantity.

According to Herington the data are consistent if

$$\text{at } dP = 0, \quad \sum \frac{100I}{T} < \frac{150Z}{T_{\min}} \quad \dots (5.2)$$

$$\begin{aligned} \text{where} \quad I &= A_1 - A_2 \\ \Sigma &= A_1 + A_2 \\ Z &= T_{\max} - T_{\min} \text{ (K}^\circ\text{)} \end{aligned}$$

$$\begin{aligned} \text{If} \quad D &= 100I / \Sigma \\ \text{and} \quad J &= 150Z / T_{\min}, \end{aligned}$$

equation 5.2 can be rewritten,

$$D < J \quad \dots (5.3)$$

In addition Herington stated that the data could still be considered correct if equation 5.3 was not obeyed but if

$$D - J < 10. \quad \dots (5.4)$$

A computer program was written to carry out this test.

Values of $\log(\gamma_A/\gamma_B)$ were computed from the literature data using the vapour pressure data of Bainbridge. A polynomial of degree four was fitted, by the method of least squares, to the $\log(\gamma_A/\gamma_B)$ versus x_A data. This polynomial was then integrated to give A_1 and A_2 , and the criteria of equations 5.3 and 5.4 were applied. The chosen data of Black were also checked by hand.

The results of this test and the Herington tests carried out by Bainbridge on systems A and B are given in Table 5.1. In all three cases equation 5.3 was found to be applicable. The equilibrium data used during this work are tabulated in Appendix III.

Table 5.1

Results of the Herington Test

System	A ₁	A ₂	D%	J
A	7.55	7.40	+1.0	0.3
B	7.20	7.56	-2.4	7.8
C	6.79	6.98	-1.4	4.9

5.2 Liquid properties5.2.1 Surface tension, σ

The surface tension at boiling point for systems A and B has been measured by Bainbridge across the concentration range. The only data available for system C, at the boiling point were those for the pure components, n-heptane and toluene. These data were also taken from Bainbridge. From knowledge of these terminal values it is possible to estimate the surface tension of mixtures using the equation

$$(\sigma_{AB})_T = x_A(\sigma_A)_T + x_B(\sigma_B)_T \quad \dots (5.5)$$

The pure component surface tensions are taken as the values at the required temperature, as suggested by Ellis (57).

The surface tensions at the boiling point of the mixtures are plotted in Fig. III.1 (Appendix III).

5.2.2 Density, ρ_L

The densities at boiling point for mixtures of system A and B, and for pure n-heptane and toluene have been reported by Bainbridge. The densities of mixtures of system C, at boiling point, were calculated from the densities of the pure components from the formula

$$\rho_L = \frac{\rho_A \rho_B M_{AB}}{M_B x_{B A} + M_A x_{A B}} \quad \dots (5.6)$$

where ρ_A, ρ_B = densities of the pure components at the desired temperature,

M_A, M_B = molecular weights of the components,

M_{AB} = average molecular weight = $M_A x_A + M_B x_B$,

x_A, x_B = mole fractions.

This assumes no volume change on mixing. The values obtained are plotted in Fig. III.2.

5.2.3 Viscosity, μ_L

Bainbridge has measured the viscosity/temperature data for systems A and B, and for pure n-heptane and toluene. The viscosity of mixtures of system C at the boiling point, have been calculated from the pure component data by the relation put forward by Kendall (58)

$$\log \mu_L = x_A \log \mu_A + x_B \log \mu_B \quad \dots (5.7)$$

where μ_A , μ_B = viscosities of the pure components at the desired temperature.

The values are plotted in Fig. III.3.

5.2.4 Diffusivity, D_L

The diffusivity for dilute solutions was calculated using the method of Wilke and Chang (59), as recommended by Gambill (60)

This is:

$$\frac{D_{L12}^0}{T} \mu_2 = \frac{(7.4 \times 10^{-8})(M_2)^{\frac{1}{2}}}{V_1^{0.6}} = \frac{1}{F_1'} \quad \dots (5.8)$$

where D_{L12}^0 = liquid diffusivity of 1 in 2, for dilute solution of 1 in 2.

k = 'association parameter' - used to define the effective molecular weight of the solvent, with respect to the diffusion process (= 1 for non-associated solvents).

M_2 = solvent molecular weight.

μ_2 = solvent viscosity (cp).

V_1 = solute molar volume at its normal boiling point (cc/g.mole)

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

Equation 5.8 gives different values for D_{L12}^0 and D_{L21}^0 which means that the diffusivity for mixtures of 1 and 2 will have to take into account the existence of bulk flow if it is assumed that the heat

transfer conditions at the interface demand equimolar counter diffusion.

As shown in Appendix IV the diffusivity of a mixture of 1 in 2, at a composition x_1 is given by:

$$D_L = D_{L12}^0 (1 - (1 - (D_{L21}^0 / D_{L12}^0)) \cdot x_1) \quad \dots (5.9)$$

The variation of the liquid diffusivity with composition at the boiling point has been plotted in Fig. III.4.

5.3 Vapour Properties

5.3.1 Vapour pressure data

The data are taken from Bainbridge. He correlates his results in terms of the Antoine (62) equation

$$\log_{10} P^0 = A - (B/(T' + C)) \quad \dots (5.10)$$

where P^0 = the vapour pressure in mm.Hg,

T' = temperature in $^{\circ}\text{C}$

A, B, C are constants. 4

The constants for the Antoine equation, calculated from the data of Bainbridge, are given in Table 5.2.

Table 5.2

Constants for the Antoine equation

	A	B	C
Benzene	6.998	1258	225.4
Cyclohexane	6.849	1215	225.4
n-Heptane	6.872	1262	217.8
Toluene	6.932	1330	217.8

5.3.2 Viscosity, μ_G

In the absence of any literature data it was necessary to calculate this property. According to Gambill (63), who reviewed the various methods available, that of Arnold (64) is the most reliable. This is

$$\mu_G^o = \frac{27 M^{\frac{1}{2}} T^{\frac{3}{2}}}{V_b^{\frac{2}{3}} (T + 1.4 T_b)} \quad \dots (5.11)$$

where V_b is the liquid molar volume at the normal boiling point.

Equation 5.11 was used to calculate the vapour viscosities of the pure components. For mixtures, the formula of Wilke (65) was used

$$\mu_G = \frac{\mu_A}{1 + \frac{x_B}{x_A} \epsilon_{AB}} + \frac{\mu_B}{1 + \frac{x_A}{x_B} \epsilon_{BA}} \quad \dots (5.12)$$

$$\text{where } \epsilon_{AB} = \frac{(1 + (\mu_A/\mu_B)^{\frac{1}{2}} (M_B/M_A)^{\frac{1}{4}})^2}{2\sqrt{2} (1 + M_A/M_B)}$$

and ϵ_{BA} is similarly defined.

The variation of vapour viscosity with composition for the three systems at the condensation point is plotted in Fig. III.3.

5.3.3 Vapour density, ρ_G

The vapour density at condensation point was calculated from the ideal gas law. The variation of this property with composition is plotted in Fig. III.2.

5.3.4 Diffusivity, D_G

The diffusion coefficient has been calculated by the equation due to Arnold (66), as recommended by Gambill (57).

This equation is

$$D_G = 0.00837 \frac{T^{5/2}}{(T + C)} \frac{(1/M_A + 1/M_B)^{1/2}}{P(V_A^{1/3} + V_B^{1/3})^2} \quad \dots (5.13)$$

where C is given by

$$C = F \sqrt{C_A \cdot C_B}, \quad F = f(V_A, V_B)$$

Arnold presents a table for F in terms of V_A/V_B . The constants C_A and C_B can be calculated from

$$C_i = 1.47 T_{bi} \quad \dots (5.14)$$

The variations of vapour diffusivity with composition for the three systems at their condensation points are plotted in Fig. III.5.

CHAPTER SIX

RESULTS

6.1 Introduction

As discussed previously the factors determining the dispersion type and corresponding mass transfer performance of a given sieve plate are,

- (i) vapour and liquid physical properties;
- (ii) vapour and liquid flow rates;
- (iii) column geometry.

In this investigation attention has been focussed on the dominant factors within each group. Thus the dispersion types, characterised by average and local values of dispersion density, liquid hold up and photographic evidence, and the mass transfer performance are discussed in terms of their dependence on,

- (i) surface tension gradients, absolute surface tension and composition effects.
- (ii) Hole vapour velocity, expressed as $F_H (=V_H \sqrt{\rho_G})$.
- (iii) Weir height.

It is evident that these factors interact strongly.

6.2 Calculation of data

6.2.1 Liquid hold-up

The liquid hold-up, expressed in cms., of clear liquid, is obtained by integration of the dispersion density/height curves,

normally termed dispersion profiles. This integration can be done by drawing the profile and counting squares, by application of the trapezoidal rule or by fitting a polynomial to the profile and integrating directly. Since the data were handled by the computer it was relatively simple to try out the two latter procedures, and use the time consuming by-hand approach as a check.

A library program incorporating the method of least squares was used to fit a polynomial to the dispersion density profile. The degree of the polynomial was varied to see how this affected the result. Table 6.1 shows the results obtained for run AIVI, and includes the values obtained from the trapezoidal rule approach, and from calculations by hand.

Table 6.1

Liquid holdup from integration of dispersion density profile		
Polynomial of degree	1	2.601 cms.
"	" 2	2.589
"	" 3	2.533
"	" 4	2.543
"	" 5	2.550
Trapezoidal rule		2.548
Counting squares		2.557

As these and other analyses indicate, the results obtained from the straightforward trapezoidal rule are acceptably close to the hand calculated values. In consequence the integrations were carried out by the trapezoidal rule, with considerable reduction in computing time. Occasional checks were made by counting squares, agreement always being within 3%.

6.2.2 Column performance

The plate performance is expressed in terms of the Murphree plate efficiency, based on the vapour phase

$$E_{MV} = (y_1 - y_2)/(y_1^* - y_2) \quad \dots (6.1)$$

which is derived from equation 2.5 with the subscript n equal to 1. At total reflux the composition of the vapour leaving the plate is the same as the composition of the liquid entering the plate, i.e. $y_1 = x_o$. Similarly, $y_2 = x_1$, and equation 6.1 can be written

$$E_{MV} = (x_o - x_1)/(y_1^* - x_1) \quad \dots (6.2)$$

where y_1^* is in equilibrium with x_1 .

The modified plate efficiency of Bainbridge (50) which was based on the midplate driving force was not used, owing to inconsistencies in the measured value of the midplate composition, x_p . Under certain conditions x_1 was found to be slightly greater than x_p , implying that some liquid by-passing was occurring. It is of interest to note that similar inconsistencies have been observed

elsewhere (Delaware Report (15), P.56). Although this behaviour was confined to the lower vapour rates it was decided to use E_{MV} , as defined by equation 6.2, throughout this work.

6.3 Factors affecting column performance

6.3.1 Composition and surface tension properties

The column performance results are presented graphically in Figs. 6.1 to 6.9, and are tabulated in Appendix VI.

The variation of efficiency with composition is seen to be influenced by weir height and hole F factor. As observed by Zuiderweg and Harmens (6) the most significant conditions for the variation of efficiency with composition are those in which the mass transfer surface is present as the lamellae between bubbles in a cellular foam dispersion. These conditions prevail at low hole velocities and with surface tension positive systems, as discussed in section 2.3.1. In Fig. 6.1 the azeotropic system A is seen to exhibit a distinct maximum in the efficiency-composition curve for the lowest F_H value, and a weir height of 1.9 cms., on the 'positive' side of the azeotrope. Similar behaviour is shown by the positive system C, Fig. 6.5.

Significantly the composition effect is almost absent when the weir height is increased to 7.6 cms., Fig. 6.2. Under these deep-pool conditions it would appear that the transfer surface is less dependent on the surface tension gradients for its effective magnitude.

Under conditions which do not favour the build^{up} of a cellular foam, the composition effect is less than in the cellular foam, but it is still in evidence. Thus the negative systems A $\ominus$ and B at all hole velocities, and the positive systems A $\oplus$ and C at the higher hole velocities, all show some composition effect; see Figs. 6.1, 6.3, 6.4, 6.5 and 6.6. The interaction between the surface tension gradient effect and the hole velocity is most significant, as can be seen from comparing Figs. 6.7 and 6.8. Thus at the low F_H values, the positive system C has an appreciably higher efficiency than the negative system B, as would be expected from the criteria of Zuiderweg and Harmens (6). However at the highest F_H values these trends appear to be reversed, in accord with the findings of Bainbridge and Sawistowski (2). Since the dispersion type at the highest F_H is spray it is relevant to apply equation 2.43, which suggests a dependence of plate performance on the absolute surface tension. Performance data for systems B and C over the composition range 20 to 100%, have been considered. The constraints applied to the composition are precautionary, since the systems are here assumed to be gas phase controlled. The importance of the liquid phase resistance is known to increase as the slope of the equilibrium line m increases (equation 2.1) and this occurs at the lowest compositions.

It is assumed that the plate is well mixed so that the measured plate efficiency E_{MV} can be taken as equal to the point efficiency E_p . Applying equation 2.4 the performance can be expressed as N_{OG} . If the liquid phase resistance is assumed to be negligible N_{OG} is well approximated by N_G , the number of gas phase transfer units. Before comparisons can be made it is necessary to apply equation 2.10, the Schmidt number correction, so the corrected performance is written

$$(N_G)_{Sc1} = N_G \cdot Sc^{\frac{2}{3}} \quad \dots (6.3)$$

where $(N_G)_{Sc1}$ is the plate performance corrected to a Schmidt number of unity. In addition since the fluid viscosities and the liquid densities are varying it is necessary to standardise $(N_G)_{Sc1}$ by applying equation 2.44. Moreover equation 2.43 permits correction for differences in superficial vapour velocity; a value of 1 m/s being taken as reference. In Figure 6.9 the standardised values N'_G are plotted against the absolute value of the surface tension. It is seen that the negative system B, gives a higher performance than the positive system C, and that for both systems the performance decreases with increase in surface tension. The actual dependence is similar for the two systems:

$$\text{Negative system B, } N'_G \propto \sigma^{-1.10}$$

$$\text{Positive system C, } N'_G \propto \sigma^{-0.90}$$

If allowance is also made for the fact that the liquid hold-up is not quite constant, as assumed when developing equation 2.43 the dependence becomes

$$\text{System B, } N_G' \propto \sigma^{-0.85} \quad (\text{at constant hold-up})$$

$$\text{C, } N_G' \propto \sigma^{-0.75} \quad (\text{at constant hold-up})$$

With or without correction for the variation in hold-up the performance appears to be considerably more dependent on surface tension than suggested by equations 2.43 or 2.45. The exponent appears to be closer to the empirical value of -1.0 found by Bainbridge and Sawistowski (2) and by de Goederen (20). It should be pointed out that in the former case no correction was made for hold-up, and in the latter the velocity correction was $V^{1.5}$ rather than the $V^{0.3}$ correction used in this work.

6.3.2 The effect of hole velocity

The influence of this parameter is dependent on surface tension properties and weir height. From Figs. 6.1 and 6.2, azeotropic system A, it is seen that the plate efficiency decreases with vapour rate for both positive and negative systems at the weir heights 7.6 and 1.9 cms. At the higher weir height, the deep pool condition, the efficiency is strongly dependent on hold F factor. It is noted that at the lower F_H values the deep pool condition tends to exhibit weeping with some diminution in plate efficiency.

With the 1.9 cm. weir and at low F_H the positive systems, A(+) and C show a marked dependence of efficiency on F_H , Figs. 6.1 and 6.5; this behaviour being closely mirrored by the liquid hold-up data and photographic observations, both of which are discussed later.

In contrast the plate efficiencies of the negative systems A(-) and B, at the intermediate weir height, are almost independent of hole F factor over the range investigated, Figs. 6.1, 6.3 and 6.4. System A(-) shows a greater fall in efficiency than system B, which may be a consequence of the higher surface tension negativity of the latter system.

Vapour rate effects are evident at the lowest weir height, 0.95 cms., and depend on the surface tension properties. For the negative system B the plate efficiency increases with F_H , for the positive system A it falls to a minimum and starts to rise again.

The behaviour is most clearly illustrated in Figs. 6.7 and 6.8, which compares the results for system B and system C at conditions of similar absolute surface tension. Over the velocity range studied the positive system shows a reduction in efficiency with an increase in F_H , the exceptions are mentioned below. On the other hand the negative system shows little or no fall off of efficiency with F_H , and a noticeable increase at the lowest weir height. The other significant fact is that at the higher values of F_H the plate efficiency tends to become independent of weir height;

a finding similar to Yoshida et al (8) working with a high capacity sieve plate column. This independence of weir height applies to both systems.

Finally it is worth noting the behaviour of the positive system C at the highest composition studied, ca 92%. From Fig. 6.5 it can be seen that at this composition and with a weir height of 1.9 cms. there is very little variation of efficiency with F_H factor. Moreover at a weir height of 0.95 cms. the efficiency is found to increase with F_H ; Fig. 6.6. In other words at this composition the positive system is behaving as a negative, or neutral system.

6.3.3 The effect of weir height

The general trends of the variation of plate efficiency with weir height have been established in sections 6.3.1 and 6.3.2. The most graphic illustrations are shown in Figs. 6.7 and 6.8. For both positive and negative systems, at the lower values of F_H , reduction in weir height produces a reduction in plate efficiency. This type of behaviour is well known (8) (41). However the magnitude of the reduction is here shown to be less for the negative system than the positive system. Of greater significance is the gradual reduction in importance of the variation of weir height as F_H increases. At the highest measured F_H the plate efficiency appears to be virtually independent of weir height, for the intermediate and low weir heights. Under these conditions of hole velocity and weir height the dispersion is in the spray regime,

and, as discussed in Section 2.3.2.ii, it is to be expected that the performance becomes almost independent of weir height. This agrees with the findings of Yoshida et al (8).

For a weir height of 7.6 cms. the dispersion type is agitated froth even at the highest F_H , which explains the non-convergence of the efficiency curve in Fig. 6.8. It would appear that once the spray regime is reached, decreasing the weir height has no effect on the plate performance.

6.4. Factors affecting the dispersion

The data taken to characterise the type of dispersion are represented in Figs. 6.10 to 6.24, and the dispersion density measurements are tabulated in Appendix VI. It is apparent that the parameters which affected the plate performance have a parallel effect on the hydrodynamic behaviour. Although a number of other workers (30) (38) (39) (69) have reported dispersion density data it is not easy to compare their findings with this work. Firstly, the bulk of the reported work was taken from sieve plates with foam baffles; the positioning of this baffle was shown to be critical by Sanadaji (69). Secondly all the investigations were carried out on surface tension neutral systems, either air/water (30) (69) or pentanes (38) (39), deliberately chosen to exclude surface tension gradient effects. Nevertheless there is some common ground which will be noted.

It is advantageous to discuss the hydrodynamic data in two sections:

- (i) Dispersion density, including profiles and mean values.
- (ii) Liquid hold-up data.

6.4.1 Composition and surface tension effects

6.4.1.i Dispersion density

It is found that positive and negative systems give markedly different dispersion density profiles for the same F factor under certain conditions of weir height and flow rate.

For deep pool conditions (Figs. 6.13 to 6.15) there is no obvious difference between the profiles of the positive and negative systems. This accords with the photographic observations and the fact that the mass transfer performance is independent of the surface tension gradients.

At the lower weir heights, and particularly at the lower flow rates the differences emerge (Figs. 6.10a to 6.12b). It is noted that the azeotropic system A, Fig. 6.10, is not very representative of the high F_H values, since the data for A(+) is not taken at a sufficiently high F_H . A clear comparison is given in Figs. 6.11a and 6.12a, which show profiles at the 1.92 cm. weir heights for the systems B and C respectively. At the lowest F_H , the positive system has a profile of almost constant slope, which is observed to coincide with the cellular foam regime (see photograph,

Fig. 6.25). This correspondence of the constant slope zone with the cellular foam has previously been made by Sanadaji (69) and Schroter (30). In contrast the negative system at the intermediate weir height tends to give a zone of constant density from the 0.5 to the 2.0 cm. level, tailing off fairly rapidly. The constant density region was not observed by Schroter at such low vapour rates. This zone coincides with a non-cellular foam, or froth (Figs. 6.29 and 6.30). Further evidence of this basic difference between the positive and negative systems is shown in Fig. 6.16, where the mean dispersion density, derived from dispersion height and liquid hold-up, is plotted against F factor. As discussed in section 2.3.1, and illustrated in Fig. 2.2, the variation of mean dispersion density with gas velocity will be different for foaming and non-foaming systems. The foaming systems will exhibit cellular foam formation at low vapour velocities, followed by froth and then spray formation as the vapour velocity is increased. The mean dispersion density will pass through a minimum followed by a maximum as the vapour velocity is increased from a low value. In Fig. 6.16 it can be seen that for the three mid-range compositions of the positive system the mean dispersion density passes through a maximum indicating the transition from cellular foam to froth. For the two extremes of composition the mean dispersion density curves are similar to those of the negative system. In this case the behaviour is as shown by curve I in Fig. 2.2, which shows the

variation of mean dispersion density with velocity for the condition in which no cellular foam is formed and the regimes change gradually from froth to spray.

These results illustrate the difference between the two systems, particularly at the low vapour rates. What is also important is that the behaviour of the positive system at the extremes of composition is that of a non-foaming system. This is in qualitative agreement with the comments on the effect of hole velocity on performance for the positive system at the highest composition, discussed at the end of section 6.3.2.

Returning to the dispersion density profiles, Figs. 6.11b and 6.12b, it appears that at the 0.95 cm. weir height and for the lowest F_H there is less obvious difference between the positive and negative systems. However the negative system does show a slight kink in the profile at the 1 cm. level, the positive system gives a smooth curve. In addition the dispersion heights differ, 2 cm. and 3 cm. respectively at this weir height and vapour rate. This difference is reflected in the mean dispersion density values (Fig. 6.16), which follow the trends already discussed.

For the higher vapour velocities the profiles become more similar, until at the highest rate the two systems both give a curve with a distinct maximum at about 2.5 cms. above the plate. As will be discussed later (7.2) this maximum is a feature to be expected in the spray regime. It is interesting to note that the

negative system tends to show maxima at the intermediate vapour velocities, whilst the positive system has a lesser tendency to do so. This is consistent with the more predominantly spraylike dispersions exhibited by negative systems.

6.4.1.ii Hold-up data - composition and surface tension effects

Hold-up data are shown in Figs. 17a to 24. The hold-up of liquid on the plate can be seen to vary with composition, particularly under the conditions of cellular foam formation, systems A(+) and C at low F_H (Figs. 6.19 and 6.21). A minimum in the liquid hold-up curve coincides with a maximum in the efficiency curve, the latter being consistent with the higher gas residence time suggested by the former. Similar observations have been made by Thorogood (40).

Apart from the minima exhibited by the positive systems, the other striking difference is that the hold-up decreases with F_H for the positive systems; see Figs. 6.19 to 6.21. This behaviour is illustrated in Figs. 17a and 17b, which also shows that the difference is not in evidence at the low weir height. Sanadaji (69) working with the air/water system noted an increase in hold-up with vapour rate when a foam baffle was present, and a decrease when the foam baffle was absent. This suggests that when cellular foam is allowed to build up, either by use of a foaming system or by introduction of a foam baffle, the liquid hold up is low, increasing as the cellular foam degenerates to froth at higher vapour rates.

At the lowest weir height this argument is apparently not valid.

Finally if the fraction of the liquid hold-up between the plate and an arbitrarily selected level is used as a criteria for comparing the two systems, results are obtained as in Fig. 6.24. In this case the fractional hold-up above the weir height level, and the fractional hold up above a level 2.5 cms. above the weir height are plotted against the absolute value of surface tension. Values are corrected to an F_H of 18.5 (spray regime) by plots similar to Fig. 6.23. It is evident from 6.24 that the fraction of liquid in the spray above the chosen levels increases with a decrease in surface tension, and is greater for the negative system. The significance of this is that it suggests the two systems possess different drop size distributions at the same value of absolute surface tension. This point is discussed more fully in Chapter 7. It suffices here to observe that other workers (36) (37) have shown that smaller drops reach greater heights in spraying systems. This implies that the negative system has a higher proportion of small drops and hence has a higher specific surface, which is in qualitative agreement with the mass transfer results for the spray regime, shown in Fig. 6.9 and discussed in section 6.3.1.

6.4.2 The effect of hole velocity

6.4.2.i Dispersion density

The characteristic effect of increasing the hole velocity

is to bring about a redistribution of the liquid hold-up; this is evident from Figs. 6.10a to 6.15, and shown quantitatively in Fig. 6.23. From this it is seen that as the vapour rate increases there is a tendency for less of the dispersion to be present near the plate floor and more in the upper regions of the dispersion. This behaviour which is in line with the findings of Sargent et al (1), prevails at all the conditions of weir height and composition investigated.

For the highest weir height (Figs. 6.13 to 6.15) the effect of increasing the hole velocity is to eliminate the maximum in the dispersion density curve and increase the slope of the constant slope zone. This type of maximum has also been observed by Sargent et al. and by Bernard (39). A local maximum in the liquid hold-up means a reduced cross-sectional area for the vapour flow. Thus at the maximum the vapour linear velocity, and hence its kinetic energy, will also be passing through a maximum. Teller and Rood (70) who studied the effect of bubble coalescence noted that this produced a release of energy which resulted in an increase in vapour kinetic energy. Thus it is possible that bubble coalescence in the early stages of bubble rise produces a local increase in the vapour kinetic energy which manifests itself in a reduction in the local vapour hold up. This behaviour would give a maximum in the dispersion density profile. It is relevant to note that the maxima are absent at the highest vapour rates, which would

suggested that coalescence is occurring at the plate floor, as observed by Schroter (30).

For the intermediate and lowest weir height the effect of increasing the hole velocity is to diminish the differences between the profiles of the positive and the negative systems, Figs. 6.11a to 12b. The system C at the lowest F_H gives a 'constant slope' profile, characteristic of a cellular foam dispersion, whilst system B gives a 'constant density' profile. As the hole F factor increases the 'constant density' type of profile prevails for both systems. At the highest F factor and for both weir heights the profiles show a maximum at a level between 2 and 2.5 cm. above the plate floor, the position depending on the precise hole F factor prevalent. It is of interest to note that this maximum occurs at intermediate F factors for the negative system.

6.4.2.ii Hold-up data - effect of hole velocity

The variation of liquid hold-up with vapour velocity is dependent not only on the surface tension properties of the systems but also on the weir height.

For the 7.6 cm. weir the hold-up was found to decrease with hole velocity for both positive and negative systems, see Fig. 6.18. Some anomalous behaviour occurs at the lowest vapour rates as a result of weeping.

At the intermediate weir height (1.9 cms.) the negative

System B shows a decrease in hold-up with F_H , and the positive System C shows an increase, Figs. 6.17a and 6.17b. These trends are present across the concentration range (Figs. 6.20 and 6.21) and are confirmed by the azeotropic System A, Fig. 6.19.

For the lowest weir height both Systems B and C show an increase in hold-up with an increase in F_H , Figs. 6.17a and 6.17b.

At the highest F_H value the hold-up is virtually independent of weir height, for the intermediate and low weir height. This is paralleled by the similar mass transfer performance at this high F_H value, and is consistent with operating in the spray regime.

6.4.3 The effect of weir height

6.4.3.1 Dispersion density

In general the effect of increasing the weir height is to shift the profile upwards, a finding in general agreement with Sargent et al (1). The higher weir height will act as a support for the dispersion, not unlike the addition of a foam baffle to a lower weir height system. As the weir height is reduced the dispersion becomes more dependent on its own self-supporting properties. In other words the foaming systems A $\oplus$ and C at the low vapour rates, will encourage a build-up of rigid cellular foam having these self supporting tendencies. The non-foaming systems, A $\ominus$ and B, are unable to sustain a bubbly dispersion above the weir height level and a spray or droplet zone will prevail in the upper levels of the dispersion.

Comparison of the profiles at low F_H in Figs. 6.11a and 6.12a indicates that the 'constant slope' zone can either be a spray zone (Fig. 6.11a) or a cellular foam zone (Fig. 6.12a). This observation is emphasised when comparing the low weir height profiles at the lowest F_H value, Figs. 6.11b and 6.12b. The negative system is not supporting a cellular foam, but is generating a noticeable droplet population, whilst the positive system is a well defined cellular foam dispersion.

As mentioned above the differences between the profiles are less obvious as the hole velocity increases; and for the two lowest weir heights the 'spray regime' profiles ($F_H \approx 18$) are indistinguishable.

From Fig. 6.16 it would appear that the mean dispersion densities are very similar for the 1.9 and 0.95 cm. weir heights, any difference diminishing as the vapour rate increases.

6.4.3.ii Hold-up data - effect of weir height

Increase in weir height increases the liquid hold-up on the tray under most conditions of weir height, hole velocity and composition. The trends are shown in Figs. 6.17a and 6.17b, which also illustrate the important exceptions. Operation in the spray regime, which is achieved at the intermediate and low weir heights at the highest vapour rate, gives an independence of hold-up on weir height. This is to be expected since the spray regime is dependent only on the atomising action occurring at the holes and

requires a minimal liquid depth on the plate. The hold-up will then comprise this shallow pool of liquid on the plate floor, sufficient to supply the holes with the liquid which is atomised, and the droplet population formed by this atomisation. Under these conditions the hold-up will be almost independent of weir height.

6.4.4 Sieve plate regimes

The range of vapour liquid dispersions, from gas phase dispersed to liquid phase dispersed have been observed for the three systems.

Deep pool contacting ($W = 7.6$ cms.) has surprisingly diminished the differences between positive and negative systems. This is thought to be a consequence of the artificial support given by the high weir for both systems. As a result the plate efficiencies for the azeotropic system $A \ominus$ and $A \oplus$ show no variation with composition at this weir height.

Cellular foam was encountered with the positive systems at the lower weir heights and the lowest F_H values. Increase in the vapour rate tends to encourage break-up of the cellular structure until the liquid phase dispersed regime (spray) is reached. This break-up is illustrated consecutively in Figs. 6.25 to 6.28.

No cellular foam was encountered with the negative systems under any conditions. Increase in the vapour rate tends

to lift more liquid into the dispersion which becomes progressively more liquid phase dispersed. At the higher vapour rates the spray regime is encountered. The behaviour can be seen in Figs. 6.29 to 6.32.

6.5 Description of the Spray regime

The fully established spray regime, which was only encountered at the highest vapour rates, proved difficult to photograph owing to the lens effect of droplets impinging on the window. However, visual observations and the strong circumstantial evidence gained from mass transfer and hydrodynamic measurements suggests certain properties which are specific to this regime.

Firstly, it would appear from the general trends of the data that the transition from mobile foam to spray is gradual. This is in agreement with the results of de Goederen(20) and the findings of Yoshida et al(8) for a high capacity sieve plate column.

Visually the mechanism of spray formation appears to be a multi-orifice atomisation, with the holes supplied by a shallow pool of liquid less than 0.3 cms. deep on the plate. Drop size varies from a fine spray to large packages of liquid. No measurement of drop size or drop size distribution has been attempted in this investigation. However it is shown in Chapter 7 that it is possible to estimate these characteristics from the dispersion density profiles.

Performance characteristics of the spray regime indicate that the plate efficiency and the liquid hold-up are independent of weir height; findings which agree with those of Yoshida.

In addition the influence of the absolute value of the surface tension is confirmed. This has been reported previously by Bainbridge and Sawistowski (2) and de Goederen (20). Only partial confirmation is given to their findings that the performance varies with the reciprocal of the surface tension. More significantly it is shown that under the conditions of these investigations the surface tension negative System B gives a higher performance than the surface tension positive System C, at the same value for the absolute surface tension. An explanation of this phenomenon is provided by the fact that the negative system, although it has a hold-up similar to the positive system, has proportionately more of its hold-up in the upper layers of the dispersion. This suggests a finer drop size at formation, since smaller drops reach the greatest heights (36) (37). A smaller drop size would give a higher mass transfer surface and a higher performance.

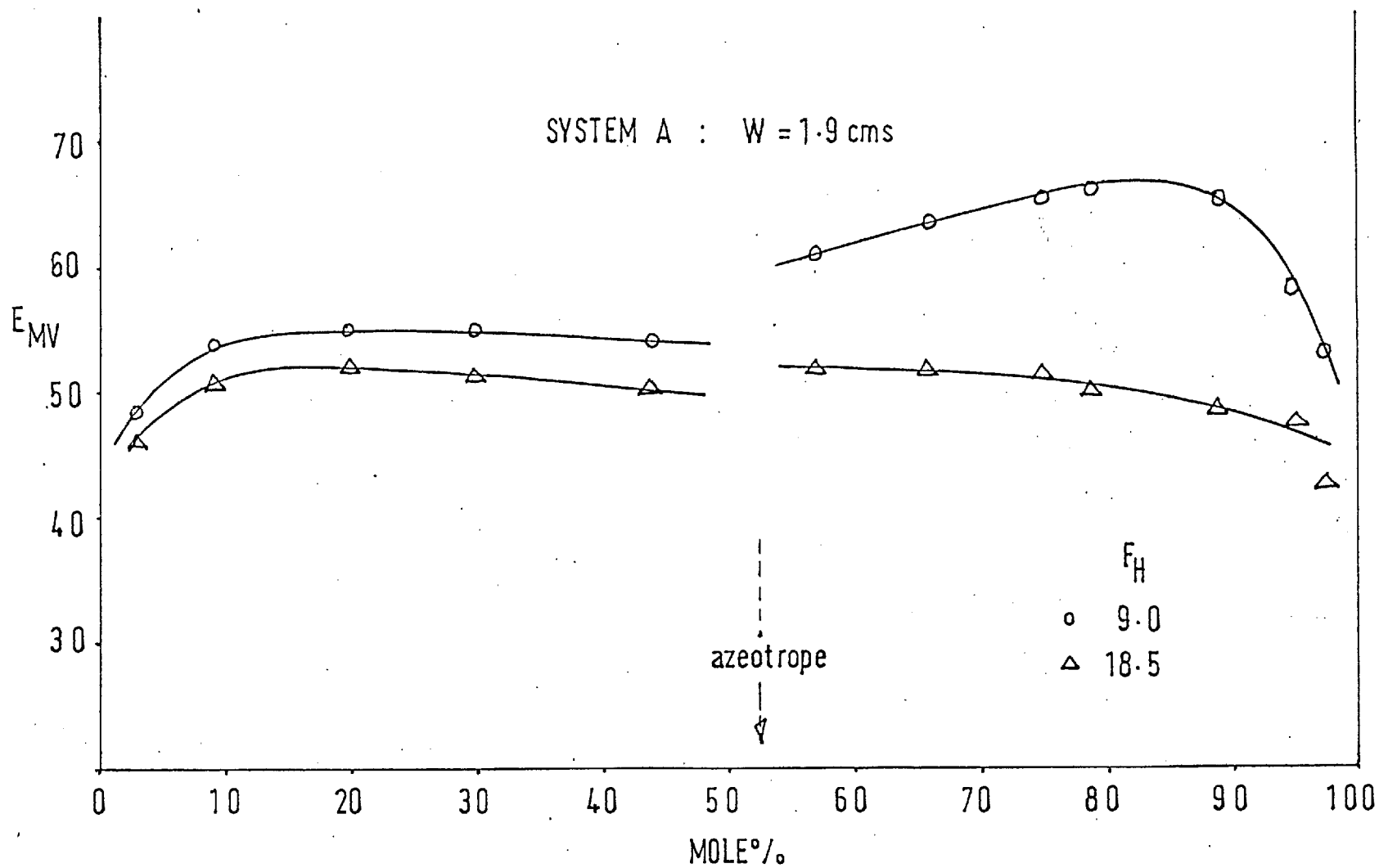


FIG 6.1 VARIATION OF EFFICIENCY WITH COMPOSITION AND FLOWRATE

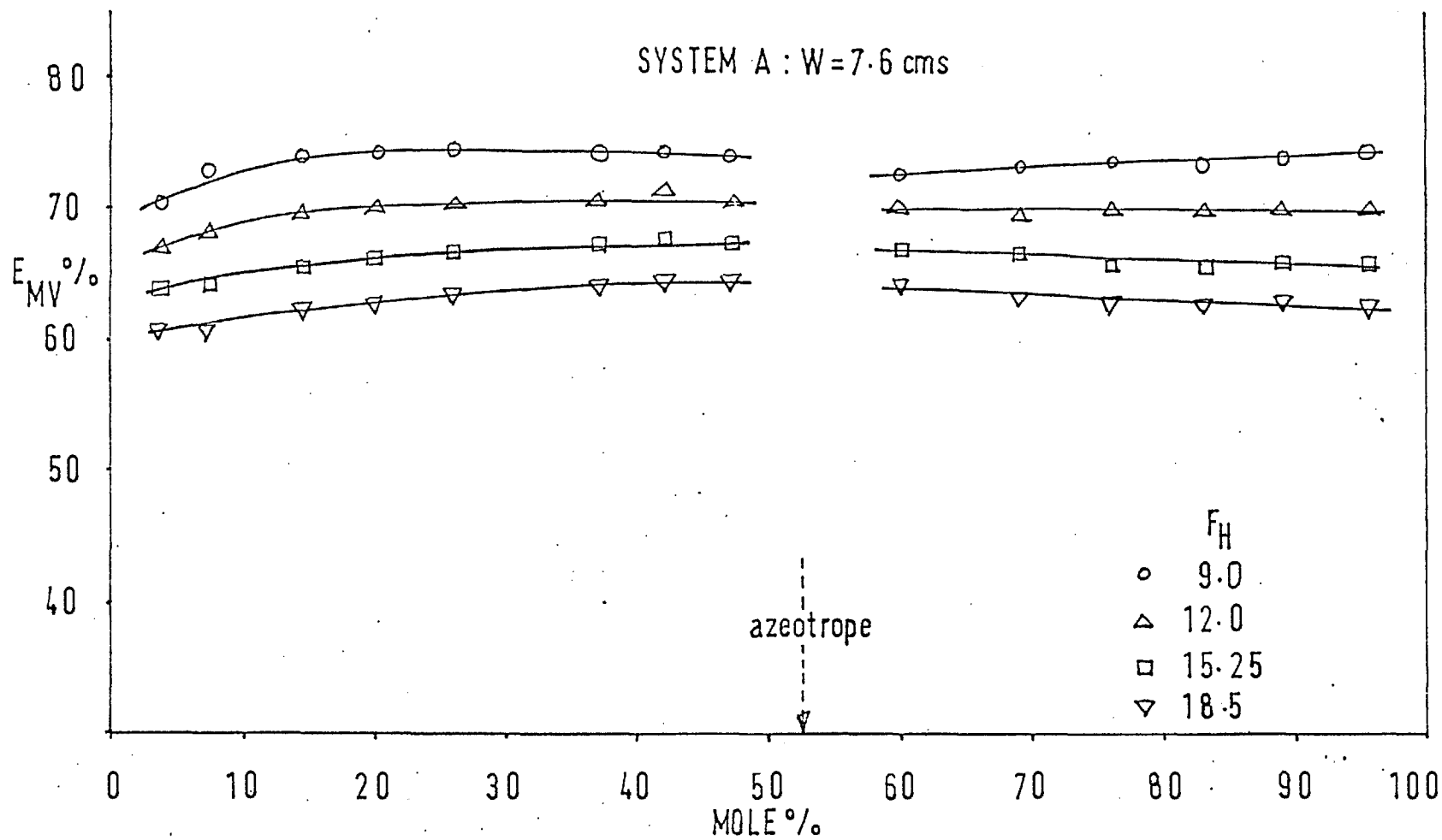


FIG 6.2 VARIATION OF EFFICIENCY WITH COMPOSITION AND F_H

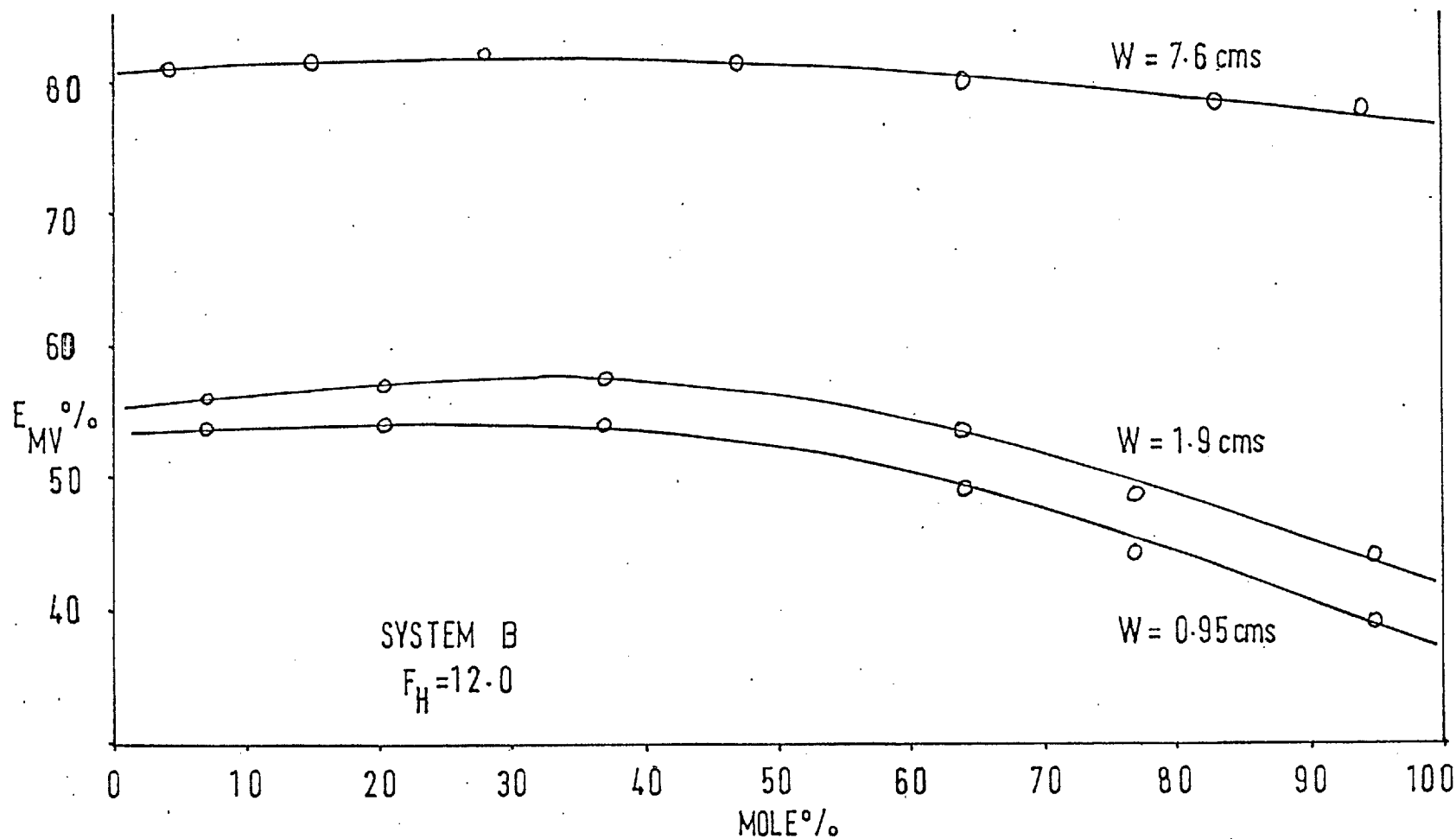


FIG 6.3 VARIATION OF EFFICIENCY WITH COMPOSITION AND W

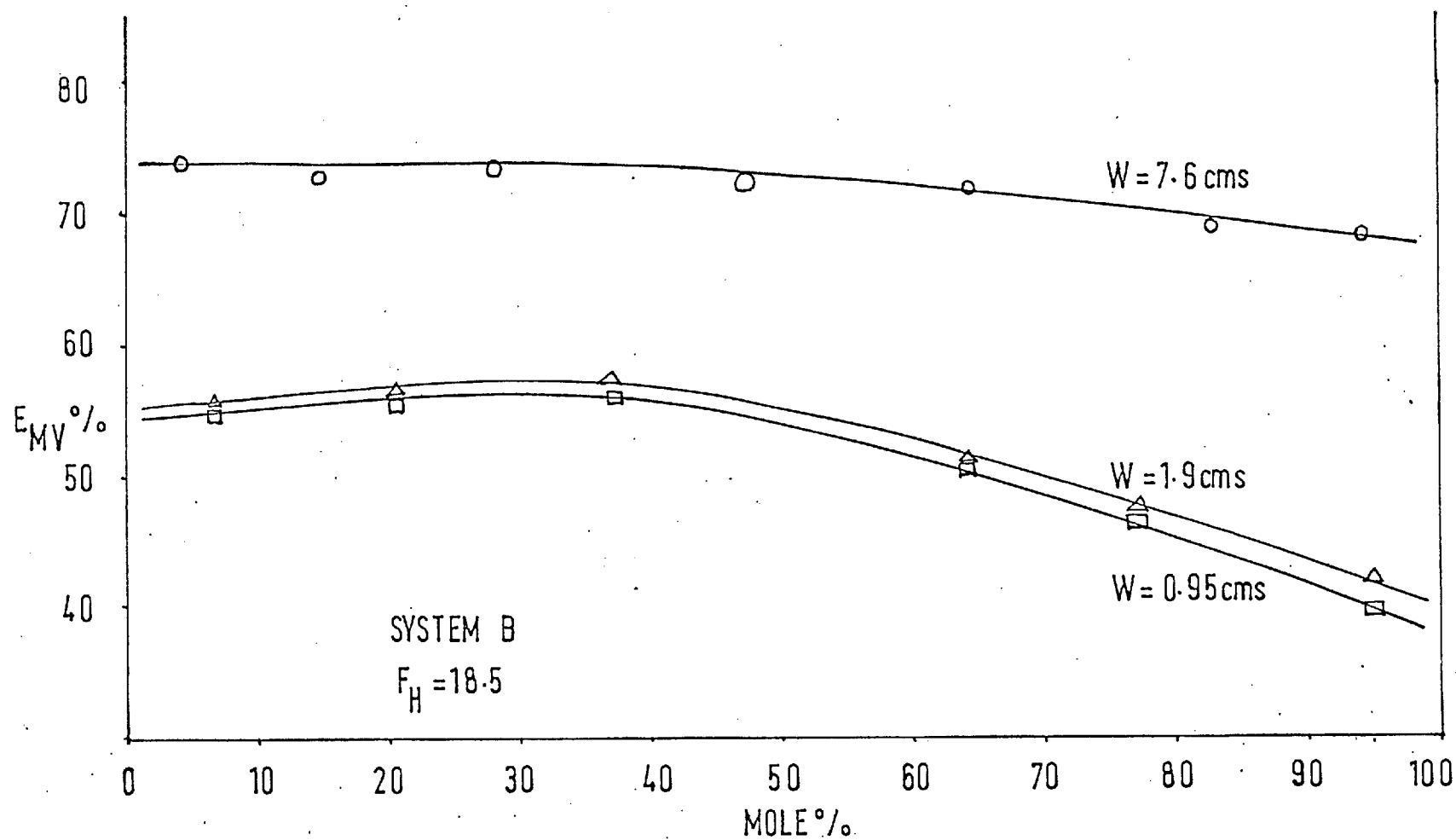


FIG 6.4 VARIATION OF EFFICIENCY WITH COMPOSITION AND W

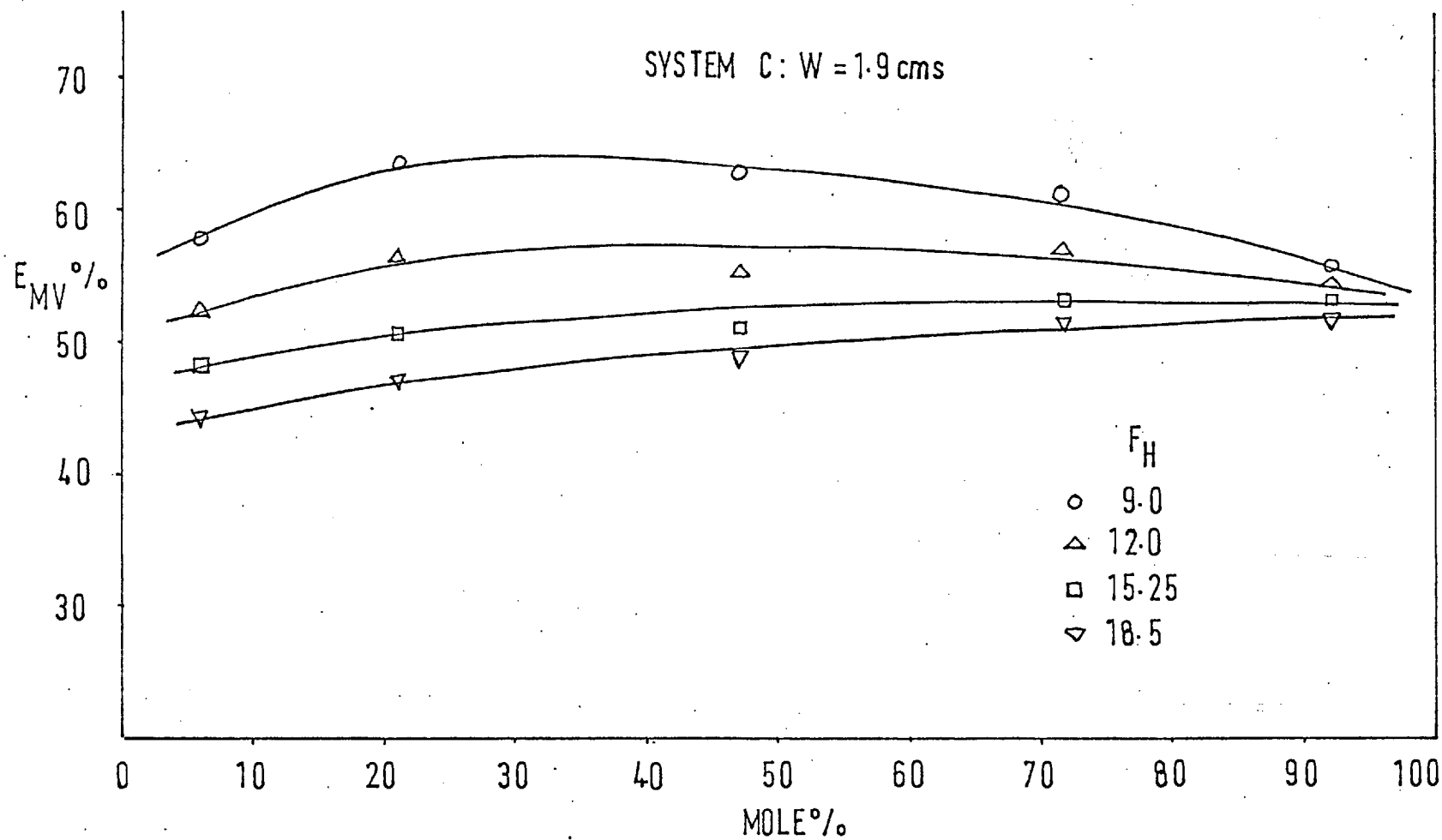


FIG 6.5 VARIATION OF EFFICIENCY WITH COMPOSITION AND F_H

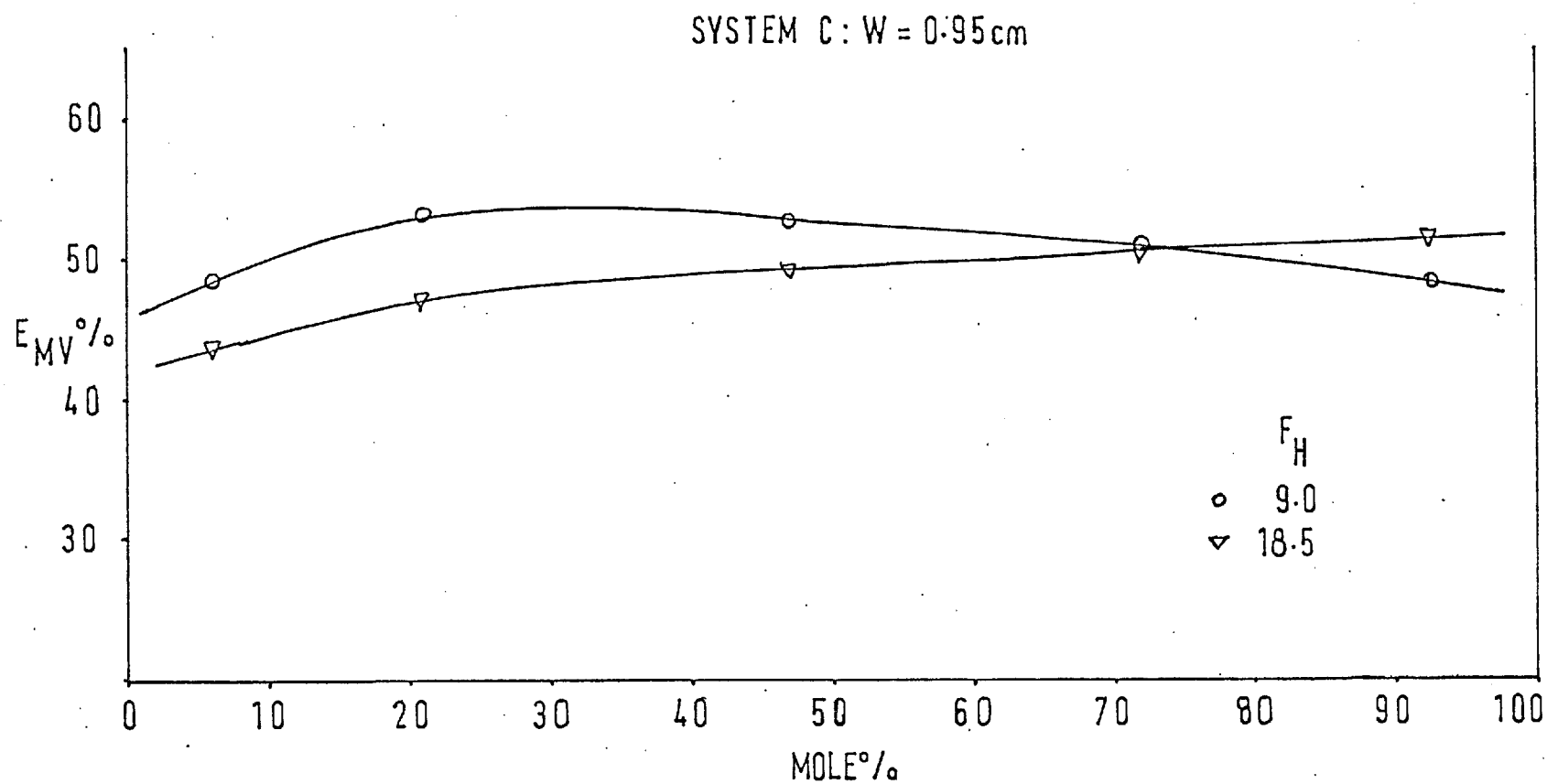


FIG 6.6 VARIATION OF EFFICIENCY WITH COMPOSITION AND F_H

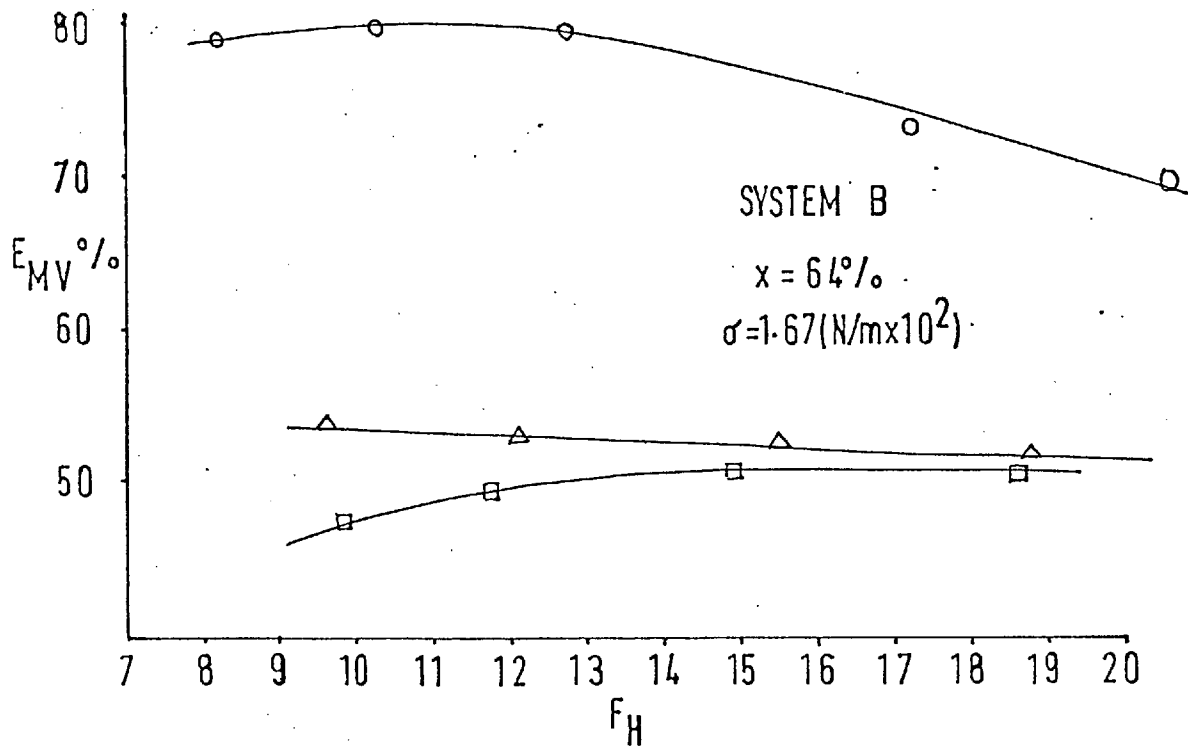


FIG 6.8 EFFECT OF VAPOUR RATE ON EFFICIENCY
 (NEGATIVE SYSTEM)

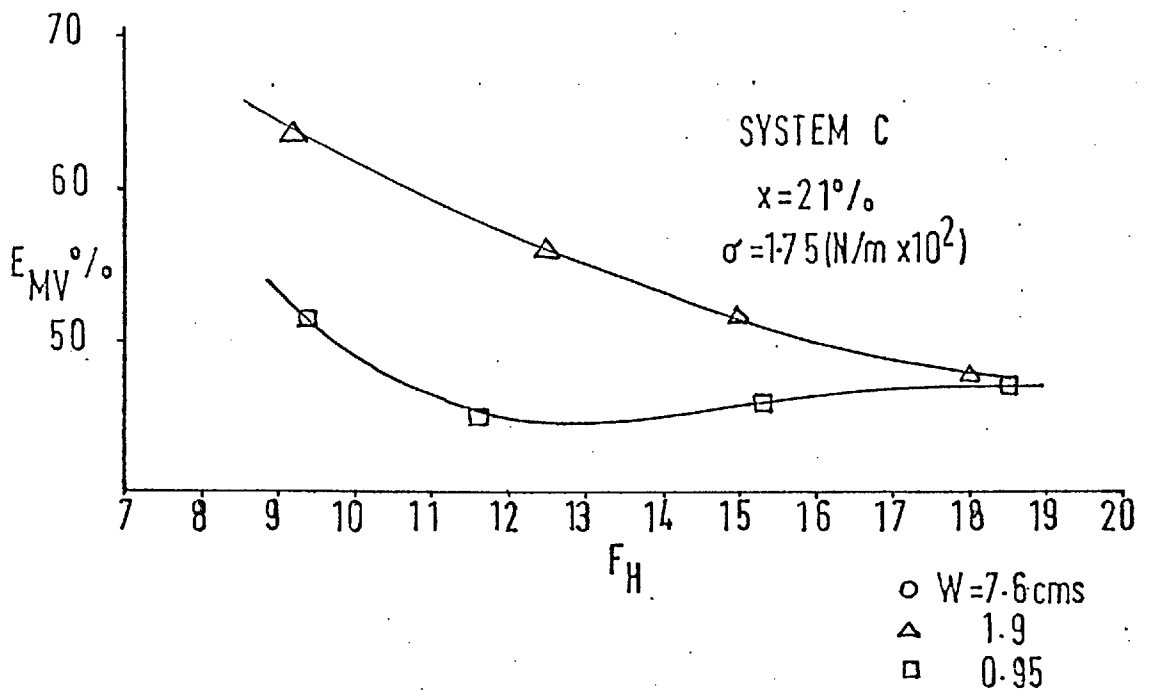


FIG 6.7 EFFECT OF VAPOUR RATE ON EFFICIENCY
 (POSITIVE SYSTEM)

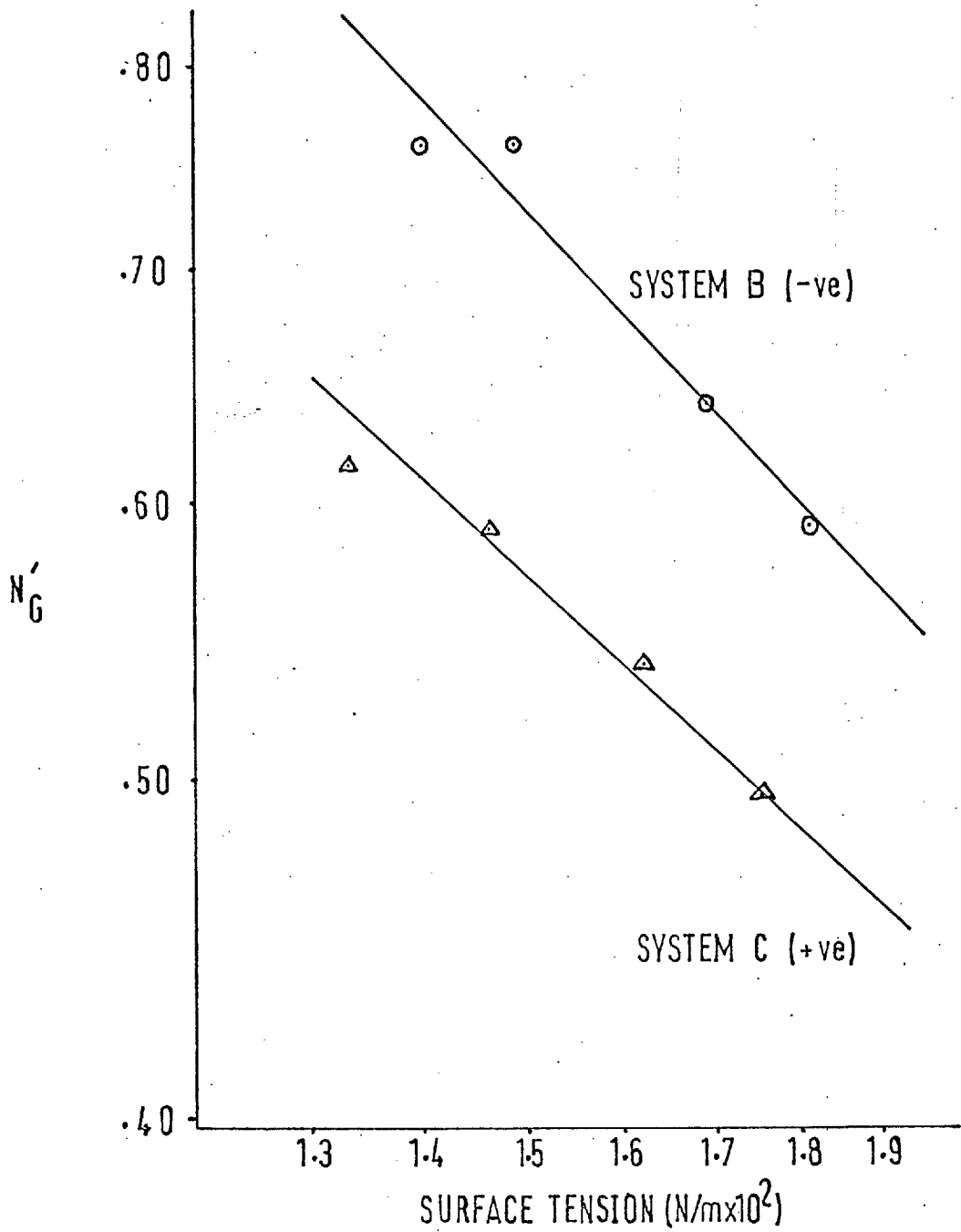
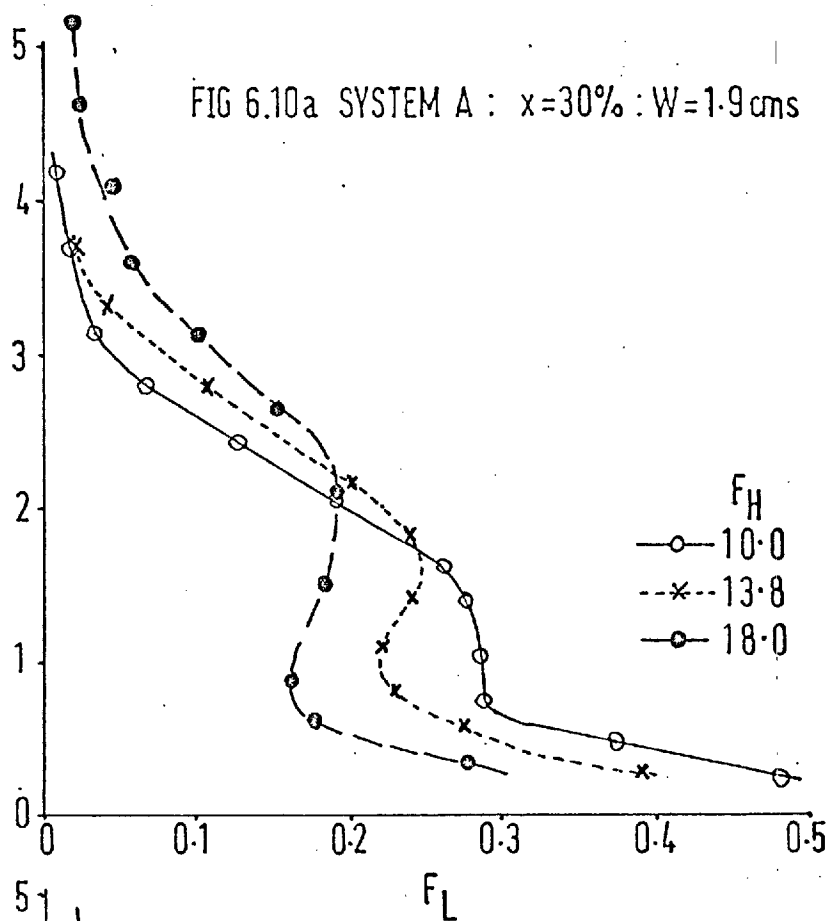
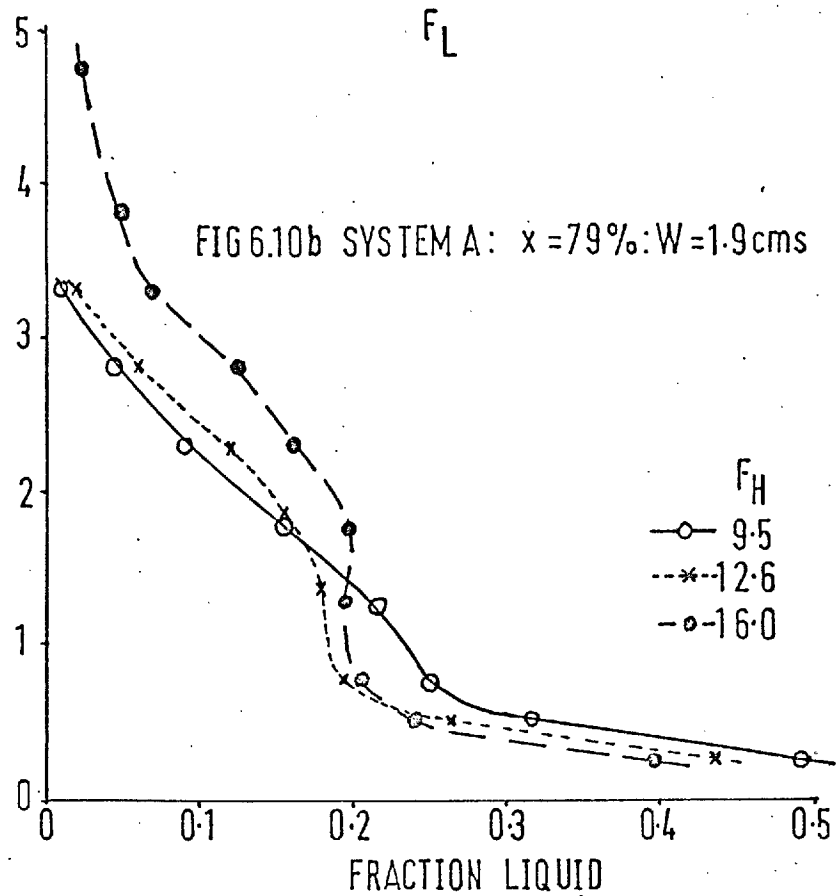


FIG 6.9 VARIATION OF PLATE PERFORMANCE WITH SURFACE TENSION IN THE
SPRAY REGIME

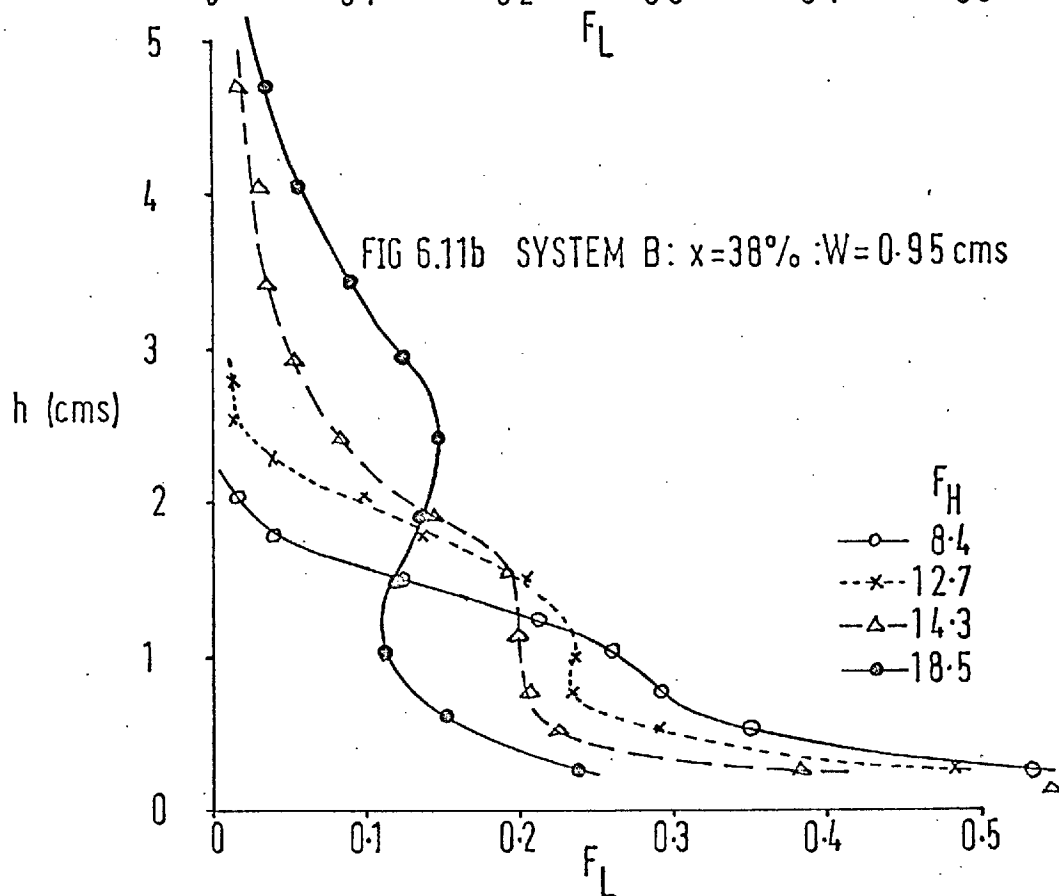
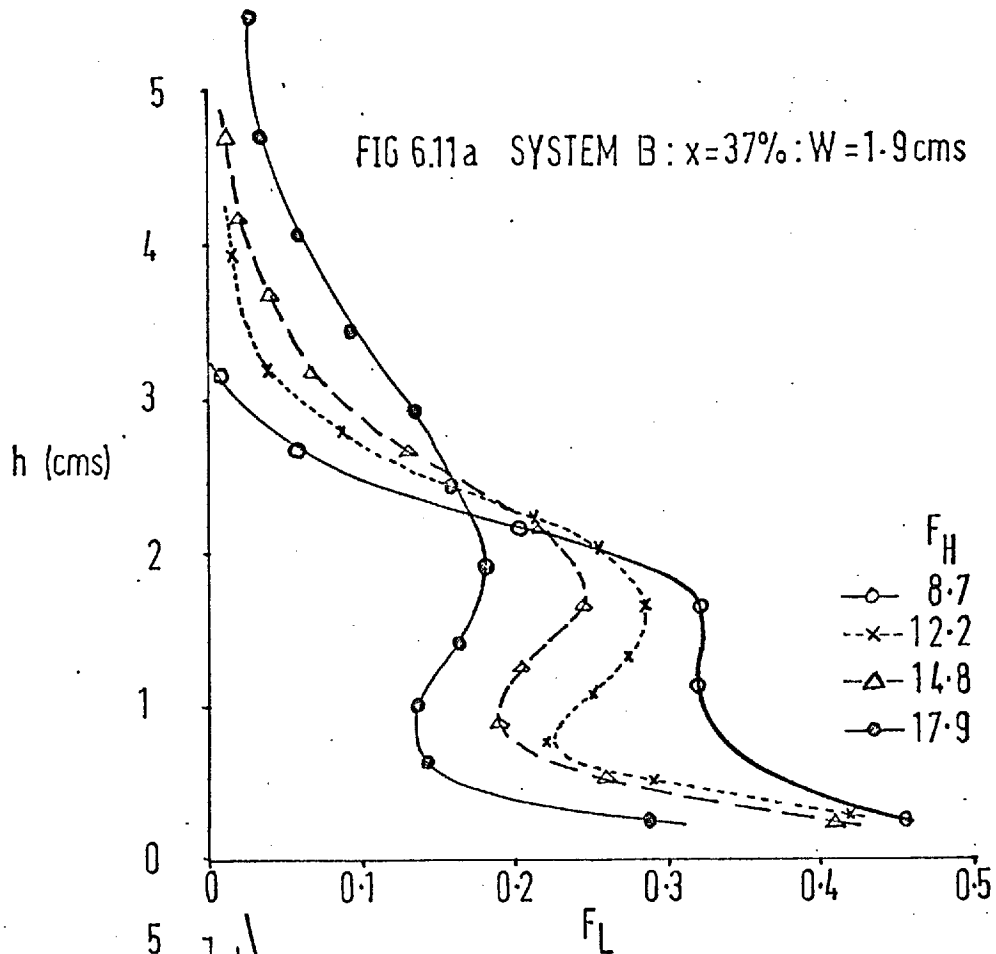
HEIGHT
ABOVE
PLATE
(cms)

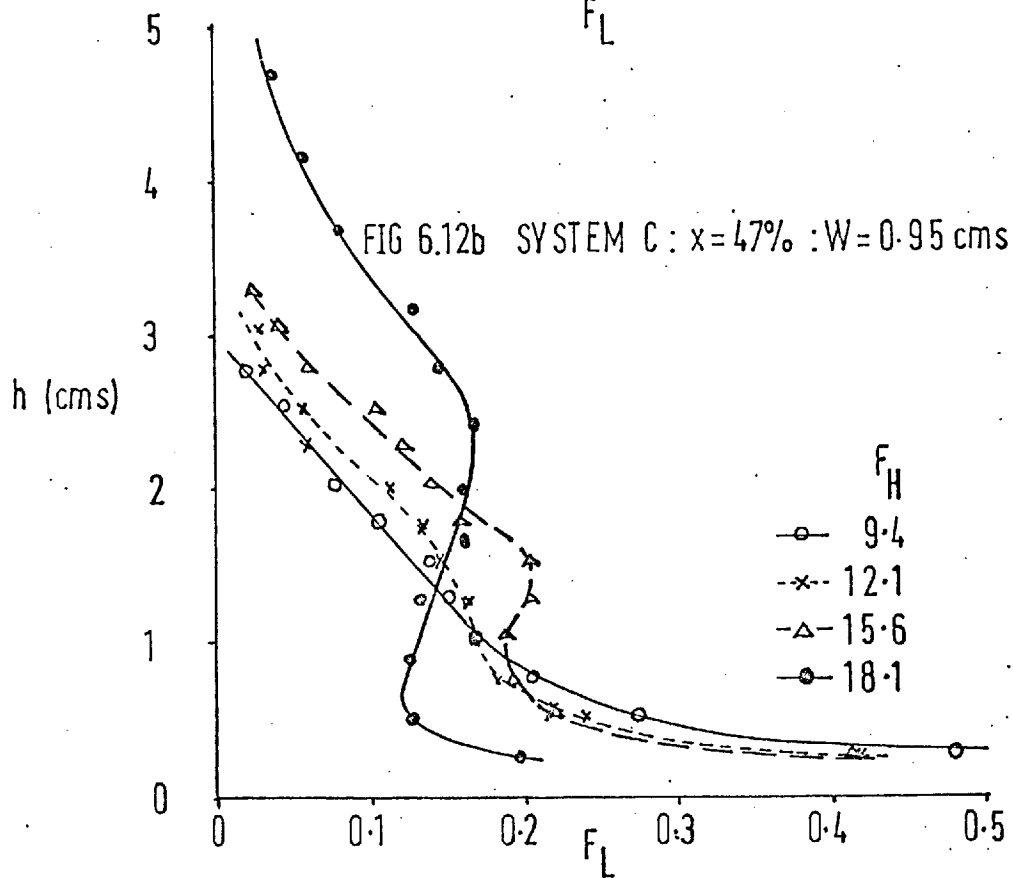
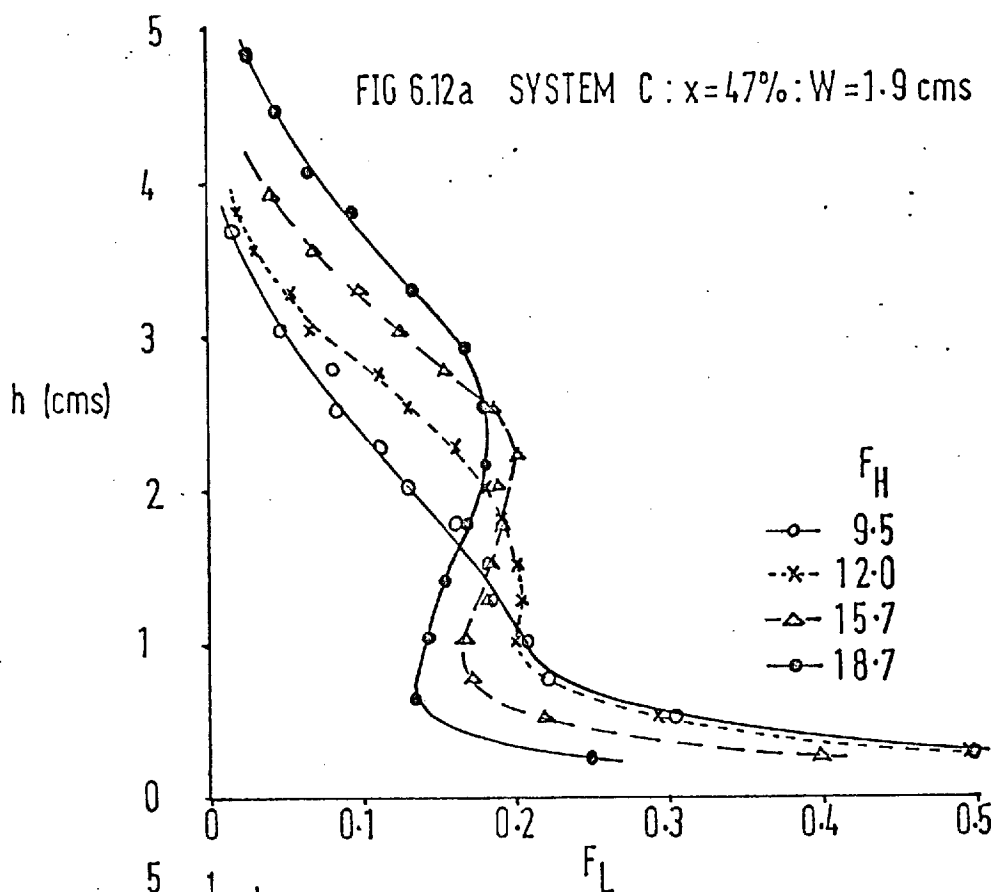


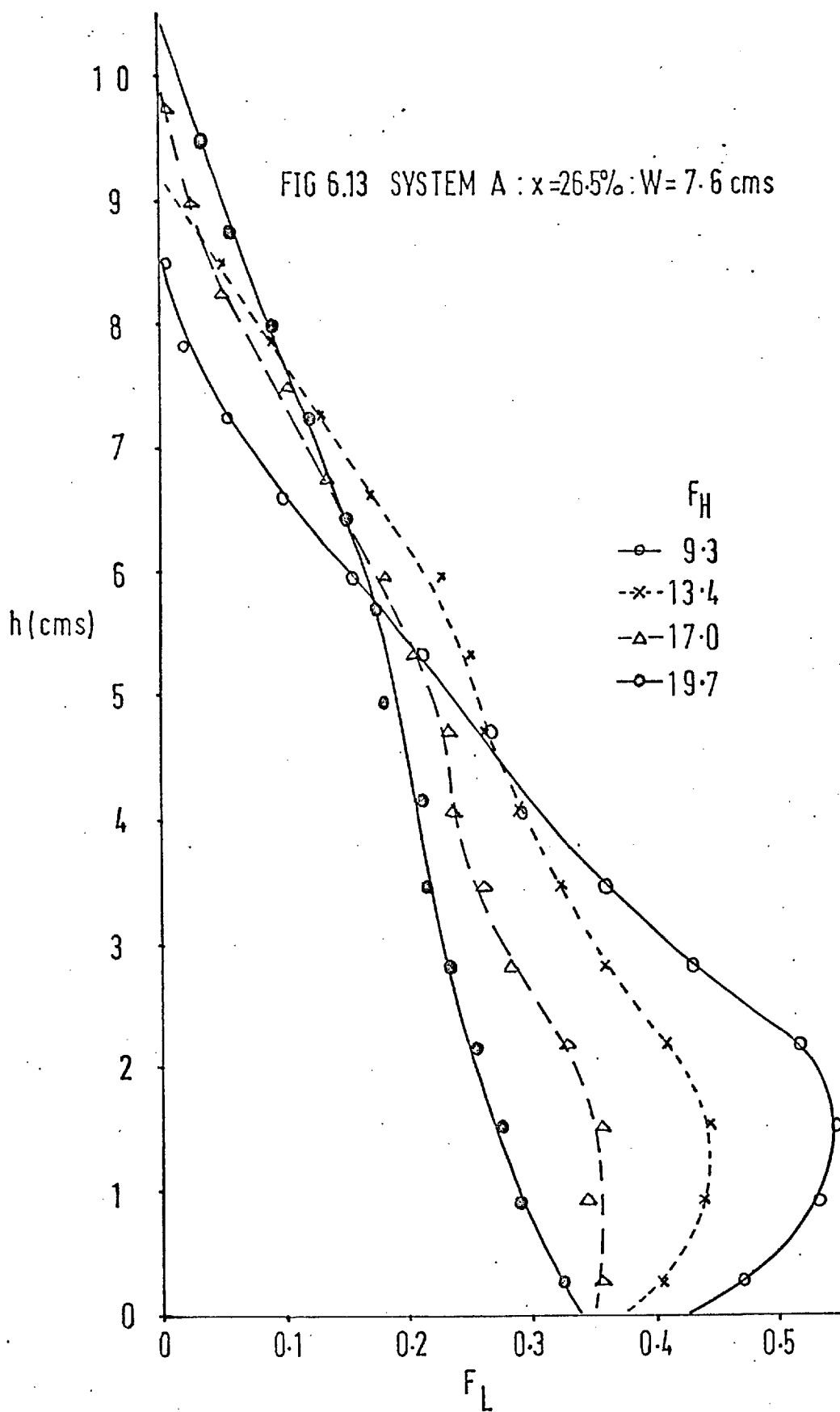
h

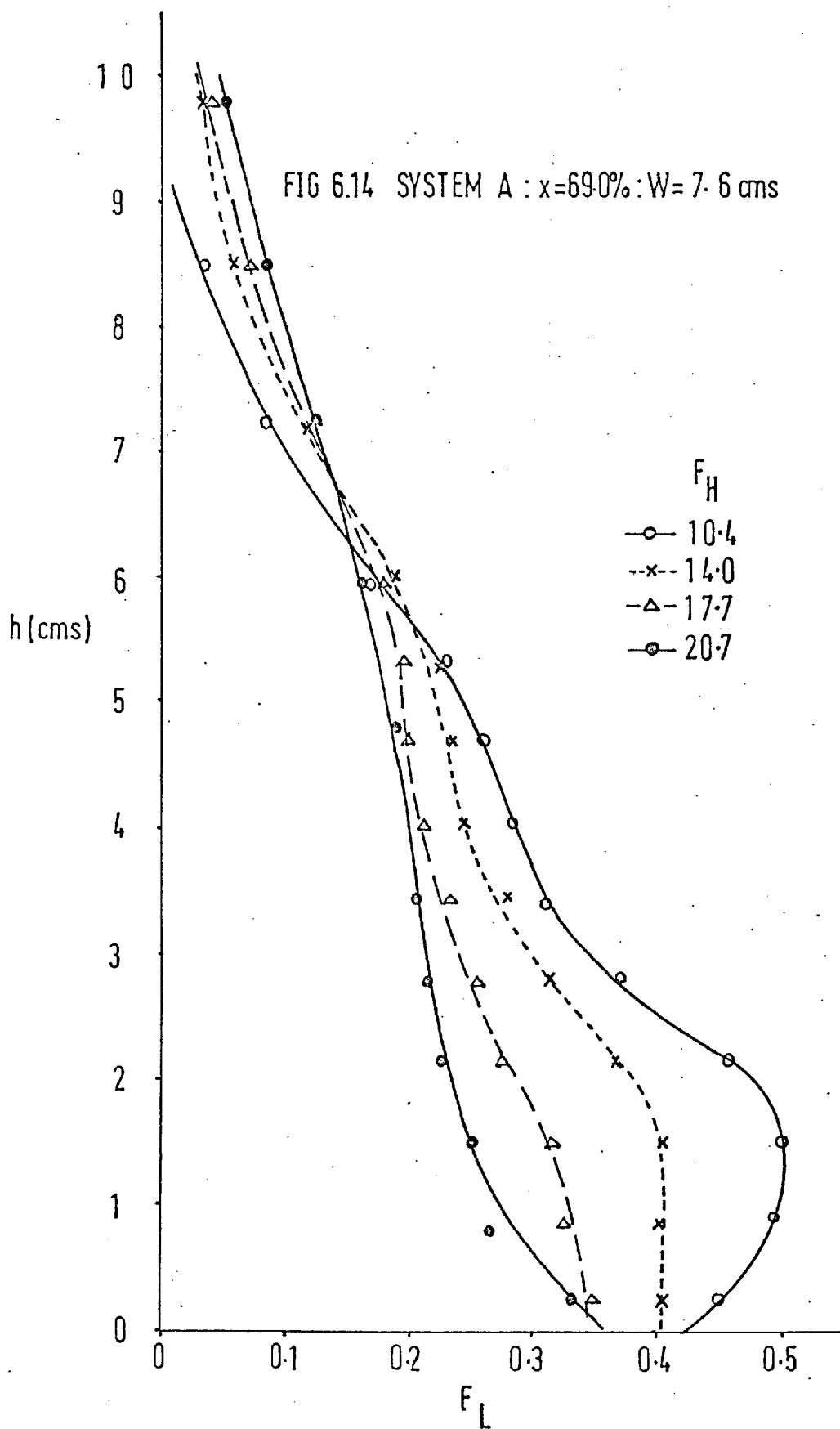


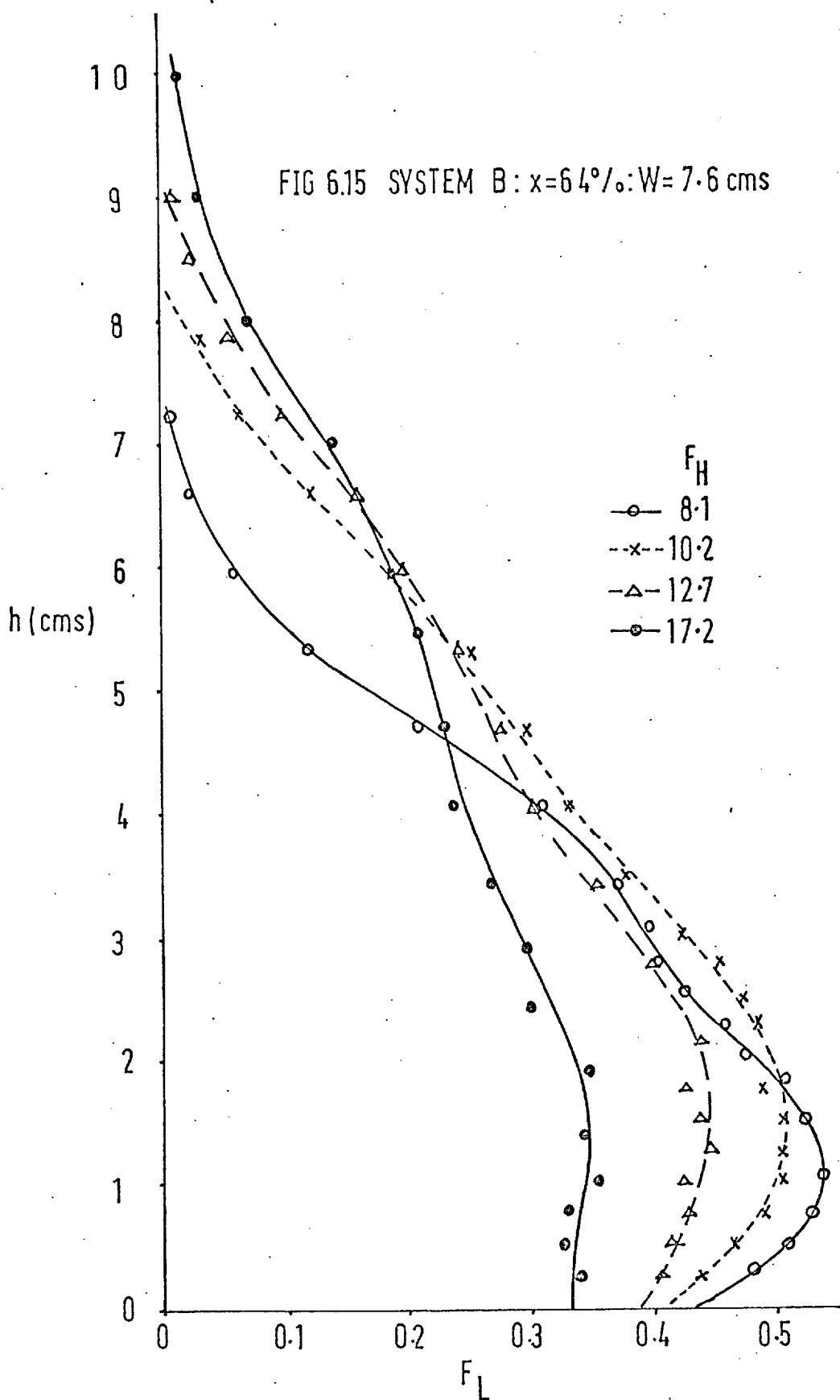
FIGS 6.10a,b DISPERSION DENSITY PROFILES











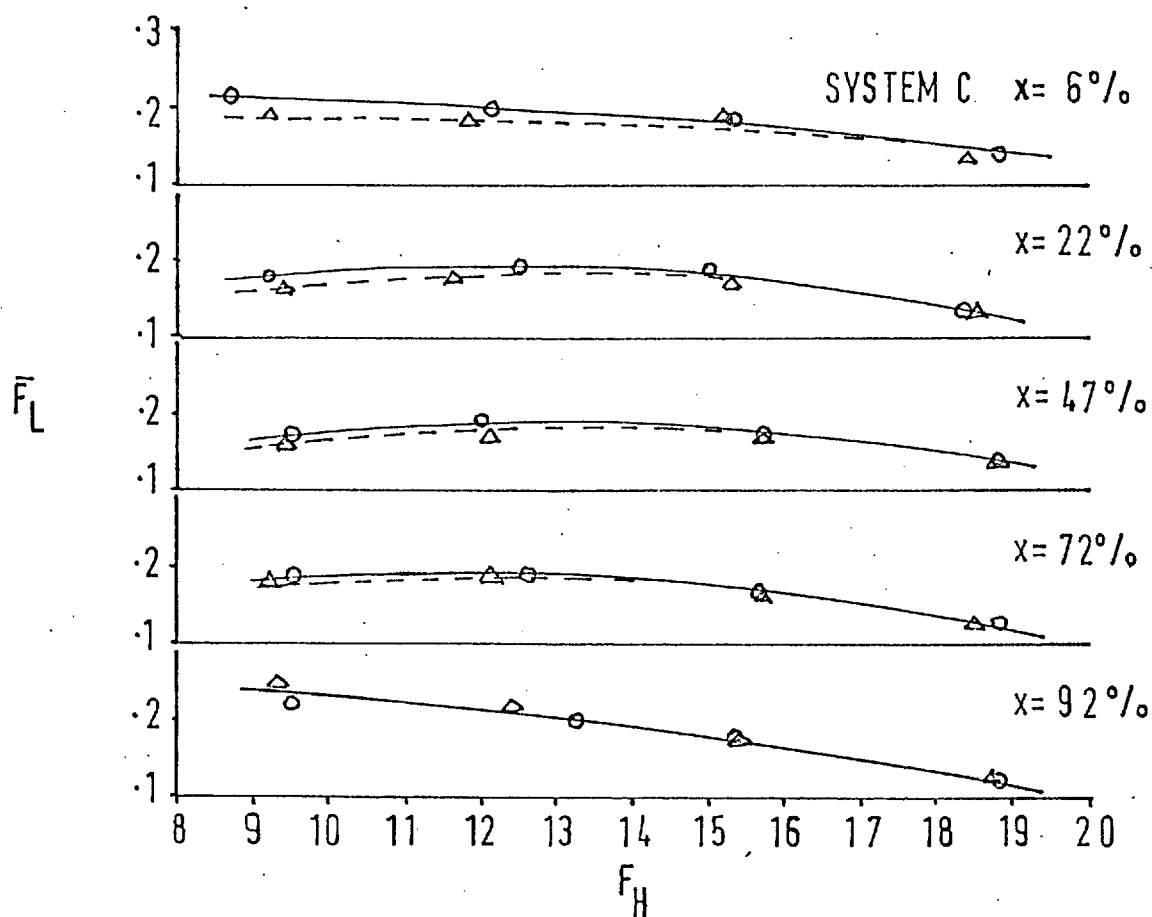
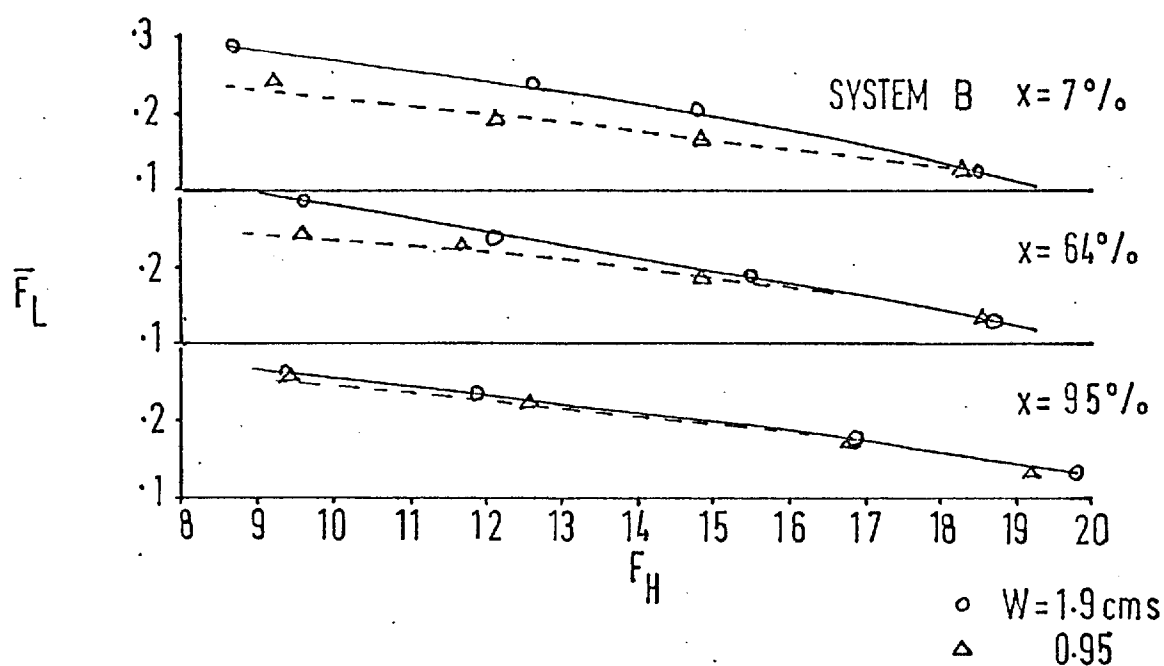
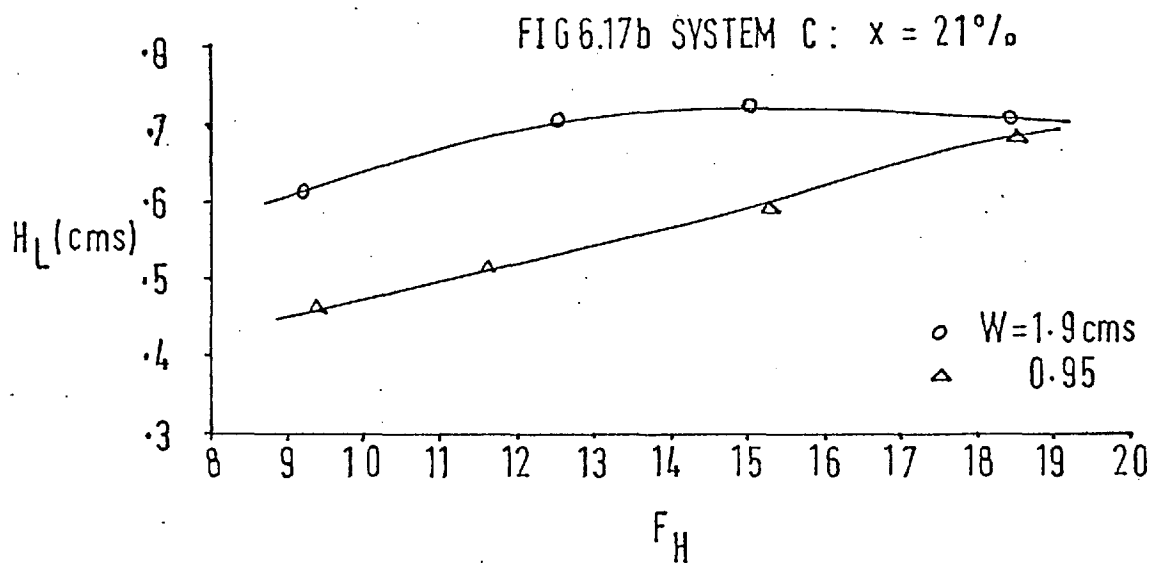
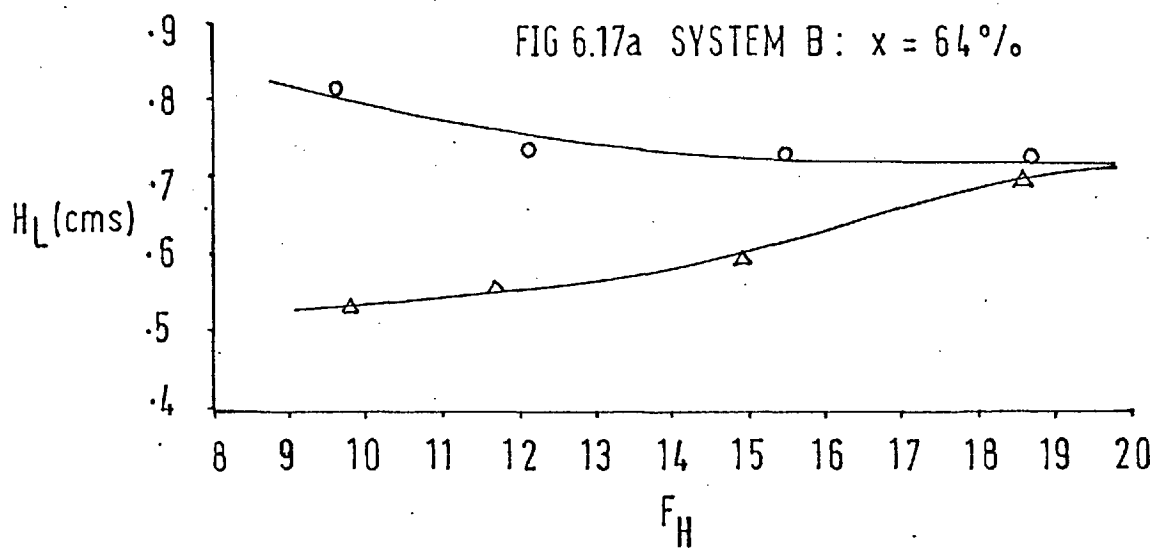
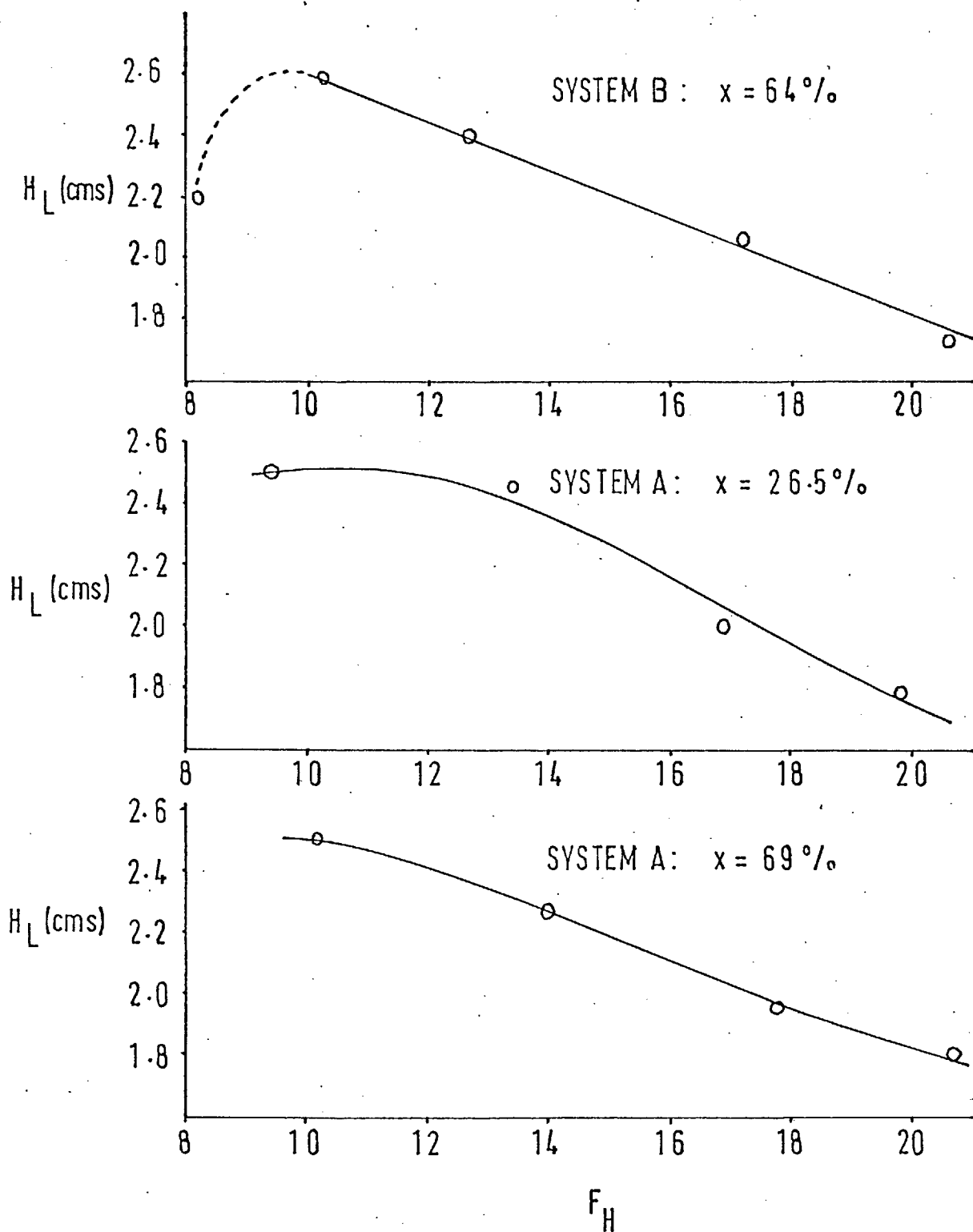


FIG 6.16 VARIATION OF MEAN DISPERSION DENSITY WITH F_H



FIGS 6.17a,b VARIATION OF LIQUID HOLD-UP WITH F_H AND W

FIG 6.18 VARIATION OF LIQUID HOLD-UP WITH F_H FOR $W=7.6$ cm

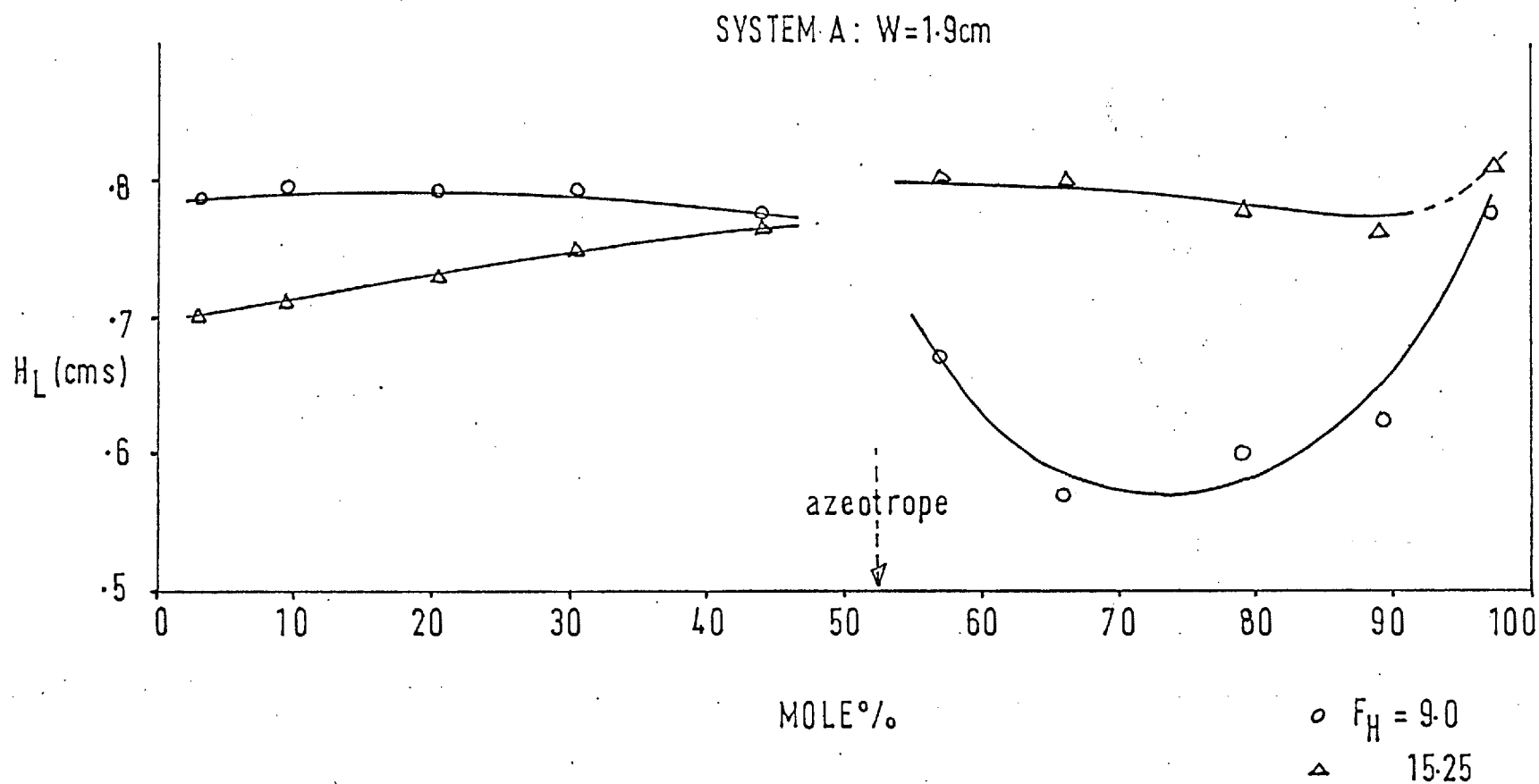


FIG 6.19 VARIATION OF LIQUID HOLD-UP WITH COMPOSITION AND FLOWRATE

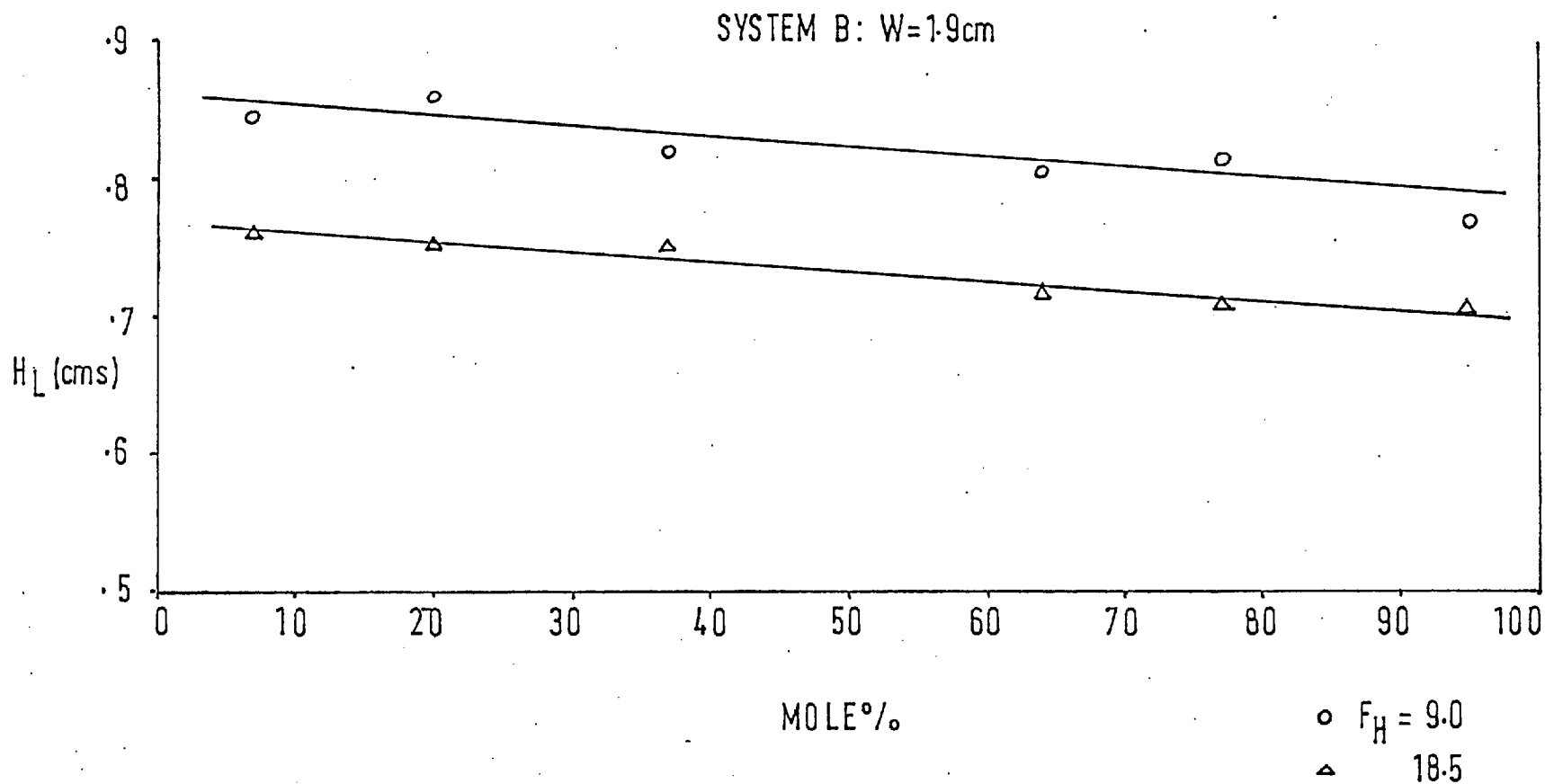


FIG 6.20 VARIATION OF LIQUID HOLD-UP WITH COMPOSITION AND FLOWRATE

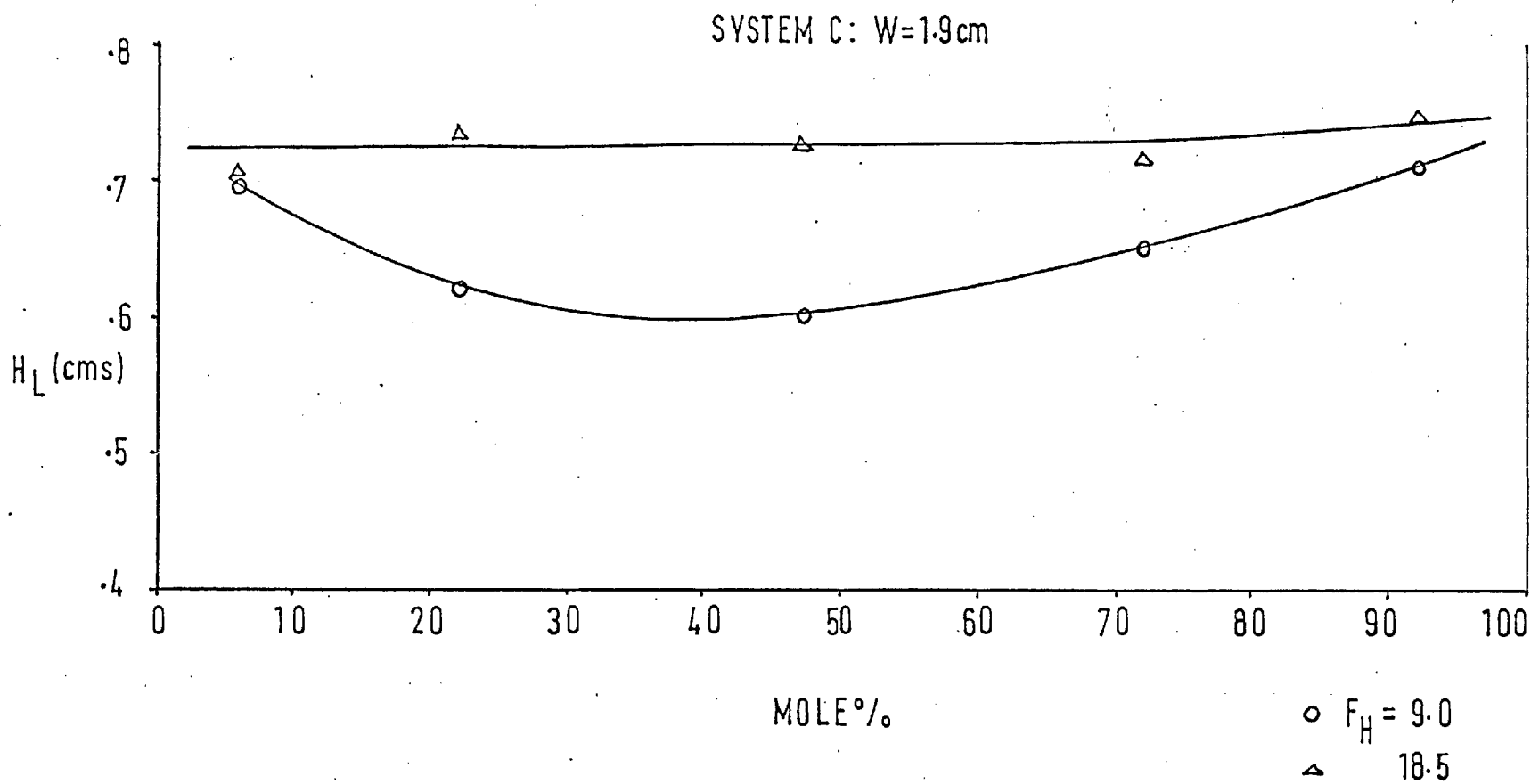


FIG 6.21 VARIATION OF LIQUID HOLD-UP WITH COMPOSITION AND FLOWRATE

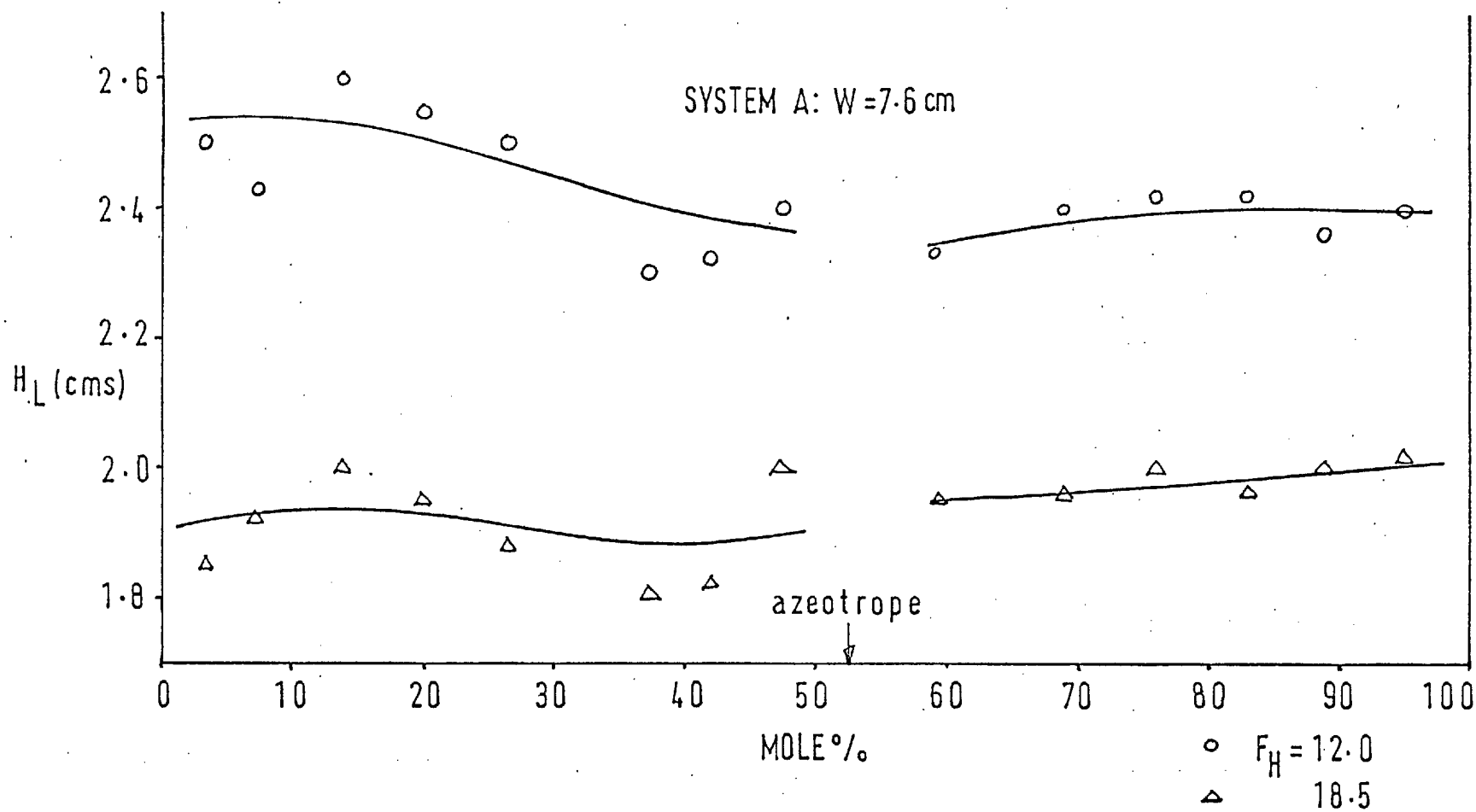
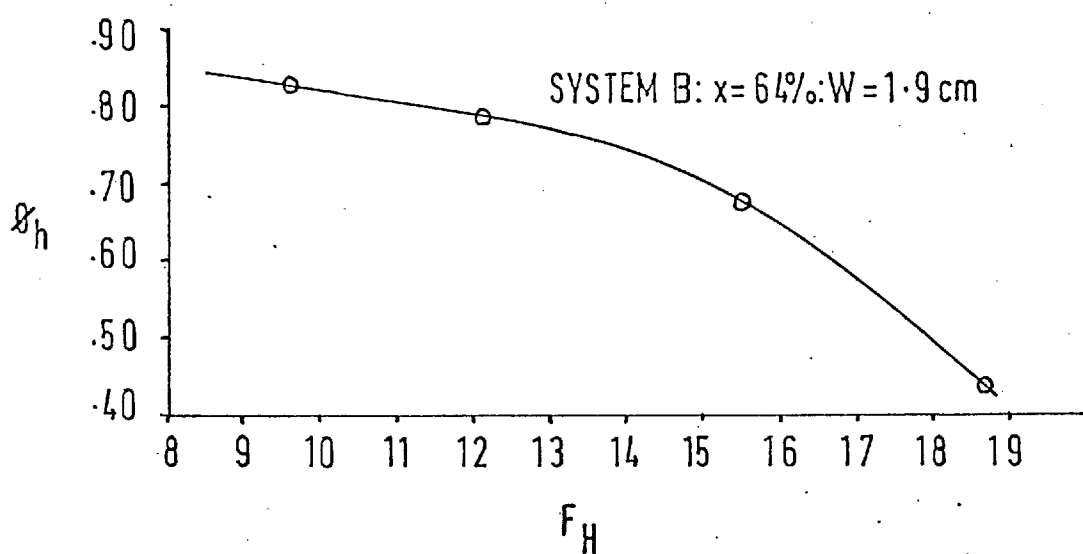
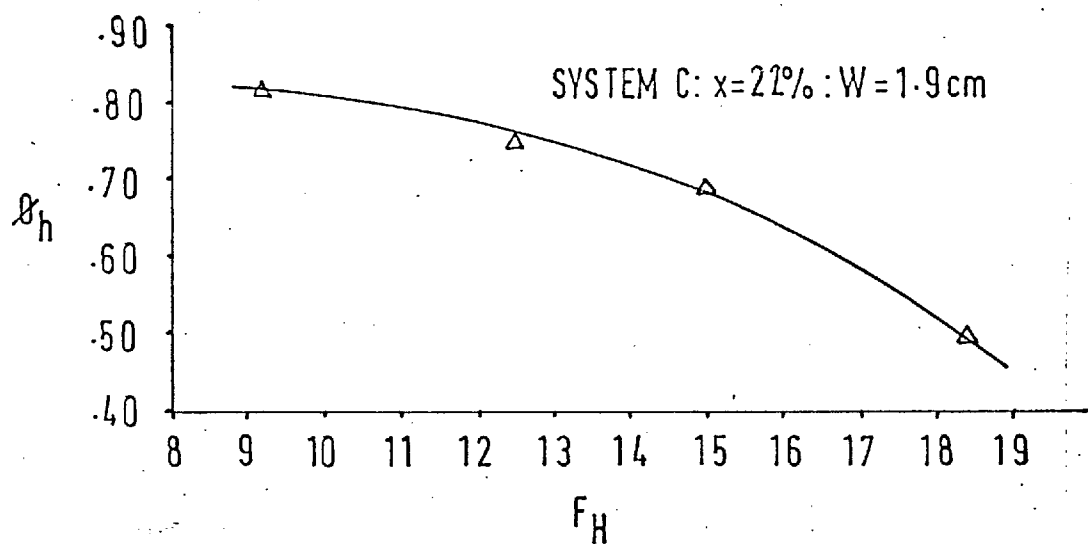


FIG 6.22 VARIATION OF LIQUID HOLD-UP WITH COMPOSITION AND FLOWRATE



$$\phi_h = \frac{\text{AMOUNT OF LIQUID IN DISPERSION UP TO LEVEL } h}{\text{TOTAL LIQUID IN DISPERSION}}$$

h = weir height W

FIG 6.23 EFFECT OF F_H ON ϕ_h , THE FRACTION OF LIQUID HOLD-UP IN THE DISPERSION UP TO LEVEL h

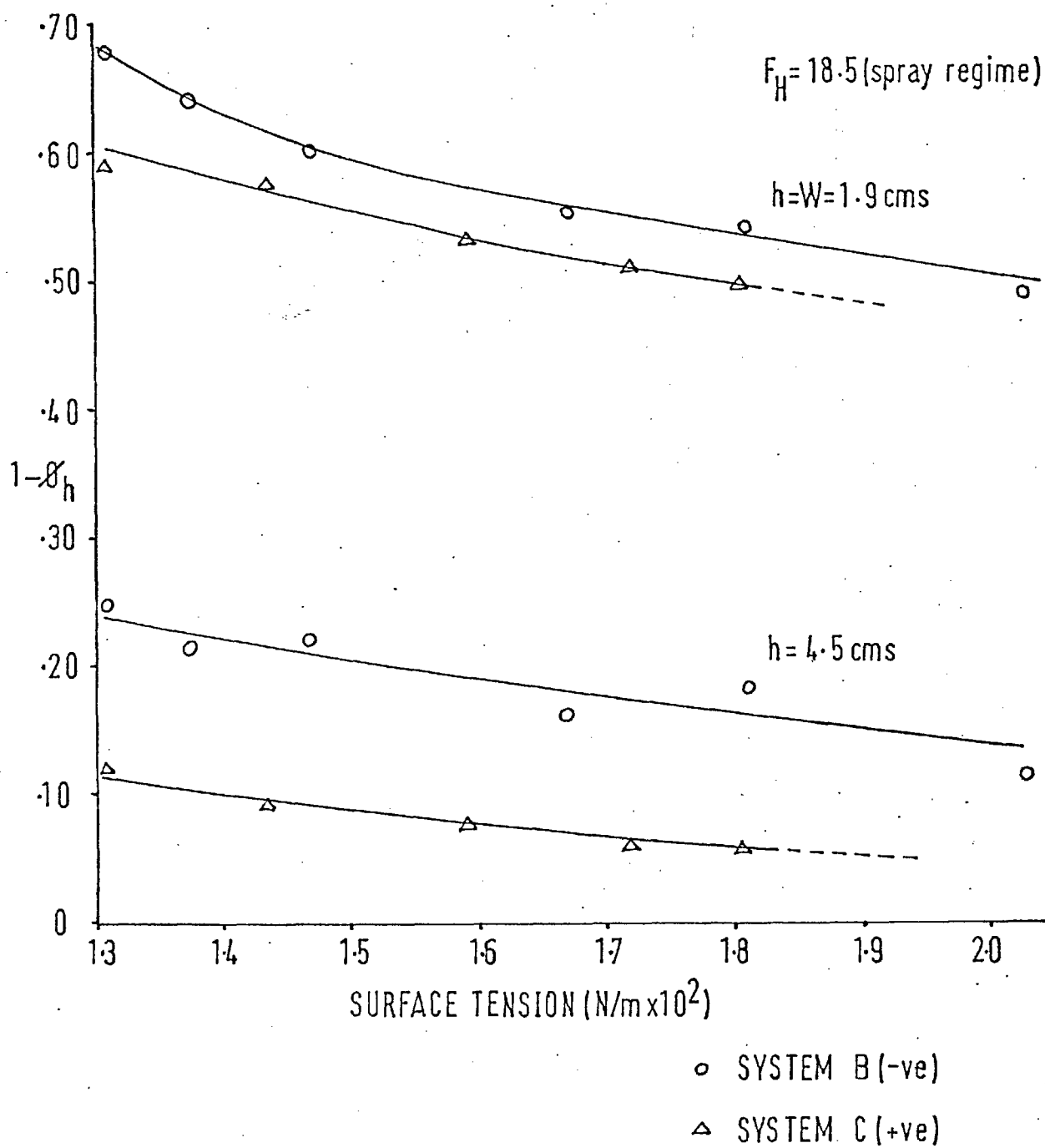


FIG 6.24 FRACTION OF LIQUID IN DISPERSION ABOVE LEVEL h
— EFFECT OF SURFACE TENSION



FIG 6.25 nHEPTANE/TOLUENE $x=24\%$ $W=1.9\text{cms}$ $F_H = 9.0$



FIG 6.26 nHEPTANE/TOLUENE $x=24\%$ $W=1.9\text{cms}$ $F_H = 12.0$



FIG 6.27 nHEPTANE/TOLUENE $x=24\%$ $W=1.9\text{cms}$ $F_H=15.0$



FIG 6.28 nHEPTANE/TOLUENE $x=24\%$ $W=1.9\text{cms}$ $F_H=18.0$



FIG 6.29 BENZENE/nHEPTANE $x=64\%$ $W=1.9\text{cms}$ $F_H=8.0$



FIG 6.30 BENZENE/nHEPTANE $x=64\%$ $W=1.9\text{cms}$ $F_H=12.0$



FIG 6.31 BENZENE/nHEPTANE $x=64\%$ $W=1.9\text{cms}$ $F_H=15.0$



FIG 6.32 BENZENE/nHEPTANE $x=64\%$ $W=1.9\text{cms}$ $F_H = 18.0$

CHAPTER SEVEN

MODEL OF TRAY OPERATION IN THE SPRAY REGIME

7.1 Introduction

From visual observations of the dispersion in the spray regime it was evident that liquid was continuously atomised at the holes by the high velocity vapour stream. The flow of liquid to the holes was supplied in part by the liquid flow onto the plate, and in part by the drops which had returned to the plate. For the conditions of weir height and flow rates used it was evident that there was very little liquid actually on the plate floor.

Inspection of the dispersion density profiles at the highest hole velocities indicated that the liquid hold-up decreased rapidly in the first 5 to 7 mm. above the plate and then increased gradually to give a maximum approximately 25 mm. above the plate. Thereafter the hold-up tailed off, approaching zero about 100 mm. above the plate.

Comparison of photographs, visual observations and the dispersion profiles points to the fact that the profiles are comprised of two separate curves, which are related to two zones in the dispersion. These two zones are (i) a shallow pool on the plate floor and (ii) the droplet population. Of the two, the droplet population predominates, the shallow pool serving as a liquid supply to the 'atomisers' (i.e. orifices) which is continually

replenished by returning droplets and the liquid feed. Figure 7.1a illustrates the two curves, where curve I is the shallow pool profile, and II is the spray profile. Since the major contribution to the mass transfer surface will be the surface area of the spray droplets, any approach towards an explanation of curve II will be of importance in quantifying the phenomena.

Before introducing the spray model it is necessary to comment on two previously observed phenomena which are central to the argument. Firstly it has been recognised that the droplet population above a sieve plate has a droplet size distribution. From their work on entrainment from multiple orifice plates, Teller and Rood (70), Cheng and Teller (71) and Aiba and Yamada (37) found that the drop-size distribution was log-normal. Nielsen (72) came to the same conclusion from his single orifice studies. In contrast, Axelrod and Yusova (36) were unable to correlate their distribution data. For the purposes of this model it is assumed that some distribution correlation does exist.

The second observation is that three of the above workers (36) (37) (71) give evidence that drops of different sizes have different initial projection velocities. In general, the smaller the drop the higher is the projection velocity. The most comprehensive coverage, in terms of hole velocity, orifice submergence, and liquid physical properties is given by Aiba and Yamada (37), and these data have been used in developing the model.

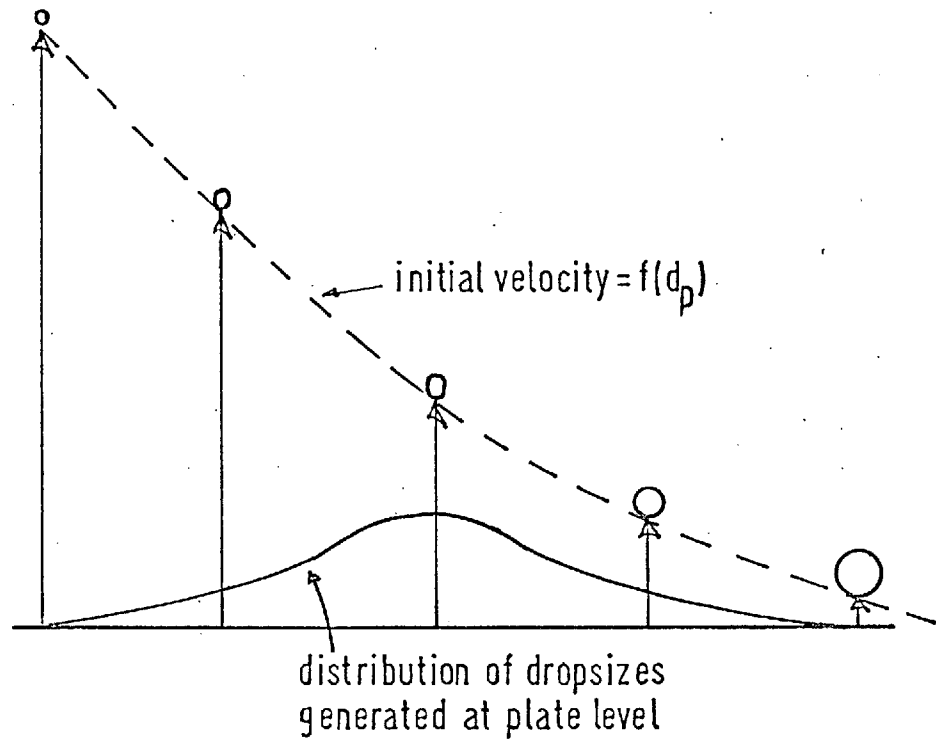
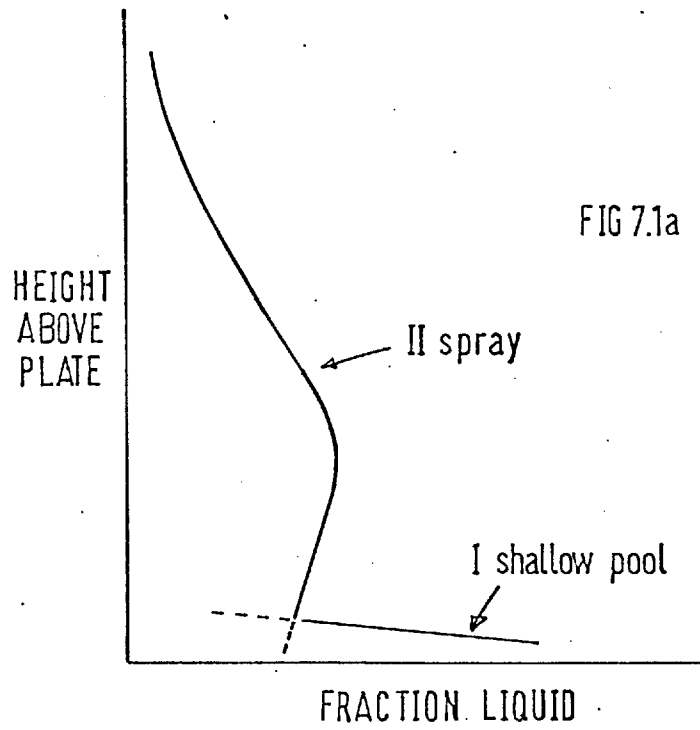


FIG 7.1b REPRESENTATION OF THE CONDITIONS AT PLATE LEVEL

Combining these two phenomena, the drops size distribution and the drop size/projection velocity relationship, with the observed maximum in the spray dispersion profile suggests an approach to 'modelling' the spray regime. The model is described and developed below.

7.2 Description of model

It is assumed that the liquid is continuously atomised at the holes at a constant rate. It is also assumed that the drop sizes are normally or log-normally distributed about a mean drop size. Each drop size has associated with it a specific projection velocity, such that this velocity is some inverse function of the drop size (the precise relationship is interpolated from Aiba's data). Bostain (73) has proposed a similar model for the prediction of entrainment.

The model is illustrated in Fig. 7.1b, which indicates a normally distributed drop size at the level of drop generation. The vertical lines are intended to represent the magnitude of the projection velocities.

Finally, it is assumed that there is no droplet interaction; that is, there is no break-up or coalescence of drops during flight.

It is suggested that only one droplet size distribution, defined by the mean drop size and the standard deviation, will be consistent with the measured spray profile. The model is used to deduce the most probable values for the distribution parameters.

It is then possible to compare the measured mass transfer performance with that expected from the calculated droplet population.

The basic requirement of the model is to provide a height/velocity map associated with each drop size. This can then be used to calculate the time spent by each drop size at all levels above the plate, which then leads to a height/hold-up relationship. This is none other than a dispersion density profile, and can be compared with the measured profile.

The first step is to describe the droplet trajectory, and in this case only the vertical components of velocity are considered. The projection velocity data of Aiba and Yamada are, in fact, the vertical components of the initial drop velocities.

7.3 Droplet trajectories

The force balance on an accelerating particle travelling vertically is;

$$\begin{aligned} \text{force on particle} = & \text{gravitational forces} + \text{buoyancy forces} \\ & + \text{drag forces.} \end{aligned}$$

$$\rho_p \frac{\pi}{6} d_p^3 \frac{dV}{dt} = -\frac{\pi}{6} d_p^3 \rho_p g + \frac{\pi}{6} d_p^3 \rho_G g + C_D \frac{\pi}{4} d_p^2 \rho_G \frac{(V - V_s)^2}{2} \quad \dots (7.1)$$

which reduces to

$$\frac{dV}{dt} = V \frac{dV}{dh} = -g \frac{(\rho_p - \rho_G)}{\rho_p} + \frac{3C_D (V - V_s)^2 \rho_G}{4\rho_p d_p} \quad \dots (7.2)$$

where V = vertical component of particle velocity (positive in upward direction)

V_s = background velocity; in this treatment the background velocity is assumed to be well approximated by the column superficial velocity, since the dispersion density is known to be low. (note: when $V > V_s$ the sign on the drag term is negative, when $V < V_s$ the sign is positive).

ρ_p, ρ_G = densities of particle (or drop) and continuous phase

d_p = particle diameter

C_D = drag coefficient

$$\text{Also, } C_D = f_1(\text{Re}), \text{ where } \text{Re} = \frac{d_p |V - V_s| \rho_G}{\mu_G} \quad \dots (7.3)$$

$$\text{Re} < 2.0, \quad C_D = 24/\text{Re} \quad \dots (7.4)$$

$$2.0 < \text{Re} < 1000, \quad C_D = 18.5/\text{Re}^{0.6} \quad \dots (7.5)$$

$$\text{Re} > 1000, \quad C_D = 0.3 \quad \dots (7.6)$$

The correlations above are taken from the data of Hughes and Gilliland (74).

Re-arranging equation 7.2

$$\int_{h_1}^{h_2} dh = \frac{-4 \rho_p d_p}{3 \rho_G} \int_{V_1}^{V_2} \frac{V dV}{\frac{4}{3} \rho_G d_p (\rho_p - \rho_G) \pm C_D (V - V_s)^2} \quad \dots (7.7)$$

The above equation provides the relationship between height and velocity for a given drop size.

By numerically integrating this equation over increments of h a height/velocity map can be obtained.

7.4 Height/Velocity map

The interplate space is divided into $N-1$ increments, as shown schematically in Fig. 7.2.

Consider a drop J . The initial velocity, V_{INIT} , is given by,

$$V_{INIT} = f_2(d_p) = 0.4 \times (0.004/d_p)^{0.93} \quad \dots (7.8)$$

which is derived from the data of Aiba and Yamada.

The right hand side of equation 7.7 is numerically integrated until the value of V_2 satisfies

$$\int_{h_1}^{h_2} dh = DH, \text{ the chosen height increment.}$$

This value of V_2 is stored, and the procedure continues to the next level, until ITERM is reached. In this increment the drop velocity changes sign, i.e. the drop reaches the top of its trajectory. In this terminal increment the procedure differs in that it is necessary to calculate the first downward increment. Equation 7.7 is integrated with $V_2 = 0$, the result being a value for the maximum height, H_{max} . Knowing the height at which $V = 0$

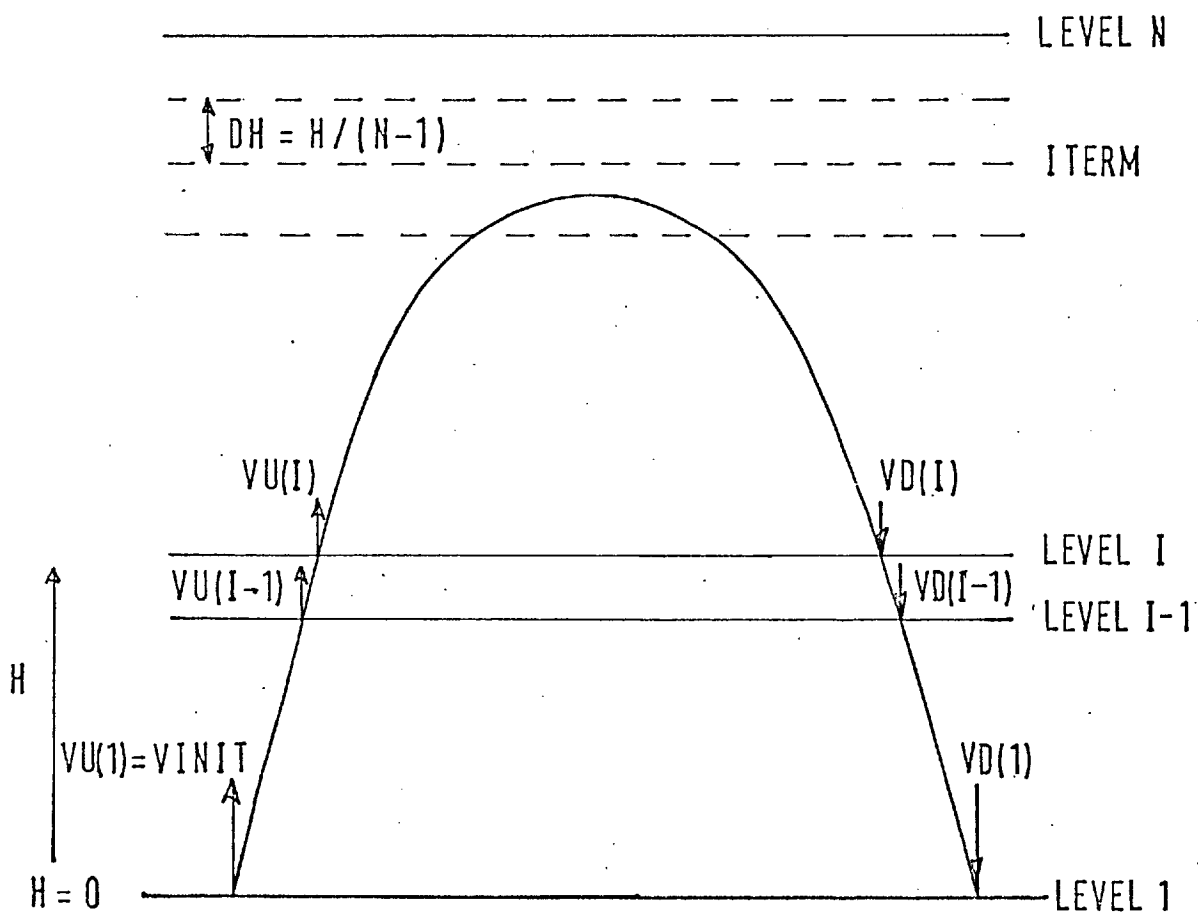


FIG 7.2 REPRESENTATION OF DROP TRAJECTORY

it is then possible to continue the integration of equation 7.7 to give downward velocities at each level, until level 1 is reached.

From this information it is possible to calculate the time spent by the drop J, in each increment;

$$TUT_{(i)} = 2DH/(W_{(i)} + VU_{(i-1)}) \quad i < ITERM \quad \dots (7.9)$$

$$TDT_{(i)} = 2DH/(VD_{(i)} + VD_{(i-1)}) \quad i < ITERM \quad \dots (7.10)$$

$$TT_{(i)} = TUT_{(i)} + TDT_{(i)}$$

where $TUT_{(i)}$ = time spent by drop in upward journey between levels I-1 and I;

and $TDT_{(i)}$ = time spent by drop in downward journey between levels I and I-1.

For terminal increment $I = ITERM$, equation 7.7 is rearranged

$$\int_0^{TT(ITERM)} dt = \frac{-4\rho_p d_p}{3\rho_G} \int_{VU(ITERM-1)}^{VD(ITERM-1)} \frac{dV}{\frac{4}{3}\rho_p (\rho_p - \rho_G) g + C_D (V - V_S)^2} \quad \dots (7.11)$$

This is numerically integrated to give $TT(ITERM)$

The result of this analysis is a height/velocity/time map for the drop J. It should be noted that any drop reaching level N is assumed to be irrevocably entrained.

In order to obtain hold-up data it is necessary to determine the height/velocity/time map for a range of drop sizes ($J = 1$ to M), which are assumed to represent the drop size distribution.

7.5 Hold-up calculations

Let rate of generation of drops = N_p (drops/sec.unit area of plate)

Assume the droplet population is expressed by a distribution function P , which is a function of d_p , such that

$$\int_{d_{p \min}}^{d_{p \max}} P(d_p) d(d_p) \approx 1 \quad \dots (7.12)$$

Fraction of drops in size range $\delta d_p = P(d_p) \delta d_p = F(j) \quad \dots (7.13)$

Rate of generation of drops in size range $\delta d_p = N_p P(d_p) \delta d_p \quad \dots (7.14)$

where;

$$P(d_p) = \frac{1}{s \sqrt{2\pi}} \exp\left(-\frac{1}{2s^2} (X - \bar{X})^2\right) \quad \dots (7.15)$$

and $X = d_p$; $\bar{X} = d_{pm}$ for a normal distribution,

$X = \ln d_p$; $\bar{X} = \ln d_{pm}$ for a log-normal distribution.

The size distribution is characterised by the mean drop size d_{pm} and the standard deviation s . In addition the values ascribed to $d_{p \max}$ and $d_{p \min}$ will have some influence. The value of $d_{p \min}$ is set to an arbitrary low value, for which the drop frequency is vanishingly small. This approach is unacceptable for $d_{p \max}$ as pointed out by Mugele (45), and his criteria are applied for estimating $d_{p \max}$; see equation 2.38. In this case the relative

velocity between the phases at drop formation is taken as the hole velocity.

The spectrum of drop sizes is taken to be represented by

$$dp(j) = d_{p(1)} \rightarrow d_{p(M)} \quad \dots (7.16)$$

$$\text{and the size range } \delta d_p = (dp_{\max} - dp_{\min})/M \quad \dots (7.17)$$

Combining equations 7.13, 7.15 and 7.17, for a normal distribution,

$$F(j) = \frac{(d_{p_{\max}} - d_{p_{\min}})}{M} \frac{1}{s\sqrt{2\pi}} \cdot \exp\left(-\frac{1}{2} \frac{(dp(j) - dp_m)^2}{(2s)^2}\right) \quad \dots (7.18)$$

To allow for the fact that equation 7.12 is inexact, the values for $F(j)$ are standardised by

$$F'(j) = F(j) / \sum_{j=1}^{j=M} F(j) \quad \dots (7.19)$$

Now rate of generation of drops in size range J

$$= N_p \cdot F'(j)$$

Holdup of drops (J) in increment I, which is bounded by levels I and I - 1

$$= \text{Time spent by drop (J) in increment} \times \text{rate of generation of drops (J)} \times \text{volume of drop (J)}$$

$$= TT_{(j,i)} \cdot N_p \cdot F'(j) \cdot \frac{\pi}{6} \cdot d_{p(j)}^3 \quad \dots (7.20)$$

Thus total hold-up of drops in I, per unit area of plate is

$$\phi_i = \sum_{j=1}^{j=M} TT(j,i) \cdot N_p \cdot F(j) \cdot \frac{\pi}{6} \cdot d_p^3(j) \quad \dots (7.21)$$

Re-arranging

$$\frac{\phi_i}{N_p} = \sum_{j=1}^{j=M} \theta_{ji} \quad \dots (7.22)$$

$$\text{where } \theta_{ji} = TT(j,i) \cdot F(j) \cdot \frac{\pi}{6} \cdot d_p^3(j)$$

Summing equation 7.22 overall all levels

$$\frac{1}{N_p} \sum_{i=2}^{i=N} \phi_i = \sum_{i=2}^{i=N} \sum_{j=1}^{j=M} \theta_{ji} = \frac{H_L}{N_p} \quad \dots (7.23)$$

where H_L is the liquid holdup per unit area of plate, and is obtained by integration of the experimentally determined dispersion density profile

$$(F_{LE})_i = f_3(H)_i \quad \dots (7.24)$$

as described in section 6.2.1.

F_{LE} is the fraction of liquid in the dispersion at level H .

Combining equations 7.23 and 7.22 to eliminate N_p

$$\phi_i = H_L \left(\sum_{j=1}^{j=M} \theta_{ji} \right) / \sum_{i=2}^{i=N} \sum_{j=1}^{j=M} \theta_{ji} \quad \dots (7.25)$$

Equation 7.25 is used to give values of ϕ_i . The dispersion density in increment I

$$= \frac{\text{volumetric liquid hold-up}}{\text{total volume of increment}}$$

$$\text{thus } (F_{LC})_i = \phi_i / (DH.1) \quad \dots (7.26)$$

where F_{LC} is the calculated fraction of liquid in the dispersion at level H. It is thus possible to obtain the profile

$$(F_{LC})_i = f_4(H)_i \quad \dots (7.27)$$

This profile is a function of the chosen drop size distribution parameters, i.e. dpm and s. Adjustment of dpm moves the position of the maximum F_{LC} , whilst adjustment of s will alter the spread. By comparing the measured and the calculated profiles an estimate can be made of the size distribution characteristics.

The values of dpm and s are chosen to minimise the residual sum of squares

$$R_s = \sum_{i=2}^{i=N} (F_{LE} - F_{LC})_i^2 \quad \dots (7.28)$$

A program has been written to carry out the analysis outlined above, and the flowcharting is presented in Appendix V. The minimisation of R_s was assisted by using a subroutine due to Powell (75).

7.6 Mass Transfer

7.6.1 Calculation of the number of transfer units

Having obtained a 'best fit' droplet population from hydrodynamic considerations it is then possible to estimate the mass transfer performance of the dispersion. This can be compared with the measured plate performance.

The number of transfer units in increment I is given by

$$(N_{OG})_i = (\Delta C')_i / (C'^* - C')_i \quad \dots (7.29)$$

where C' is the vapour phase composition (in moles per unit volume), and C'^* is the vapour composition in equilibrium with the bulk liquid composition.

This can be written

$$(\Delta N_{OG})_i = ((K_{OG})_{Ui} \Delta S_{Ui} + (K_{OG})_{Di} \Delta S_{Di}) / V_s \quad \dots (7.30)$$

which assumes that the overall driving force is the same for ascending and descending drops. The subscripts U and D, denote the ascending and descending values of the properties.

The mass transfer coefficients K_{OGU} and K_{OGD} are obtained from equation 2.1.

$$\frac{1}{K_{OG}} = \frac{1}{K_G} + \frac{m}{K_L} \quad \dots 2.1$$

As discussed in Section 7.6.2 the values of the individual coefficients will depend either on the local value of the Reynolds number, or on the age of the drop, both of which differ for the ascending and descending conditions.

Summing equation 7.30 over all levels

$$N_{OG} = \sum_{i=2}^{i=N} \frac{((K_{OG})_{Ui} \Delta S_{Ui} + (K_{OG})_{Di} \Delta S_{Di})}{V_s} \quad \dots (7.31)$$

Now

$$(K_{OG})_{Ui} \Delta S_{Ui} = \sum_{j=1}^{j=M} (K_{OG})_{U(j,i)} \cdot a_j \cdot N_p \cdot F'(j) \cdot TUT(j) \quad \dots (7.32)$$

where a_j is the surface area of drop J. A similar equation can be written for the descending term.

Putting 7.32 into equation 7.31, and eliminating N_p by using equation 7.23 the number of transfer units is

$$N_{OG} = \frac{H_L}{V_s} \frac{\sum_{j=1}^{j=M} \sum_{i=2}^{i=N} (Z_{U(j,i)} + Z_{D(j,i)})}{\sum_{j=1}^{j=M} \sum_{i=2}^{i=N} \theta_{ji}} \quad \dots (7.33)$$

where $Z_{Uji} = (K_{OG})_{U(j,i)} \cdot a_j \cdot F(j) \cdot TUT(j,i)$

Equation 7.33 has been used as a basis for the mass transfer calculations carried out after analysis of the dispersion drop size distribution (Section 7.5).

7.6.2 Individual mass transfer coefficients

i Gas phase

A number of correlations are available for convective mass transfer from a spherical drop. The relationship due to Ranz and Marshall (76) has been used. This is

$$Sh_G = 2.0 + 0.6 Re_G^{\frac{1}{2}} Sc_G^{\frac{1}{3}} \quad \dots (7.34)$$

where Re is defined by equation 7.4.

It is evident from equation 7.34 that the individual gas phase mass transfer coefficient is dependent on drop size and on relative velocity; this latter property is itself a property of height, direction of travel and drop size (equation 7.7).

ii Liquid phase

The individual liquid phase mass transfer coefficient to be introduced into equation 2.1 is obtained from consideration of

the unsteady state mass transfer into a rigid sphere. For the condition of constant surface concentration and constant diffusivity, the solution of the diffusion equation by Carslaw and Jaeger (77) can be applied.

$$\frac{C - C_o}{C_s - C_o} = \frac{R}{r} \sum_{n=0}^{\infty} \left\{ \operatorname{erfc} \frac{(2n+1)R - r}{2\sqrt{Dt}} - \operatorname{erfc} \frac{(2n+1)R + r}{2\sqrt{Dt}} \right\} \quad \dots (7.35)$$

where C is the liquid concentration at radius r in the sphere of radius R . C_s is the surface concentration and C_o is the initial bulk concentration. This solution converges rapidly for small values of time.

Now the mass flux at the surface of the sphere at time t is

$$N_s(t) = -D \left(\frac{\partial C}{\partial r} \right)_{r=R} \quad \dots (7.36)$$

In terms of the liquid mass transfer coefficient

$$N_s(t) = K_L(t) \cdot (C_o - C_s) \quad \dots (7.37)$$

Equation 7.35 is differentiated with respect to r at $r = R$.

The liquid phase mass transfer coefficient is found by equating equations 7.36 and 7.37, which gives

$$K_L(t) = \sqrt{\frac{D}{\pi t}} \left\{ 1 + 2 \exp\left(-\frac{R^2}{Dt}\right) \right\} - \frac{D}{R} \quad \dots (7.38)$$

The negative term compensates for the fact that the composition at the centre of the drop will not remain at C_o if t becomes sufficiently large. Once this happens the assumption of constant driving force

$C_o - C_s$, will be invalidated.

For small values of t , this is well approximated by

$$K_L(t) = \sqrt{\frac{D}{\pi t}} - \frac{D}{R} \quad \dots (7.39)$$

According to Hughes and Gilliland (61) (74) internal motion tends to occur in droplets due to oscillations initiated at formation or caused by drag or turbulence effects. Garner (78) observed internal motion in liquid droplets caused by drop deformation and internal circulation set up by viscous drag. He noted that internal circulation occurred at lower flow rates if diffusion was taking place. According to Davies (79) deformation occurs when the Bond number. $(\Delta \rho d^2 g / \sigma)$ is greater than 0.4. This condition is met in the majority of cases considered in this analysis.

As a result it is to be expected that the droplets in the spray dispersion will deviate from the rigid sphere model, and that deformation and internal motion will occur. This momentum transport within the drops will augment the molecular diffusion process. To allow for this an effective diffusivity is used in equation 7.39, rather than the molecular diffusivity. Calderbank and Korschinsky (80) have tabulated values of the ratio of the effective to the molecular diffusivity from 1.9 to 70, for a wide range of conditions. The results of Hughes and Gilliland (61) for the absorption of CO_2 into falling drops of water indicate a diffusivity ratio of approximately 10.

Equation 7.39 becomes

$$K_L(t) = (D_E/(\pi t))^{1/2} - 2D_E/d_p \quad \dots (7.40)$$

It is found that the ratio D_E/D_M has a marked effect on the calculated performance at low values of the ratio. This is discussed more fully in section 7.8.

7.7 Testing the model

In the initial tests on the model attention was focused on the choice of step lengths to be used, since these affected not only the accuracy of the procedure but also the computing time. Thus it was found that the number of height increments, N , could be taken as 30, a value of 50 giving no difference in the results but almost doubling the number of calculations. On the other hand the initially chosen value for M , the number of drop size increments, was 21 and this yielded an uneven dispersion profile; the value of M used was 41 which gave smooth profiles.

The height increment chosen was 0.5 cm. A value of 0.0001 m. was chosen as the minimum drop size, dp_{min} , and the relationship of Mugele (45) was used to estimate the maximum, dp_{max} . The value chosen for this latter property was not critical since a comparison was made with the results obtained by using both the estimated maximum value and twice this value; the differences were negligible.

Numerical integration of equation 7.7 required a means of incrementing the velocity term. For the ascending drop the velocity increment for level I was taken as

$$\delta V_i = (V_{i-1} - V_{i-2})/12 \quad \dots (7.41)$$

This value was stored and used later for the descending drop calculation in level I. The initial velocity increment was dependent on the projection velocity (a function of the drop size) and found by experience to be reasonably estimated by

$$\delta V_{i=2} = -0.008 - (6.4d_p) \quad \dots (7.42)$$

Both normal and log-normal distributions were checked as possible descriptions of the spray population at generation. There was little to choose between the two, although the log-normal model was marginally more successful for the profile fitting. Since other workers (37) (71) (72) have used the log-normal distribution to correlate entrainment data, it was decided to use this distribution throughout.

7.8 Results from the model

The model was used to analyse the spray regime data for systems B and C.

7.8.1 Dispersion parameters

Values for the mean drop size and the standard deviation have been derived from the dispersion profiles by minimising the residual sum of squares, R_s , obtained from a comparison of the experimental profile with the calculated profile. Figure 7.3 shows a comparison of experimental and calculated profiles, and it can be

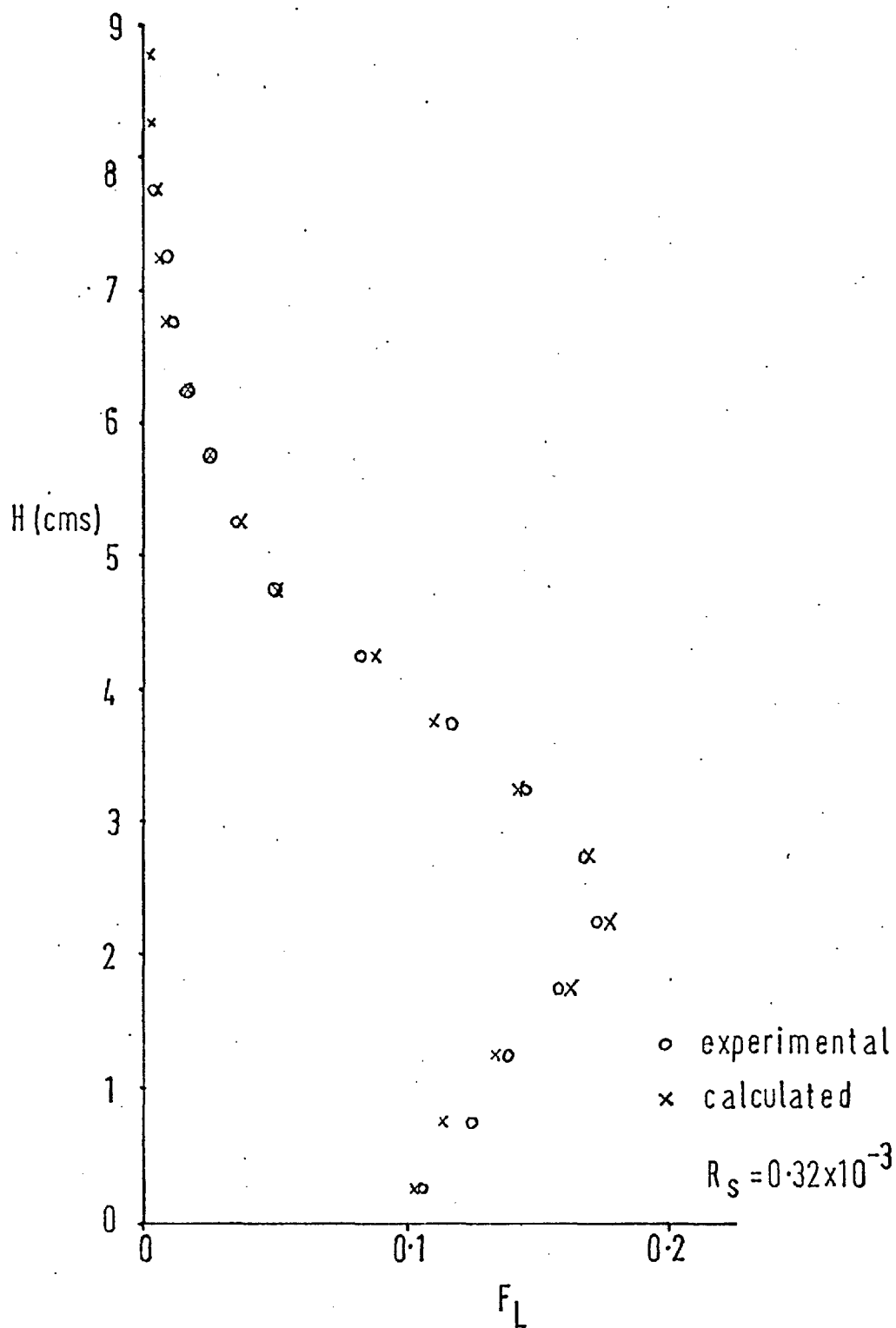


FIG 7.3 COMPARISON OF THE EXPERIMENTAL AND THE CALCULATED SPRAY PROFILES (RUN C4V4)

seen that the 'fit' is very successful.

As an illustration of the sensitivity of R_s to variation in the chosen parameters, the results of the initial minimisation leading to Fig. 7.3, are plotted in Fig. 7.4. It should be noted that the final value for R_s is one tenth the value obtained in this initial minimisation.

The calculated mean drop sizes, d_{pm} , have been corrected for variation of viscosities, liquid density and gas velocity as suggested by equations 2.41 and 2.42. The reference values are for pure benzene and a superficial gas velocity of 1 m/s. In Fig. 7.5 the corrected drop sizes d' are plotted against surface tension. Considering the deductive nature of the estimating method the scatter is not great. It can be seen that the negative system has a smaller drop size than the positive system at a given value of surface tension. This finding accords with the results shown in Fig. 6.24 of the previous chapter, which suggested that the negative system had proportionally more of the hold-up in the upper zones of the dispersion than the positive system. In addition it is evident that for both systems the drop size increases with surface tension. The dependence of drop size on surface tension is different for the two systems.

$$\text{System B (negative), } d' \propto \sigma^{0.25}$$

$$\text{System C (positive), } d' \propto \sigma^{0.55}$$

According to the spray analogy the exponent would be

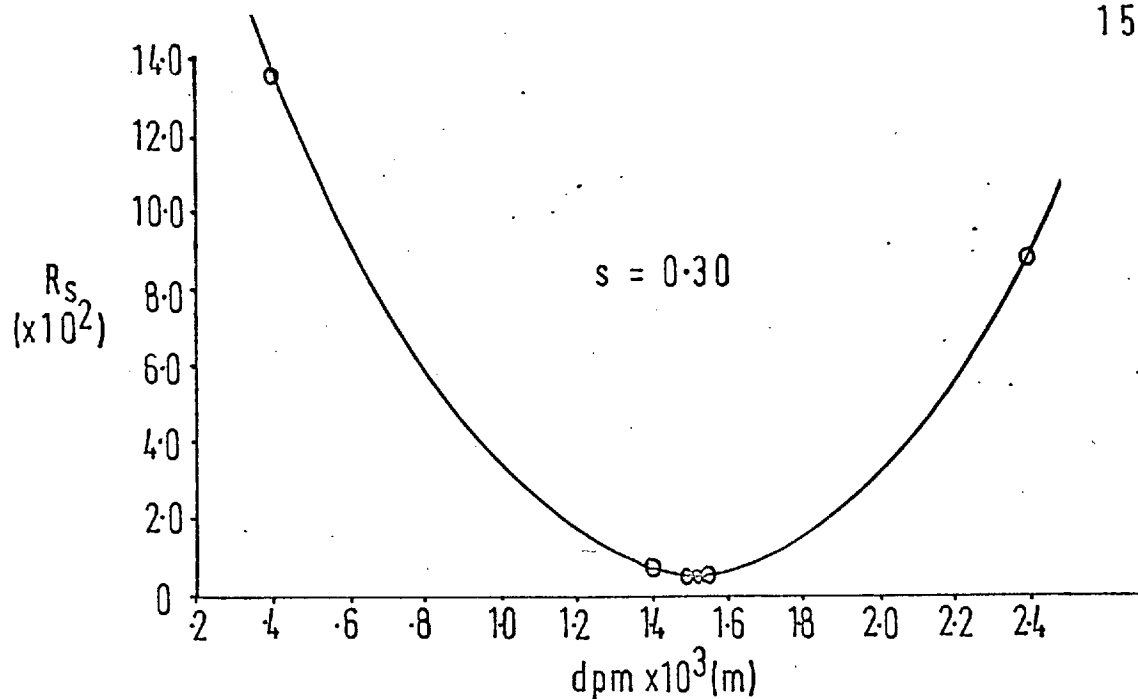


FIG 7.4a EFFECT OF THE MEAN DROPSIZE ON THE RESIDUAL SUM OF SQUARES FOR THE PROFILE FITTING OF RUN C4V4

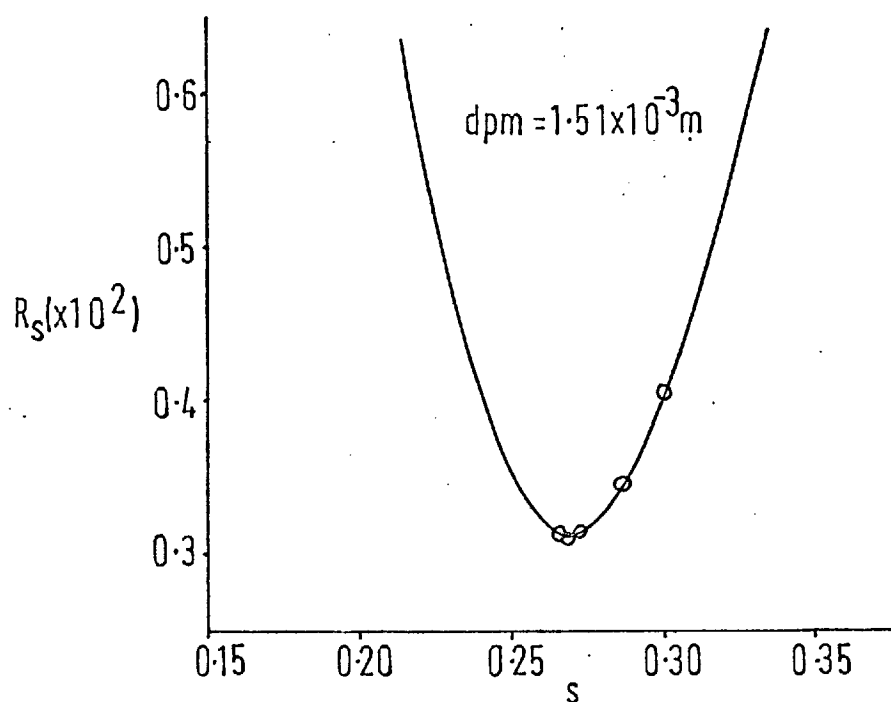


FIG 7.4b EFFECT OF THE LOG NORMAL STANDARD DEVIATION ON R_s

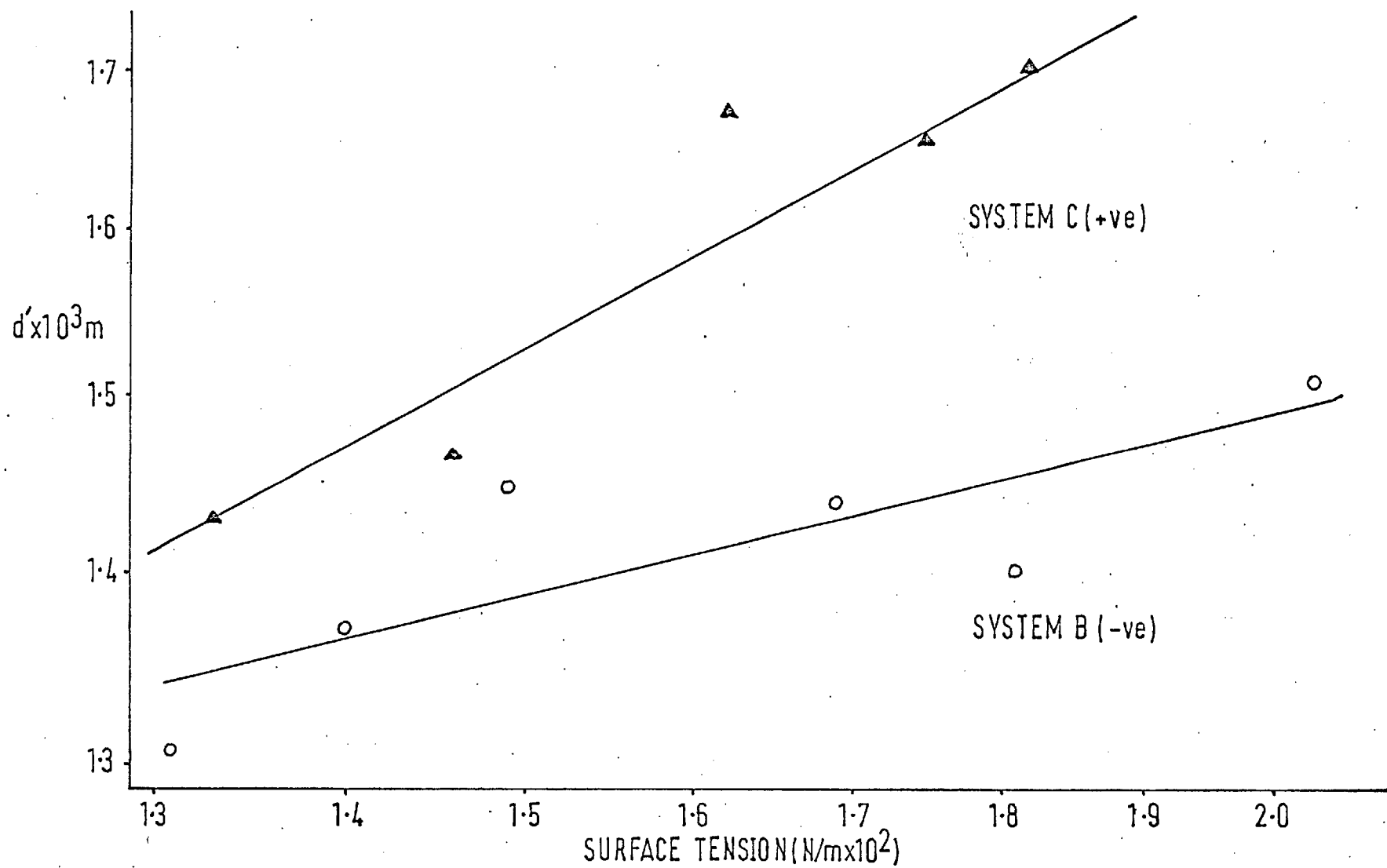


FIG 7.5 VARIATION OF MEAN DROPSIZE WITH SURFACE TENSION.

expected to be in the range 0.16 to 0.3 (46)(47). It appears that the negative system is adequately covered by this, but the positive system is outside the range. It is noteworthy that if the mean drop size is corrected only for the variation in the velocity, and the variation in the physical properties is neglected the dependence becomes,

$$\text{System B, } d' \propto \sigma^{0.23}$$

$$\text{System C, } d' \propto \sigma^{0.32}$$

The other distribution parameter is the standard deviation. No pattern emerged to indicate a dependence of this property on the other parameters. However the values for the negative system were all higher than those for the positive system 0.30 ± 0.03 and 0.19 ± 0.01 respectively. This indicates a narrower spectrum of sizes for the positive system.

7.8.2 Mass transfer

Mass transfer calculations were carried out on the deduced droplet populations. It was first necessary to obtain an estimate of the effective diffusivity. Fig. 7.6 shows the variation of the calculated performance N_{OG} with diffusivity ratio $(D_E/D_M)_L$. A maximum is reached as the negative term in equation 7.40 acquires increasing significance. Evidently N_{OG} is critically dependent on the chosen value of D_E/D_M for values of the ratio less than 5. Since the calculated performance is always less than the measured performance the ratio has been taken

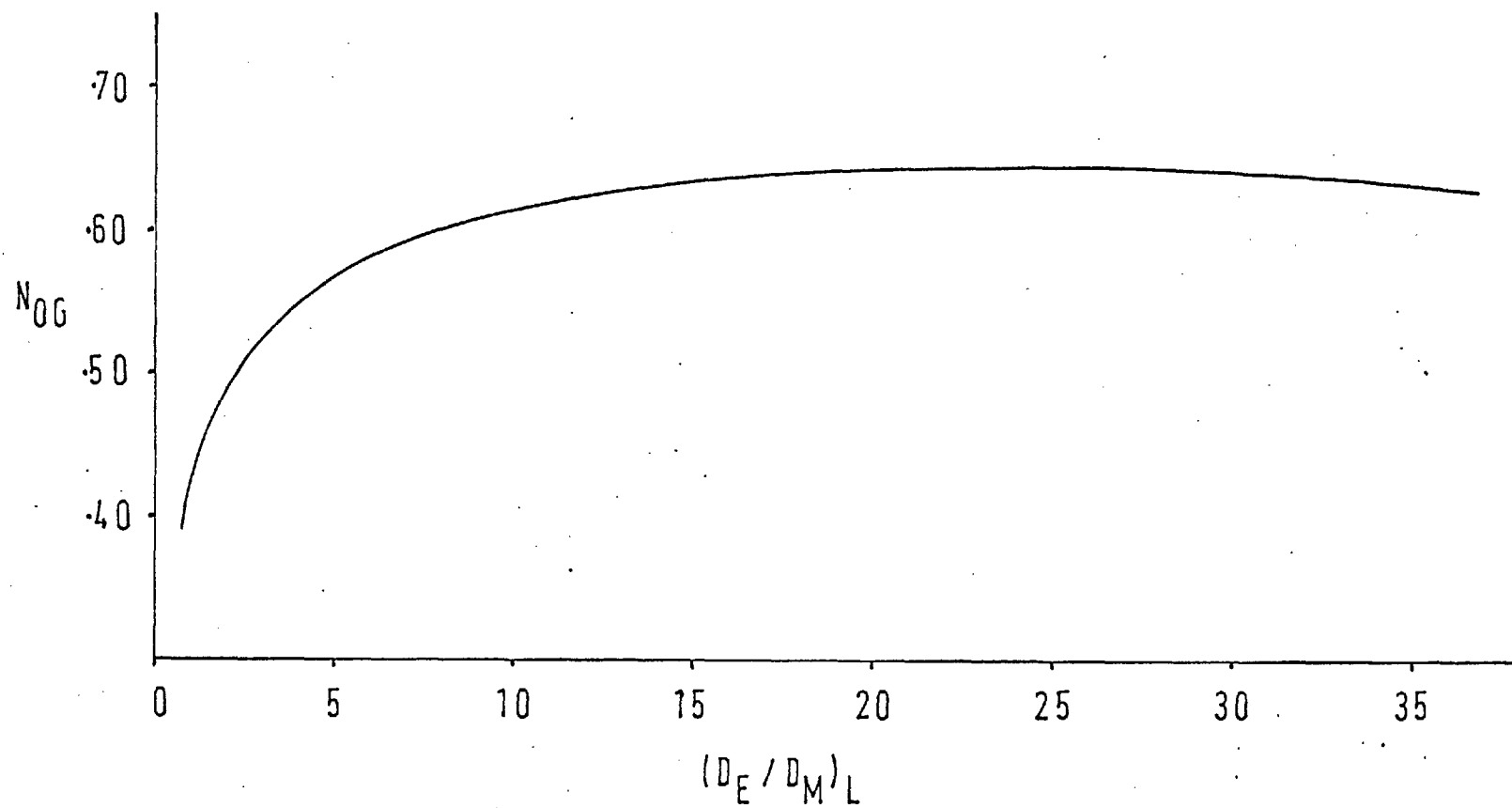


FIG 7.6 EFFECT OF VARIATION IN THE LIQUID DIFFUSIVITY RATIO ON THE CALCULATED PERFORMANCE

as 20, which gives the closest agreement between the two values. The experimental and the calculated performance, expressed as plate efficiency E_{MV} , are plotted in Figs. 7.7 and 7.8. In both cases the calculated performance is slightly lower than the measured value (ca 10% for system B and 5% for System C); although the variations with composition are followed quite closely.

This underestimation of the plate performance is no doubt a consequence of the simplifications made in developing the model. As discussed in the next chapter the most significant omissions may be the effect of mass transfer in the shallow pool (Fig. 7.1a) and the effect of drop coalescence and break-up during flight.

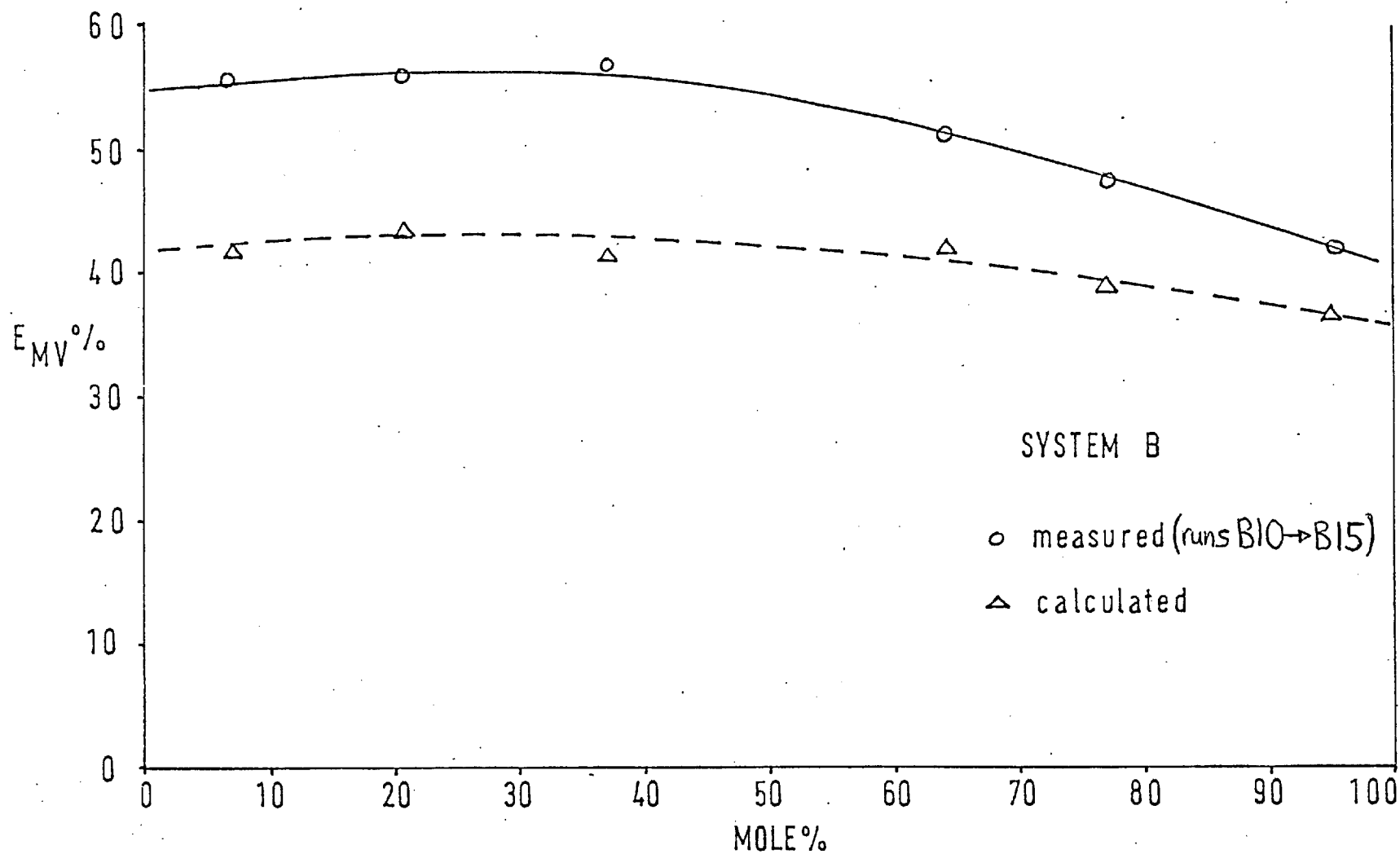


FIG 7.7 COMPARISON OF EXPERIMENTAL AND CALCULATED PERFORMANCE

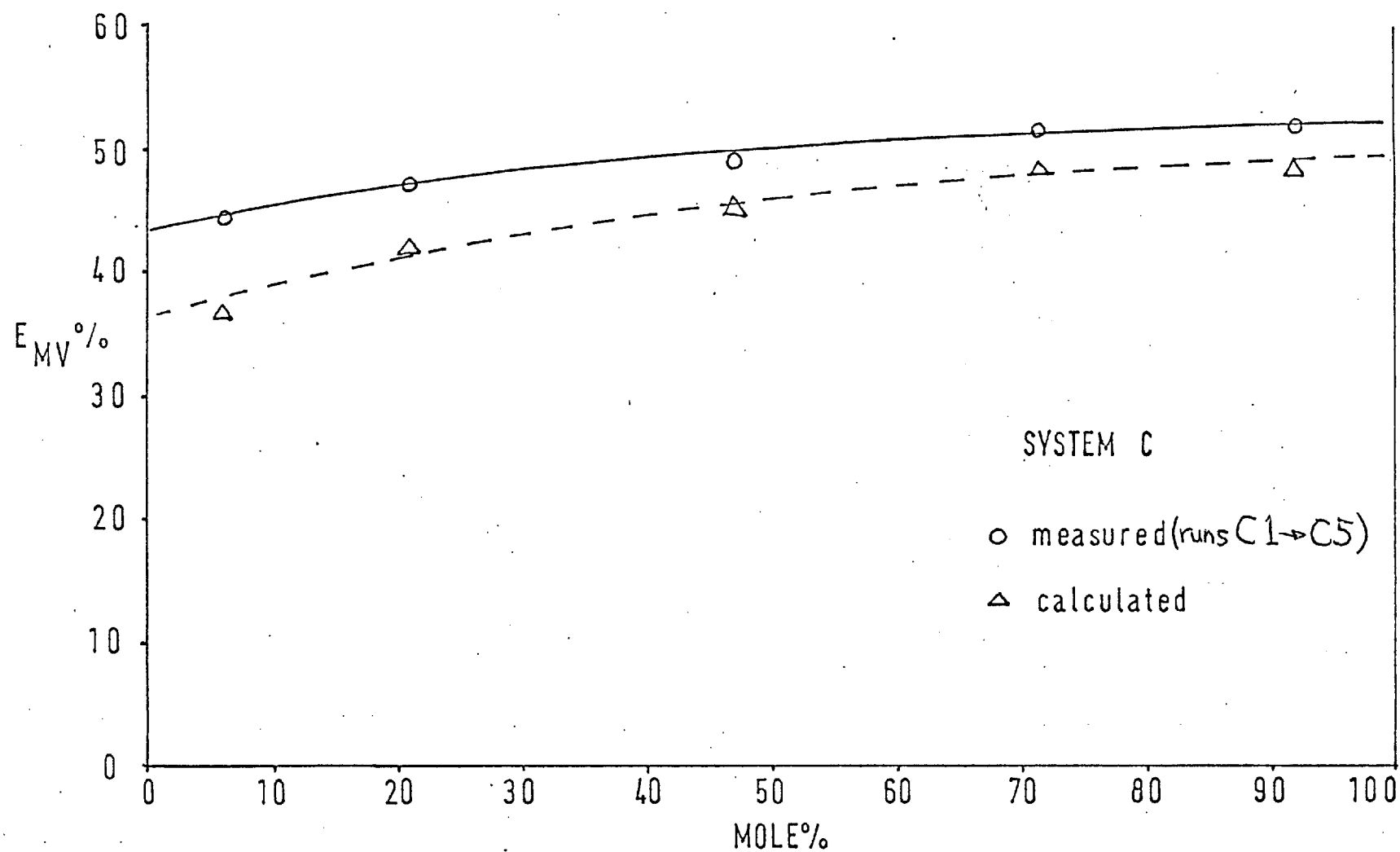


FIG 7.8 COMPARISON OF EXPERIMENTAL AND CALCULATED PERFORMANCE

CHAPTER EIGHT

DISCUSSION

8.1 Positive and Negative systems in sieve plate distillation

8.1.1 Regimes of operation

It has been shown that a variety of gas/liquid dispersions can be generated on a sieve plate, from cellular foam to mobile foam to spray. These dispersions, or regimes, are characterised by the nature of the dispersed phase, and this in turn defines the type of gas/liquid interface. Cellular foam is composed of discrete bubbles of gas within a liquid, spray consists predominantly of liquid droplets in contact with a rising gas stream, and mobile foam describes the type of dispersion which lies between these two extremes. It should be noted that mobile foam cannot be termed a transition regime, since cellular foam is not always encountered.

The basic difference between surface tension positive and surface tension negative systems is the ability of the former to promote cellular foam. According to Zuideweg and Harmens (6) this is because for positive systems the mass transfer process encourages the Marangoni (81) effect. This effect creates forces stabilising the interface and thereby aids the formation of rigid (or cellular) foams. The opposite behaviour applies to negative systems, and the rupture of the lamellae between bubbles is encouraged, giving a non-cellular dispersion.

This difference in behaviour occurs at low gas rates, as has been shown in this work. An increase in the gas rate counters the stabilization of the interface in a positive system and coalescence occurs. There is a gradual transition from cellular to mobile foam, and the plate performance falls as the gas residence time and the magnitude of interfacial area are reduced. It is apparent from the dispersion density profiles that as the gas rate is increased more of the liquid is transferred to the upper zones of the dispersion. Similar behaviour pertains to the negative system although the type of dispersion changes from a mobile foam to a highly agitated foam with an appreciable droplet population. For the negative system the variation of plate performance with gas rate depends on the weir height. Over the range of gas loadings used in these investigations the performance falls for the 7.6 cm. weir, is constant for the 1.9 cm. weir and increases for the 0.96 cm. weir. For the two lower weirs it would appear that the increase in surface area due to the droplet population counter balances or outweighs the reduced gas residence time.

At high gas rates and low weir heights the shear forces between gas and liquid are such as to atomise the liquid, and give a dispersion composed of liquid droplets, i.e. the spray regime. This applies to both positive and negative systems. Once this regime is encountered the height of the weir is of minor significance

in determining the plate performance, as observed by Yoshida et al (8).

8.1.2 The criteria of Zuiderweg and Harmens

As pointed out by Bainbridge and Sawistowski (2) the criteria of Zuiderweg and Harmens are of limited applicability. That positive systems exhibit cellular foam behaviour and consequent high performance, and negative systems give a more spray like dispersion with lower performance is not disputed. However it is clear that this behaviour occurs at low gas rates and low weir heights, and for other conditions of flow rate and plate geometry the phenomena differ. Thus at high gas rates and low weir heights both types of system give a liquid phase dispersed spray regime. Moreover as observed by Bainbridge and Sawistowski, and confirmed in this work, the positive system gives a lower performance than the negative system in the spray regime. It has also become evident that the method of comparison of system performance should not be limited to a Schmidt number correction (equation 2.10) but should include parameters influencing the formation of the dispersion. An analogy with gas/liquid atomisation has yielded a means of correcting for fluid viscosity and density effects (equation 2.44) as well as allowing for the contribution due to the hole velocity (equation 2.43). It is appreciated that this analogy is only a first step towards allowing for the factors affecting the drop size, and hence the mass transfer surface, but

it is felt that some correction is valid. With or without the correction the negative system gives a higher performance than the positive.

A further aspect of the analogy is that it indicates an increase in performance with a decrease in surface tension, consistent with a reduction in drop size and corresponding increase in surface area. This accords with the findings in this work and that of other workers (2) (20). Both positive and negative systems show this behaviour.

Summarising, it would appear that the criteria of Zuiderweg and Harmens only apply when the conditions are such as to promote cellular foam behaviour in a positive system and mobile foam or spray in a negative system.

8.1.3 Practical applications

The type of dispersion generated on a sieve plate is a function not only of surface tension properties but also of flow rates and tray design, and its choice is therefore one of the design decisions.

Positive systems exhibit their highest efficiency when allowed to operate in the cellular foam regime. As noted in section 2.3.1. this is favoured by low gas flow rates and low hole Reynolds numbers. In other words for a given free area small holes will have a greater tendency to encourage cellular foam formation than large holes. In addition the presence of the side

walls and a foam baffle will aid the build up of cellular foam. A precautionary note should be added here since the results of Bernard (39), on a neutral system, indicate that the presence of a foam baffle had a detrimental effect on the plate performance. A further observation is that cellular foam occurs at low gas loadings, and to achieve this a rather larger than normal column diameter will be required. In general design for cellular foam behaviour is probably only feasible for relatively small column diameters (0.5 m. and below).

For negative systems there is no point in attempting to design a tray to promote cellular foam. On the other hand high capacity columns such as those of Yoshida et al (8) and Manning (10), which operate in the spray regime will be advantageous. Thus high throughput should be encouraged when distilling negative systems, the intention being to form a liquid phase dispersed mass transfer zone. Two precautions must be added. Firstly, the problem of entrainment must not be overlooked since this is aggravated as the gas rate is increased. For this reason the design of spray tray due to Manning, which incorporates a coalescence zone, deserves more attention. Secondly it should be noted that systems with a high liquid phase resistance may not be favoured by the formation of a dispersion composed of droplets, since eddy transport can be very low within small drops (78).

The observed effect of increase in plate performance with decrease in the absolute value of surface tension suggests that spray regime operation is suitable for low surface tension systems, whether negative or positive. This includes both liquid air distillation and certain high pressure distillations in which high temperature gives a low surface tension.

Finally, the study of the spray regime seems particularly relevant to vacuum distillation in which the gas velocities are high and the volumetric liquid rates are low, these being the conditions most favouring spray formation.

8.2 The Spray regime

8.2.1 Mechanism of dispersion formation

In the spray regime a high velocity gas stream is passed up through a shallow pool of liquid. Atomisation of liquid takes place close to the plate floor and liquid droplets are projected upwards. Some of the droplets have an unimpeded trajectory from the floor through the dispersion zone and back to the floor. Others will collide with other droplets, either coalescing or subdividing, and some will strike the side walls.

In a fundamental study of the mechanism of atomisation of liquids Castleman (82) assumed that a necessary step in the formation of discrete droplets from a mass of liquid is the formation of ligaments. He used the photographic evidence of Scheubel (83) to illustrate this, and it has subsequently been

confirmed by other workers employing a variety of atomisers (88) (89) (90).

Bainbridge and Sawistowski (2) in their work on distillation suggested that drops form by the rupture, or "necking-down", of ligaments. In a preliminary study in this work the necking-down stage was observed by flash photography. Subsequent work by Nielsen (72) and Lindsey (84) has clarified the picture somewhat.

Nielsen studied entrainment from a single orifice, using flash photography to gain evidence of the mechanism. He observed that at moderate to high vapour rates there is a continuous vapour channel, of fluctuating diameter, from the orifice to the liquid surface. As a result of momentum exchange between the vapour and liquid, some of the latter is drawn up above the level of the liquid surface in the shape of an inverted hollow cone of narrow angle. Thinning and disturbances in the surface of the cone cause it to disintegrate vertically in several locations. According to Nielsen the forces of surface tension cause these unattached fragments to form into vertical ligaments having nearly circular cross sections. These ligaments neck down into drops as a result of disturbances and surface tension forces. The high local vapour velocity imparts considerable momentum to the inverted cone, or sheet, which leads to the formation of droplets having an upward projection velocity. A small horizontal vector is evident, but the

projection is almost vertical. It should be noted that the projection velocities of Aiba and Yamada (37) refer to the vertical vector only.

The observations of Lindsey (84) are in qualitative agreement with those of Nielsen, although the continuous vapour channel is not evident at his lower gas rates. This does not conflict with the basic argument that drops are formed from ligaments which themselves are formed from sheets.

Ligaments of different sizes will be formed from a given sheet. Each ligament will tend to give a particular size of drop according to Rayleigh (85), although surface tension gradients may accentuate or dampen disturbances of small amplitude. The behaviour of surface tension positive and surface tension negative systems may differ at this necking stage, as suggested by Bainbridge and Sawistowski (2). However it is possible that the major difference occurs during the ligament formation stage, which depends on the rupture of a liquid sheet. A negative system will be more prone to assist disturbances leading to sheet disintegration, and tend to give a higher proportion of thin filaments. In other words the negative system will tend to give a range of drop sizes with predominantly more small drops than the positive. Additionally since the surface tension gradients of the negative system will encourage all disturbances it is possible that a greater range of ligament sizes, and hence a greater range of drop sizes will prevail.

This analysis depends on the importance of the Marangoni (81) effect on the rapid events occurring in drop formation. According to Linde (86) Marangoni instabilities are established in less than 0.01 sec, which is a time scale of similar order of magnitude to that encountered in ligament formation.

8.2.2 Effect of surface tension

Variation of the absolute value of surface tension has been shown to alter both the directly measured plate performance and the indirectly estimated mean drop size of the dispersion.

In Chapter Two equations were developed, from a spray analogy, for predicting the dependence of mean drop size and plate performance on surface tension. The empirical relationship of Mugele (45) for atomisers was used as a basis of the analysis. Equation 2.42, developed for drop size can be written as

$$\text{corrected mean drop diameter} \propto \sigma^{0.20}$$

where the drop size is corrected for variations in vapour velocity, fluid viscosity and liquid density. If the work of Dombrowski et al (46)(47) on atomisers is applied, the exponent on σ varies from 0.16 to 0.33.

According to this investigation the exponent varies from 0.25 to 0.55 for the negative and positive systems respectively, when the drop size is corrected for all the above parameters. However if the correction is limited to vapour velocity alone the

exponents are found to be 0.23 and 0.32. Thus it would appear, from this investigation over a narrow range of surface tension, that equation 2.38 developed by Mugele for atomisers may not be applicable to sieve plates operating in the spray regime.

Mugele tested his equation against results taken from a grid tray. He obtained good agreement with the characteristic linear dimension taken as twice the slot width. In Fig. 8.1 a comparison has been made with the deduced drop sizes, corrected for velocity only, and the values predicted from equation 2.38. As can be seen the mean drop size of the positive system is well estimated if L' is taken as twice the hole diameter, d_H . The value for the negative system is not so well estimated. As suggested in section 8.2.1 the negative system can be expected to give somewhat lower drop sizes than a positive system. The atomisation data used by Mugele to develop equation 2.38 was for pure liquids, which behave as surface tension neutral systems. Neutral and positive systems can be expected to behave in a similar fashion in not encouraging the growth of small disturbances, and would therefore tend to give similar drop sizes.

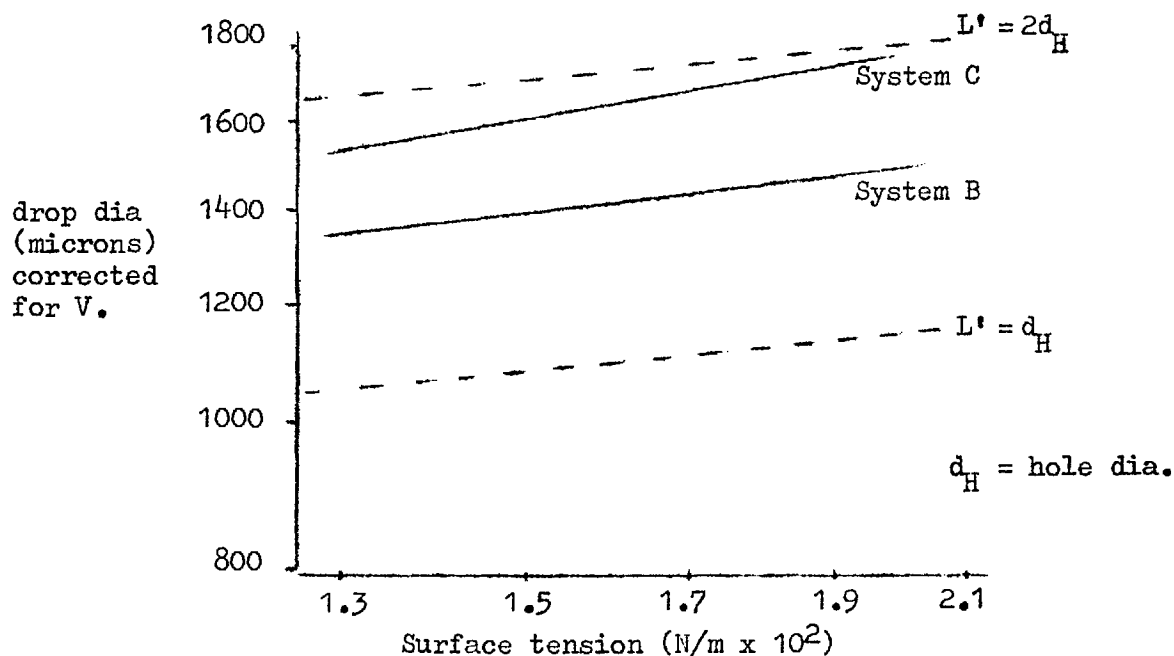
The dependence of plate performance on the absolute value of the surface tension is predicted by equation 2.45 which, for constant vapour rate, can be written

$$N_G' \propto \sigma^m$$

where $m = -0.25$ to -0.5 .

Figure 8.1

Comparison of deduced drop size with value from equation 2.38



The performance, N_G' , is corrected by equations 2.10 and 2.44 for the effect of other physical properties. This relationship is based on constant liquid hold-up and Mugele's empirical equation for mean drop size.

Experimentally determined values of m are -0.85 and -0.75 for the negative and positive systems respectively. This implies a stronger dependence on surface tension than expected. As discussed in the next section the model assumes that mass transfer only occurs between a droplet population and a rising vapour stream. This simplification neglects certain end effect contributions to mass transfer, which will themselves be dependent on flow rates and physical properties. If these dependencies are

not adequately expressed by equations 2.43 and 2.44, it is to be expected that there will be some divergence between the actual and the anticipated correlating equations.

8.2.3 Assessment of the model

The theoretical model developed in Chapter seven has proved to be a plausible description of the spray regime. In this work it has been successfully employed as a deductive method of assessing the drop size distribution parameters. Dispersion density profiles which compare well with the measured profiles, have been generated by the model.

Although the effect of entrainment has not been investigated in this work, results from the model indicate that it is of negligible significance for the conditions employed. It should be noted that the model, as formulated, can be used to estimate entrainment if it is assumed that all droplets reaching the level of the next plate are irrevocably entrained; it is similar to the entrainment model proposed by Bostain(73).

Estimates of plate efficiency from the model have been shown to be a little conservative by 5 to 10 percent. However even if the drop population is accurately determined there are two assumptions which may lead to an underestimation of plate performance. These assumptions are that the plate performance is adequately approximated by the performance in the droplet dispersion, and that no interaction (coalescence or break up) occurs between drops.

As described in Section 8.2.1 there is an initial contact stage (in the shallow pool) which consists of the passage of high velocity, and hence highly turbulent, vapour through a channel and hollow cone of liquid. For completeness some allowance should be made for this zone.

If it is assumed that the shallow pool contact zone can be approximated by a wetted wall column of diameter d_{ch} and height H_{ch} , the relationship of Gilliland and Sherwood (87) can be applied

$$K_G = 0.023 \frac{D}{d_{ch}} Re_{ch}^{0.83} Sc^{0.44}$$

where Re_{ch} is the vapour Reynolds number in the channel. This gives a shallow pool performance, N_{GO} of 0.08 to 0.04 for H_{ch} varying from 1.0 to 0.5 cms. if d_{ch} is taken as equal to the hole diameter. With d_{ch} equal to twice the hole diameter N_{GO} changes from 0.18 to 0.09 over the same range of H_{ch} . For the values of plate performance encountered 0.10 transfer units is equivalent to 6% added to the plate efficiency.

It is difficult to allow for drop coalescence and break-up. One consequence of droplet interaction is to disturb the unsteady state mass transfer process into the drop, and the result of this may be an enhancement of D_E or alteration of $K_L(t)$ owing to a change in the relative age of the drop, t . The correction involved if it is assumed that each drop has an age zero at generation, and zero at the top of its trajectory, due

to some interaction, is approximately 0.1 to 0.05 transfer units. It is appreciated that this is an unreal situation, but it serves as an illustration of the influence of the time dependent liquid phase mass transfer coefficient on the plate performance.

In conclusion it appears that the model is a suitable basis for future work, recommendations for which are given in Chapter nine.

CHAPTER NINE

CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK

9.1 Conclusions

Investigations carried out during the course of this work have confirmed the findings of Zuiderweg and Harmens (6) for sieve plate distillation at low flow rates. These findings are that surface tension positive systems tend to give higher efficiencies than negative systems as a result of the greater interfacial area presented by the cellular foam of the former compared with the mobile foam of the latter. However at higher vapour rates the cellular foam is replaced by mobile foam and the two types of system give similar performance. At high vapour rates the liquid phase becomes the dispersed phase (spray regime) and the negative system then gives a higher efficiency than the positive system, contrary to the criteria of Zuiderweg and Harmens. This is caused by the finer dispersion of the negative system.

Performance in the spray regime increases with decrease in surface tension for the range of parameters investigated. An analogy with spray atomisers is partially successful in predicting this dependence, although it slightly underestimates the significance of surface tension.

A model has been developed based on the atomisation of a continuously replenished shallow pool of liquid on the tray floor, by high velocity vapour passing through the holes. It is assumed that

there is a distribution of drop sizes generated and that each drop size has associated with it a particular projection velocity. This model has been shown to be a plausible description of the spray regime.

The model can be used as a deductive method for determining the drop size distribution parameters from dispersion density profiles. In this investigation it has indicated that for the negative system the mean drop size is smaller and the standard deviation larger than for the positive system. For both systems the drop size decreases with surface tension, as predicted from the spray analogy.

Performance data calculated by the model from known distribution data are lower than the measured values by 5 to 10 percent. This discrepancy is partly eliminated by allowing for the mass transfer taking place in the shallow pool.

9.2 Future work

Considerable research effort has been applied to sieve plate distillation in the foam regime. On the other hand the spray regime has received scant attention. Further investigation of the spray regime should prove fruitful for the following reasons:

- (i) operation in the spray regime implies high throughput, without necessarily impairing the performance.
- (ii) Since the physical picture of the spray regime is somewhat clearer than the mobile foam regime, it should be easier to model; as attempted in this thesis.

At this stage the model is dependent on certain externally assessed data. A more complete model requires information on the influence of fluid flow rates, physical properties (including surface tension gradients) and plate geometry, on the following,

- (a) the projection velocity/drop size relationship,
- (b) the drop size distribution parameters,
- (c) the liquid holdup in the spray regime,
- and (d) the end effects, namely the shallow pool transfer
and the interaction of droplets.

An alternative approach to prediction of spray regime performance lies in the development of the spray analogy.

APPENDICES

APPENDIX I

SUBSTANCE SPECIFIC CORRECTION FACTORS

The results are shown in Fig. I-1. As can be seen the curves are broken. This behaviour was unexpected when the first calibration was attempted and considerable effort was spent in perfecting the calibration techniques. This effort increased the reproducibility of the method, which is outlined below.

I.1 Calibration Technique

One of the basic problems of making up known mixtures of volatile substances is ensuring that there are negligible evaporative losses between weighings and before analysis. A special calibration bottle, Fig. I-2, was devised to minimise this evaporation and to allow mixing before analysis. The bottle was carefully dried before weighing, on a Metler semi-automatic balance. Component A, which had been chilled, was added by removing the top. After weighing, the stopper was removed, Component B, also pre-chilled, was added from a syringe, and the stopper replaced. Rapid re-weighing was possible with the automatic balance. Once made up, the mixture was cooled in an ice bath before sampling and analysis. Complete mixing was ensured by use of the stirrer. Losses were further reduced by working in an ambient temperature of less than 10°C.

A further check was obtained by making up some of the mixtures in reverse order, i.e. adding A to B, rather than B to A. No difference could be detected by the two methods.

I.2 The importance of errors

It is of interest to assess the relationship between errors in f_j and the subsequent errors in w_j , the weight fraction of the j component (the i component has f_i set = 1, by definition). From Kaiser (52).

$$(\text{Wt ratio})_j = f_j \cdot \text{Area } j / \text{Area } i \quad \dots (\text{I.1})$$

then weight fraction

$$w = A' \cdot f_j / (1 + (f_j \cdot A')) \quad \dots (\text{I.2})$$

where A' is the ratio of the peak areas $\text{Area } j / \text{Area } i$.

Differentiating with respect to f_j .

$$dw/df_j = A' / (1 + (f_j \cdot A'))^2 \quad \dots (\text{I.3})$$

From (I.3) and (I.2)

$$\frac{dw}{w} = \frac{df_j}{f_j} \left(\frac{1}{1 + f_j \cdot A'} \right) \quad \dots (\text{I.4})$$

For $f_j \approx 1$, and a small finite change in f_j , equation I.4 can be written

$$\frac{\Delta w}{w} = \frac{\Delta f_j}{f_j} \left(\frac{1}{1 + A'} \right) \quad \dots (\text{I.5})$$

For $A' \gg 1$

$$\frac{\Delta w}{w} \rightarrow \text{zero} \quad \dots (\text{I.6})$$

and $A' \ll 1$

$$\frac{\Delta w}{w} \rightarrow \frac{\Delta f_j}{f_j} \quad \dots (\text{I.7})$$

This means that for low concentrations of j , the relative errors in w will be the same magnitude as errors in the value of f_j . However, it follows from equation I.6, that at high concentrations of j , the errors in w caused by errors in f_j are of diminishing significance.

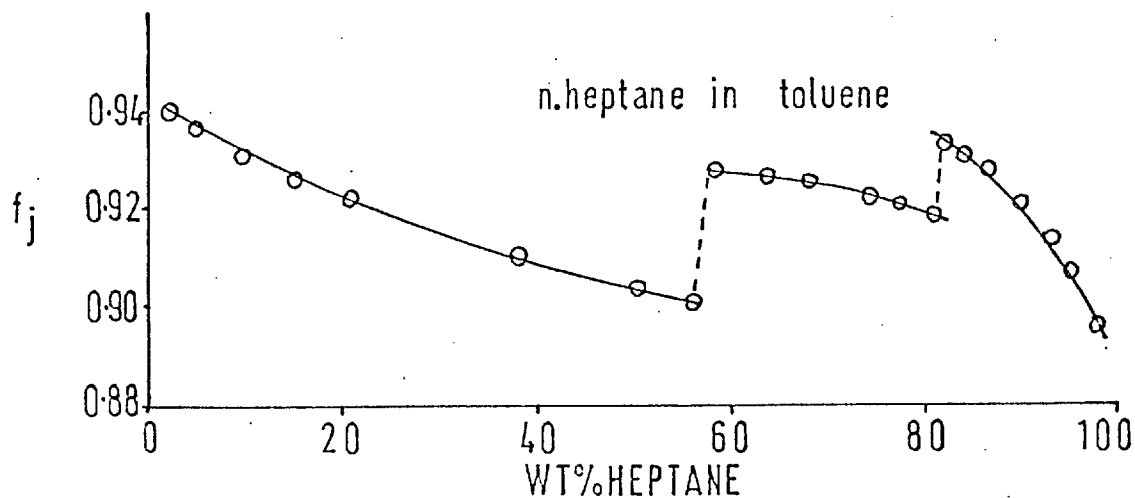
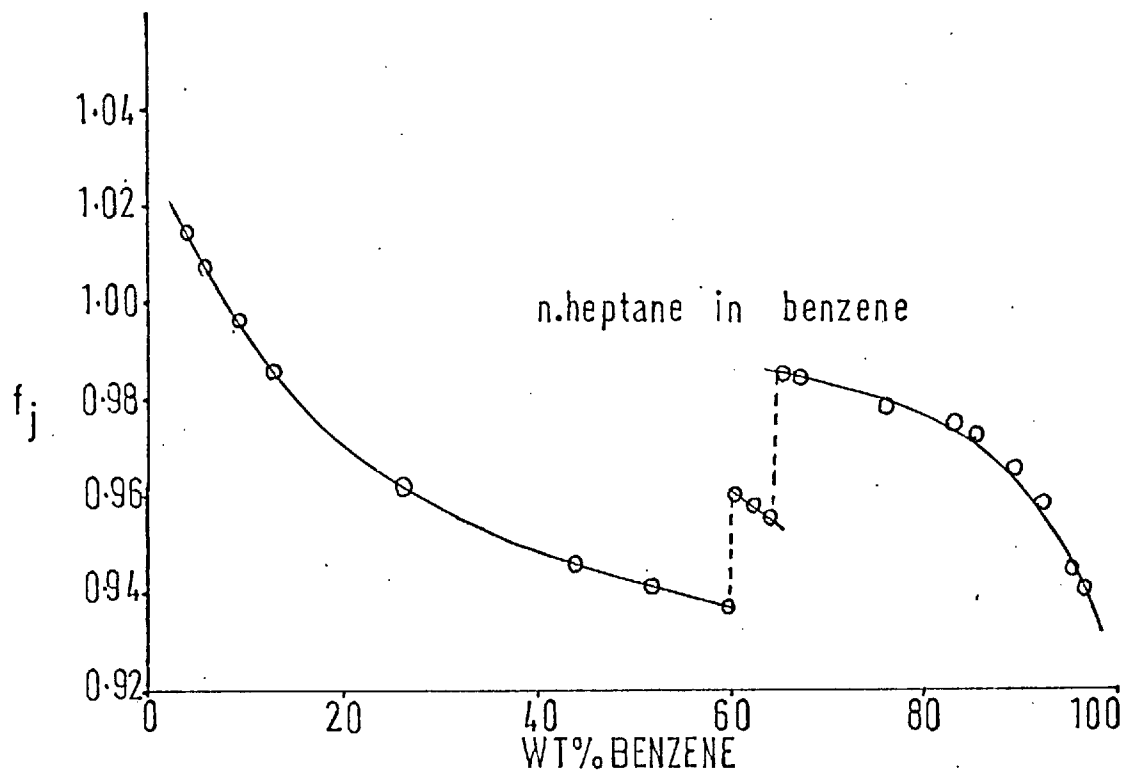
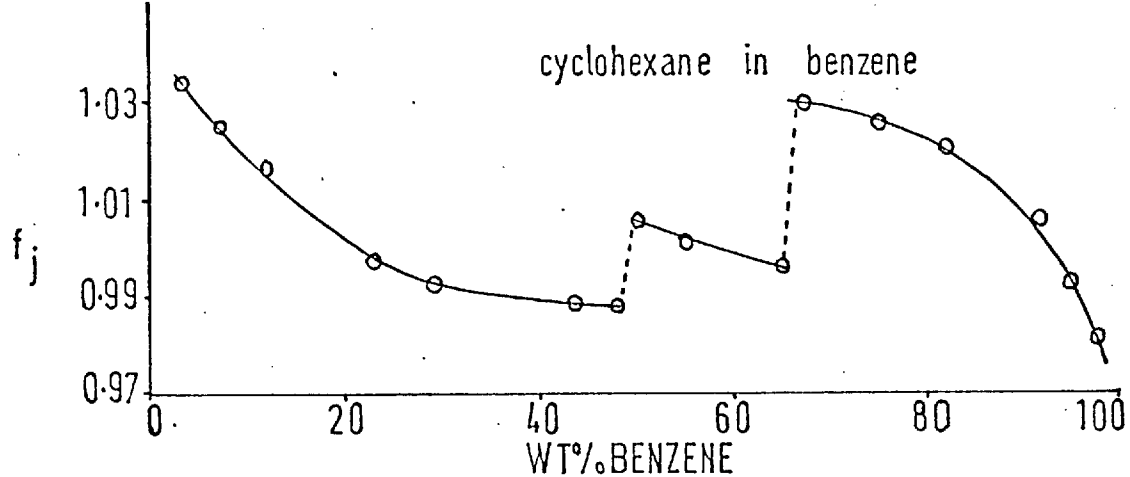


FIG I.1 SUBSTANCE SPECIFIC CORRECTION FACTORS VS COMPOSITION(wt%)

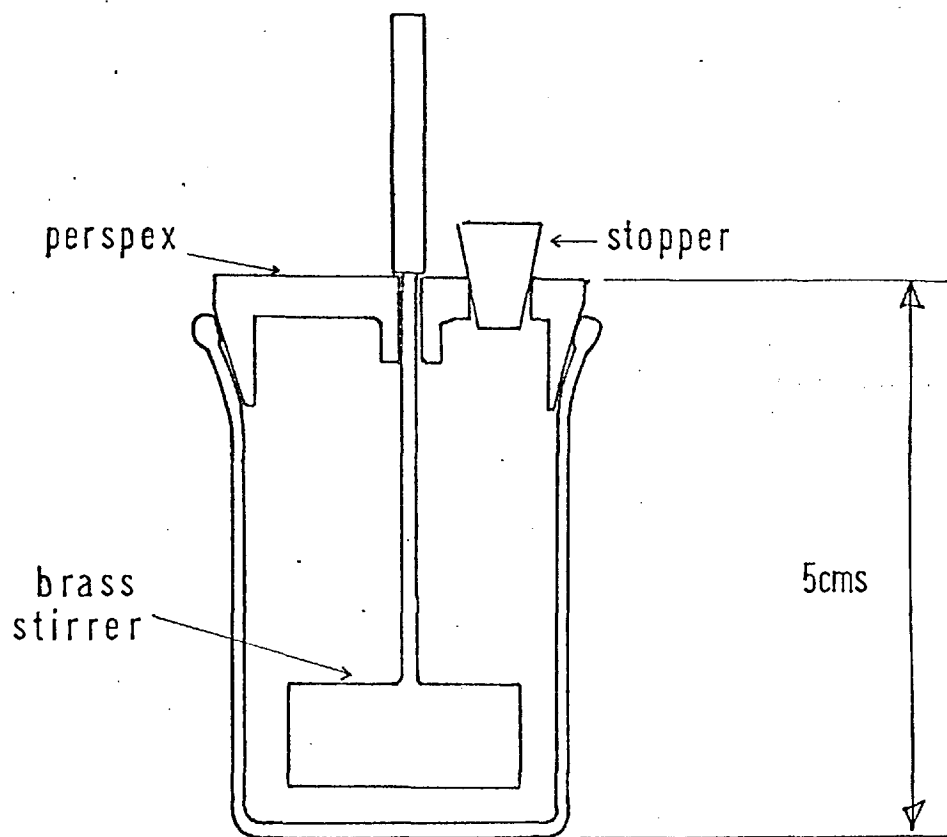
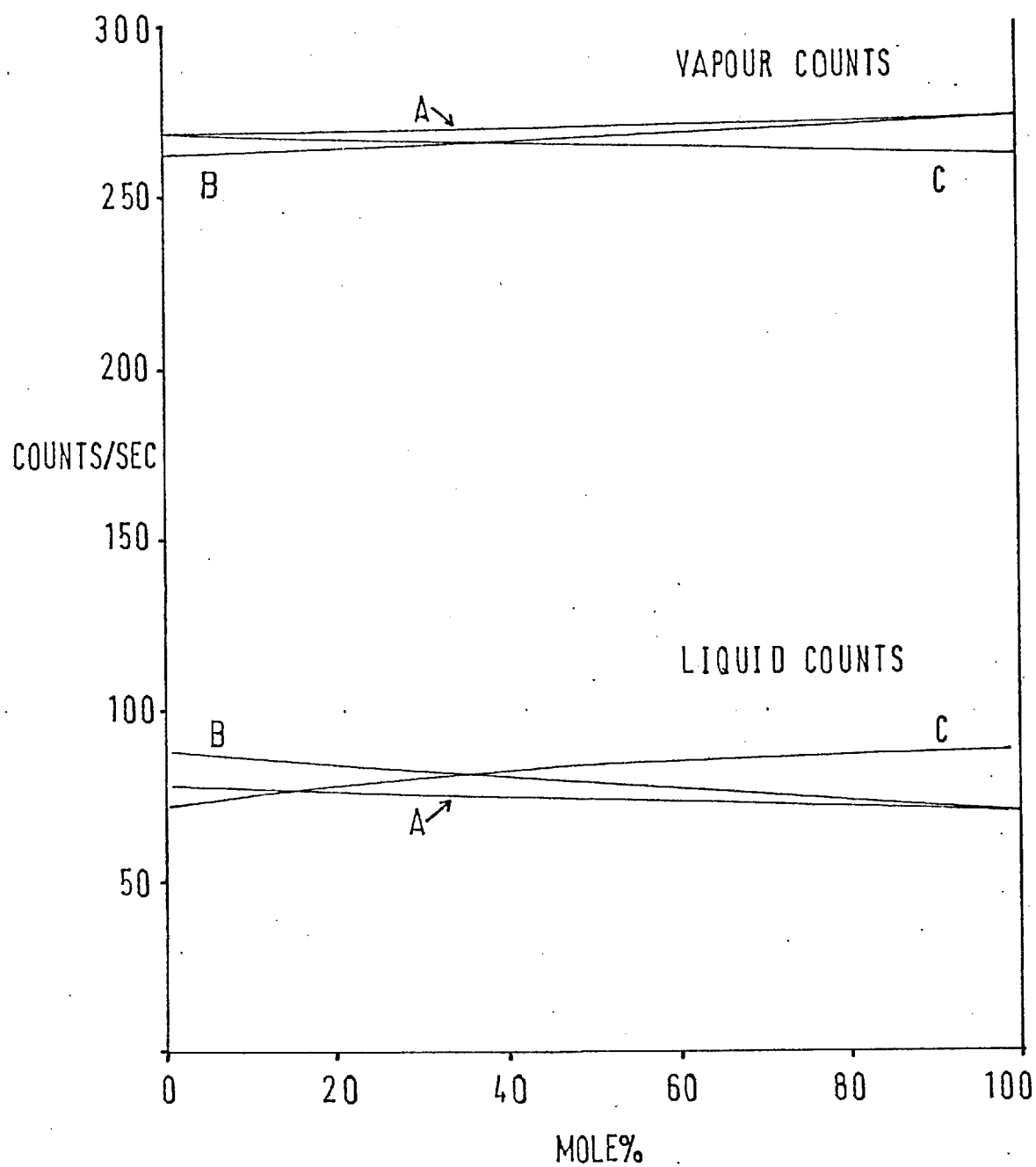


FIG I.2 SAMPLE BOTTLE FOR CALIBRATION MIXTURES

APPENDIX IICalibration charts for the gamma-ray traverse system

Fig. II-1 Vapour and Liquid count rate plotted against composition.

Fig. II-2 Typical semilog plot of $\log n$ against F_L
(fraction liquid). System A.



FIGII.1 VAPOUR AND LIQUID COUNT RATE VS. COMPOSITION

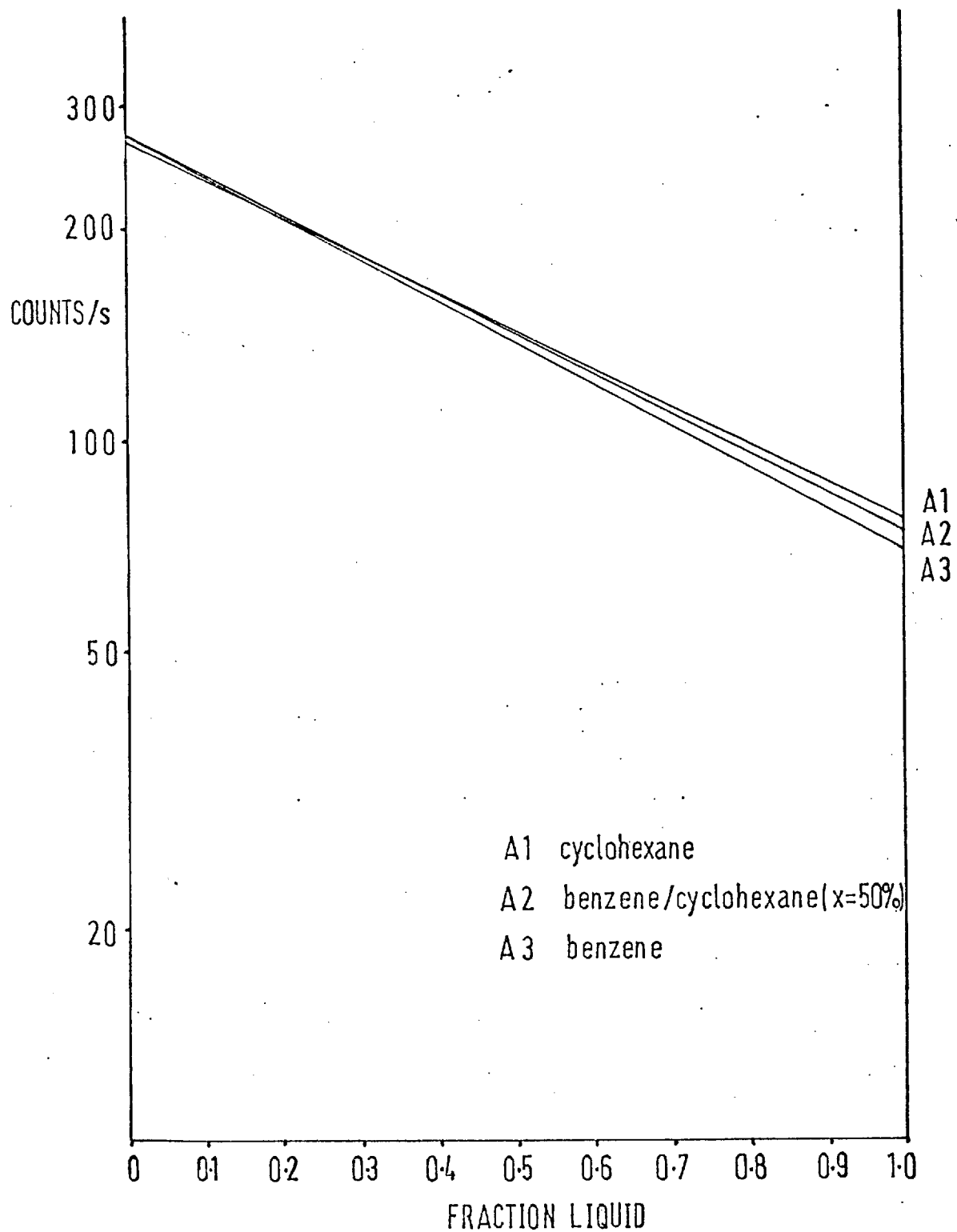


FIG II.2 CALIBRATION CHART FOR SYSTEM A

APPENDIX IIIPHYSICAL PROPERTIES OF THE BINARY SYSTEMS

Table III.1	Benzene/Cyclohexane equilibrium data
.2	Benzene/n-Heptane equilibrium data
.3	n-Heptane/Toluene equilibrium data.
Figure III.1	Surface tension at boiling point vs. composition
.2	Liquid density at boiling point))vs. composition Vapour density at condensation point)
.3	Liquid viscosity at boiling point))vs. composition Vapour viscosity at condensation point)
.4	Liquid diffusivity at boiling point. vs. composition.
.5	Vapour diffusivity at condensation point vs. composition.

Nomenclature

System A	Benzene/Cyclohexane
B	Benzene/n-Heptane
C	n-Heptane/Toluene
x_A	mole fraction of Benzene (Systems A and B) and of n-Heptane (System C).

TABLE III.1Equilibrium data: Benzene/Cyclohexane (ref 50)

B.P °C	x mole %	y mole %
80.80	0	0.00
80.46	2	2.84
79.99	5	6.87
79.30	10	13.06
78.30	20	23.84
77.75	30	33.39
77.40	40	42.18
77.25	50	50.45
77.25	60	58.62
77.45	70	67.03
77.85	80	76.52
78.60	90	87.20
79.20	95	93.18
79.71	98	97.12
80.15	100	100.00

TABLE III.2Equilibrium data: Benzene/n.Heptane (ref 50)

B.P °C	x mole %	y mole %
98.40	0	0.00
97.75	2	3.80
96.75	5	9.80
95.10	10	18.91
91.95	20	33.53
89.25	30	45.60
87.05	40	55.79
85.10	50	64.16
83.50	60	71.46
82.35	70	78.15
81.45	80	84.88
80.75	90	91.72
80.44	95	95.62
80.27	98	98.30
80.15	100	100.00

TABLE III.3Equilibrium data; n.Heptane/Toluene (ref 54)

B.P °C	x mole %	y mole %
110.63	0	0.0
110.30	1	1.84
109.10	5	8.70
107.90	10	16.30
105.70	20	29.20
104.00	30	40.00
102.60	40	49.44
101.50	50	58.10
100.60	60	66.40
99.86	70	74.40
99.25	80	82.60
98.77	90	91.10
98.58	95	95.50
98.46	99	99.10
98.42	100	100.00

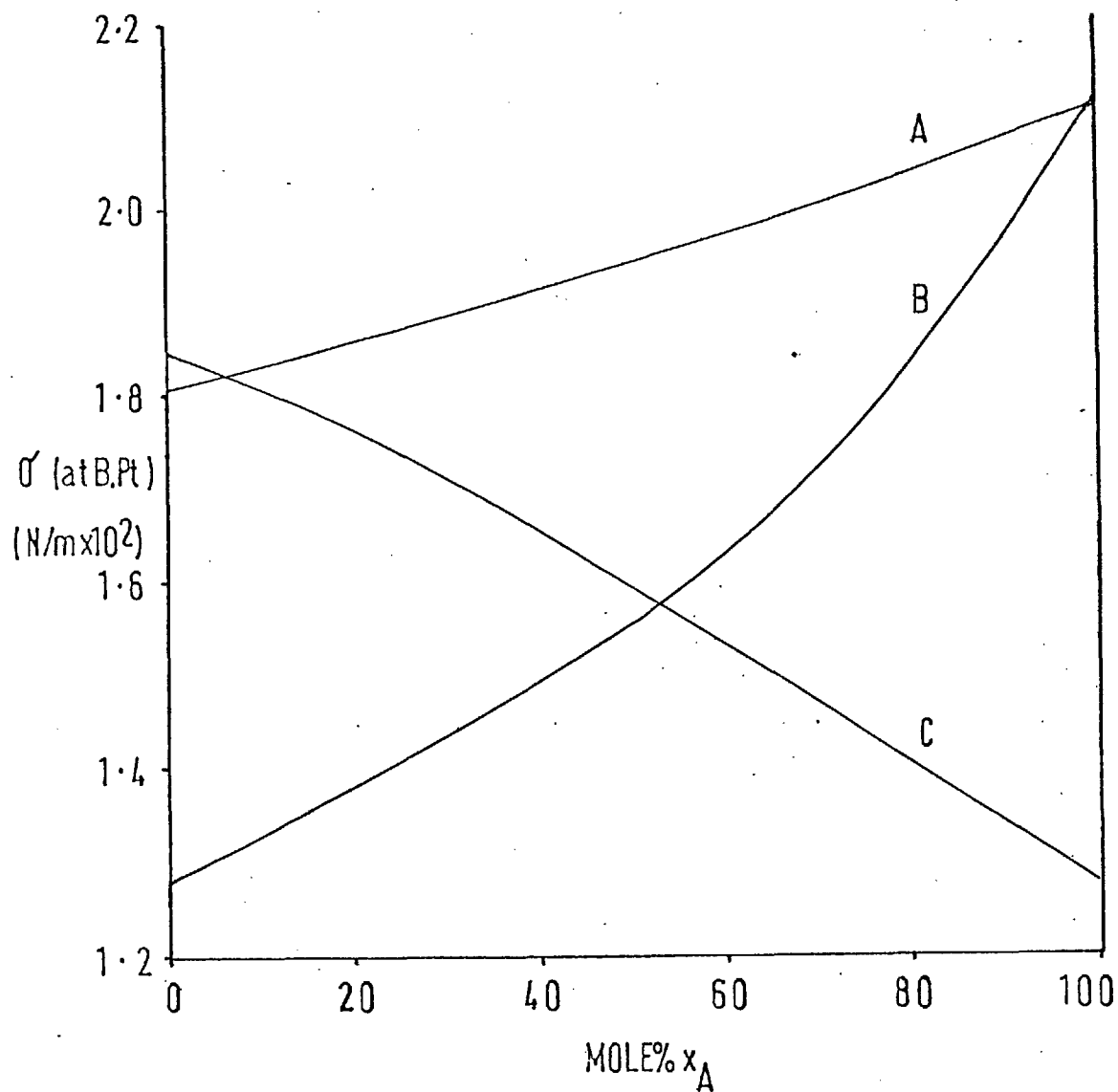


FIG III.1 SURFACE TENSION VS. COMPOSITION

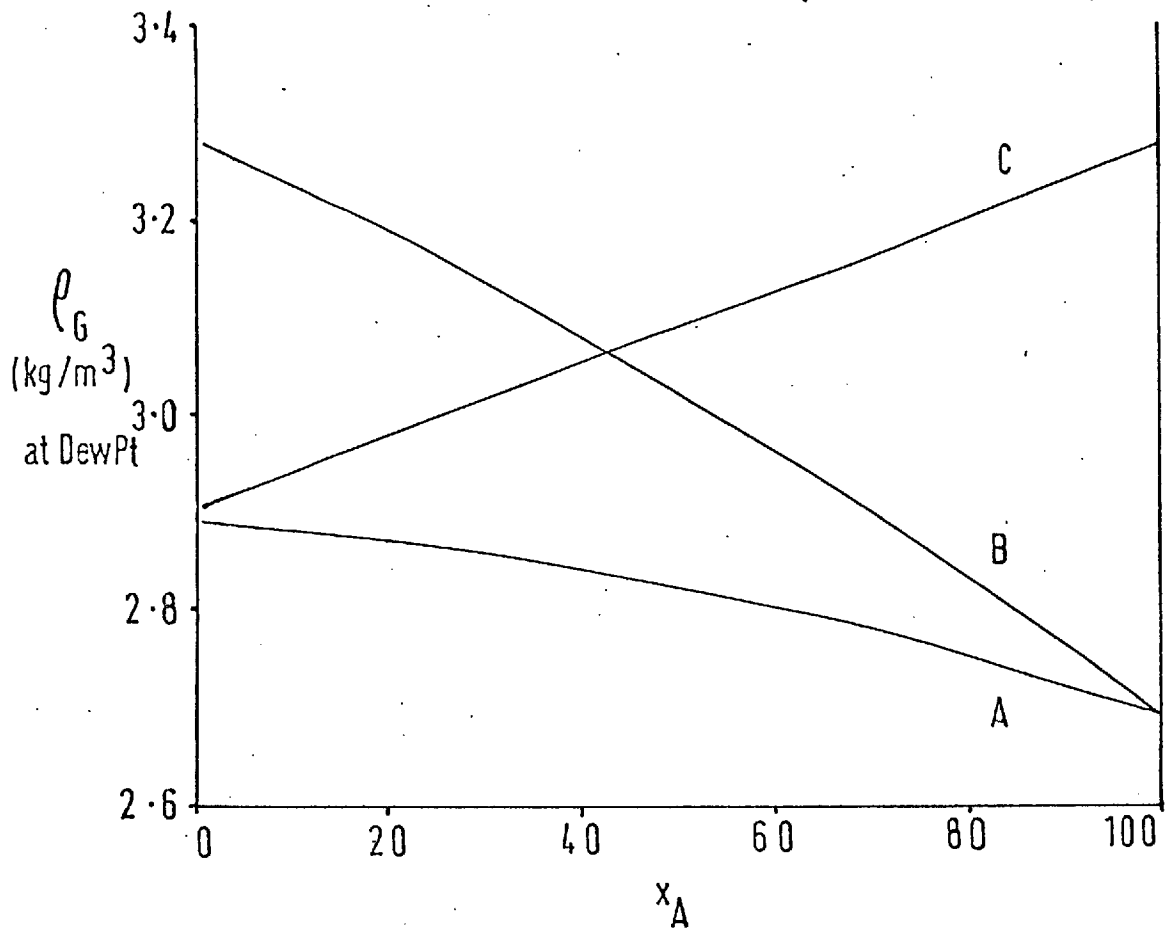
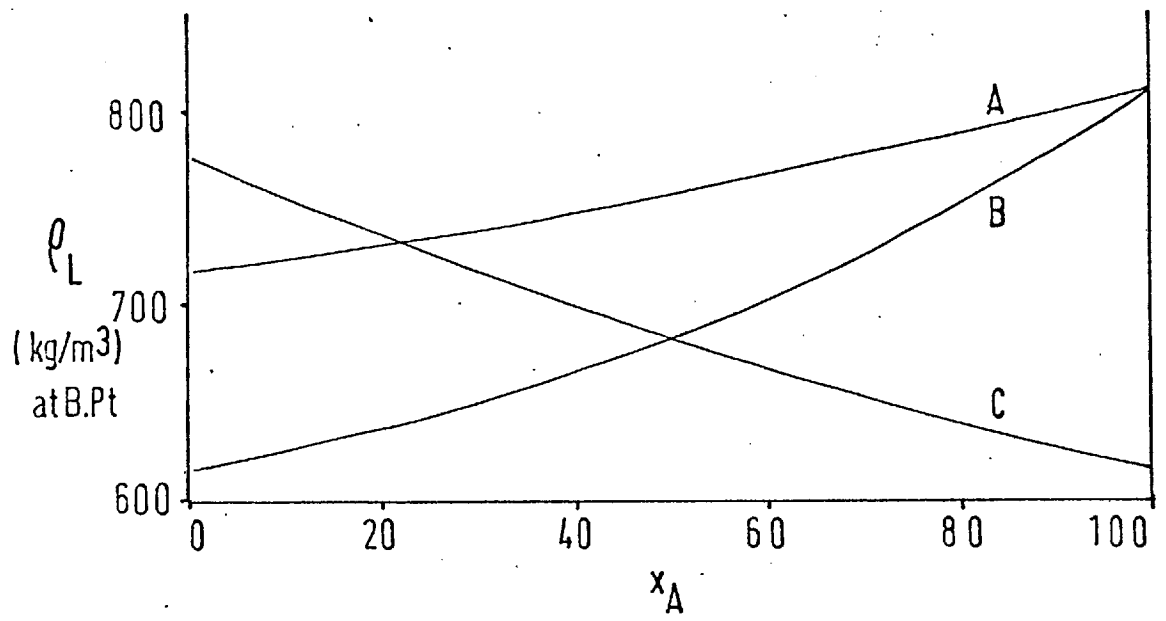
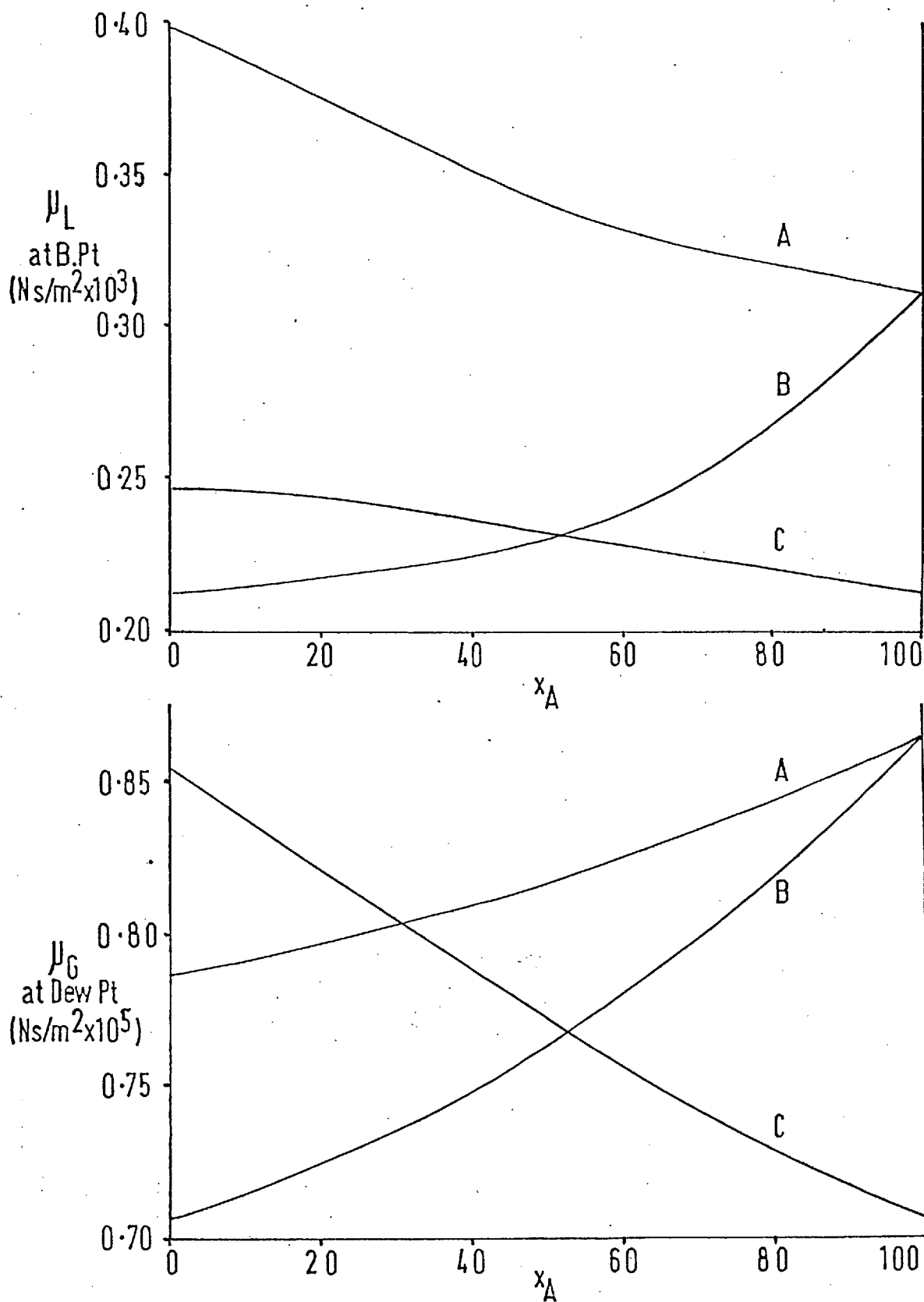


FIG. 3.2 DENSITY VS. COMPOSITION



FIGIII.3 VISCOSITY VS COMPOSITION

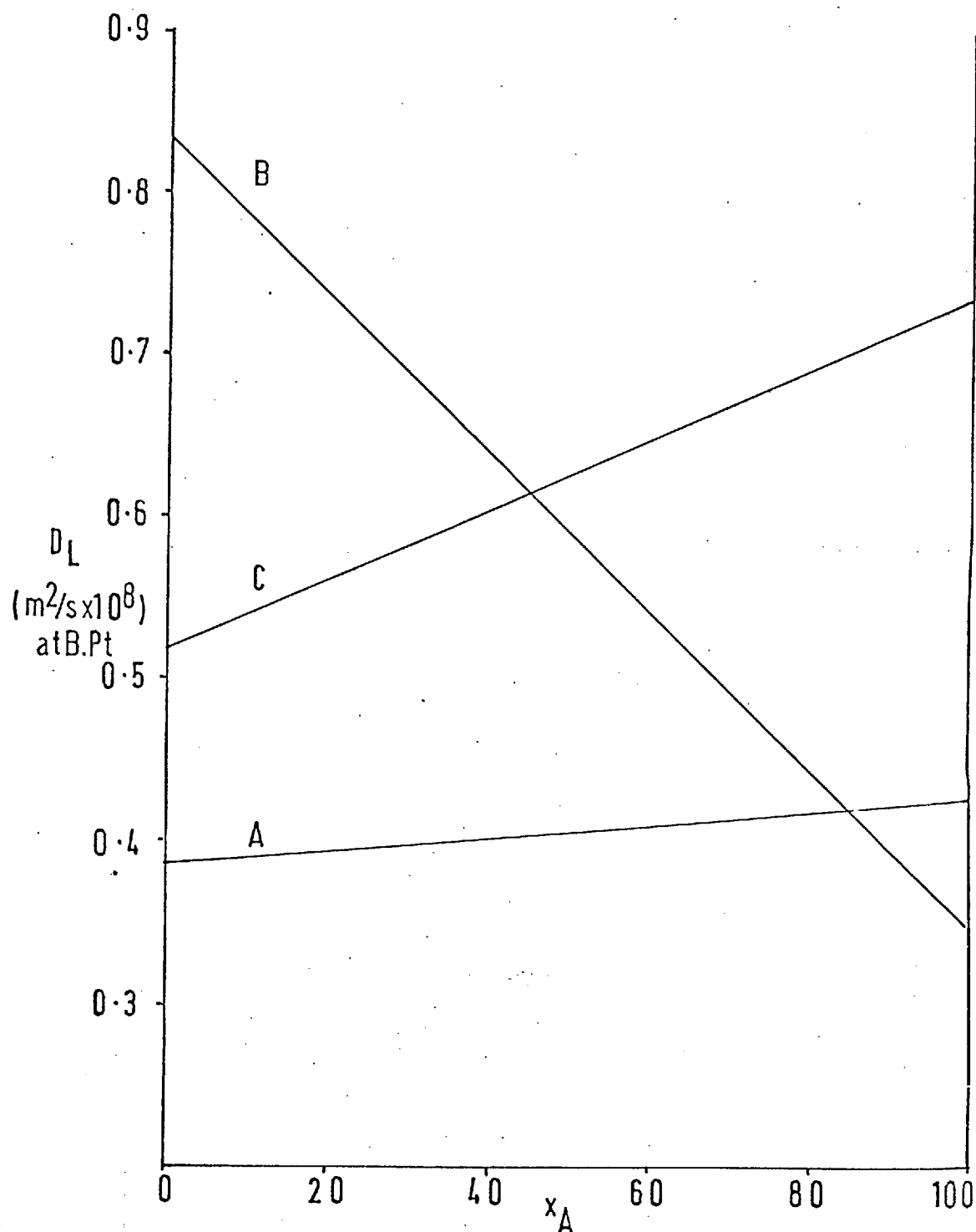


FIG III.4 LIQUID DIFFUSIVITY VS COMPOSITION

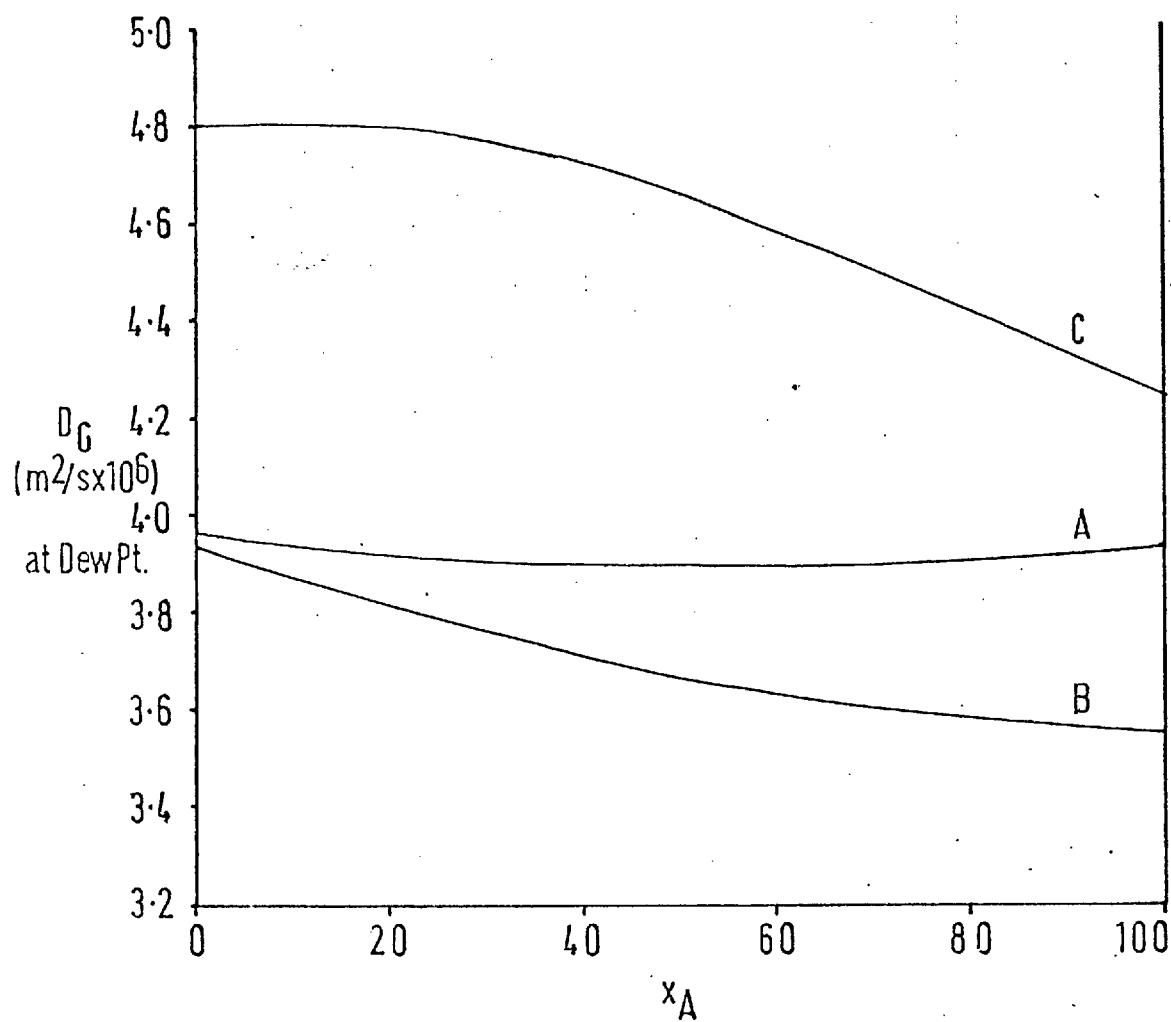


FIG III.5 VAPOUR DIFFUSIVITY VS COMPOSITION

APPENDIX IV
THE DIFFUSIVITY OF LIQUID MIXTURES

The heat transfer requirements at the interface in the distillation process impose, in general, the condition of zero nett flow, thus

$$N_1 + N_2 = 0 \quad \dots \text{IV.1}$$

where N_1 and N_2 are the molar fluxes of land 2.

If the bulk flow rate is v , and the total molar density C_T (moles/unit volume), the molar fluxes can be written

$$N_1 = vC_T x_1 - D_{L12} \frac{dx_1}{dz} C_T \quad \dots \text{IV.2}$$

$$N_2 = vC_T x_2 - D_{L21} \frac{dx_2}{dz} C_T \quad \dots \text{IV.3}$$

From IV.1, IV.2 and IV.3

$$v = \frac{dx_1}{dz} (D_{L12} - D_{L21}) \quad \dots \text{IV.4}$$

Hence if the diffusivity of species 1 in species 2 is different from the diffusivity of 2 in 1, the condition of zero nett flux implies the existence of bulk flow.

Thus

$$N_1 = -(D_{L12} - (D_{L12} - D_{L21})x_1) \frac{dx_1}{dz} C_T \quad \dots \text{IV.5}$$

If it is assumed that the molar flux can be expressed

$$N_1 = -D_L \frac{dx_1}{dz} C_T \quad \dots \text{IV.6}$$

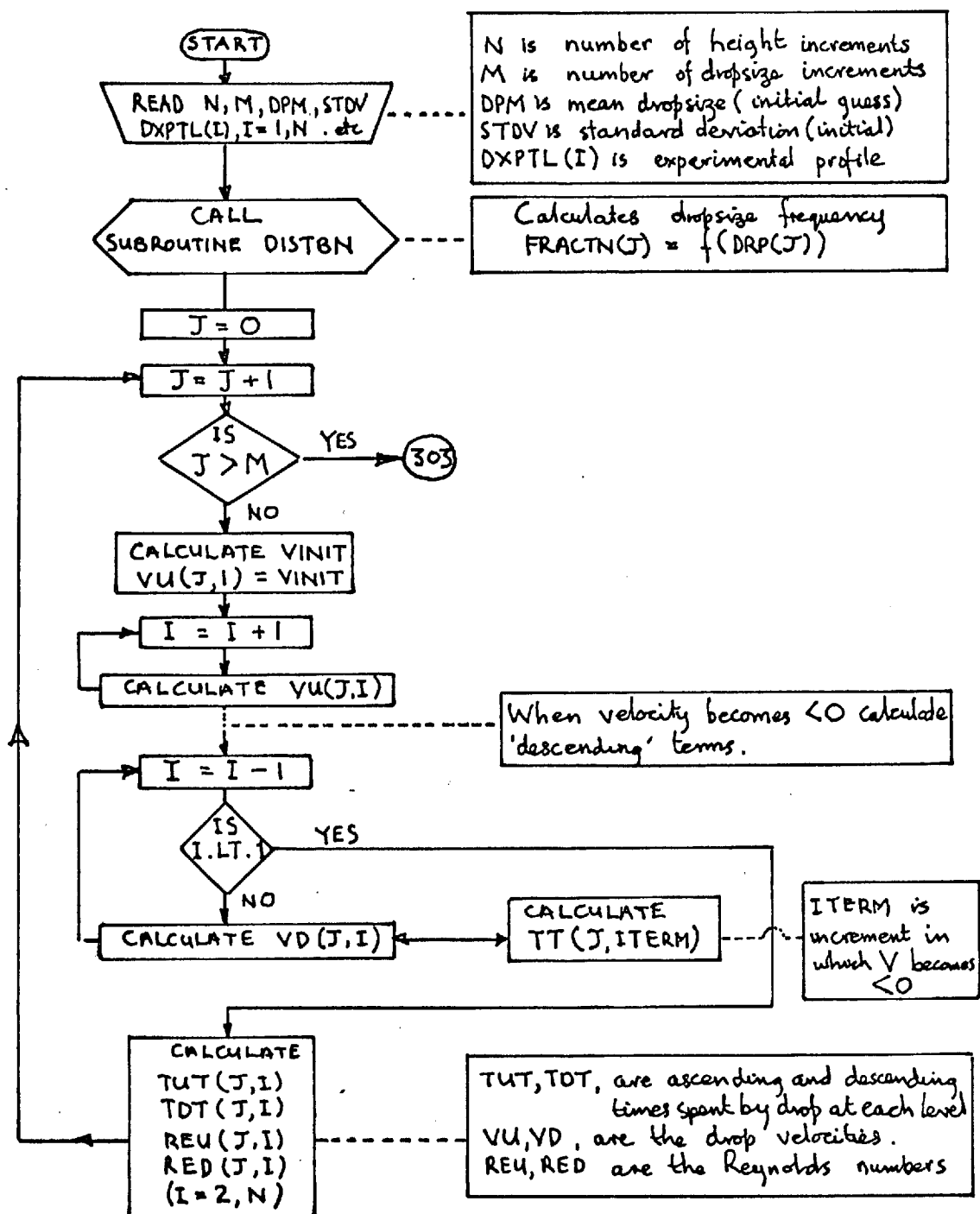
where D_L is the diffusivity of 1 in a mixture of 1 and 2. Thus from IV.5 and IV.6 the diffusivity of the mixture is

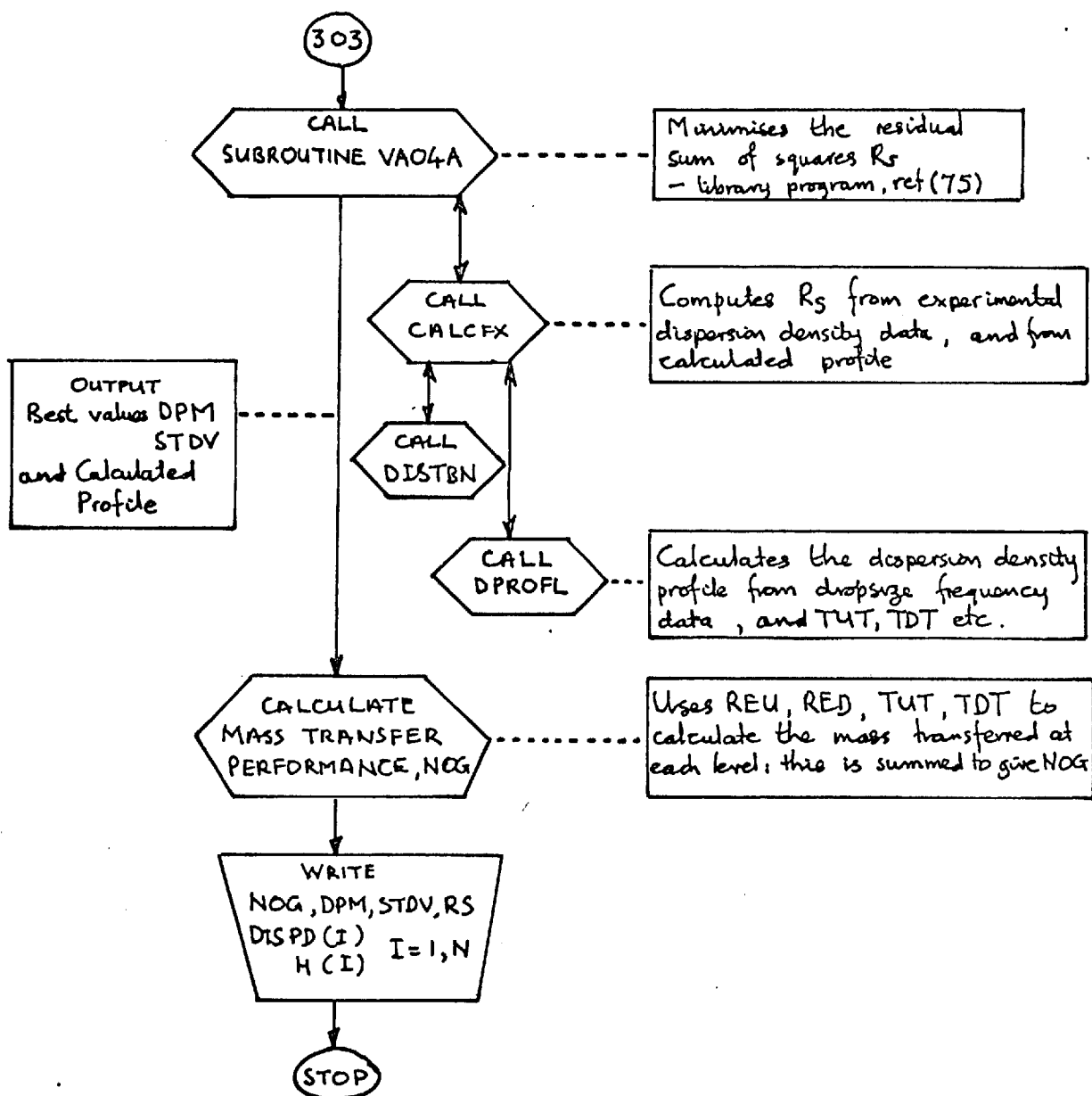
$$D_L = D_{L12} \left(1 - \left(1 - \frac{D_{L12}}{D_{L21}}\right)x_1\right) \quad \dots \text{IV.7}$$

APPENDIX V

FLOW CHARTS OF THE COMPUTER PROGRAM FOR
THE MODEL OF THE SPRAY REGIME.

FLOW CHARTS FOR THE SPRAY MODEL





EXPERIMENTAL RESULTS

Table VI.1. Dispersion density data System A W = 7.6 cms.

2		W = 1.9
3	System B	W = 7.6
4		W = 1.9
5		W = 0.95
6	System C	W = 1.9
7		W = 0.95
8	Performance data	System A W = 7.6
9		W = 1.9
10	System B	W = 7.6
11		W = 1.9
12		W = 0.95
13	System C	W = 1.9
14		W = 0.95

System A Benzene/Cyclohexane

System B Benzene/n.Heptane

System C n.Heptane/Toluene

TABLE VI.1

Dispersion Density Data: System A, W = 7.6 cms.

Run A1V1 x = 3.4%; F_H = 10.0; Holdup = 2.55 cms.

Level (cms.)	0.25	1.55	2.24	2.79	3.43	4.70	5.97	7.24	8.50	9.78
Dispersion density	0.512	0.538	0.479	0.401	0.334	0.262	0.181	0.083	0.190	0.007

Run A1V2 x = 3.23%; F_H = 13.7; Holdup = 2.43 cms.

Level (cms.)	0.25	0.89	1.52	2.24	3.43	4.70	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.406	0.422	0.426	0.396	0.282	0.267	0.211	0.139	0.062	0.034	0.020

Run A1V3 x = 3.18%; F_H = 17.1; Holdup = 1.99 cms.

Level (cms.)	0.25	0.89	1.52	2.16	3.43	4.70	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.357	0.340	0.326	0.310	0.242	0.218	0.171	0.118	0.065	0.021	0.006

Run A1V4 x = 3.17%; F_H = 20.1; Holdup = 1.74 cms.

Level (cms.)	0.25	0.89	1.52	2.16	3.43	4.70	5.97	7.24	8.51	9.78	11.05	13.6
Dispersion density	0.322	0.282	0.256	0.250	0.204	0.186	0.152	0.111	0.064	0.034	0.011	0.005

Run A2V1 x = 7.05%; F_H = 10.0; Holdup = 2.52 cms.

Level (cms.)	0.18	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51
Dispersion density	0.454	0.516	0.547	0.499	0.414	0.339	0.253	0.187	0.075	0.019

Run A2V2 x = 7.09%; F_H = 13.9; Holdup = 2.25 cms.

Level (cms.)	0.20	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.401	0.402	0.429	0.382	0.332	0.291	0.240	0.191	0.110	0.045	0.010	0.001

Run A2V3 x = 7.04%; F_H = 16.6; Holdup = 2.00 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.360	0.337	0.345	0.327	0.294	0.251	0.216	0.174	0.113	0.053	0.010	0.001

Run A2V4 x = 6.97% F_H = 19.8; Holdup = 1.85 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05	12.32
Dispersion density	0.313	0.287	0.272	0.260	0.236	0.220	0.190	0.174	0.129	0.071	0.028	0.025	0.013

TABLE VI.1 (cont'd.)

Run A3V1 $x = 14.57\%$; $F_H = 8.7$; Holdup = 2.54 cms.

Level (cms.)	0.25	0.89	1.52	2.24	2.79	3.43	4.19	4.95	5.72	7.24	7.87	8.51	9.78
Dispersion density	0.478	0.537	0.574	0.530	0.417	0.331	0.278	0.235	0.188	0.049	0.040	0.040	0.001

Run A3V2 $x = 14.31\%$; $F_H = 10.0$; Holdup = 2.64 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.60	7.24	8.51	9.78
Dispersion density	0.460	0.510	0.558	0.527	0.430	0.340	0.325	0.300	0.269	0.211	0.153	0.086	0.46	0.033

Run A3V3 $x = 14.15\%$; $F_H = 13.6$; Holdup = 2.56 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.73	7.49	8.26	9.78	12.3
Dispersion density	0.396	0.430	0.439	0.399	0.363	0.305	0.279	0.254	0.233	0.190	0.173	0.133	0.090	0.044	0.011

Run A3V4 $x = 14.08\%$; $F_H = 17.0$; Holdup = 2.11 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.73	7.49	8.51	9.78	11.05
Dispersion density	0.358	0.357	0.351	0.332	0.289	0.252	0.237	0.227	0.209	0.191	0.150	0.112	0.054	0.027	0.015

Run A4V1 $x = 20.56\%$; $F_H = 9.8$; Holdup = 2.50 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.73	7.49	8.26	9.02
Dispersion density	0.450	0.521	0.542	0.506	0.436	0.344	0.296	0.258	0.230	0.170	0.100	0.046	0.024	0.006

Run A4V2 $x = 20.27\%$; $F_H = 13.7$; Holdup = 2.46 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.73	7.49	8.26	9.02	9.78
Dispersion density	0.383	0.412	0.425	0.398	0.349	0.317	0.282	0.245	0.220	0.222	0.220	0.118	0.083	0.054	0.034

Run A4V3 $x = 20.04\%$; $F_H = 19.7$; Holdup = 1.85 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.19	4.95	5.72	6.48	7.24	8.51	9.78	11.05	13.6
Dispersion density	0.310	0.285	0.264	0.241	0.222	0.192	0.199	0.175	0.196	0.170	0.137	0.094	0.046	0.034	0.029

Run A5V1 $x = 26.39\%$; $F_H = 9.4$; Holdup = 2.48 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.60	7.24	7.87	8.51
Dispersion density	0.468	0.533	0.546	0.517	0.429	0.356	0.295	0.270	0.211	0.154	0.098	0.055	0.019	0.008

TABLE VI.1 (Cont'd.)

Run A5V2 $x = 26.18\%$; $F_H = 13.4$; Holdup = 2.48 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.60	7.24	7.87	8.51	9.02
Dispersion Density	0.403	0.436	0.446	0.409	0.364	0.324	0.292	0.264	0.252	0.228	0.169	0.129	0.087	0.047	0.030

Run A5V3 $x = 26.11\%$; $F_H = 16.9$; Holdup = 2.00 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.73	7.49	8.26	9.02	9.78
Dispersion density	0.364	0.343	0.356	0.330	0.286	0.259	0.237	0.233	0.201	0.179	0.136	0.105	0.051	0.027	0.006

Run A5V4 $x = 25.93\%$; $F_H = 19.8$; Holdup = 1.78 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.19	4.95	5.72	6.48	7.24	8.00	8.76	9.53	10.29
Dispersion density	0.321	0.288	0.276	0.257	0.231	0.210	0.208	0.175	0.171	0.150	0.120	0.080	0.056	0.033	0.019

Run A6V1 $x = 37.81\%$; $F_H = 10.5$; Holdup = 2.35 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78
Dispersion density	0.435	0.487	0.500	0.480	0.403	0.325	0.284	0.252	0.206	0.164	0.059	0.001	0.001

Run A6V2 $x = 37.63\%$; $F_H = 13.6$; Holdup = 2.13 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78
Dispersion density	0.368	0.395	0.412	0.374	0.325	0.274	0.253	0.240	0.209	0.184	0.103	0.029	0.003

Run A6V3 $x = 37.58\%$; $F_H = 19.7$; Holdup = 1.68 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.314	0.287	0.279	0.243	0.217	0.202	0.186	0.186	0.169	0.149	0.095	0.054	0.019	0.018

Run A7V1 $x = 42.22\%$; $F_H = 9.9$; Holdup = 2.40 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78
Dispersion density	0.460	0.494	0.522	0.485	0.399	0.332	0.290	0.254	0.211	0.157	0.057	0.010	0.005

Run A7V2 $x = 42.05\%$; $F_H = 13.2$; Holdup = 2.18 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78
Dispersion density	0.390	0.429	0.423	0.393	0.332	0.287	0.248	0.237	0.214	0.165	0.090	0.040	0.008

TABLE VI.1 (Cont'd.)

Run A7V3 $x = 42.02\%$; $F_H = 15.8$; Holdup = 2.09 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78	11.05	12.30
Dispersion density	0.365	0.355	0.351	0.340	0.298	0.259	0.238	0.217	0.193	0.167	0.092	0.044	0.028	0.018	0.008

Run A7V4 $x = 41.24\%$; $F_H = 19.9$; Holdup = 1.64 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78	11.05	12.20
Dispersion density	0.301	0.259	0.250	0.227	0.226	0.209	0.188	0.172	0.171	0.150	0.105	0.055	0.018	0.010	0.005

Run A8V1 $x = 47.31\%$; $F_H = 10.0$; Holdup = 2.41 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.419	0.473	0.490	0.434	0.374	0.307	0.246	0.169	0.104	0.042	0.024	0.010

Run A8V2 $x = 47.3\%$; $F_H = 13.2$; Holdup = 2.37 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05	12.30
Dispersion density	0.411	0.447	0.452	0.388	0.312	0.280	0.229	0.180	0.122	0.060	0.033	0.025	0.008

Run A8V3 $x = 47.20\%$; $F_H = 16.7$; Holdup = 2.22 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05	12.30
Dispersion density	0.382	0.380	0.368	0.327	0.289	0.272	0.254	0.194	0.124	0.067	0.038	0.024	0.007

Run A8V4 $x = 47.72\%$; $F_H = 20.4$; Holdup = 1.94 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05	12.30	13.6
Dispersion density	0.325	0.292	0.285	0.246	0.246	0.260	0.205	0.160	0.101	0.078	0.044	0.033	0.025	0.008

Run A9V1 $x = 59.76\%$; $F_H = 10.6$; Holdup = 2.38 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78
Dispersion density	0.441	0.484	0.499	0.458	0.354	0.304	0.278	0.249	0.218	0.173	0.080	0.033	0.022

Run A9V2 $x = 59.91\%$; $F_H = 13.2$; Holdup = 2.23 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78
Dispersion density	0.409	0.430	0.426	0.385	0.322	0.287	0.272	0.231	0.212	0.187	0.099	0.037	0.019

TABLE VI.1 (Cont'd.)

Run A9V3 $x = 59.92\%$; $F_H = 17.7$; Holdup = 2.03 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78	11.05	12.30
Dispersion density	0.332	0.315	0.287	0.292	0.263	0.232	0.225	0.213	0.192	0.178	0.117	0.068	0.050	0.036	0.008

Run A10V1 $x = 68.92\%$; $F_H = 10.2$; Holdup = 2.48 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.447	0.497	0.500	0.457	0.369	0.311	0.283	0.261	0.228	0.172	0.085	0.035	0.025	0.018

Run A10V2 $x = 68.94\%$; $F_H = 14.0$; Holdup = 2.29 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.402	0.400	0.408	0.366	0.318	0.278	0.244	0.236	0.225	0.194	0.118	0.061	0.036	0.027

Run A10V3 $x = 68.96\%$; $F_H = 17.8$; Holdup = 1.95 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.24	8.51	9.78	11.05
Dispersion density	0.347	0.324	0.315	0.277	0.253	0.233	0.213	0.198	0.194	0.178	0.122	0.071	0.038	0.021

Run A10V4 $x = 69.14\%$; $F_H = 20.7$; Holdup = 1.86 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.70	5.97	7.24	8.51	9.78	11.05	12.3	13.6
Dispersion density	0.331	0.266	0.249	0.223	0.215	0.205	0.187	0.163	0.123	0.081	0.052	0.035	0.029	0.018

Run A11V1 $x = 75.72\%$; $F_H = 9.9$; Holdup = 2.50 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05
Dispersion density	0.471	0.514	0.529	0.465	0.361	0.316	0.277	0.254	0.217	0.148	0.075	0.037	0.028	0.018

Run A11V2 $x = 76.13\%$; $F_H = 13.8$; Holdup = 2.28 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05
Dispersion density	0.400	0.429	0.423	0.387	0.303	0.279	0.245	0.241	0.220	0.189	0.111	0.050	0.035	0.020

Run A11V3 $x = 76.44\%$; $F_H = 20.4$; Holdup = 1.89 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05
Dispersion density	0.331	0.284	0.205	0.237	0.234	0.224	0.198	0.189	0.171	0.162	0.127	0.068	0.044	0.035

TABLE VI.1 (Cont'd.)

Run A12V1 $x = 82.68\%$; $F_H = 10.4$; Holdup = 2.46 cms.

Levels (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05
Dispersion density	0.446	0.500	0.497	0.450	0.351	0.303	0.268	0.247	0.214	0.184	0.093	0.069	0.030	0.021

Run A12V2 $x = 82.85\%$; $F_H = 14.4$; Holdup = 2.38 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05	12.3
Dispersion density	0.392	0.406	0.383	0.401	0.361	0.301	0.273	0.243	0.236	0.193	0.089	0.072	0.044	0.036	0.020

Run A12V3 $x = 83.03\%$; $F_H = 16.9$; Holdup = 2.03 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05	12.3
Dispersion density	0.344	0.344	0.337	0.315	0.270	0.255	0.225	0.223	0.199	0.173	0.112	0.058	0.031	0.022	0.015

Run A13V1 $x = 89.07\%$; $F_H = 10.5$; Holdup = 2.31 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78
Dispersion density	0.441	0.488	0.484	0.438	0.354	0.301	0.261	0.218	0.224	0.162	0.073	0.026	0.011

Run A13V2 $x = 89.33\%$; $F_H = 15.5$; Holdup = 2.27 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05
Dispersion density	0.399	0.397	0.411	0.356	0.303	0.273	0.254	0.240	0.219	0.201	0.120	0.058	0.042	0.027

Run A13V3 $x = 89.56\%$; $F_H = 20.4$; Holdup = 1.93

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05	12.3
Dispersion density	0.339	0.279	0.279	0.253	0.220	0.215	0.204	0.185	0.184	0.175	0.133	0.082	0.054	0.040	0.033

Run A14V1 $x = 95.41\%$; $F_H = 10.5$; Holdup = 2.43 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78
Dispersion density	0.447	0.501	0.523	0.409	0.395	0.328	0.286	0.261	0.231	0.162	0.063	0.026	0.012

Run A14V2 $x = 95.53\%$; $F_H = 13.8$; Holdup = 2.37 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05
Dispersion density	0.412	0.448	0.448	0.404	0.338	0.302	0.272	0.251	0.236	0.191	0.108	0.042	0.030	0.019

Run A14V3 $x = 95.60\%$; $F_H = 17.1$; Holdup = 2.13 cms.

Level (cms.)	0.25	0.89	1.52	2.16	2.79	3.43	4.06	4.70	5.33	5.97	7.23	8.51	9.78	11.05	12.30
Dispersion density	0.356	0.344	0.342	0.316	0.274	0.253	0.240	0.230	0.219	0.185	0.120	0.060	0.032	0.020	0.020

TABLE VI.2
Dispersion Density Data: System A; W = 1.9 cms.

Run A2IV1 $x = 2.93\%$; $F_H = 8.8$; Holdup = 0.790 cms.

Level (Cms)	0.25	0.56	0.76	1.27	1.73	2.34	2.80	3.30	3.80	4.32	4.83
Dispersion density	0.513	0.278	0.271	0.240	0.136	0.069	0.016	0.018	0.010	0.010	0.008

Run A2IV2 $x = 2.88\%$; $F_H = 11.7$; Holdup = 0.749 cms.

Level (Cms)	0.25	0.54	0.71	1.37	1.88	2.39	2.80	3.38	3.76	4.40	4.90	5.33
Dispersion density	0.484	0.290	0.246	0.242	0.214	0.134	0.079	0.025	0.018	0.022	0.013	0.003

Run A2IV3 $x = 2.85\%$; $F_H = 18.1$; Holdup = 0.668 cms.

Level (Cms)	0.25	0.51	0.76	1.40	1.88	2.41	3.00	3.43	3.94	4.42	4.95	5.33	5.97	7.24
Dispersion density	0.314	0.198	0.147	0.150	0.147	0.163	0.123	0.087	0.047	0.031	0.019	0.019	0.015	0.008

Run A22VI $x = 9.55\%$; $F_H = 9.5$; Holdup = 0.785 cms.

Level (cms)	0.25	0.48	0.76	1.27	1.83	2.31	2.87	3.25	3.76	4.31	4.93
Dispersion density	0.503	0.358	0.300	0.272	0.238	0.135	0.040	0.019	0.024	0.014	0.012

Run A22V2 $x = 9.41\%$; $F_H = 13.1$; Holdup = 0.719 cms.

Level (cms)	0.25	0.56	0.74	1.24	1.78	2.39	2.92	3.46	4.06	4.62	5.20
Dispersion density	0.443	0.248	0.209	0.200	0.229	0.163	0.074	0.027	0.023	0.015	0.009

Run A22V3 $x = 9.24\%$; $F_H = 18.6$; Holdup = 0.701 cms.

Level (cms)	0.25	0.61	0.81	1.27	1.75	2.24	2.80	3.38	4.06	4.70	5.35	6.0	6.6	7.25	7.85	8.5
Dispersion density	0.256	0.171	0.151	0.159	0.175	0.170	0.137	0.088	0.041	0.029	0.027	0.021	0.019	0.014	0.010	0.005

Run A23VI $x = 20.64\%$ $F_H = 9.8$; Holdup = 0.775 cms.

Level (cms)	0.25	0.54	0.74	0.97	1.35	1.65	2.13	2.54	2.94
Dispersion density	0.526	0.370	0.314	0.296	0.293	0.265	0.181	0.673	0.015

Run A23V2 $x = 20.42\%$ $F_H = 13.3$; Holdup = 0.744 cms.

Level (cms)	0.28	0.51	0.79	1.22	1.65	2.10	2.50	2.90	3.43
Dispersion density	0.493	0.327	0.268	0.245	0.237	0.189	0.123	0.057	0.012

TABLE VI-2 (Cont'd)

Run A23V3 $x = 20.08\%$ $F_H = 18.4$; Holdup = 0.711 cms.

Level (cms)	0.25	0.51	0.86	1.27	1.62	2.10	2.50	2.84	3.30	3.86	4.26	4.70	5.33	5.95
Dispersion density	0.371	0.199	0.157	0.168	0.186	0.184	0.166	0.134	0.088	0.045	0.043	0.025	0.017	0.013

Run A24VI $x = 30.57\%$ $F_H = 10.0$; Holdup = 0.785 cms.

Level (cms)	0.25	0.48	0.76	1.02	1.40	1.70	2.08	2.42	2.80	3.18	3.70	4.20
Dispersion density	0.478	0.373	0.285	0.285	0.276	0.262	0.186	0.123	0.063	0.033	0.019	0.007

Run A24V2 $x = 30.40\%$; $F_H = 13.8$; Holdup = 0.770 cms.

Level (cms)	0.33	0.59	0.84	1.09	1.42	1.80	2.16	2.70	3.33	3.74	4.15	4.75	5.35	5.85	6.20
Dispersion density	0.392	0.272	0.227	0.219	0.243	0.238	0.199	0.106	0.037	0.029	0.027	0.021	0.019	0.014	0.009

Run A24V3 $x = 30.26\%$; $F_H = 18.0$; Holdup = 0.709 cms.

Level (cms)	0.38	0.64	0.89	1.53	2.11	2.62	3.10	3.60	4.10	4.62	5.13	5.70	6.30	6.75	7.25
DDispersion density	0.272	0.176	0.159	0.177	0.190	0.150	0.099	0.057	0.048	0.028	0.021	0.021	0.017	0.016	0.011

Run A25VI $x = 44.14\%$; $F_H = 8.9$; Holdup = 0.749 cms.

Level (cms)	0.25	0.48	0.76	1.09	1.32	1.62	1.98	2.36	2.74	3.00	3.30
Dispersion density	0.483	0.390	0.300	0.283	0.280	0.251	0.195	0.111	0.053	0.036	0.018

Run A25V2 $x = 44.00\%$; $F_H = 10.6$; Holdup = 0.772 cms.

Level (cms)	0.25	0.51	0.69	0.94	1.35	1.62	1.95	2.18	2.54	2.79	3.30	3.80	4.45	4.95
Dispersion density	0.441	0.324	0.286	0.263	0.263	0.248	0.198	0.164	0.091	0.067	0.027	0.024	0.017	0.012

Run A25V3 $x = 43.92\%$; $F_H = 18.0$; Holdup = 0.701 cms.

Level (cms)	0.25	0.51	0.76	1.32	1.78	2.28	2.79	3.43	4.06	4.70	5.35	5.95	6.60	7.24
Dispersion density	0.303	0.186	0.160	0.167	0.173	0.169	0.127	0.072	0.039	0.029	0.019	0.018	0.015	0.010

Run A26V1 $x = 56.96\%$; $F_H = 9.7$; Holdup = 0.699 cms.

Level (cms)	0.25	0.56	0.71	1.22	1.85	2.41	2.87	3.30	3.81	4.31	4.70
Dispersion density	0.495	0.309	0.257	0.213	0.175	0.043	0.061	0.041	0.028	0.020	0.008

TABLE VI-2 Cont'd)

Run A26V2 $x = 56.82\%$; $F_H = 12.7$; Holdup = 0.785 cms

Level (cms)	0.25	0.48	0.72	1.27	1.78	2.28	2.72	3.30	3.81	4.31	4.82
Dispersion density	0.448	0.316	0.233	0.236	0.218	0.170	0.099	0.065	0.043	0.028	0.018

Run A26V3 $x = 56.87\%$; $F_H = 15.2$; Holdup = 0.798 cms.

Level (cms)	0.28	0.20	0.87	1.27	1.88	2.28	2.80	3.30	3.81	4.31	4.82	5.35
Dispersion density	0.402	0.258	0.206	0.212	0.219	0.176	0.124	0.081	0.061	0.048	0.037	0.020

Run A27V1 $x = 66.38\%$; $F_H = 9.5$; Holdup = 0.599 cms.

Level (cms)	0.25	0.54	0.81	1.29	1.78	2.36	2.80	3.30
Dispersion density	0.441	0.301	0.224	0.195	0.164	0.096	0.048	0.025

Run A27V2 $x = 66.53\%$; $F_H = 13.1$; Holdup = 0.767 cms.

Level (cms)	0.28	0.54	0.81	1.27	1.78	2.24	2.80	3.30	3.81	4.31	4.82	5.35
Dispersion density	0.452	0.294	0.224	0.212	0.206	0.175	0.109	0.068	0.049	0.037	0.021	0.013

Run A27V3 $x = 66.54\%$; $F_H = 15.9$; Holdup = 0.79 cms.

Level (cms)	0.25	0.51	0.76	1.27	1.75	2.28	2.80	3.30	3.80	4.32	4.82	5.35
Dispersion density	0.439	0.259	0.209	0.193	0.209	0.179	0.120	0.079	0.047	0.038	0.025	0.018

Run A29V1 $x = 79.23\%$; $F_H = 9.5$; Holdup = 0.620 cms.

Level (cms)	0.25	0.51	0.76	1.27	1.78	2.28	2.70	3.30
Dispersion density	0.487	0.316	0.250	0.214	0.155	0.089	0.044	0.011

Run A29V2 $x = 79.45\%$; $F_H = 12.6$; Holdup = 0.759 cms.

Level (cms)	0.25	0.51	0.76	1.32	1.83	2.28	2.80	3.30	3.81
Dispersion density	0.563	0.342	0.248	0.234	0.202	0.152	0.077	0.027	0.003

Run A29V3 $x = 79.39\%$; $F_H = 15.9$; Holdup = 0.798 cms.

Level (cms)	0.25	0.51	0.76	1.27	1.78	2.28	2.80	3.30	3.81	4.31	4.82	5.35	5.96	7.25	8.50
Dispersion density	0.396	0.239	0.204	0.194	0.198	0.164	0.126	0.069	0.050	0.045	0.032	0.028	0.020	0.014	0.006

TABLE VI-2 (Cont'd)

Run A30V1 $x = 88.66\%$; $F_H = 9.8$; Holdup = 0.645 cms.

Level (cms)	0.25	0.76	1.27	1.78	2.28	2.80	3.30	3.81
Dispersion density	0.450	0.238	0.215	0.164	0.105	0.056	0.033	0.015

Run A30V2 $x = 88.82\%$; $F_H = 12.7$; Holdup = 0.696 cms.

Level (cms)	0.25	0.76	1.27	1.78	2.28	2.80	3.30	3.81	4.31	5.35
Dispersion density	0.420	0.220	0.206	0.189	0.135	0.079	0.045	0.028	0.022	0.010

Run A30V3 $x = 88.84\%$; $F_H = 16.0$; Holdup = 0.780 cms.

Level (cms)	0.25	0.76	1.27	1.78	2.28	2.80	3.30	3.81	4.31	4.82	5.35	5.96	7.5
Dispersion density	0.378	0.193	0.200	0.215	0.171	0.122	0.077	0.046	0.048	0.037	0.023	0.016	0.007

Run A31V1 $x = 94.98\%$; $F_H = 9.4$; Holdup = 0.803 cms.

Level (cms)	0.25	0.81	1.27	1.78	2.24	2.80	3.30	3.81	4.31	4.82
Dispersion density	0.521	0.263	0.264	0.227	0.134	0.073	0.041	0.033	0.025	0.014

Run A31V2 $x = 94.97\%$; $F_H = 12.9$; Holdup = 0.805 cms.

Level (cms)	0.25	0.76	1.27	1.78	2.28	2.80	3.30	3.81	4.37	4.88	5.33
Dispersion density	0.451	0.253	0.250	0.227	0.172	0.098	0.069	0.039	0.027	0.015	0.007

Run A31V3 $x = 95.03\%$; $F_H = 18.5$; Holdup = 0.828 cms.

Level (cms)	0.25	0.76	1.27	1.78	2.28	2.80	3.30	3.81	4.31	4.82	5.33	7.25	8.5
Dispersion density	0.339	0.181	0.178	0.199	0.190	0.152	0.097	0.062	0.056	0.045	0.030	0.023	0.010

TABLE VI.3

Dispersion density data System B $W = 7.6$ cms.Run B3V1 $x = 28.22\%$; $F_H = 8.1$; Holdup = 2.408 ins.

Level (cms.)	0.25	0.51	0.76	1.04	1.27	1.52	1.78	2.16	2.67	3.18	3.68	4.45	5.21	5.97	6.73
Dispersion density	0.505	0.531	0.544	0.558	0.580	0.555	0.549	0.538	0.495	0.405	0.331	0.269	0.194	0.080	0.017
	7.49	8.51													
	0.012	0.002													

Run B3V2 $x = 27.67\%$; $F_H = 10.8$; Holdup = 2.910 cms.

Level (cms.)	0.25	0.51	0.76	1.07	1.27	1.57	1.78	2.16	2.67	3.18	3.94	4.70	5.46	6.22	6.99
Dispersion density	0.456	0.469	0.505	0.528	0.525	0.541	0.550	0.527	0.504	0.443	0.361	0.313	0.267	0.214	0.144
	7.87	8.51	9.78												
	0.074	0.053	0.031												

Run B3V3 $x = 27.44\%$; $F_H = 12.8$; Holdup = 2.50 cms.

Level (cms.)	0.25	0.51	0.76	1.14	1.52	1.91	2.41	2.92	3.43	4.19	4.95	5.72	6.48	7.49	8.51
Dispersion density	0.348	0.342	0.380	0.423	0.460	0.442	0.427	0.377	0.351	0.305	0.283	0.232	0.187	0.106	0.040
	9.53	10.54													
	0.025	0.010													

Run B3V4 $x = 26.74\%$; $F_H = 18.4$; Holdup = 2.17 cms.

Level (cms.)	0.25	0.51	0.76	1.11	1.52	1.91	2.41	2.92	3.43	4.19	4.95	5.72	6.73	7.75	8.76
Dispersion density	0.331	0.313	0.318	0.320	0.319	0.303	0.279	0.259	0.247	0.228	0.226	0.202	0.174	0.115	0.080
	9.78	11.05													
	0.043	0.029													

Run B5M1 $x = 65.81\%$; $F_H = 8.1$; Holdup = 2.192 cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.43	4.06	4.70
	0.483	0.508	0.527	0.538	0.494	0.525	0.507	0.476	0.459	0.427	0.404	0.398	0.371	0.310	0.208
	5.33	5.97	6.60	7.24											
	0.120	0.060	0.025	0.010											

TABLE VI.3 (Cont'd.)

Run B5V2 $x = 64.90\%$; $F_H = 10.2$; Holdup = 2.590 cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.30	2.54	2.79	3.05	3.43	4.06	4.70	5.33
Dispersion density	0.438	0.464	0.479	0.499	0.499	0.499	0.491	0.485	0.475	0.457	0.425	0.384	0.343	0.300	0.256
	5.97	6.60	7.24	7.87	8.51										
	0.190	0.123	0.064	0.035	0.021										

Run B5V3 $x = 64.29\%$; $F_H = 12.7$; Holdup = 2.40 cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.60
Dispersion density	0.406	0.414	0.428	0.425	0.447	0.439	0.425	0.440	0.401	0.355	0.302	0.278	0.243	0.183	0.162
	7.24	7.87	8.51	9.02											
	0.100	0.055	0.027	0.010											

Run B5V4 $x = 63.87\%$; $F_H = 17.2$; Holdup = 2.07 cms.

Level (cms.)	0.25	0.51	0.78	1.02	1.40	1.91	2.41	2.92	3.43	4.06	4.70	5.46	5.97	6.99	8.00
	0.339	0.326	0.340	0.355	0.343	0.348	0.301	0.298	0.268	0.239	0.231	0.210	0.189	0.141	0.072
	9.02	10.03	11.05												
	0.032	0.017	0.012												

TABLE VI-4
Dispersion density data: System B; W = 1.9 cms.

Run B10V1 $x = 7.55\%$; $F_H = 8.7$; Holdup = 0.850 cms.

Level (cms)	0.13	0.25	0.38	0.51	0.76	1.04	1.32	1.52	1.78	2.03	2.29	2.54	2.80	3.05	3.56
Dispersion density	0.541	0.494	0.443	0.352	0.302	0.286	0.302	0.332	0.308	0.235	0.188	0.114	0.060	0.025	0.015

Run B10V2 $x = 6.99\%$; $F_H = 12.6$; Holdup = 0.797 cms.

Level (cms)	0.10	1.25	1.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.80	3.05	3.30	4.06
Dispersion density	0.581	0.440	0.277	0.225	0.201	0.212	0.235	0.244	0.226	0.188	0.150	0.091	0.058	0.034	0.011

Run B10V3 $x = 6.94\%$; $F_H = 14.8$; Holdup = 0.797 cms.

Level (cms)	0.13	0.23	0.51	0.76	1.04	1.39	1.78	2.16	2.54	2.92	3.43	3.94	4.45	4.95	5.46
Dispersion density	0.514	0.421	0.238	0.185	0.181	0.205	0.226	0.219	0.183	0.122	0.068	0.031	0.019	0.016	0.004

Run B10V4 $x = 6.44\%$; $F_H = 18.5$; Holdup = 0.760 cms.

Level (cms)	0.25	0.51	0.89	1.40	1.91	2.67	3.68	4.45	5.21	5.97	6.73	7.60	8.25	9.02	10.54
Dispersion density	0.191	0.117	0.128	0.136	0.145	0.158	0.109	0.082	0.052	0.035	0.026	0.022	0.018	0.014	0.007

Run B11V1 $x = 21.25\%$; $F_H = 8.9$; Holdup = 0.878 cms.

Level (cms)	0.25	0.51	0.76	1.04	1.27	1.52	1.78	2.29	2.79	3.30	3.81
Dispersion density	0.485	0.385	0.340	0.327	0.332	0.335	0.299	0.190	0.055	0.017	0.015

Run B11V2 $x = 20.41\%$; $F_H = 11.7$; Holdup = 0.809 cms.

Level (cms)	0.13	0.25	0.51	0.76	1.14	1.52	1.91	2.29	3.05	3.43	3.94
Dispersion density	0.570	0.453	0.338	0.262	0.276	0.280	0.275	0.208	0.058	0.035	0.021

Run B11V3 $x = 20.30\%$; $F_H = 15.0$; Holdup = 0.777 cms.

Level (cms)	0.13	0.25	0.51	0.76	1.14	1.52	1.91	2.41	2.92	3.43	4.06	4.70
Dispersion density	0.524	0.378	0.238	0.179	0.190	0.208	0.207	0.180	0.105	0.057	0.024	0.015

Run B11V4 $x = 20.25\%$; $F_H = 17.3$; Holdup = 0.758 cms.

Level (cms)	0.25	0.51	0.76	1.14	1.65	2.16	2.67	3.18	3.81	4.44	5.08	6.15	7.23	8.30
Dispersion density	0.274	0.168	0.137	0.152	0.176	0.187	0.167	0.117	0.074	0.056	0.049	0.032	0.021	0.010

TABLE VI - 4 (Cont'd)

Run B12V1 $x = 38.75\%$; $F_H = 8.7$; Holdup = 0.823 cms.

Level (cms)	0.25	0.64	1.14	1.65	2.16	2.67	3.18	3.68
Dispersion density	0.454	0.349	0.320	0.322	0.204	0.058	0.010	0.001

Run B12V2 $x = 37.25\%$; $F_H = 12.2$; Holdup = 0.776 cms.

Level (cms)	0.28	0.51	0.76	1.07	1.32	1.65	2.03	2.41	2.79	3.17	3.56	3.94
Dispersion density	0.420	0.291	0.220	0.252	0.275	0.285	0.255	0.160	0.088	0.039	0.024	0.017

Run B12V3 $x = 33.37\%$; $F_H = 14.8$; Holdup = 0.745 cms.

Level (cms)	0.25	0.51	0.89	1.27	1.65	2.16	2.67	3.17	3.68	4.19	4.70
Dispersion density	0.410	0.261	0.192	0.205	0.245	0.219	0.132	0.068	0.041	0.020	0.012

Run B12V4 $x = 36.64\%$; $F_H = 17.9$; Holdup = 0.767 cms.

Level (cms)	0.25	0.64	1.02	1.40	1.91	2.41	2.92	3.43	4.06	4.70	5.35	7.23	8.30	9.14
Dispersion density	0.301	0.150	0.144	0.173	0.182	0.166	0.142	0.098	0.061	0.049	0.030	0.023	0.021	0.015

Run B13V1 $x = 66.23\%$; $F_H = 9.6$; Holdup = 0.813 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.504	0.388	0.335	0.299	0.300	0.303	0.279	0.229	0.158	0.082	0.041	0.023	0.010

Run B13V2 $x = 64.50\%$; $F_H = 12.1$; Holdup = 0.736 cms.

Level (cms)	0.28	0.51	0.76	1.04	1.27	1.52	1.78	2.03	2.26	2.54	2.79	3.05	3.30
Dispersion density	0.434	0.312	0.256	0.256	0.269	0.267	0.255	0.224	0.181	0.123	0.075	0.023	0.018

Run B13V3 $x = 63.93\%$; $F_H = 15.5$; Holdup = 0.723 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.52	2.03	2.41	2.79	3.18	3.56	3.94	4.45
Dispersion density	0.404	0.249	0.193	0.204	0.234	0.207	0.164	0.125	0.071	0.046	0.021	0.015

Run B13V4 $x = 63.17\%$; $F_H = 18.7$; Holdup = 0.736 cms.

Level (cms)	0.25	0.51	0.76	1.14	1.52	1.91	2.41	2.92	3.43	4.06	4.70	5.97	7.24	8.51	9.78
Dispersion density	0.282	0.151	0.118	0.123	0.135	0.157	0.156	0.134	0.122	0.065	0.045	0.029	0.021	0.017	0.012

TABLE VI-4 (cont'd)

Run B14V1 $x = 78.17\%$; $F_H = 9.7$; Holdup = 0.792 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.27	1.52	1.75	2.03	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.463	0.358	0.307	0.291	0.307	0.303	0.280	0.227	0.154	0.106	0.053	0.026	0.013

Run B14V2 $x = 77.66\%$; $F_H = 12.6$; Holdup = 0.787 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.68	4.06
Dispersion density	0.440	0.311	0.246	0.240	0.249	0.255	0.261	0.228	0.171	0.147	0.097	0.072	0.061	0.045	0.011

Run B14V3 $x = 77.38\%$; $F_H = 15.4$; Holdup = 0.737 cms.

Level (cms)	0.25	0.51	0.76	1.14	1.52	1.91	2.29	2.67	3.18	3.68	4.70
Dispersion density	0.384	0.248	0.192	0.193	0.205	0.223	0.124	0.132	0.053	0.038	0.019

Run B14V4 $x = 76.90\%$; $F_H = 18.2$; Holdup = 0.720 cms.

Level (cms)	0.25	0.51	0.76	1.14	1.91	2.41	2.95	3.43	4.06	4.70	5.33	5.97	6.61	7.23	7.87
Dispersion density	0.280	0.177	0.130	0.147	0.161	0.154	0.119	0.073	0.046	0.033	0.030	0.029	0.024	0.018	0.013

Run B15V1 $x = 94.77\%$; $F_H = 9.4$; Holdup = 0.757 cms.

Level (cms)	0.25	0.51	0.79	0.99	1.27	1.52	1.83	2.03	2.29	2.51	2.79	3.05	3.3
Dispersion density	0.506	0.377	0.278	0.271	0.290	0.262	0.237	0.183	0.131	0.085	0.049	0.030	0.024

Run B15V2 $x = 94.68\%$; $F_H = 11.9$; Holdup = 0.764 cms.

Level (cms)	0.25	0.51	0.79	1.02	1.27	1.52	1.78	1.98	2.29	2.62	2.92	3.30	3.68
Dispersion density	0.507	0.338	0.287	0.257	0.250	0.258	0.230	0.198	0.151	0.094	0.068	0.024	0.016

Run B15V3 $x = 94.62\%$; $F_H = 17.2$; Holdup = 0.716 cms.

Level (cms)	0.25	0.51	0.76	1.12	1.52	1.91	2.41	2.92	3.43	4.06	4.70	5.33	5.97
Dispersion density	0.355	0.219	0.164	0.165	0.183	0.216	0.193	0.121	0.066	0.036	0.032	0.621	0.010

Run B15V4 $x = 94.58\%$; $F_H = 19.8$; Holdup = 0.746 cms.

Level (cms)	0.25	0.51	0.76	1.14	1.65	2.16	2.79	3.43	4.06	4.70	5.33	6.60	7.87	9.14
Dispersion density	0.261	0.453	0.124	0.127	0.137	0.148	0.141	0.113	0.066	0.047	0.036	0.027	0.016	0.007

TABLE VI-5

Dispersion density data: System B; W = 0.95 cms

Run B20V1 x = 8.86 %; F_H = 9.2; Holdup = 0.546 cms.

Level (cms)	0.13	0.28	0.38	0.51	0.74	1.10	1.26	1.50	1.76	2.07	2.29	2.79
Dispersion density	0.619	0.499	0.405	0.337	0.266	0.220	0.202	0.155	0.085	0.024	0.012	0.011

Run B20V2 x = 8.39%; F_H = 12.1; Holdup = 0.554 cms.

Level (cms)	0.13	0.25	0.38	0.51	0.76	1.16	1.27	1.52	1.78	2.06	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.586	0.437	0.330	0.239	0.202	0.297	0.190	0.169	0.149	0.099	0.068	0.043	0.025	0.022	0.013

Run B20V3 x = 8.13%; F_H = 14.8; Holdup = 0.621 cms.

Level (cms)	0.13	0.25	0.38	0.51	0.76	1.16	1.27	1.52	1.78	2.06	2.29	2.54	2.92	3.30	3.68
Dispersion density	0.404	0.263	0.200	0.162	0.160	0.152	0.164	0.163	0.175	0.153	0.121	0.115	0.077	0.049	0.038
	4.06	4.95													
	0.034	0.020													

Run B20V4 x = 7.80%; F_H = 18.3; Holdup = 0.725 cms.

Level (cms)	0.25	0.38	0.51	0.76	1.16	1.40	1.78	2.16	2.54	2.92	3.43	3.94	4.45	5.46	5.97
Dispersion density	0.170	0.125	0.105	0.104	0.109	0.117	0.127	0.134	0.140	0.120	0.103	0.083	0.069	0.047	0.039
	6.99	7.49	8.51												
	0.029	0.026	0.020												

Run B21V1 x = 21.40%; F_H = 8.6; Holdup = 0.526 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.27	1.50	1.78	2.16	2.54
Dispersion density	0.510	0.330	0.256	0.245	0.201	0.136	0.076	0.019	0.013

Run B21V2 x = 20.01%; F_H = 14.4; Holdup = 0.608 cms.

Level (cms)	0.15	0.25	0.51	0.76	1.02	1.37	1.78	2.16	2.54	3.94	4.95	2.92
Dispersion density	0.470	0.379	0.196	0.174	0.173	0.189	0.158	0.124	0.182	0.028	0.015	0.034

TABLE VI-5 (Cont'd)

Run B22V1 $x = 40.30\%$; $F_H = 8.4$; Holdup = 0.541 cms.

Level (cms)	0.13	0.23	0.51	0.76	1.02	1.24	1.50	1.78	2.03
Dispersed density	0.598	0.534	0.351	0.294	0.259	0.212	0.124	0.040	0.018

Run B22V2 $x = 38.57\%$; $F_H = 12.7$; Holdup = 0.582 cms.

Level (cms)	0.13	0.25	0.51	0.76	0.99	1.27	1.50	1.78	2.03	2.29	2.54	2.79
Dispersed density	0.596	0.484	0.291	0.232	0.237	0.214	0.206	0.137	0.098	0.039	0.015	0.015

Run B22V3 $x = 38.42\%$; $F_H = 14.3$; Holdup = 0.625 cms.

Level (cms)	0.13	0.25	0.51	0.76	1.14	1.54	1.90	2.41	2.92	3.43	4.06	4.70
Dispersed density	0.543	0.383	0.225	0.206	0.198	0.193	0.145	0.084	0.053	0.037	0.033	0.018

Run B22V4 $x = 37.29\%$; $F_H = 18.5$; Holdup = 0.728 cms.

Level (cms)	0.25	0.63	1.02	1.40	1.90	2.41	2.92	3.42	4.06	4.70	5.33	5.96	6.73	7.49	8.26
Dispersed density	0.238	0.152	0.112	0.121	0.137	0.148	0.126	0.093	0.057	0.038	0.031	0.025	0.026	0.022	0.011

Run B23V1 $x = 64.48\%$; $F_H = 9.8$; Holdup = 0.530 cms.

Level (cms)	0.13	0.25	0.38	0.51	0.79	1.02	1.27	1.50	1.78	2.03	2.29
Dispersed density	0.620	0.533	0.425	0.342	0.264	0.224	0.180	0.112	0.065	0.024	0.018

Run B23V2 $x = 63.78\%$; $F_H = 11.7$; Holdup = 0.551 cms.

Level (cms)	0.13	0.25	0.38	0.51	0.76	1.02	1.27	1.55	1.78	2.01	2.26	2.54	2.79
Dispersed density	0.621	0.503	0.384	0.285	0.232	0.219	0.208	0.156	0.088	0.066	0.026	0.019	0.017

Run B23V3 $x = 63.02\%$; $F_H = 14.9$; Holdup = 0.590 cms.

Level (cms)	0.13	0.25	0.38	0.51	0.76	1.02	1.30	1.52	1.78	2.03	2.29	2.54	2.79	3.30	4.06
Dispersed density	0.513	0.403	0.305	0.205	0.186	0.176	0.195	0.190	0.160	0.114	0.078	0.065	0.036	0.033	0.020

4.70
0.013

TABLE VI- 5 (cont'd)

Run B23V4 $x = 62.09\%$; $F_H = 18.6$; Holdup = 0.691 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.67	3.18	3.81	4.19	5.33	6.60
Dispersion density	0.258	0.145	0.127	0.128	0.132	0.132	0.145	0.151	0.151	0.149	0.125	0.081	0.070	0.038	0.028
	7.87														
	0.017														

Run B24V1 $x = 78.57\%$; $F_H = 9.5$; Holdup = 0.530 cms.

Level (cms)	0.25	0.48	0.76	1.02	1.29	1.52	1.78	2.03	2.29	2.54	2.79
Dispersion density	0.475	0.338	0.288	0.249	0.178	0.113	0.083	0.046	0.040	0.029	0.014

Run B24V2 $x = 78.14\%$; $F_H = 12.2$; Holdup = 0.555 cms.

Level (cms)	0.25	0.53	0.71	1.02	1.32	1.52	1.78	2.06	2.29	2.54	2.79	3.05
Dispersion density	0.438	0.290	0.263	0.244	0.212	0.151	0.106	0.057	0.043	0.033	0.033	0.017

Run B24V3 $x = 77.76\%$; $F_H = 15.9$; Holdup = 0.615 cms.

Level (cms)	0.25	0.53	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.92	3.30	3.68	4.06	4.70
Dispersion density	0.341	0.231	0.197	0.206	0.213	0.214	0.183	0.150	0.104	0.082	0.054	0.040	0.022	0.020	0.018

Run B24V4 $x = 77.39\%$; $F_H = 18.6$; Holdup = 0.690 cms.

Level (cms)	0.25	0.51	0.89	1.27	1.70	2.16	2.67	3.18	3.94	4.70	5.71	6.73	7.61	8.49
Dispersion density	0.201	0.145	0.139	0.156	0.160	0.150	0.115	0.092	0.061	0.048	0.039	0.035	0.025	0.011

Run B25V1 $x = 94.71\%$; $F_H = 9.4$; Holdup = 0.568 cms.

Level (cms)	0.25	0.51	0.76	1.04	1.24	1.52	1.80	2.03	2.29	2.79
Dispersion density	0.529	0.383	0.296	0.249	0.213	0.144	0.083	0.029	0.034	0.013

Run B25V2 $x = 94.66\%$; $F_H = 12.53$; Holdup = 0.591 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.24	1.50	1.70	2.03	2.29	2.54	3.05
Dispersion density	0.549	0.319	0.366	0.209	0.211	0.188	0.141	0.091	0.052	0.037	0.014

TABLE VI-5 (cont'd)

Run B25V3 $x = 94.55\%$; $F_H = 16.8$; Holdup = 0.637 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	4.19
Dispersion density	0.407	0.198	0.162	0.170	0.200	0.212	0.203	0.153	0.148	0.114	0.109	0.070	0.048	0.031

Run B25V4 $x = 94.57\%$; $F_H = 19.2$; Holdup = 0.701 cms.

Level (cms)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	3.18	3.68	4.19	4.70	5.33
Dispersion density	0.245	0.139	0.110	0.102	0.132	0.125	0.138	0.119	0.135	0.159	0.134	0.103	0.077	0.060	0.044
	6.60	7.87													
	0.028	0.016													

TABLE VI.6

Dispersion density data System C $W = 1.9$ cms.Run C1V1 $x = 6.27\%$; $F_H = 8.7$; Holdup = 0.683 cms.

Level (cms.)	0.28	0.53	0.76	0.96	1.32	1.63	2.16	2.67	3.18	3.81
Dispersion density	0.486	0.288	0.225	0.221	0.243	0.214	0.136	0.073	0.042	0.012

Run C1V2 $x = 6.04\%$; $F_H = 12.1$; $H_L = 0.764$ cms.

Level (cms.)	0.25	0.51	0.76	1.14	1.52	1.91	2.41	2.92	3.43	3.94	4.45
Dispersion density	0.539	0.313	0.228	0.214	0.234	0.203	0.137	0.091	0.056	0.022	0.016

Run C1V3 $x = 5.95\%$; $F_H = 15.3$; $H_L = 0.728$ cms.

Level	0.25	0.51	0.76	1.02	1.40	1.78	2.16	2.67	2.92	3.68	4.19	4.70
Dispersion density	0.439	0.226	0.186	0.192	0.180	0.207	0.202	0.151	0.092	0.045	0.036	0.020

Run C1V4 $x = 5.86\%$; $F_H = 18.6$; $H_L = 0.701$ cms.

Level	0.25	0.51	0.76	1.14	1.65	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.60
Dispersion density	0.290	0.160	0.136	0.149	0.165	0.186	0.158	0.113	0.044	0.030	0.024	0.014	0.010

Run C2V1 $x = 24.12\%$; $F_H = 9.2$; $H_L = 0.625$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79
Dispersion density	0.539	0.292	0.219	0.198	0.184	0.160	0.143	0.117	0.099	0.074	0.056

Run C2V2 $x = 23.02\%$; $F_H = 12.5$; $H_L = 0.704$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56	3.94
Dispersion density	0.518	0.295	0.200	0.202	0.209	0.199	0.187	0.178	0.157	0.108	0.081	0.072	0.044	0.030	0.014

Run C2V3 $x = 22.60\%$; $F_H = 15.0$; $H_L = 0.732$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	4.19
Dispersion density	0.529	0.270	0.190	0.168	0.173	0.191	0.202	0.195	0.180	0.149	0.122	0.102	0.052	0.020

TABLE VI.6 (Cont'd.)

Run C2V4 $x = 22.06\%$; $F_H = 18.4$; $H_L = 0.705$ cms.

Level	0.25	0.51	0.89	1.27	1.65	2.03	2.41	2.79	3.18	3.68	4.19	4.70	5.21	6.47
Dispersion density	0.307	0.163	0.143	0.148	0.168	0.189	0.186	0.166	0.121	0.091	0.050	0.037	0.014	0.010

Run C3V1 $x = 48.74\%$; $F_H = 9.5$; $H_L = 0.629$ cms.

Level	0.28	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56
Dispersion density	0.499	0.304	0.221	0.207	0.185	0.183	0.160	0.130	0.112	0.083	0.081	0.048	0.033	0.024

Run C3V2 $x = 47.82\%$; $F_H = 12.0$; $H_L = 0.720$ cms.

Level	0.28	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56	3.81
Dispersion density	0.498	0.291	0.221	0.200	0.205	0.202	0.193	0.181	0.161	0.130	0.112	0.069	0.055	0.031	0.018

Run C3V3 $x = 47.00\%$; $F_H = 15.7$; $H_L = 0.722$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56	3.94
Dispersion density	0.397	0.281	0.171	0.166	0.182	0.180	0.192	0.189	0.202	0.187	0.153	0.126	0.100	0.069	0.041
	4.45														
	0.026														

Run C3V4 $x = 46.13\%$; $F_H = 18.7$; $H_L = 0.721$ cms.

Level	0.25	0.64	1.02	1.40	1.78	2.16	2.54	2.92	3.30	3.81	4.06	4.45	4.83	5.21	5.77
Dispersion density	0.249	0.133	0.142	0.153	0.167	0.183	0.180	0.167	0.134	0.095	0.066	0.045	0.031	0.027	0.024
	5.97														
	0.016														

Run C4V1 $x = 72.37\%$ $F_H = 9.5$; $H_L = 0.633$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.81
Dispersion density	0.495	0.313	0.223	0.219	0.205	0.191	0.165	0.157	0.118	0.104	0.078	0.049	0.038	0.017

Run C4V2 $x = 71.83$; $F_H = 12.1$; $H_L = 0.702$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56	3.81
Dispersion density	0.448	0.280	0.197	0.195	0.203	0.204	0.186	0.184	0.173	0.132	0.105	0.079	0.050	0.053	0.034

TABLE VI.6 (Cont'd.)

Run C4V2

(cont'd.) 4.06
 0.017

Run C4V3 $x = 71.54\%$; $F_H = 15.7$; $H_L = 0.705$

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.68	4.06
Dispersion density	0.333	0.189	0.153	0.159	0.167	0.197	0.191	0.208	0.186	0.172	0.160	0.131	0.114	0.067	0.048
	4.45	4.83													
	0.026	0.020													

Run C4V4 $x = 71.25\%$; $F_H = 18.8$; $H_L = 0.713$ cms.

Level	0.25	0.64	1.02	1.40	1.78	2.16	2.67	3.18	3.68	4.19	4.70	5.20	5.72	6.22	6.73
Dispersion density	0.197	0.136	0.118	0.148	0.150	0.170	0.166	0.152	0.122	0.090	0.048	0.036	0.025	0.020	0.013

Run C5V1 $x = 92.13\%$; $F_H = 9.5$; $H_L = 0.719$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56
Dispersion density	0.463	0.323	0.241	0.245	0.231	0.249	0.227	0.192	0.159	0.120	0.074	0.037	0.024	0.012

Run C5V2 $x = 92.09\%$; $F_H = 13.2$; $H_L = 0.748$ cms.

Level	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56	3.68
Dispersion density	0.452	0.293	0.209	0.208	0.214	0.217	0.223	0.211	0.200	0.168	0.120	0.098	0.073	0.033	0.017

Run C5V3 $x = 92.04\%$; $F_H = 15.3$; $H_L = 0.756$ cms.

Level	0.25	0.53	0.76	1.14	1.52	1.91	2.29	2.67	3.04	3.43	3.81	4.19	4.57	4.95	5.33
Dispersion density	0.346	0.201	0.175	0.174	0.210	0.221	0.209	0.187	0.130	0.072	0.045	0.040	0.031	0.024	0.014

Run C5V4 $x = 91.89\%$; $F_H = 18.8$; $H_L = 0.740$ cms.

Level	0.25	0.64	1.02	1.40	1.78	2.03	2.41	2.79	3.18	3.56	3.94	4.32	4.70	5.08	5.97
Dispersion density	0.203	0.129	0.119	0.142	0.161	0.165	0.172	0.164	0.162	0.127	0.102	0.085	0.058	0.049	0.024
	6.48	6.98													
	0.011	0.010													

TABLE VI.7

Dispersion density data System C W = 0.95 cms.

Run C11V1 x = 6.57%; $F_H = 9.2$; Holdup = 0.511 cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05
Dispersion density	0.528	0.298	0.205	0.185	0.157	0.120	0.083	0.062	0.043	0.026	0.019	0.012

Run C11V2 x = 6.47%; $F_H = 11.9$; $H_L = 0.549$ cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.54	2.79	3.05	3.30	3.56
Dispersion density	0.503	0.285	0.218	0.186	0.177	0.156	0.139	0.053	0.044	0.031	0.033	0.025	0.015

Run C11V3 x = 6.30%; $F_H = 15.2$; $H_L = 0.620$ cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30	3.56	3.81
Dispersion density	0.475	0.253	0.192	0.182	0.184	0.180	0.189	0.148	0.117	0.085	0.051	0.048	0.032	0.030	0.014

Run C11V4 x = 6.21%; $F_H = 18.4$; $H_L = 0.672$ cms.

Level (cms.)	0.25	0.64	1.02	1.52	2.16	2.79	3.33	4.06	4.70	5.33	5.97	6.60
Dispersion density	0.252	0.152	0.139	0.166	0.176	0.135	0.085	0.046	0.026	0.021	0.016	0.011

Run C12V1 x = 22.08%; $F_H = 9.4$; $H_L = 0.462$ cms.

Level (cms.)	0.28	0.56	0.74	1.04	1.40	1.75	2.08	2.41	2.79	3.18
Dispersion density	0.411	0.262	0.206	0.166	0.132	0.090	0.076	0.033	0.011	0.018

Run C12V2 x = 21.67%; $F_H = 11.6$; $H_L = 0.516$ cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05
Dispersion density	0.505	0.267	0.187	0.177	0.163	0.152	0.108	0.076	0.051	0.031	0.033	0.027

Run C12V3 x = 21.06%; $F_H = 15.3$; $H_L = 0.591$ cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.30	3.56	3.94
Dispersion density	0.416	0.226	0.168	0.174	0.178	0.183	0.161	0.146	0.116	0.087	0.055	0.022	0.026	0.020

TABLE VI.7 (Cont'd.)

Run C12V4 $x = 20.76\%$; $F_H = 18.5$; $H_L = 0.685$ cms.

Level (cms.)	0.25	0.63	1.14	1.65	2.16	2.79	3.43	4.06	4.70	5.33	5.97	6.60
Dispersion density	0.222	0.148	0.148	0.169	0.188	0.160	0.100	0.063	0.042	0.027	0.019	0.011

Run C13V1 $x = 48.71\%$; $F_H = 9.4$; $H_L = 0.481$ cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.30	1.52	1.78	2.03	2.54	2.79
Dispersion density	0.456	0.262	0.195	0.162	0.144	0.132	0.101	0.074	0.042	0.020

Run C13V2 $x = 47.95\%$; $F_H = 12.1$; $H_L = 0.521$ cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.417	0.238	0.182	0.166	0.165	0.138	0.134	0.113	0.058	0.057	0.032	0.033	0.024

Run C13V3 $x = 47.16\%$; $F_H = 15.6$; $H_L = 0.589$ cms.

Level (cms.)	0.25	0.53	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.414	0.218	0.194	0.187	0.203	0.203	0.158	0.138	0.123	0.102	0.060	0.041	0.025

Run C13V4 $x = 46.03\%$; $F_H = 18.1$; $H_L = 0.691$ cms.

Level (cms.)	0.25	0.51	0.89	1.27	1.65	2.03	2.41	2.79	3.18	3.68	4.19	4.70	5.21	5.72	6.22
Dispersion density	0.203	0.130	0.129	0.136	0.166	0.165	0.172	0.149	0.132	0.081	0.060	0.039	0.028	0.026	0.015

Run C14V1 $x = 71.89\%$; $F_H = 9.2$; $H_L = 0.511$ cms.

Level (cms.)	0.13	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.605	0.462	0.258	0.198	0.174	0.147	0.126	0.098	0.074	0.046	0.028	0.020	0.029	0.010

Run C14V2 $x = 71.61\%$; $F_H = 12.1$; $H_L = 0.541$ cms.

Level (cms.)	0.13	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.03	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.580	0.400	0.238	0.174	0.184	0.167	0.172	0.138	0.103	0.074	0.051	0.019	0.018	0.016

Run C14V3 $x = 71.15\%$; $F_H = 15.7$; $H_L = 0.599$ cms.

Level (cms.)	0.13	0.25	0.64	1.02	1.40	1.78	2.16	2.54	2.92	3.30	3.68	4.06	4.45
Dispersion density	0.430	0.253	0.145	0.153	0.159	0.169	0.160	0.114	0.090	0.057	0.040	0.031	0.012

TABLE VI.7 (Cont'd.)

Run C14V4 $x = 70.74\%$; $F_H = 18.5$; $H_L = 0.692$ cms.

Level (cms.)	0225	0.63	1.02	1.40	1.78	2.16	2.67	3.18	3.68	4.19	4.70	5.21	5.72	6.22	6.73
Dispersion density	0.159	0.112	0.104	0.129	0.144	0.181	0.170	0.141	0.117	0.079	0.056	0.042	0.035	0.024	0.018
	7.24														
	0.012														

Run C15V1 $x = 92.66\%$; $F_H = 9.3$; $H_L = 0.610$ cms.

Level (cms.)	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.06	2.29	2.54	2.79
Dispersion density	0.579	0.317	0.244	0.227	0.215	0.184	0.123	0.079	0.038	0.019	0.015

Run C15V2 $x = 92.61\%$; $F_H = 12.4$; $H_L = 0.651$ cms.

Level (cms.)	0.13	0.25	0.51	0.76	1.02	1.27	1.52	1.78	2.02	2.29	2.54	2.79	3.05	3.30
Dispersion density	0.625	0.415	0.241	0.199	0.206	0.208	0.196	0.177	0.129	0.087	0.055	0.027	0.025	0.018

Run C15V3 $x = 92.47\%$; $F_H = 15.4$; $H_L = 0.672$ cms.

Level (cms.)	0.13	0.25	0.38	1.02	1.40	1.78	2.16	2.54	2.92	3.30	3.68	4.06	4.45	4.83
Dispersion density	0.437	0.264	0.152	0.152	0.169	0.194	0.171	0.134	0.086	0.072	0.035	0.031	0.026	0.026

TABLE VI.8Performance data System A W = 7.6 cms.

<u>Run No.</u>	<u>F_H</u>	<u>x₀</u>	<u>x₁</u>	<u>E_{MV}</u>
A1V1	10.0	4.28	3.35	69.6
V2	13.7	4.07	3.23	65.2
V3	17.1	3.97	3.175	62.3
V4	20.1	3.92	3.165	59.0
A2V1	10.0	8.81	7.05	71.9
V2	13.9	8.70	7.09	65.8
V3	16.6	8.56	7.04	62.2
V4	19.8	8.40	6.97	59.1
A3V1	8.7	17.18	14.57	70.2
V2	10.0	17.04	14.31	73.9
V3	13.6	16.65	14.15	68.1
V4	17.0	16.42	14.04	63.8
A4V1	9.8	23.49	20.56	72.9
V2	13.7	23.01	20.27	68.2
V3	19.7	22.52	20.04	61.9
A5V1	9.4	29.21	26.39	74.0
V2	13.4	28.84	26.18	69.5
V3	16.9	28.61	26.11	65.0
V4	19.8	28.32	25.93	61.9
A6V1	10.5	39.65	37.81	72.3
V2	13.6	39.41	37.63	69.2
V3	19.7	39.19	37.58	62.6
A7V1A	9.9	43.61	42.22	74.8
V1B	10.7	42.96	41.55	72.1
V2	13.2	43.38	42.05	70.5
V3	15.8	43.30	42.02	68.0
V4	19.9	42.51	41.24	63.0
A8V1	10.0	48.02	47.31	72.5
V2	13.2	47.98	47.30	69.0
V3	16.7	47.88	47.20	67.1
V4	20.4	48.28	47.72	62.4

TABLE VI.8 (Cont'd.)

<u>Run No.</u>	<u>F_H</u>	<u>x₀</u>	<u>x₁</u>	<u>E_{MV}</u>
A9V1	10.6	58.89	59.76	71.2
V2	13.2	59.05	59.91	69.1
V3	17.7	59.11	59.92	64.8
A10V1	10.2	67.10	68.92	72.1
V2	14.0	67.23	68.94	67.6
V3	17.8	67.34	68.96	64.0
V4	20.7	67.58	69.14	61.1
A11V1	9.9	73.47	75.73	72.6
V2	13.8	74.00	76.12	67.9
V3	20.4	74.51	76.44	61.3
A12V1	10.4	80.35	82.68	71.9
V2	14.4	80.69	82.85	66.8
V3	16.9	80.97	83.03	63.8
A13V1	10.5	87.05	89.07	72.4
V2	15.5	87.52	89.33	65.5
V3	20.4	87.89	89.56	61.2
A14V1	10.5	94.29	95.41	72.9
V2	13.8	94.48	95.53	67.4
V3	17.1	94.62	95.60	64.0

TABLE VI.9

Performance data System A $W = 1.9$ cms.

Run.No.	F_H	x_0	x_1	E_{MV}
A21V1	8.8	3.51	2.93	49.0
V2	11.7	3.42	2.88	46.4
V3	18.1	3.40	2.86	46.3
A22V1	9.5	11.16	9.55	53.6
V2	13.1	10.97	9.41	52.2
V3	18.6	10.74	9.24	50.8
A23V1	9.8	22.82	20.64	54.7
V2	13.3	22.54	20.42	53.1
V3	18.4	22.19	20.08	52.8
A24V1	10.0	32.46	30.57	54.4
V2	13.8	32.22	30.40	52.2
V3	18.0	32.04	30.26	50.9
A25V1	8.9	44.95	44.14	53.2
V2	10.6	44.84	44.00	53.8
V3	18.0	44.70	43.92	50.5
A26V1	9.7	56.52	56.96	60.8
V2	12.7	56.41	56.82	57.8
V3	15.2	56.49	56.87	54.5
A27V1	9.5	64.99	66.38	62.0
V2	13.1	65.26	66.53	56.3
V3	15.9	65.33	66.54	53.5
A28V1	9.8	72.61	74.56	64.0
V2	13.1	72.99	74.69	56.3
V3	15.5	73.03	74.68	54.1
A29V1	9.5	77.02	79.23	65.7
V2	12.6	77.41	79.45	60.5
V3	15.9	77.56	79.39	54.4
A30V1	9.8	87.03	88.66	64.5
V2	12.7	87.33	88.82	58.5
V3	16.0	87.51	88.84	52.5
A31V1	9.4	94.01	94.98	57.5
V2	12.9	94.07	94.97	52.8
V3	18.5	94.23	95.03	47.5
A32V1	9.1	96.54	97.10	53.1
V2	12.2	96.76	97.26	48.5

TABLE VI.10Performance data System B W = 7.6 cms.

Run No .	F_H	x_0	x_1	E_{MV}
B1V1	7.5	7.41	4.22	79.3
V2	10.0	7.22	4.03	82.7
V3	13.1	6.94	3.93	80.0
B2V1	7.3	25.44	15.46	83.8
V2	10.9	24.21	14.61	83.4
V3	13.4	23.30	14.31	79.2
V4	17.4	22.26	13.98	73.9
B3V1	8.1	41.61	28.22	87.4
V2	10.8	40.45	27.67	83.9
V3	12.8	39.78	27.44	81.2
V4	18.4	38.03	26.74	74.9
B4V1	8.1	60.41	48.22	83.8
V2	10.9	59.54	47.45	82.2
V3	13.3	58.64	46.87	79.5
V4	18.3	56.89	45.95	73.0
B5V1	8.1	73.32	65.81	78.9
V2	10.2	72.70	64.90	79.5
V3	12.7	72.24	64.29	79.4
V4	17.2	71.31	63.87	73.2
V5	20.6	70.49	63.27	69.7
B6V1	8.5	86.16	82.99	83.1
V2	11.8	86.04	83.03	79.4
V3	14.4	85.62	82.69	74.8
V4	18.2	85.17	82.40	69.0
B7V1	11.1	94.83	94.21	79.7
V2	19.2	94.62	94.09	67.4

TABLE VI.11Performance data System B W = 1.9 cms.

Run No.	F_H	x_0	x_1	E_{MV}
B10V1	8.7	11.47	7.55	55.7
V2	12.6	10.69	6.99	56.2
V3	14.8	10.53	6.94	55.0
V4	18.5	9.84	6.44	55.7
B11V1	8.9	29.27	21.25	57.7
V2	11.7	28.10	20.41	56.3
V4	17.3	27.91	20.25	56.2
B12V1	8.7	48.01	38.78	58.3
V2	12.2	46.35	37.25	57.4
V3	14.8	46.52	37.37	57.7
V4	17.9	45.69	36.64	57.1
B13V1	9.6	71.26	66.23	53.6
V2	12.1	69.78	64.50	53.0
V3	15.5	69.27	63.93	52.6
V4	18.7	68.55	63.17	51.7
B14V1	9.7	80.83	78.17	48.5
V2	12.6	80.37	77.66	48.0
V3	15.4	80.10	77.38	47.3
V4	18.2	79.73	76.90	47.9
B15V1	9.4	95.07	94.77	44.2
V2	11.9	94.98	94.68	45.3
V3	17.0	94.91	94.62	42.5
V4	19.8	94.86	94.58	40.1

TABLE VI.12

Performance data System B W = 0.95 cms.

Run No.	F_H	x_0	x_1	E_{MV}
B20V1	9.2	13.07	8.86	52.2
V2	12.1	12.62	8.39	54.8
V3	14.8	12.24	8.13	54.3
V4	18.3	11.71	7.80	54.0
B21V1	8.6	28.64	21.40	52.0
V2	14.4	27.36	20.01	54.3
B22V1	8.4	48.56	40.30	52.3
V2	12.7	47.15	38.57	54.0
V3	14.3	47.18	38.42	55.1
V4	18.5	46.18	37.29	56.0
B23V1	9.8	69.22	64.48	47.5
V2	11.7	68.85	63.78	49.6
V3	14.9	68.31	63.02	50.6
V4	18.6	67.50	62.09	50.3
B24V1	9.5	80.81	78.57	41.8
V2	12.2	80.55	78.14	43.9
V3	15.9	80.34	77.76	46.0
V4A	18.6	80.07	77.39	46.7
V4B	18.6	79.87	77.22	45.7
B25V1	9.4	94.97	94.71	38.8
V2	12.53	94.93	94.66	39.5
V3	16.8	94.86	94.59	39.4
V4	19.2	94.85	94.57	39.9

TABLE VI.13

Performance data System C $W = 1.9$ cms.

Run No.	F_H	x_0	x_1	E_{MV}
C1V1	8.7	8.85	6.27	58.6
V2	12.1	8.27	6.04	52.1
V3	15.3	7.99	5.95	48.4
V4	18.6	7.70	5.86	44.1
C2V1	9.2	30.31	24.12	63.4
V2	12.5	28.76	23.02	55.6
V3	15.0	27.54	22.60	51.4
V4	18.4	26.59	22.06	47.5
C3V1	9.5	53.81	48.74	61.3
V2	12.0	52.48	47.82	55.4
V3	15.7	51.32	47.00	50.6
V4	18.7	50.38	46.13	49.0
C4V1	9.5	74.75	72.37	60.6
V2	12.1	74.18	71.83	58.2
V3	15.7	73.73	71.54	53.4
V4	18.8	73.39	71.25	51.3
C5V1	9.5	92.58	92.13	55.7
V2	13.2	92.55	92.09	54.4
V3	15.3	92.49	92.04	53.5
V4	18.8	92.33	91.89	51.4

TABLE VI.14Performance data System C W = 0.95 cms.

Run No.	F_H	x_0	x_1	E_{MV}
C11V1	9.2	8.78	6.57	48.1
V2	11.9	8.46	6.47	44.0
V3	15.2	8.21	6.30	43.4
V4	18.4	8.10	6.21	43.6
C12V1	9.4	26.96	22.08	51.0
V2	11.6	25.94	21.67	44.9
V3	15.3	25.36	21.06	45.8
V4	18.5	25.16	20.76	47.1
C13V1	9.4	52.98	48.71	51.4
V2	12.1	51.93	47.95	47.4
V3	15.6	51.29	47.16	48.5
V4	18.1	50.27	46.03	49.0
C14V1	9.2	73.95	71.89	50.9
V2	12.1	73.60	71.41	48.9
V3	15.7	73.19	71.15	49.0
V4	18.5	72.89	70.74	50.5
C15V1	9.3	93.04	92.66	48.7
V2	12.4	93.02	92.61	51.1
V3	15.4	92.88	92.47	51.9

APPENDIX VIINOMENCLATURE

C	Concentration (Kg moles/m ³)
C_D	Drag coefficient
D	Diffusivity
D_E	Effective diffusivity
D_M	Molecular diffusivity
d	Bubble or drop diameter
d'	Corrected drop diameter from equation 2.41
d_c	Column diameter
d_p	Particle diameter
E_{MV}	Murphree plate efficiency based on gas phase
E_p	Murphree point efficiency
F_b	F-factor based on bubbling area
F_H	F-factor based on hole area
F_G	Volumetric fraction of gas in the dispersion
F_L	Volumetric fraction of liquid in the dispersion
$f_{i,j}$	Substance specific correction factor
G	Volumetric gas rate
g	Acceleration due to gravity
H_L	Liquid holdup per unit area of plate
h	Height of dispersion
K_G, K_L	Gas, liquid phase mass transfer coefficient
K_{OG}, K_{OL}	Overall mass transfer coefficient based on gas, liquid phase
L	Volumetric liquid rate

L'	Characteristic linear dimension, equation 2.38
L_W	Liquid flow per unit column width
M	Molecular weight
m	Slope of the equilibrium line
N_G, N_L	Number of gas, liquid phase transfer units
N_{OG}, N_{OL}	Number of overall transfer units based on gas, liquid phase
P	Pressure
Q_L	Liquid holdup in dispersion
R	Bubble or drop radius
R_S	Residual sum of squares
r	Radius
S	Surface area of mass transfer
s	Standard deviation
T	Absolute temperature
t	Time
V	Gas velocity
V_H	Hole velocity
V_o	Actual gas velocity
W	Weir height
x	Mole fraction of component in the liquid phase
y	Mole fraction of component in the vapour phase
α	Relative volatility
γ	Activity coefficient
Γ_h	Hydrodynamic factor in equation 2.27a
μ	Viscosity

ρ	Density
σ	Surface tension

Subscripts

A,B	Components A,B
b	Bulk condition
G	Gas phase
H	Hole
i	Interface value
L	Liquid
T	At temperature T
1,2	Conditions 1 and 2

Superscripts

*	Equilibrium
o	Reference state

Dimensionless groups

F_{OD}	Diffusional Fourier number
G_a	Galileo number
Pe	Peclet number
Re	Reynolds number
Sc	Schmidt number
Sh	Sherwood number

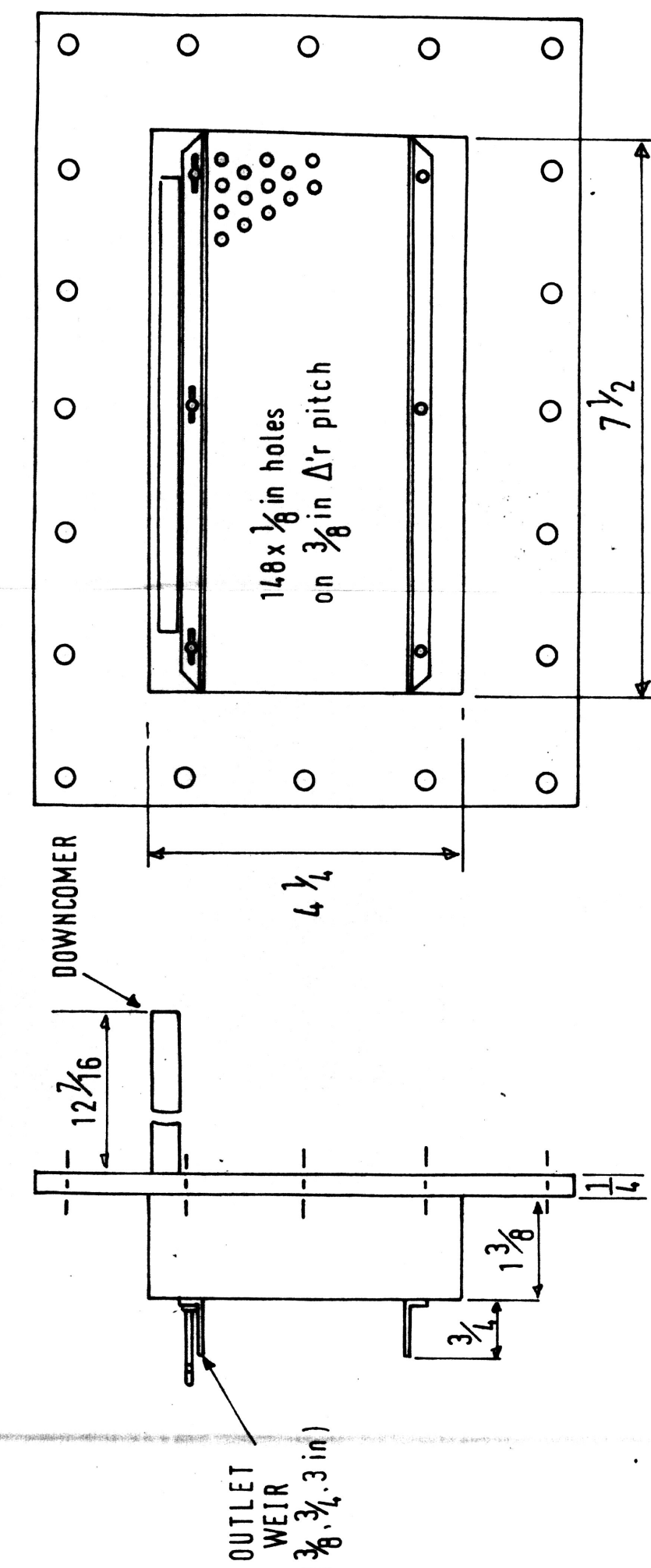
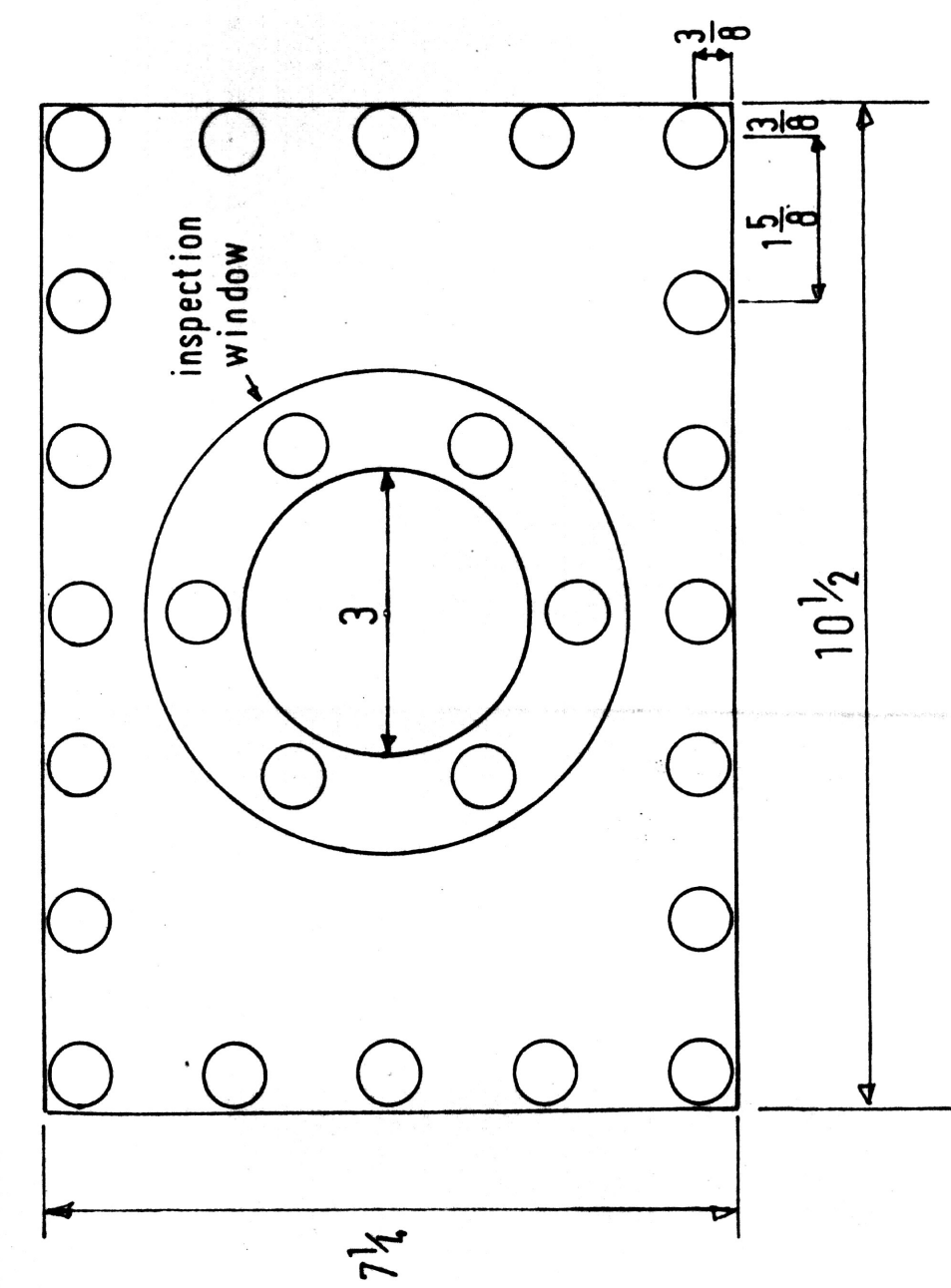
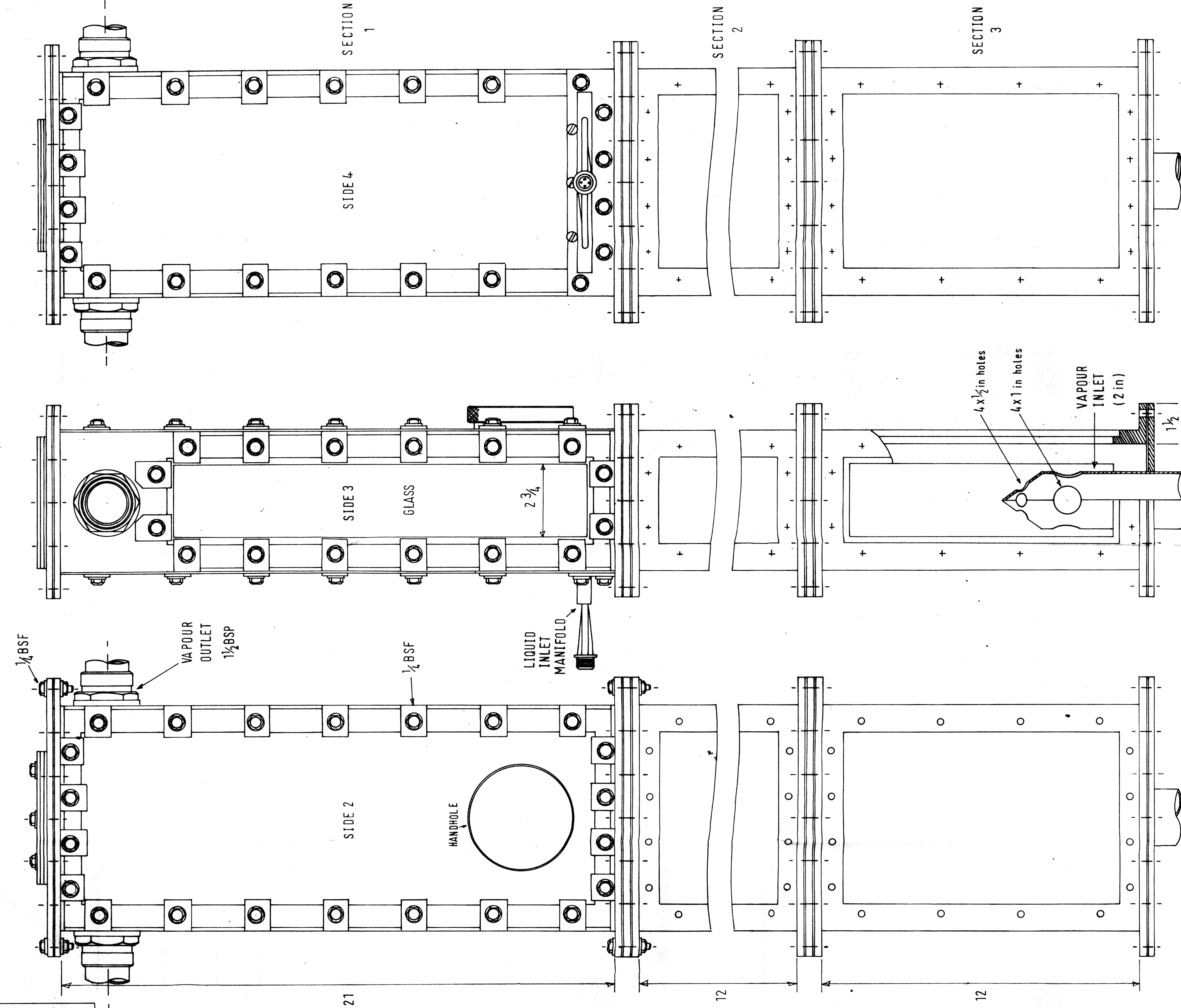
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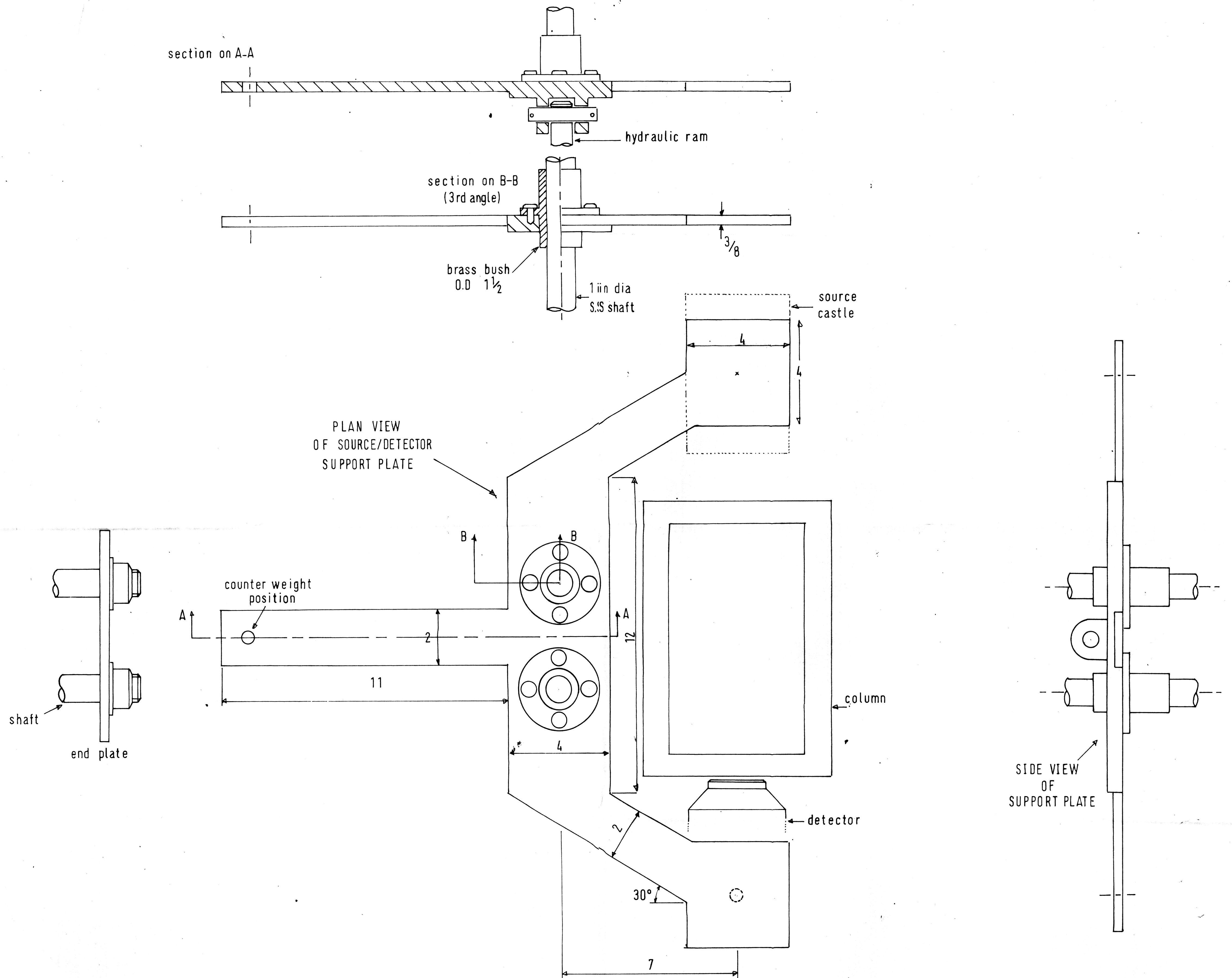
DRG No.1

SIEVE PLATE COLUMN

ALL BRASS CONSTRUCTION

HALF SCALE
all dimensions in inches.

drawn by A.G.FANE



DRG No. 2

GAMMA-RAY TRAVERSE SYSTEM

HALF SCALE
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