

THE PHYSICAL AND MASS TRANSFER
PROPERTIES OF BUBBLES IN A FLUIDISED BED

by

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The work described in this thesis is the original work of the author except where specific reference is made to other sources. It has not been submitted, in whole or in part, for any degree at any other University.

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ABSTRACT OF THESIS

A rapid technique has been developed to measure the size, shape and velocity of bubbles in a gas fluidised bed of electrically conducting particles together with the simultaneous gas exchange rates of these bubbles with the dense phase. The technique uses a compound probe which comprises a five element conductivity sensor to detect the bubble parameters and a hot wire to sense the concentration of a tracer both in the dense phase and within the bubbles, the probe being coupled to a high speed digital computer processing system which operates in real time.

The system of measurement was unambiguous and of adequate speed of response and was applied to the cases where single bubbles, or bubble doublets, were injected in an air fluidised bed of carbon maintained at incipient fluidisation.

The rates of mass transfer from individual bubbles rising through an incipiently air fluidised bed of carbon have been measured under conditions where the bubble rising velocity was very much greater than the dense-phase gas velocity.

Helium concentration recorded near to and within a single rising bubble showed that the principal concentration difference resided close to the frontal interface of the bubble. The overall steady-state rates of mass transfer could be predicted using a single surface renewal over the unimpeded frontal surface of the bubble.

The experiments conducted in which two helium bubbles were caused to coalesce, as they rise through the air fluidised bed, show that the resulting single bubble almost instantaneously

assumes a concentration very close to that of the adjacent dense phase. This observation makes it clear that bubble coalescence is responsible for a significant contribution to the overall mass-transfer rate between the bubbles and the dense phase.

CHAPTER 1

INTRODUCTION

INTRODUCTION

Brötz (5), in 1952, brought to general knowledge the earliest recorded application of fluidisation. A fluidised-like ore processing operation has been described by Agricola in 1556. However, the first use of a fluidised bed as an industrial chemical reactor was not until 1921 when BASF developed the Winkler gas generator and it was not until 1942 that scientific interest in fluidised bed processes began. The increased demand for oil derivatives during the war led to the successful operation of a fluidised bed as a catalytic cracking reactor and stimulated interest in its scientific study. Since then a great effort has been expended, especially in the study of the fundamental aspects governing the mechanism of gas flow within the bed, in order to facilitate scale-up operations.

Numerous works in the literature report small-scale experiments conducted in ideal laboratory conditions, avoiding gas by-passing. Results reported are not reproduced in large-scale industrial equipment, where it is known that a considerable portion of the gas may by-pass the bed in the form of bubbles. This regime is referred to as aggregative fluidisation and is distinct from the previous particulate regime.

The above mentioned discrepancies between laboratory-scale experiments and industrial beds can be mainly attributed to differences in "hold-up" time of the gas in the bed when one or other of the fluidisation regimes is prevailing. Bubbles or slugs rise in gas-fluidised beds with velocities much higher than those of the gas in the bulk or dense phase. It is apparent that in the aggregative regime two distinct hold-up times of the gas in the bed

exist, whereas, in the particulate regime all the gas has the same hold-up time.

This simple two-phase flow model of gas flow in the bed explains, in the first instance, the change in the degree of conversion when a catalytic reaction is carried out under particulate or aggregative fluidisation. Since conversion depends on the gas-solid contact time, the degree of conversion in a fluidised bed will depend on the relative distribution of gas between the bubble and dense phases, the bubble size distribution, and the interaction between both phases.

The measurement of all these variables will be necessary to rationalise the design of fluidised bed reactors.

Several years of research in this department led to the development of a probe technique for the simultaneous measurement of bubble velocity, shape and size, and its distribution in fluidised beds of electrically conductive particles. The present work was undertaken to extend such a technique to the measurement of a tracer gas concentration within the bubble and in the neighbouring dense phase as the bubble undergoes gas transfer with its surroundings.

The technique will thus permit the simultaneous measurement of all the variables necessary for the evaluation of mass transfer rates between bubbles and dense phase.

The technique was applied to cases where streams of single bubbles or bubble doublets were introduced at various frequencies, in an incipiently-fluidised bed of coke.

Mass transfer rates between bubbles and dense phase were obtained in the first set of experiments by measuring bubble parameters and gas concentration at different levels. The experiments in which a stream of bubble doublets were introduced

in the bed, permit an assessment of the effect of bubble coalescence upon the gas exchange between bubbles and dense phase. Detailed comparisons of the data and existing theoretical models relating to bubble motion and mass transfer were carried out. Finally, gas exchange between a single rising bubble and the dense phase was found to be well represented by the Higbie "surface renewal" model and complete mixing between bubbles and dense phase occurs when two identical bubbles coalesce.

CHAPTER 2

BUBBLES IN FLUIDISED BEDS

2.1 BUBBLE VELOCITY

One of the most remarkable observations of single bubbles rising in fluidised beds is the fact that their behaviour is similar to that of larger air bubbles in liquids of low viscosity, such as water.

Davies and Taylor (21), studying the rising of large bubbles in water, found that the rising velocity, U_b , of a bubble was proportional to the square root of the diameter of the equivalent spherical bubble, D_e . They proposed the semi-empirical formula

$$U_b = K \sqrt{g \cdot D_e} \quad (2.1)$$

where K is an empirical constant found to be

$$K = 0.711 \quad (2.2)$$

Experimental measurements of bubble rising velocity in fluidised beds were made by several workers.

Davidson et al. (19) and Harrison and Leung (29), injecting bubbles in incipiently air-fluidised beds found that bubbles in fluidised beds rise more slowly than in an inviscid liquid, although the apparent difference between their experimental findings and the predicted velocity from Davies-Taylor equation was not significant. This led Davidson and Harrison (17) to assume that the rising velocity of bubbles in fluidised beds could be obtained from equation (2.1).

Rowe and Partridge (69), measuring the rise velocity of single three-dimensional bubbles in fluidised beds using different materials and particle sizes, found a scattered relationship between bubble size and bubble velocity, although a semi-empirical equation

similar to the Davies-Taylor equation was proposed

$$U_b = K' \sqrt{g \cdot r_b} \quad (2.3)$$

where r_b is the bubble frontal radius, and K' has values between 0.835 and 1.210. From their experimental findings of bubble shapes, as will be shown below, the relation between r_b and D_e is

$$r_b = \frac{1}{2} \frac{D_e}{\sqrt[3]{1 - f_w}} \quad (2.4)$$

where f_w is the wake fraction. The above equation (2.3) in terms of D_e will become

$$U_b = \frac{K'}{1.41 \sqrt[6]{1 - f_w}} \sqrt{g \cdot D_e} \quad (2.5)$$

or as proposed by Davies-Taylor

$$U_b = K \sqrt{g \cdot D_e} \quad (2.6)$$

where

$$K = \frac{K'}{1.41 \sqrt[6]{1 - f_w}} \quad (2.7)$$

the values of K ranging now from 0.62 to 0.88 for fluidised beds.

They found that the constant K' tends to increase as the particle size decreases. Although not finding any possible relationship between the constant and particle size, they proposed the following empirical equation

$$K' = 1.34 \epsilon_{mf} + 0.31 \quad (2.8)$$

which relates the constant K' with the bed porosity ϵ_{mf} at incipient

fluidisation.

Kunii and Levenspiel (41) summarized the major experimental findings for the rise velocity of single bubbles and from their work one can notice that all the authors proposed a Davies-Taylor type of equation (21), the values of the constant being always confined within the limits found by Rowe and Partridge (69).

The remarkable agreement between the Davies-Taylor equation (21) and the scatter of K values for fluidised beds obtained by Rowe and Partridge (69) is that the value of K given for bubbles in water ($K = 0.711$) is the median for those found in fluidised beds, and until further light is thrown upon this subject the Davies-Taylor equation (21) is certainly the best representation of bubble rise velocity in fluidised beds. Its main advantages are

- a) Its independence of the fluidised solid properties.
- b) Its independence of the fluidised gas.
- c) Bubble velocity is obtained from the knowledge of a single parameter characteristic of the bubble - its equivalent spherical diameter.

2.2 BUBBLE SHAPE

Unlike bubbles in gas-liquid systems, where bubble shapes range from spheres through oblate spheroids, irregular ellipsoids to spherical caps, as reported by several workers (2, 10, 15, 35, 44, 75) the bubble shape depending on its volume, bubbles in fluidised beds present a single shape independent of volume. Bubble shapes in fluidised beds are generally spherical with an indented rear surface, which comprises a wake of particles travelling with the

bubble.

The best description of bubble shape in fluidised beds was given by Rowe and Partridge (69) analysing three-dimensional bubbles in fluidised beds of different materials, using X-ray photography.

Figure 2.1 presents an idealised description of their findings. The wake angle, θ_w , was found to increase with diminishing particle size. For glass spheres they found $\theta_w = 134^\circ$ at $d_p = 550 \mu\text{m}$ and $\theta_w = 151^\circ$ at $d_p = 60 \mu\text{m}$. Figure 2.2 shows the bubble outlines as obtained by Rowe and Partridge (69) for different materials.

The variation of the wake angle with particle size was analysed by Harrison (28) by considering the variation of the bed viscosity with particle size. This has been measured by Kramers (39) and Matheson et al. (46) who found that bed viscosity decreases with decreasing particle size. Thus, as particle size decreases bubbles in fluidised beds will tend to have the same shape as the idealised shape proposed by Davies and Taylor (21) for bubbles in inviscid liquids, a trend which is reflected in figure 2.2.

2.3 DAVIDSON MODEL OF BUBBLE MOTION

The first theoretical interpretation of gas and solid movement in the vicinity of a rising bubble in a fluidised bed was given by Davidson and Harrison (17). In proposing their model the following assumptions were made

- a) The bubble is a solid free spherical void.
- b) The dense phase is a homogeneous incompressible fluid having the same properties as at incipient fluidisation.
- c) Flow of gas in the dense phase obeys Darcy's Law.

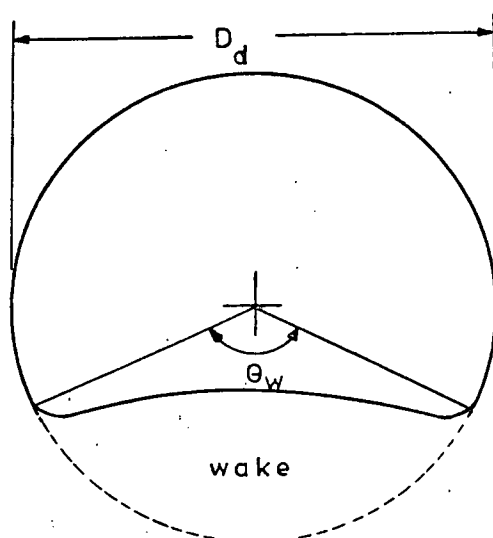


Figure 2.1 Ideal gas-fluidised bed bubble.
(after Rowe and Partridge (69))

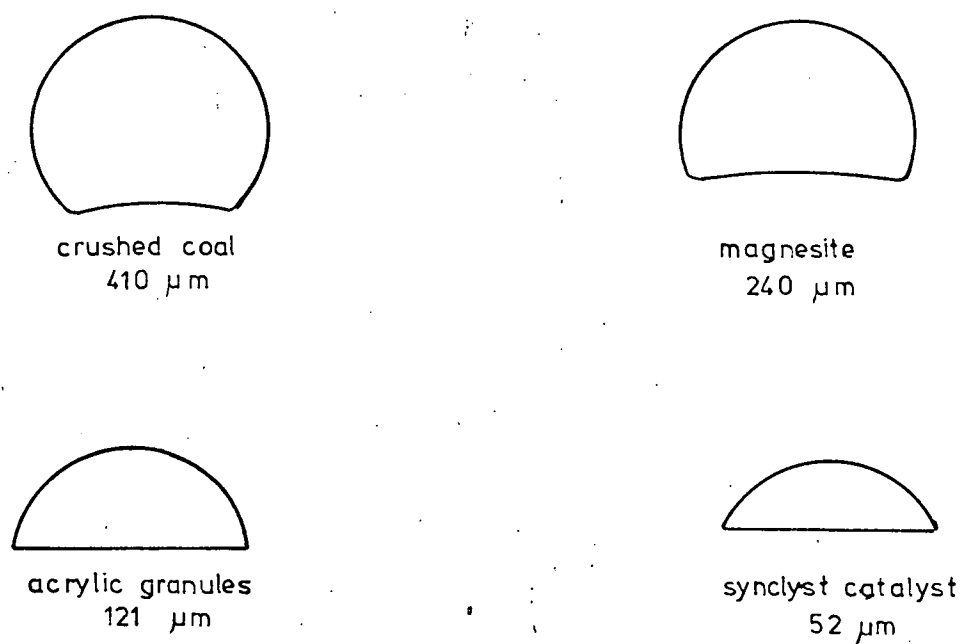


Figure 2.2 Bubble shapes in various materials.
(after Rowe and Partridge (69)).

d) Particles flowing around the bubble move as an inviscid fluid.

e) The pressure within the bubble is constant.

From the above assumptions the flow of particles around the bubble was described by the potential flow theory, and is given by the velocity potential

$$\phi = U_b \left(r + \frac{a^3}{2r^2} \right) \cos \theta \quad (2.9)$$

where U_b is the rising velocity of the bubble, r and θ polar co-ordinates and a the radius of the bubble. The pressure distribution within the fluidised fluid will satisfy the Laplace equation

$$\nabla^2 P = 0 \quad (2.10)$$

from which

$$P = - \frac{U_{mf}}{\epsilon_{mf} K_p} \left(r - \frac{a^3}{r^2} \right) \cos \theta \quad (2.11)$$

where U_{mf} and ϵ_{mf} are, respectively, the superficial velocity of fluidising fluid and bed porosity, both at incipient fluidisation and K_p is a permeability constant.

Davidson and Harrison (17) showed that the pressure in the lower part of the bubble is lower than the pressure in the surrounding bed but the converse is true in the upper part of the bubble. Since they postulate that the pressure within the bubble is constant, fluidising fluid must flow into the bubble from below and out from the top. Evidence of this flow was obtained by Wace and Burnett (78) by measuring the pressure distribution round

artificial gauze bubbles. Also, Rowe and Wace (71) showed that a visible trace of nitrogen dioxide gas passed up through a bubble rising in a two-dimensional bed, for bubbles with a rising velocity less than that of the fluidising gas, and gas circulation within the bubble was observed by Rowe et al. (70) for fast moving bubbles.

The stream function which describes the fluidising fluid motion is obtained by superimposing the potential field of the particle motion and the pressure gradient within the fluidising fluid. The stream function is then

$$\psi = (U_b - V_o) \left(1 - \frac{A^3}{r^3}\right) \frac{r^2 \sin^2 \theta}{2} \quad (2.12)$$

where

$$A^3 = a^3 \left(\frac{U_b + 2V_o}{U_b - V_o} \right) \quad (2.13)$$

and

$$V_o = \frac{U_{mf}}{\epsilon_{mf}} \quad (2.14)$$

is the interstitial velocity of the fluidising fluid.

A close examination of equations (2.12) and (2.13) reveals two different patterns of gas flow in the vicinity of the bubble.

a) Slow moving bubbles ($V_o > U_b$)

As $V_o > U_b$ it follows from equation (2.13) that in this case A is negative, and therefore equation (2.12) represents the stream function of fluid percolation through a void of radius A .

The gas in the dense phase moves upwards faster than the

bubble and percolates through the bubble as a preferential path.

b) Fast moving bubbles ($V_o < U_b$)

In this case it is clear from the above equations, that there exists a sphere concentric with the bubble and with radius $A > a$, where the radial component of fluidising fluid velocity is zero. The region between the bubble and this surface being named the bubble cloud.

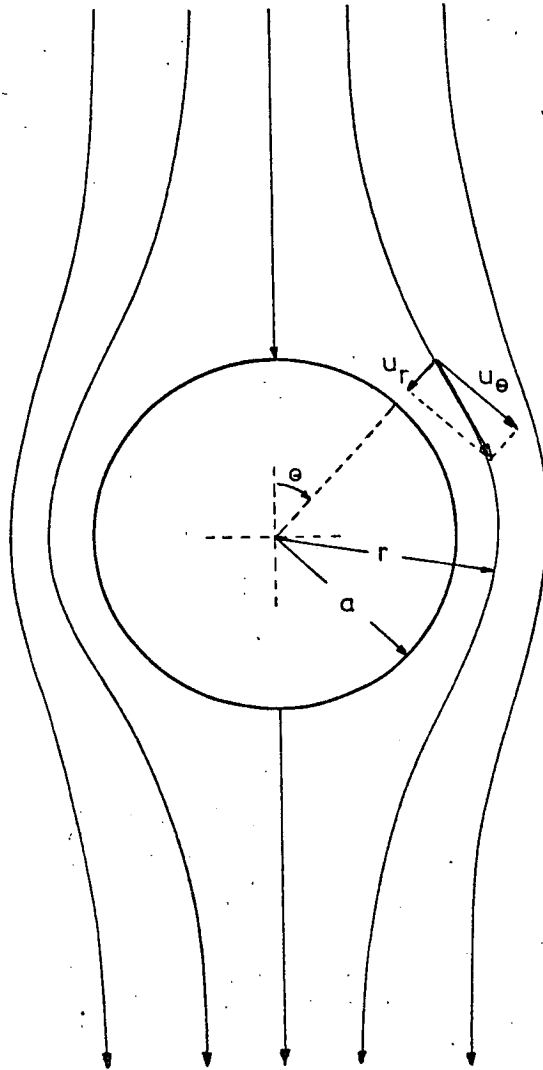
As explained above, gas leaving the bubble from the top will be dragged down by the fast moving particles in this region re-entering the bubble at the base. No mixing between the gas in the dense phase and the gas in the cloud exists. Thus, the surface of radius A is the limit of penetration of the circulating gas in and out of the bubble.

It is noteworthy from equation 2.1 that the ratio A/a decreases as the ratio U_b/V_o increases.

$$A \rightarrow a \quad \text{as} \quad \frac{U_b}{V_o} \rightarrow \infty$$

Figures 2.3 and 2.4 show schematically the model of Davidson.

The earlier experiments of Wace and Burnett (78), de Kock (22) and the more recent experiments of Rowe and Partridge (68) (69) and Rowe et al. (70), present experimental evidence of cloud formation around bubbles. In this latest experimental work the injection of NO_2 bubbles in a two-dimensional air fluidised bed, at incipient fluidisation, enabled the authors to describe the shape and size of clouds as they form around bubbles. Bubble shapes were discussed earlier in this chapter. In relation to the cloud Rowe et al. (70) found that it also has a spherical shape with an indented rear surface



$$\phi = U_b \left(r + \frac{a^3}{2r^2} \right) \cos \theta$$

$$u_r = \frac{\partial \phi}{\partial r}$$

$$u_\theta = \frac{1}{r} \frac{\partial \phi}{\partial \theta}$$

Figure 2.3

Davidson model - Particle trajectories near
a fast, single three dimensional bubble.

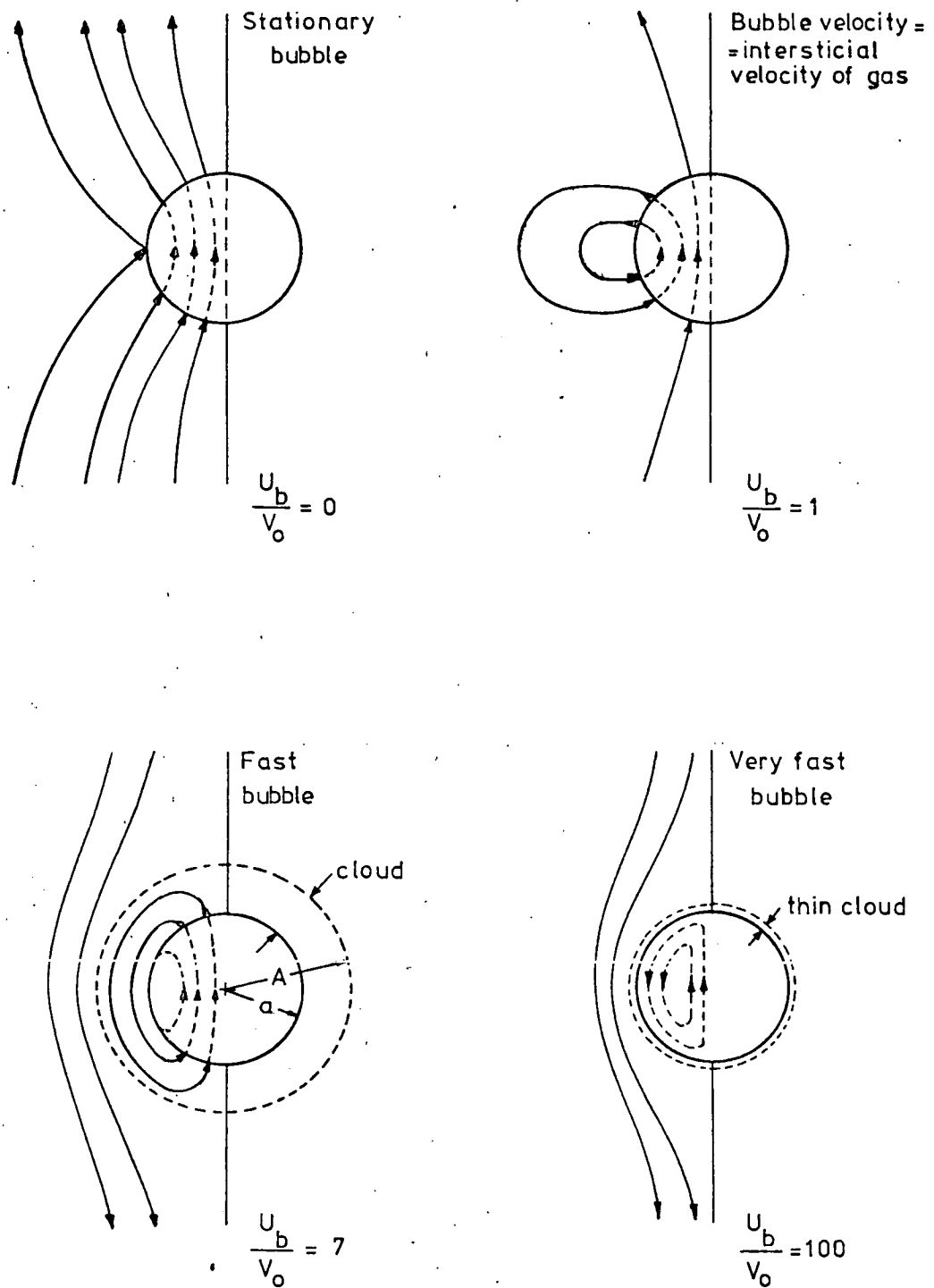


Figure 2.4

Davidson model - Gas streamlines near single
three dimensional gas-fluidised bed bubbles.

which does not extend beyond the lower edge of the bubble. The bubble-cloud, as found by them, is presented in figure 2.5.

2.4 OTHER MODELS

The elegant model of Davidson (16), based on a spherical bubble, gives a clear understanding of bubble and cloud behaviour, and because of its simple physical basis has been widely used in the literature.

More elaborate models for fluidising fluid flow around bubbles have been proposed by Jackson (33) and Murray (51) (52).

Jackson (33) assumed that constant pressure conditions are satisfied only approximately and locally at the bubble surface, and also that the dense phase voidage varies in the vicinity of the bubble. Jackson's (33) model leads to a smaller cloud than predicted by Davidson (16) and is displaced ahead of the bubble.

Jackson (33) regards his solution as complementary to Davidson's (16) in establishing the validity of their common features and by no means more complete.

Murray (52), in his model, saw no bubble curvature directly related to the flow of gas and solids, and for the solution of the equations proposed three bubble shapes. One of the shapes is spherical and the other two are related to experimentally observed bubble shapes, as obtained by Rowe et al. (70).

As in Jackson's (33) model, Murray's (52) model predicts also an eccentric cloud which is displaced above the bubble and is also smaller than that predicted from Davidson's (16) model.

Figure 2.6 compares bubble-cloud size as given by

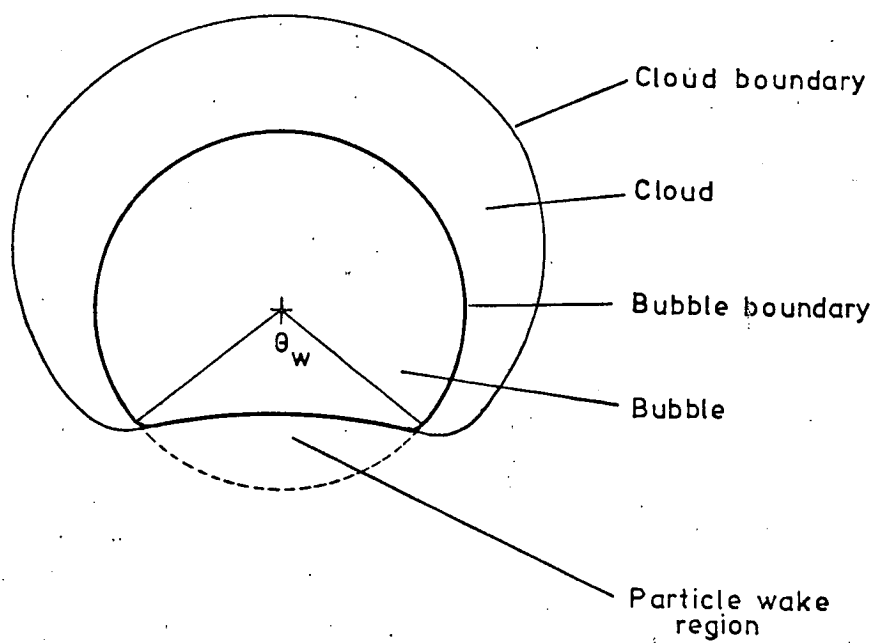


Figure 2.5 Ideal gas-fluidised bed bubble-cloud.
(after Rowe et al. (70))

Davidson's (16) model and bubble cloud size at the bubble nose,
as given by Murray's (52) model.

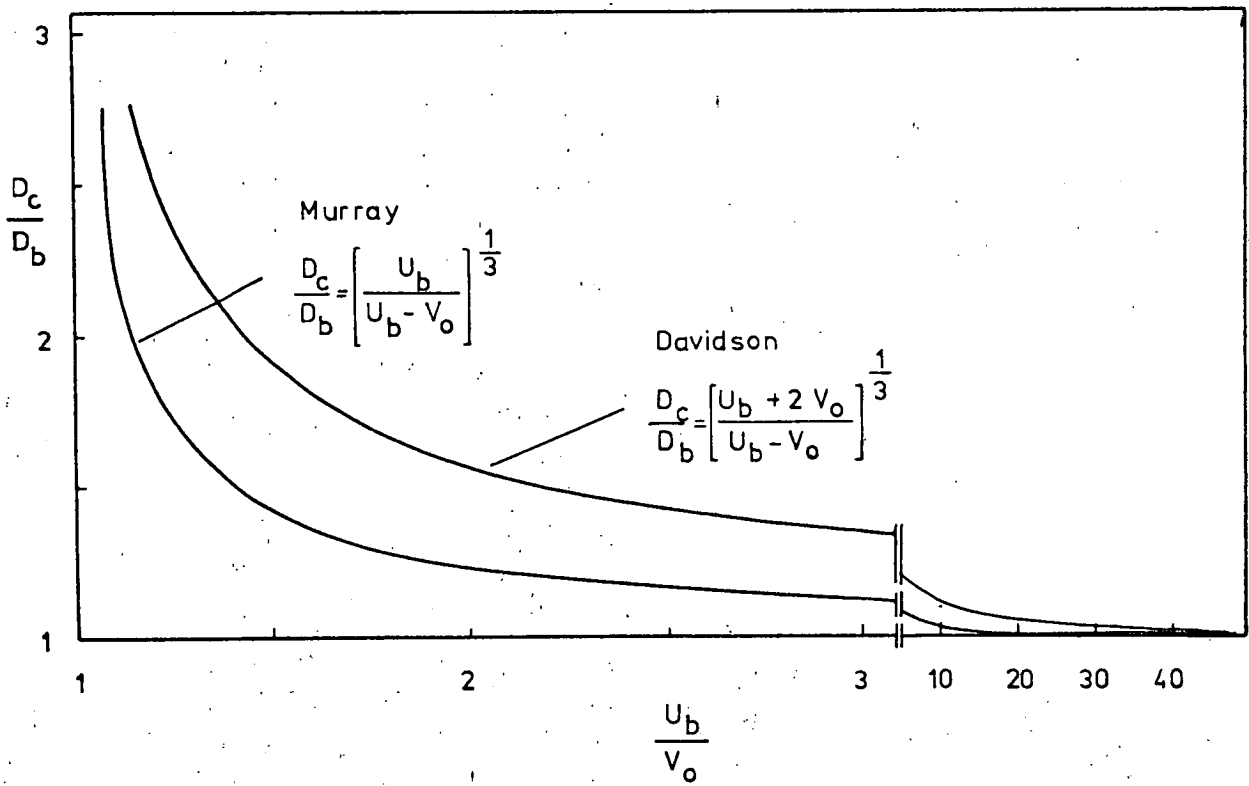


Figure 2.6 Ratio of cloud diameter to bubble diameter on vertical axis above three dimensional bubbles. Comparison between Davidson's and Murray's models. (D_c -diameter of cloud D_b -diameter of bubble)

CHAPTER 3

GAS EXCHANGE BETWEEN BUBBLES AND DENSE PHASE IN THREE DIMENSIONAL BEDS

3.1 INTRODUCTION

Fluidised beds in large industrial applications are operated at flow rates where aggregative fluidisation prevails. In these conditions bubbles exist and short-circuiting of gas through bubbles, without contacting the particles, should occur.

A number of mathematical models have been presented (11, 36, 40, 55, 58, 59) in order to predict conversion efficiency in fluidised bed reactors. Most of these models are based on the two phase concept of fluidisation, postulated by Toomey and Johnstone (76). They differ mainly in the assumptions regarding gas exchange between the two phases.

The extent of gas exchange between the two phases has been the subject of both theoretical and experimental work and this chapter is devoted to the analysis of such published investigations.

3.2 THEORETICAL MODELS FOR GAS-TRANSFER BETWEEN BUBBLES AND DENSE PHASE

3.2.1 Davidson and Harrison Model

A - Flow of gas through the bubble

In the foregoing section it has been shown that potential flow describes both the flow of gas and solids around a spherical bubble. From this analysis two distinct gas flow patterns result, when the ratio of bubble rising velocity to the interstitial gas velocity in the dense phase is less than, or greater than, unity. When $U_b/V_o < 1$ gas percolates through the bubble and when $U_b/V_o > 1$

there is a circulation of gas between the bubble and the cloud. Pyle and Rose (60) obtained a stream function for gas flow within a spherical bubble and the flow patterns within the bubble are described in figure 2.4 for different ratios U_b/V_o .

Davidson and Harrison (17) demonstrated that the upward gas flow rate across the equatorial plane of a spherical bubble of radius a is

$$Q_t = 3 U_{mf} \pi a^2 \quad (3.1)$$

although, previously, Zenz (84) suggested the value for Q_t was given by

$$Q_t = 2 U_{mf} \pi a^2 \quad (3.2)$$

As explained above, the mode of gas flow through the bubbles as given by equation (3.1) is strongly dependent upon the ratio U_b/V_o . For a stationary bubble all the gas flowing through the bubble percolates from the dense phase. When U_b/V_o has values between 0 and 1 some of the gas flowing through the bubble is fresh gas coming in from the dense phase and some is recirculating gas. For U_b/V_o greater than 1 all the gas flow through the bubble is due to the gas circulating between bubble and dense phase.

B - Gas exchange due to through-flow

The above analysis led Davidson and Harrison (17) to propose a convective mechanism for gas exchange between real bubbles and dense phase.

The rate of gas exchange is given by the equation

$$Q_t = \frac{3}{4} U_{mf} \pi D_e^2 \quad (3.3)$$

where $2a$ is replaced by D_e for a real bubble. D_e is the equivalent diameter of a sphere having the same volume as the bubble.

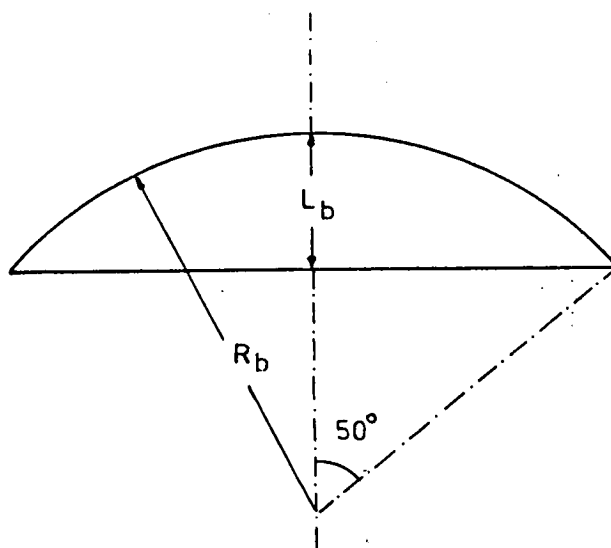
This equation is certainly valid for a stationary bubble, since all the gas flowing through the bubble is fresh gas coming from the dense phase. For fast moving bubbles, as is seen in figure 2.4, this analysis is rather insubstantial since the through-flow of gas is due to circulation between bubble and cloud, and no mixing exists across the cloud boundary between the circulating gas and that in the dense phase, as results from the potential flow model of Davidson. Its application to fast moving bubbles can only be accepted if gas in the cloud is assumed to be in streamline flow until it reaches the wake level, where complete mixing with the dense phase occurs before re-entering the bubble.

C - Gas exchange due to diffusion across the bubble boundary

Davidson and Harrison (17) also considered that, simultaneous to the above convective gas transfer contribution, a diffusional contribution exists. They then estimated the rate of gas exchange by diffusion across the bubble boundary by analogy with diffusion across the frontal surface of large bubbles rising in liquids.

Thus, considering a well mixed bubble with a spherical cap shape, having a flat base and a 50° angle between its rim and polar axis (figure 3.1), and assuming all the resistance to mass transfer is within the bubble, the following expression for the mass transfer coefficient was obtained

$$K_g = 0.975 D_g^{\frac{1}{2}} D_e^{-\frac{1}{4}} g^{\frac{1}{4}} \quad (3.4)$$



Bubble volume $V_b = \frac{\pi}{3} L_b (3 L_b R_b - L_b)$

Bubble frontal surface area $S_b = 2 \pi R_b L_b$

Figure 3.1 Idealised bubble shape with a flat base and a 50° angle between its rim and polar axis.

The mass transfer coefficient K_g in equation (3.4) is defined in terms of the surface area of an equivalent sphere of diameter D_e having the same volume as the bubble; D_g is the gas phase diffusion coefficient.

The rate of gas transfer by molecular diffusion is then

$$Q_d = K_g S_t \quad (3.5)$$

Since

$$S_t = \pi D_e^2 \quad (3.6)$$

Equation (3.5) becomes

$$Q_d = 0.975 \pi D_g^{\frac{1}{2}} D_e^{\frac{7}{4}} g^{\frac{1}{4}} \quad (3.7)$$

The above equation (3.5) has been previously obtained by Baird and Davidson (3) for liquid-film controlled diffusion for a rising bubble.

For the bubbles in fluidised beds, it is note worthy that the above equation gives only an estimate for diffusional gas transfer from a spherical bubble since it considers neither gas transfer from the rear of the bubble (9), which certainly increases the rate of mass transfer, nor the existence of a concentration gradient in the shell surrounding the bubble, which has an opposite effect (41).

D - Overall mass transfer rate between the bubble and dense phase

As explained above, Davidson and Harrison (17) considered that both mechanisms, through-flow and molecular diffusion, are independent and simultaneous. The overall mass transfer rate being then

$$Q = Q_t + Q_d \quad (3.8)$$

$$Q = \frac{3}{4} U_{mf} \pi D_e^2 + 0.975 \pi D_g^{\frac{1}{2}} D_e^{\frac{7}{4}} g^{\frac{1}{4}} \quad (3.9)$$

The rate of gas exchange per unit volume of bubble is

$$K_v = 4.5 \frac{U_{mf}}{D_e} + 5.85 D_g^{\frac{1}{2}} D_e^{-\frac{5}{4}} g^{\frac{1}{4}} \quad (3.10)$$

3.2.2 Kunii and Levenspiel model

A - In order to solve some of the difficulties of Davidson's model in the case of fast moving bubbles, as explained in section 3.2.1-B, Kunii and Levenspiel (41) presented a model for gas transfer between the bubble-cloud system and the dense phase. They considered that the above two contributions for gas transfer between bubble and dense phase only account for gas exchange between the bubble void and bubble cloud. They further proposed that gas is also exchanged between the bubble cloud and dense phase.

B - Because there is no gas flow across the cloud/dense phase interface they assumed that the only mechanism for gas exchange between these two regions is by molecular diffusion. They then estimated the coefficient of gas exchange between cloud and dense phase on the basis of the surface-renewal model of Higbie (31), which supposes that unsteady diffusion occurs for the time the cloud takes to rise its own diameter, t_c ,

$$K_c = \left[\frac{4 D_g \epsilon_{mf}}{\pi t_c} \right]^{\frac{1}{2}} \quad (3.11)$$

where D_c is the cloud diameter. The contact time t_c is then

$$t_c = \frac{D_c}{U_b - V_o} \quad (3.12)$$

Considering now the cloud dimensions, as given by the Davidson and Harrison model. (equation 2.13),

$$A^3 = a^3 \left[\frac{U_b + 2V_o}{U_b - V_o} \right] \quad (3.13)$$

then

$$D_c = 2 A \quad (3.14)$$

and

$$D_e = 2 a \quad (3.15)$$

the above equation (3.11) becomes

$$K_c = 1.13 \left[\frac{D_g \epsilon_{mf} (U_b - V_o)}{D_e} \right]^{\frac{1}{2}} \left[\frac{U_b - V_o}{U_b + 2V_o} \right]^{\frac{1}{6}} \quad (3.16)$$

The mass transfer rate due to diffusion across the cloud/dense phase interface per unit of bubble void is

$$K_{cd} = \frac{K_c \cdot S_c}{V_b} \quad (3.17)$$

where S_c is the cloud surface area and V_b the bubble void volume.

From the above equations K_{cd} reduces simply to

$$K_{cd} = 6.78 \left[\frac{D_g \epsilon_{mf} U_b}{D_e^3} \right]^{\frac{1}{2}} \quad (3.18)$$

for

$$U_b \gg V_o \quad (3.19)$$

C - Kunii and Levenspiel (41) considered mass transfer between the bubble and the dense phase to be divided into two steps, with mass transfer coefficients

$$K_{bc} = 4.5 \frac{U_{mf}}{D_e} + 5.85 D_g^{\frac{1}{2}} D_e^{-\frac{5}{4}} g^{\frac{1}{4}} \quad (3.20)$$

for the transfer between bubble and cloud, and

$$K_{cd} = 6.78 \left[\frac{D_g \epsilon_{mf} U_b}{D_e^3} \right]^{\frac{1}{2}} \quad (3.21)$$

For the transfer between cloud and dense phase. The overall mass transfer coefficient between bubble and dense phase, K_{bd} , will be

$$\frac{1}{K_{bd}} = \frac{1}{K_{bc}} + \frac{1}{K_{cd}} \quad (3.22)$$

3.2.3 Chiba and Kobayashi model

In Kunii and Levenspiel's model, cloud composition is considered as different from bubble composition. Chiba and Kobayashi (13), analysing the above model, assumed that bubble and cloud composition would be essentially the same. Therefore, in the above equation (3.22), the first term in the right hand side is not significant as compared with the second term. They then proposed a model considering that gas transfer between the bubble-cloud system

and the dense phase takes place only by molecular diffusion across the spherical boundary of the cloud.

Their analysis for gas diffusion is similar to that of Davidson (section 3.2.1-C) applied to a sphere and assuming gas and particle flow around the bubble as described by Murray's stream function (52).

They obtained the following transfer coefficient per unit of bubble void between the bubble-cloud system and the dense phase

$$K_v = \frac{6.76}{1-f_w} \left[\frac{D_g \epsilon_{mf}^2 U_b}{D_b^3} \right]^{\frac{1}{2}} \quad (3.23)$$

where f_w is the volume fraction of the wake and D_b is the bubble diameter.

$$\frac{\pi D_e^3}{6} = (1 - f_w) \frac{\pi D_b^3}{6} \quad (3.24)$$

3.2.4 Partridge and Rowe Model

Partridge and Rowe concluded from their experiments that the bubble is surrounded by a spherical cloud (58) and that gas exchange takes place by a phenomenon of cloud shedding (70). The rate of gas exchange is the same order of magnitude as might be expected from diffusive exchange across the bubble boundary.

Thus, they proposed a model in which the spherical cloud has a volume of gas with uniform composition, completely mixed with the gas in the bubble void, which is rising through another gas in largely streamline flow. From a mass transfer point of view, the process of gas transfer between cloud and dense phase is somewhat

similar to the mass transfer between a drop of an immiscible liquid rising through another, or the transfer from a solid sphere in an extensive flowing fluid.

The mass transfer coefficient is then

$$h' = \frac{Sh_c D_g}{D_c} \quad (3.25)$$

Sh_c is the Sherwood number for transfer from a sphere with a diameter D_c and uniform surface composition in the Reynolds number (Re) range of 30 - 2000

$$Sh_c = 2 + 0.69 Sc^{\frac{1}{3}} Re^{\frac{1}{2}} \quad (3.26)$$

Sc is the Schmidt number

$$Sc = \frac{\mu}{\rho D_g} \quad (3.27)$$

$$Re = \frac{U_r \rho D_c}{\mu} \quad (3.28)$$

D_c is the diameter of the cloud, μ and ρ the viscosity and density of the gas, D_g the gas diffusivity and U_r the relative velocity between cloud and interstitial gas

$$U_r = U_b - V_o \quad (3.29)$$

3.3 EXPERIMENTAL MEASUREMENTS OF GAS TRANSFER BETWEEN BUBBLES AND DENSE PHASE - LITERATURE SURVEY

3.3.1 Davidson and Harrison (17) calculated the equivalent diameter,

D_e , of bubbles from the fraction of reactant by-passing the bed for various catalytic reactions, using the experimental data of several workers (45, 47, 49, 54, 56), assuming that the rate of gas exchange is given by equation (3.9). Then, comparing the above computed values of D_e with values of D_e estimated by other methods, they concluded from the agreement that the rate of gas exchange between the two phases in a fluidised bed is well represented by their model, equation (3.9).

3.3.2 Székely (74) pioneered the direct experimental determination of gas transfer between bubbles and dense phase. The experiments were conducted by injecting single bubbles of known volume, containing a known concentration of carbon tetrachloride, into an incipiently air-fluidised bed of highly adsorbent silica alumina catalyst.

The experimental results were expressed in terms of a transfer coefficient assuming that the area for gas transfer between the bubble and the dense phase is the curved area of a spherical cap having a flat base and a 50° angle between its rim and polar axis, figure 3.1.

Szekely (74) found little gas by-passing and suggested that most of the transfer takes place during bubble formation. He also found that the mass transfer coefficient decreases with increasing size of bubbles.

Davidson and Harrison (17) compared the above experimentally determined mass transfer coefficients with those predicted from equation (3.9), and found that gas by-passing, as measured by Székely (74), is much less than predicted (up to 33% less) in most of the

experiments; the most likely explanation for such a difference, as pointed out by Davidson and Harrison (17), being due to end effects.

3.3.3 Davies and Richardson (20) injected bubbles of CO_2 continuously at a steady frequency into air-fluidised beds of non-adsorbent particles. They found that bubbles increase in size as they rise through the bed, when injected into beds operating at gas velocities above the incipient fluidisation velocity. Their mass transfer experiments were carried out, however, under minimum fluidisation conditions for which bubble volume remains constant.

They also found that the steady injection of bubbles in the bed produces a flat radial profile of concentration in the dense phase, and that the concentration over the cross section increases with the bed height.

The rate of gas transfer between bubbles and dense phase was then obtained by sampling the gas in the dense phase at different levels outside the bubble path. If dQ/dh is the rate of gas exchange between the bubble and dense phase, then

$$\frac{dQ}{dh} = \frac{V_b}{h - h_0} \cdot \frac{F_1}{F_1 + F_2} \cdot \ln \left[\frac{1}{1 + \frac{F_1 + F_2}{F_2} C_h} \right] \quad (3.30)$$

where V_b is the bubble volume, F_1 and F_2 the bed and bubble flow rates, h_0 an integration constant corresponding to the height above the injector at which transfer can be regarded as commencing, h is the height above the injector where the concentration in the dense phase is C_h . They then determined a gas transfer coefficient between bubble and dense phase which they expressed in terms of an

equivalent gas velocity, v_x , through the curved surface of a spherical cap having a 50° angle between its rim and polar axis.

The results are of the same order of magnitude as those of Szekely (74), although they found an opposite variation of mass transfer coefficient with bubble size. In table I, the results of Davies and Richardson (20) and Szekely (74) are represented versus D_e . The Szekely data were taken from figure 5 of the published work and correspond to the case in which $z - D_e$ is a maximum; z being the bed height.

TABLE I

D_e (cm)	Szekely (74) h_d (cm/sec)	Davies and Richardson (20) v_x (cm/sec)
3.85	2.21	-
4.85	2.17	-
5.28	-	2.12
5.34	2.00	-
6.39	-	2.05
7.17	-	2.30
7.79	-	2.34

In Table II, Davies and Richardson's (20) results are compared with the prediction from Davidson's model (equation 3.4), since they used particles of low incipient fluidisation velocity. As $U_{mf} = 0.142$ cm/sec and, from Davidson's model, the diffusional term predominates over the convective term

TABLE II

D_e (cm)	Davies and Richardson (74) v_x (cm/sec)	Davidson model (20) K_g (cm/sec)
5.28	2.12	1.26
6.39	2.05	1.21
7.17	2.30	1.17
7.79	2.34	1.13

Equation (3.4) suggests that the transfer coefficient decreases with increasing bubble size. Tables I and II show that the trend detected by Szekely (74) agrees with the predicted trend but Davies and Richardson (20) show an opposite effect.

The authors (20) also observed that the rate of gas transfer decreased with increasing particle size. Whereas, equation (3.4) is insensitive to variation in particle size, equation (3.9) shows that the rate of transfer will increase with particle size for bubbles with the same volume. This is due to the increase in through-flow across the bubble, which increases with particle size, as shown from equation (3.3), since U_{mf} increases with increasing particle size.

They also found that the bubble has to travel some distance above the injection orifice before transfer commenced, that is they found that the integration constant, H_o , in equation (3.30), has positive values. This finding is rather surprising when compared with Szekely's work (74) which found that most of the gas transfer takes place as bubbles are being formed.

3.3.4 Stephens, Sinclair and Potter (72) continuously injected air containing mercury vapour as a tracer into fluidised beds maintained at incipient fluidisation by tracer-free air. A stream of bubbles containing tracer was produced and, in order to measure the rate of gas exchange between bubbles and dense phase, the gas in the dense phase was sampled. They used particles with sizes ranging from 50 to 380 μm , and found that for the larger particles (290 μm) marked radial concentration gradients exist, even at 14 inches above the injector with a peak in the profile at the axis. For the smaller particles a uniform radial concentration was readily obtained at a distance 3 to 5 inches above the injector. They found that the mean concentration over a cross section increases with bed height in all cases, as did Davies and Richardson (20) who used particles of lower incipient fluidisation velocity.

They then calculated, for each run, the "bubble diameter" in the bed from the "bubble-breaking" frequency on the top of the bed. Assuming the two phase flow model, they calculated gas transfer coefficients between bubbles and dense phase using the axial profiles of mean tracer concentration over the cross section of the bed.

They compared their results with equation (3.3) and (3.9) from the Davidson and Harrison model (17) and equation (3.2) proposed by Zenz (84). They showed that for the smaller particles, with low incipient gas velocity, their data compares well with equation (3.9), the diffusional term predominating over the convective term. For the larger particles they found that Zenz's equation fits their data better than the equations due to Davidson and Harrison which predict higher gas transfer rates than those experimentally determined.

From their experiments good evidence of two different

contributions for gas transfer can be visualised, as proposed by Davidson and Harrison (17). Gas transfer, in the case of large particle beds, is mainly due to through-flow across the bubble and due to diffusion in the case of small particle beds.

Although such a conclusion can be drawn from the comparison of their data with equations (3.2), (3.3) and (3.9), it should be treated with care. Because the bubble diameter in the bed was obtained from the gas flow rate and visual determination of "bubble breaking" frequency on the top of the bed, no account is taken of coalescence of bubbles near the injection point and, consequently, the existence of small bubbles near the injector, which have different gas transfer rates than those bubbles on the top, is ignored.

3.3.5 Chiba and Kobayashi (13) injected, at known frequencies, bubbles containing ozone as a tracer in 10 cm diameter air fluidised beds of glass particles. Bubble size, frequency and velocity was determined using an opto-electronic detector which was inserted in the bed. Tracer concentration within the bubble was measured using a spectrophotometric technique. The detector was made out of two glass tubes fitted with quartz windows at their tips. By placing the two tubes in the bed from opposite sides of the column and leaving a distance of 1 to 2 cm between their tips, a cell for UV absorption was created. The composition of the gas forming the bubbles was continuously monitored before its insertion in the bed. Gas composition within the bubble was measured at four different levels above the injector.

From a knowledge of the above variables, gas transfer

between bubbles and dense phase was calculated. Further, a comparison between the experimental results and the model presented by these authors (see section 3.2.3) shows that the experimental data compare well with the prediction from the model.

The experimental set-up, although ingenious, did not account for the showering of particles within the bubble (64) and so the concentration measurement of ozone within the bubble was subject to errors. Chavarie and Grace (12), using a similar technique in a two-dimensional bed, overcame such a problem by using a dual amplifier system and a reference beam not absorbed by any of the gaseous components in the bed. Also, bubble size was inferred from a bubble length measurement but because bubbles can pass the probe off-centre, bubble size cannot be unambiguously inferred from a bubble chord measurement.

In order to eliminate possible erroneous data from their experiments, Chiba and Kobayashi (13) used only the best 4 or 5 out of 300 signals for the measurement of concentration of tracer, although no criterion for the discrimination of "bad" signals was given and the selection of the 4 or 5 best signals out of 300 seems to be quite arbitrary, thus jeopardizing the reliability of their findings.

3.3.6 Drinkenburg and Rietema (23), (24) injected single bubbles of air containing hydrocarbons (methane, ethane, ethene and propane) as tracer gas, in incipiently-fluidised beds of cracking catalyst free of tracer gas. Bubble velocity was calculated from equation (2.1), D_e being obtained from the measurement of the bubble volume using a water displacement technique.

Gas concentration within the bubble was detected at different levels above the distributor. Gas was drawn continuously through a capillary tube and analysed by flame ionization. When a traced bubble strikes the sample probe, a higher concentration is detected.

Mass transfer coefficients, K^* , were then calculated assuming spherical bubbles having the measured volume

$$-U_b \frac{d(V_b C_b)}{dh} = K^* S_t V_b (C_b - C_d) \quad (3.31)$$

They found that the interchange coefficient increases with increasing bubble diameter, except only for propane, where a maximum for the mass transfer coefficient was found.

Simultaneously with their experimental work, they proposed a numerical model for the computation of the transfer coefficients, which predicts their decrease with increase in bubble diameter. The experimentally determined mass transfer coefficients were always much higher than those predicted by the model with the exception of propane and then only for very high values of D_e ($D_e \approx 15$ cm).

Their results are difficult to interpret, due to the bubble shape, as explained in chapter 2, and also due to gas adsorption on the solids.

3.3.7 Rietema and Hoeibink, (62) used a similar technique to that of Drinkenburg and Rietema (24) both for bubble injection and gas concentration detection using only ethene as a tracer. Bubble velocity and size were measured at the same level as the gas detector probe, using Wherter and Molerus (80) capacitance probes,

which consist of two capacitances vertically separated, thus permitting the calculation of local bubble velocity and a vertical chord of the bubble.

The use of such a probe arrangement was discussed by Burgess and Calderbank (6) (7). This measurement of bubble size and velocity is of doubtful reliability. Due to the fact that a vertical chord is used to estimate the bubble size and also due to the fact that a lateral movement of the bubble between the making of the leading contact and the making of the upper contact, severe errors will be introduced in the estimation of the bubble local velocity, which in turn will affect the calculation of the detected vertical chord.

In order to overcome such difficulties Rietema and Hoebink (62) used only those bubbles whose local velocity was the same as that obtained by calculating the mean velocity between injection and detection points.

Bubble volume was calculated from the rising velocity according to the equation

$$U_b = 0.71 \sqrt{g v_b^{\frac{1}{3}}} \quad (3.32)$$

In order to calculate mass transfer rates the authors used the same method as Drinkenburg and Rietema (24), (equation 3.31). Although now, bubbles were assumed to have a spherical cap shape, as obtained by Rowe and Partridge (69), S_t being the total surface area of the bubble in equation (3.31).

They discussed different processes for gas transfer between bubble and dense phase, such as diffusive and convective mass transfer via the cloud-bubble boundary; diffusive transfer via the cloud-dense phase boundary; transfer via cloud shedding; transfer

to solids in the cloud and solids raining through a bubble in the case of porous adsorbing solids, but they assumed and calculated that all these processes are part of a lumped or overall mass transfer coefficient, which they found to increase with increasing bubble diameter.

Their results are rather inconclusive and for the same tracer, ethene, they differ from Drinkenburg's results (23) (24).

3.4 CONCLUSION

As presented above, while different models exist for predicting the mass transfer rate between bubbles and dense phase in three-dimensional fluidised beds, experimental work obtained by several authors yields rates of mass transfer which do not permit sufficient discrimination between the existing models.

The evaluation of such models through direct measurement of the rate of gas transfer between the two phases requires a knowledge of bubble parameters such as volume, surface area and velocity, as well as the concentration of the tracer in both phases and its time derivative.

In the above discussed experimental work, none of the authors were able to measure simultaneously all the mentioned variables.

The present work was undertaken to develop a probe technique for the simultaneous measurement of all variables which are relevant to the computation of mass transfer coefficient between bubbles and dense phase.

In the following chapters such a technique is presented and discussed together with experimental determination of mass transfer rates between bubbles and dense phase.

CHAPTER 4

DESIGN AND DEVELOPMENT OF A
PROBE TECHNIQUE FOR GAS TRANSFER
STUDIES BETWEEN BUBBLES AND
DENSE PHASE

4.1 INTRODUCTION

In the past, several different probe techniques for the detection of bubble velocity and size in three dimensional fluidised beds have been reported. Optical probes were reported by Yasui and Johanson (83) and Whitehead and Young (82); capacitance probes were used by Morse and Ballou (50), Angelino (1), Lanneau (43), Bakker (4), Fukuda et al. (25), Kunii et al. (42), Geldart and Kelsey (26) and Werther and Molerus (80). Conductivity probes were used by Park et al. (57) and Rigby et al. (63).

A comprehensive review of the above probe techniques and their limitations in the correct prediction of bubble velocity, size and shape in three dimensional fluidised beds, is given by Burgess and Calderbank (6) (7) (8), who proposed a three dimensional resistivity probe with five channels, which was coupled to a high speed digital computer with analogue-digital conversion facilities (PDP-8e Digital Corporation).

In the previous chapter, the techniques presented for the measurement of mass transfer rates between bubbles and dense phase were discussed and none of them was able to discriminate between the existing models for mass transfer.

The direct measurement of the rate of gas transfer between the phases requires knowledge of bubble parameters such as volume, surface area and velocity as well as the concentration of gas in both phases and the time derivative of the latter. The techniques described in the previous chapter failed to measure simultaneously all the relevant variables and are based on one or other model of bubble behaviour in order to estimate those parameters not measured directly.

4.2 PROBE DESIGN

Design formulation

In order to overcome the problems of the previous technique when measuring rates of gas transfer between bubbles and dense phase in three dimensional beds, the following objectives must be upheld in the probe design.

1. The accurate measurement of bubble parameters: velocity, shape and size.
2. The measurement of gas composition within the bubble and in the dense phase nearby as the bubble undergoes mass transfer with its surroundings.
3. The probe should permit the discrimination of gas when sampled from the bubble phase and from the dense phase around it.

4.3 BUBBLE PARAMETERS

In order to obtain the bubble parameters in a three dimensional fluidised bed it was decided to use the simple and elegant method proposed by Burgess and Calderbank, which permits the accurate description of bubble size and velocity in fluidised beds of conducting particles.

Although this technique is described elsewhere (6) (7) (8) a brief summary is presented below.

The probe is composed of an array of five contacts. The contacts are connected so that each contact forms an electrical circuit whereby current flows from an external DC power supply. Whilst the contact is resident in the dense phase which is chosen to be conducting (vide infra) a voltage is developed and when a bubble

approaches the probe and breaks the contacts the voltage becomes zero. Pulses are obtained as the circuits are broken and made by the passage of a bubble through the array.

Figure 4.1 shows the probe and the contacts arrangement as used by Burgess and Calderbank (6) (7) (8).

A leading contact (C1) senses the arrival of a rising bubble at the probe. In a plane above and symmetrically ordered around the leading contact are three contacts (C2, C3, C4) which serve a discriminating and velocity measuring function. The discrimination is achieved by requiring the near-coincidence of the probe and bubble centre lines within given limits, and is obtained in practice by comparing the delay times as positive going pulses move from the leading contact to the upper contacts and fixing the ratio minimum delay - maximum delay between 0.85 and 1.

In order to ensure that the bubble has no lateral movement when in contact with the probe a second discrimination function was invoked by requiring that the duration of the pulses on the three upper contacts were nearly identical, such that the ratio minimum-maximum pulse duration was between 0.70 and 1.

When these criteria are satisfied the data are stored for computer processing as below.

Bubble velocity and central length

For a successful bubble encounter the bubble rising velocity, U_b , is then given by

$$U_b = \frac{3 \cdot S_d}{T_{12} + T_{13} + T_{14}} \quad (4.1)$$

where S_d is the vertical separation distance between the leading contact and the three upper contacts, and T_{12} , T_{13} and T_{14} are the pulse delay times between the leading contact and the three upper

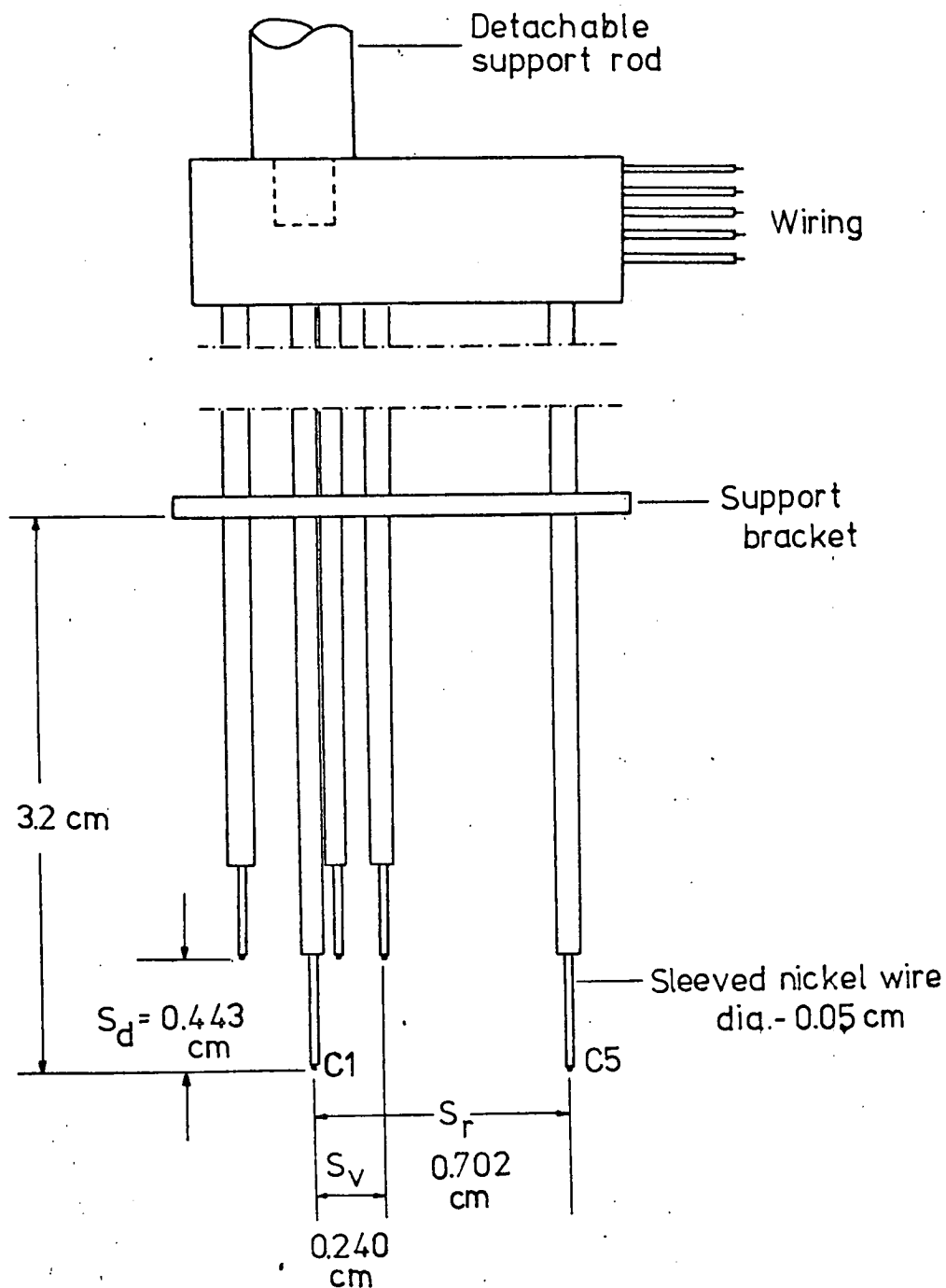


Figure 4.1

Schematic diagram of the bubble detector probe reported by Burgess and Calderbank (6) (7).

contacts. The bubble central length, L_b , is obtained from the bubble velocity and the pulse duration on the leading channel, T_1

$$L_b = U_b \cdot T_1 \quad (4.2)$$

Bubble nose curvature and shape

Since the bubble hits the probe successfully its axis will be almost coincident with the leading contact. If a fifth contact (C5) is placed radially distant from the leading one and, at the same level, the delay time between the arrival of the bubble to the leading channel and the arrival to the off-centre channel will permit determination of the vertical distance, L_d , between the bubble nose and bubble interface at the radial distance, S_r , where S_r is the distance between C1 and C5,

$$L_d = U_b \cdot T_{15} \quad (4.3)$$

T_{15} being the delay time.

Now, simple trigonometric considerations permit one to determine the radius of curvature of the bubble nose, R_b

$$R_b = \frac{S_r^2 + L_d^2}{2 L_d} \quad (4.4)$$

Also, the pulse duration on the fifth contacts T_5 , permits the determination of an off-centre vertical chord length of the bubble,

L_o

$$L_o = U_b \cdot T_5 \quad (4.5)$$

The measurement of bubble axial height, off-centre chord length and nose curvature radius, as described above, are certainly insufficient to give an accurate description of the bubble shape. A complete picture could only be obtained using a whole ensemble of off-centre contacts. This situation, although ideal, will certainly make the probe a bulky one, damaging severely the bubble, and the provision of extra channels would make the data acquisition too slow for accurate measurements, specially in the case of fast moving bubbles.

Experimentally determined bubble shapes can only be obtained from the probe output if some ideal shapes are related with the probe output.

Bubble shapes have been previously determined by Rowe and Partridge (69) who, using x-ray photography, found that single bubbles injected in three dimensional beds, have a spherical frontal surface and a spherical cap shape with an indented rear surface. Also, in a given fluidised bed when bubble size increases in such a way that wall effects begin to be important a transition from spherical cap shaped bubbles to elongated bubbles in a vertical direction is obtained, as was found by Goldsmith and Rowe (27) using photography. Due to the shape description given by Goldsmith and Rowe it was decided to treat this shape as a parabolic one.

In Appendix A the relations between probe output and proposed shape is discussed, which are

Spherical cap	$L_b < R_b$
Hemisphere	$L_b = R_b$
Low eccentricity cap	$L_b > R_b$ and $L_b < 2.R_b$
Sphere	$L_b = 2.R_b$ and $L_d + L_o < 2.R_b$

Paraboloid

$$L_b \geq 2.R_b \text{ and } L_d + L_o \geq 2.R_b$$

In figure 4.2 is shown schematically the probe recorded parameters and the criteria of shape selection.

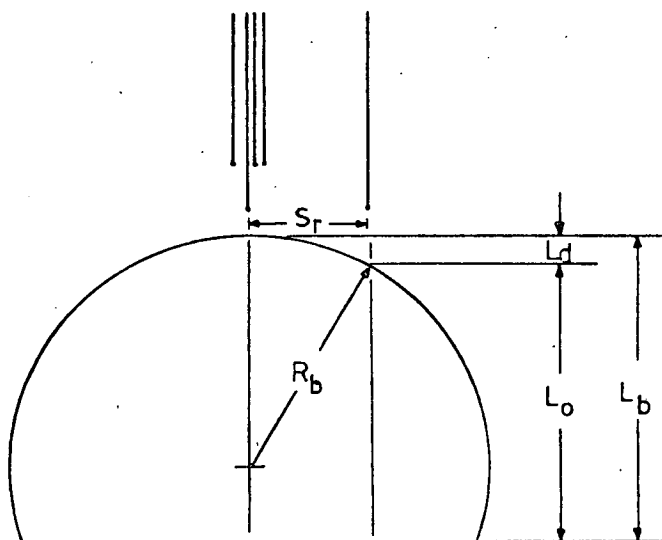
Bubble volume and surface area

Bubble volume and surface area were calculated using the equation for each one of the solids of revolution corresponding to the proposed shape. The mathematical formulation is also described in Appendix A.

4.4 GAS COMPOSITION DETECTION

Since the main objective of this work is the study of mass transfer between bubbles and dense phase, the gas composition in both phases should be accurately detected, and gas composition should be unambiguously obtained when gas is sampled from the dense phase or from the bubble phase. Also, detailed information of gas composition profiles within the bubble and in its surroundings have not yet been published, and such detailed information is necessary to assess some of the assumptions made by the various authors when proposing their models, as seen in chapter 3. Provision for this information should be made in the design of the gas sensor and detection technique.

An ideal gas sensor would be a thermal conductivity detector. Such a sensor implies the use of a tracer with a different thermal conductivity from that of the fluidising gas. The injection of bubbles containing helium as a tracer in an air fluidised bed would



$$R_b = \frac{S_r^2 + L_d^2}{2 \cdot L_d}$$

Spherical cap	$L_b < R_b$
Hemisphere	$L_b = R_b$
Low eccentricity cap	$L_b > R_b$ and $L_b < 2 \cdot R_b$
Sphere	$L_b = 2 \cdot R_b$ and $L_d + L_o < 2 \cdot R_b$
Paraboloid	$L_b \geq 2 \cdot R_b$ and $L_d + L_o \geq 2 \cdot R_b$

Figure 4.2 Probe recorded parameter and criteria of shape selection.

then be a suitable gas system for our purpose. In table 3.1 the thermal conductivity of helium and air at 20°C are compared.

Table 3.1

gas	$k - \left[\text{cal/sec cm}^2 \left(^\circ\text{C/cm} \right) \right] \cdot 10^{-6}$
Helium	333.50
Air	56.24

The ideality of the thermal conductivity sensor is due to its miniturisation, its high frequency of response and sensitivity.

Detector design

The mechanical features of the thermal conductivity detector to be used in our studies are shown in figure 4.3.

A hot-wire detector is placed within a chamber, diameter 0.22 cm, at the tip of a perspex tube, diameter (O.D.) 0.48 cm, which is terminated with a sintered plate.

The gas to be analysed was drawn from the fluidised bed through the sintered plate of brass, 0.02 cm thick with 25 μm pore diameter, and passed over the hot-wire detector which is 0.5 - 1 mm distant from the sintered plate. The hot-wire sensor is made out of 90% Pt - 10% Ir wire, $1.27 \cdot 10^{-3}$ cm thick and 0.18 cm long, and was supported between two needles fixed to a mount which was interfaced with a DISA constant temperature hot-wire anemometer bridge (CTHWB). The plastic cap in which the detector and mount was encapsulated was screwed into a vertical supporting tube.

In order to sample the gas from the fluidised bed the above supporting tube was connected to a vacuum pump.

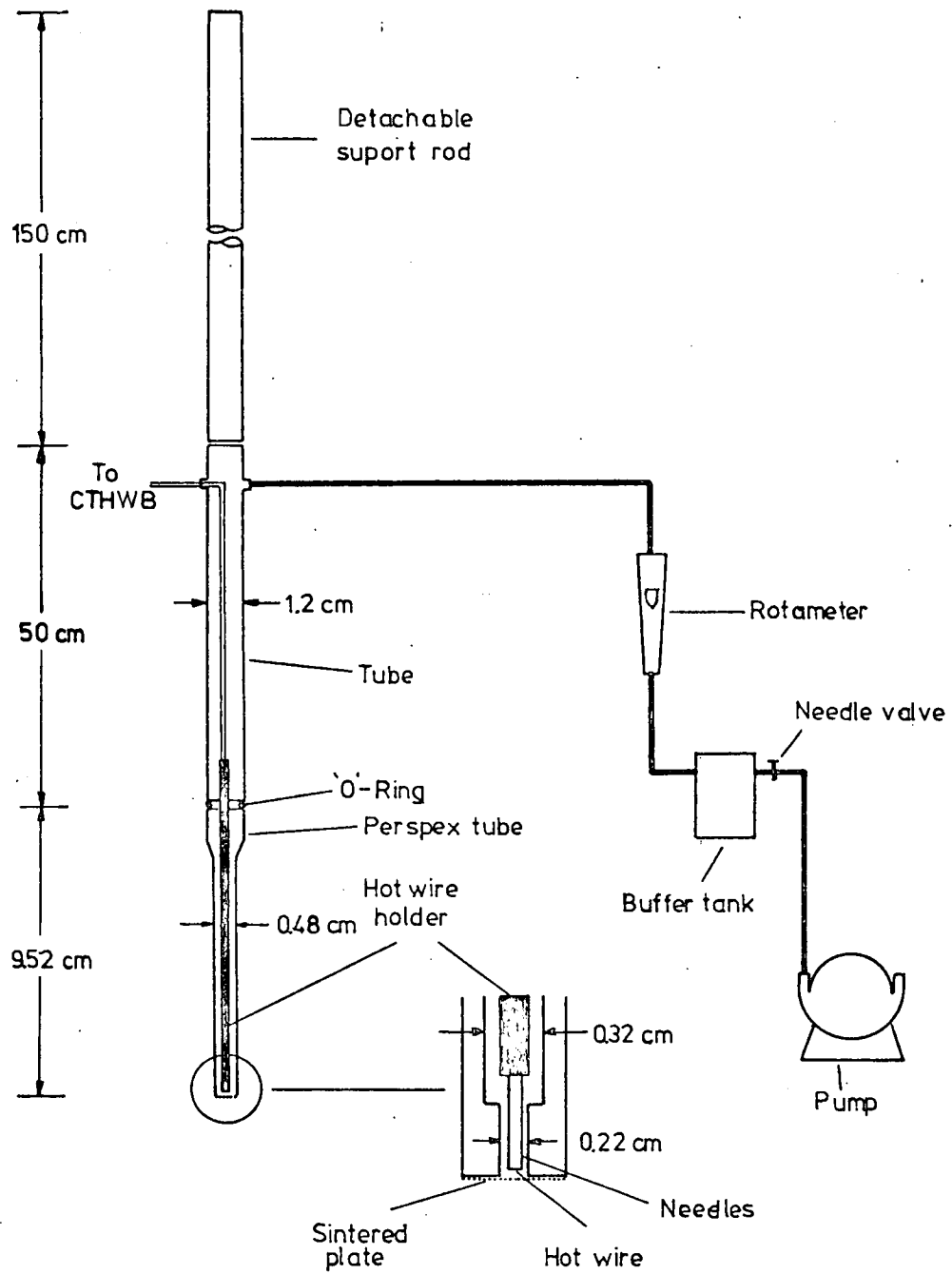


Figure 4.3

Schematic diagram of the thermal conductivity probe assembly.



As is shown in Appendix B (detector calibration and testing), the hot-wire anemometer is sensitive to variations of the thermal conductivity of the gas as well as to variations of the gas velocity around it. In order to ensure a constant rate of gas sampling a rotameter was placed between the supporting tube and the vacuum pump. To eliminate gas sampling oscillations, a buffer tank, volume 650 cm³, was introduced between the rotameter and the pump, as shown also in figure 4.3.

4.5. THE COMPOUND PROBE

The probe for bubble parameter detection was described in Section 4.3 and in 4.4 the gas composition detector head.

As is shown in figure 4.1, the bubble parameter sensor is comprised of five contacts. If the gas detector head is placed coaxially and at the same level between the three upper contacts (velocity contacts) a gas composition discriminator is obtained if the DISA - CTHWB output is related with the pulsing mode of these three upper contacts.

Let us assume a bubble is rising in the bed and approaching the probe. When the bubble interface of the approaching bubble envelops the leading contact the three upper contacts are still in dense phase mode. As gas is sampled continuously, the gas composition in the dense phase ahead of the bubble is obtained if the DISA - CTHWB output is recorded during the time that it takes the bubble to rise from the leading contacts to the level of the three upper contacts. Then the three upper contacts will be in bubble phase mode and bubble gas composition will be obtained until

these contacts return to the dense phase mode, when gas composition in the dense phase behind the bubble will be obtained.

In figure 4.4 a detailed diagram of the compound probe and its dimensions is presented.

4.6 COMPOUND PROBE OUTPUT ANALYSIS

The analysis of the compound probe output and the discrimination function referred to in 4.3 and 4.5 contain a logical sequence of decision criteria which are undertaken in real time, and which should be made very rapidly since events are very fast and small time durations are examined.

Such difficulties are overcome by the use of a computer (PDP-8e, 16K, 12 bit work computer, from Digital Corporation), which permits a rapid and accurate conversion of analog voltage signals to its discrete digital representation, which takes typically 26.2 microseconds and requires that the analog voltage must be between 0 and 1 volt.

The interfacing arrangement between the five contacts of the bubble parameter detector and the computer is shown in figure 4.5. The voltage pulses on these five channels were adjusted to between 0 and 1 volt by means of five variable resistors, R.

As the CTHWB has a voltage output always higher than 1 volt; even for pure air, and it will increase with increasing concentration of helium as the hot-wire tends to cool, it was necessary to condition the CTHWB output signal in order to suit the characteristics of the analog-digital converter of the computer. This was done by means of an off-set computer line matching circuit with variable

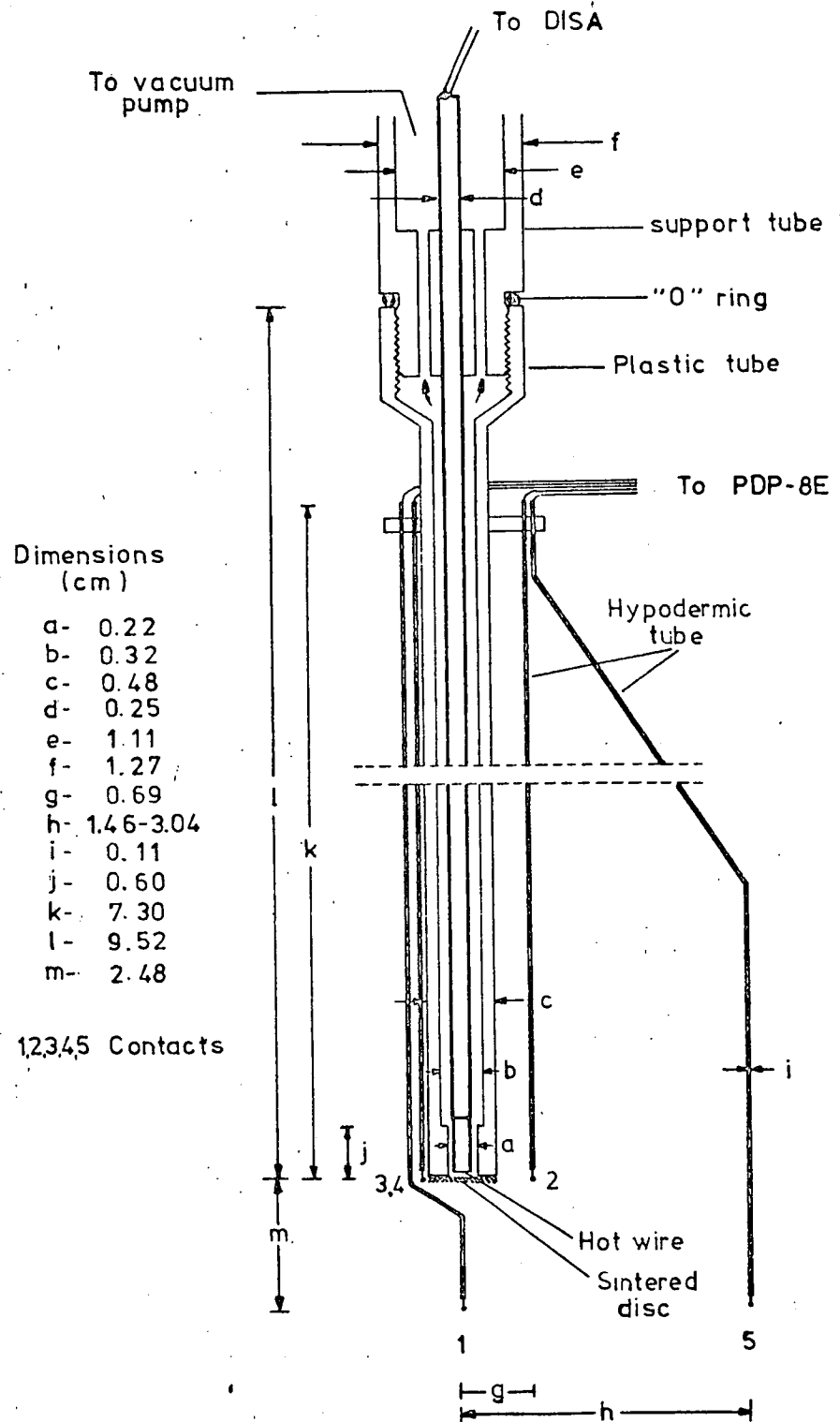


Figure 4.4

Compound probe details.

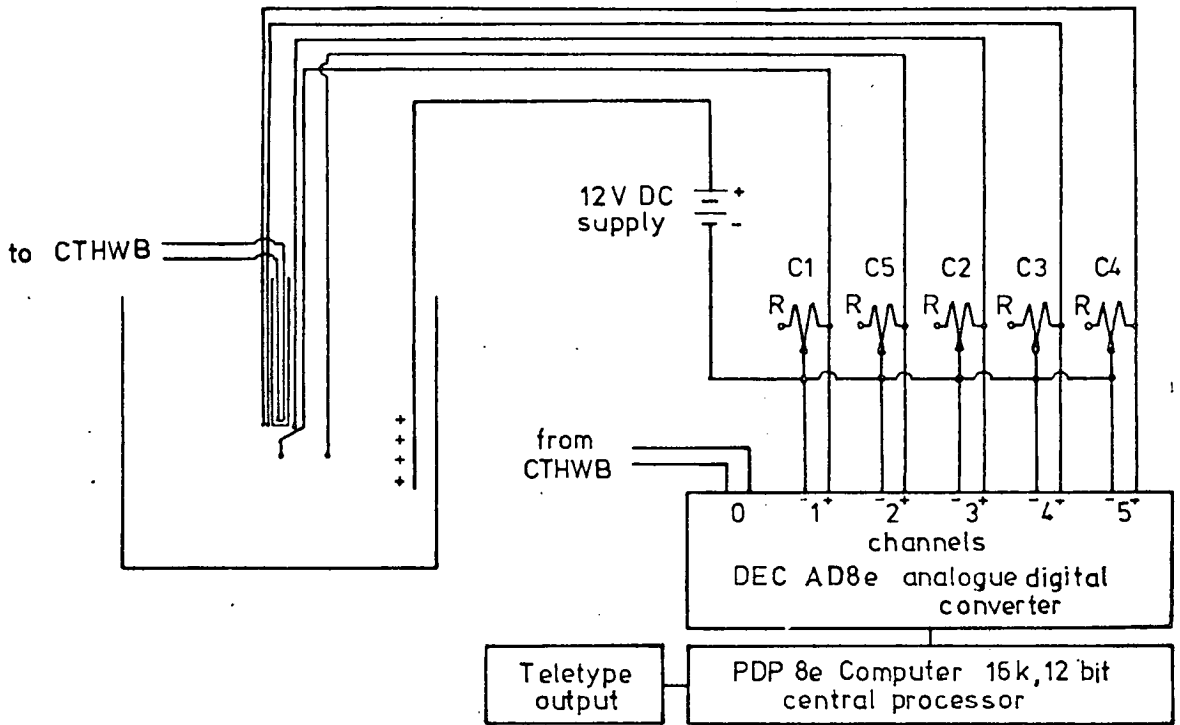


Figure 4.5 Schematic illustration of the relationship between the compound probe and the PDP-8e computer.

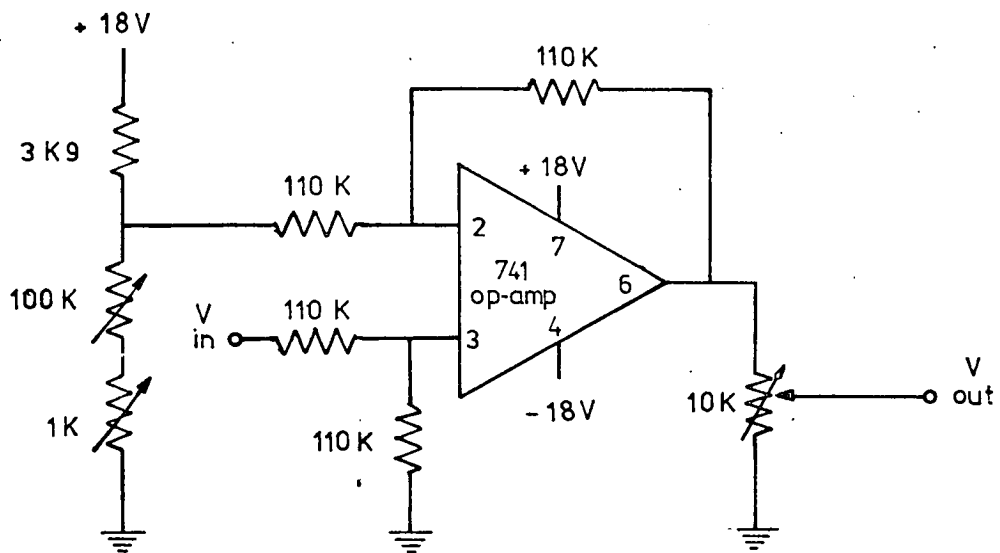


Figure 4.6 Scheme of the off-set computer line matching circuit.

gain in order to obtain 0 volts for pure air and a voltage between 0.5 and 1 for any concentration of tracer between 0.5 and 1 $\text{cm}^3 \text{He}/\text{cm}^3$. The line matching circuit is described in figure 4.6.

Bubble parameter channels

From the previous paragraphs it is clear that an ideal pulse sequence on all the five channels will be similar to the one shown in figure 4.7.

The discriminating function for acceptance of a bubble has been described in 4.3 and makes use of pulse delay time $T_{1,i}$ ($i = 2, 3, 4$) as well as pulse duration time T_i ($i = 2, 3, 4$).

Although figure 4.7 shows an ideal pulse sequence on all five channels, pulses emanating from the contacts are not ideal in shape, that is, they are not squarely shaped, and a finite time is required for the voltage to fall from the maximum voltage (dense phase mode) to the minimum voltage (bubble phase mode). This will require the definition of a threshold voltage, V_t , from which the pulsing mode of the channels can be identified.

$$\begin{array}{ll} V_i > V_t & \text{Dense phase mode} \\ V_i \leq V_t & \text{Bubble phase mode} \end{array}$$

Ideally V_t should be 1V, but due to the fact that the fluidised dense phase is a non-continuous media, the probe tips being intermittently in contact with conducting particles, thus leading to a high level of noise, it was decided to choose $V_t = 0.5 \text{ V}$ as the threshold voltage, below which the probe is considered in bubble phase mode.

The above process, although very useful for the decision criteria to be implemented in computer language, will make the

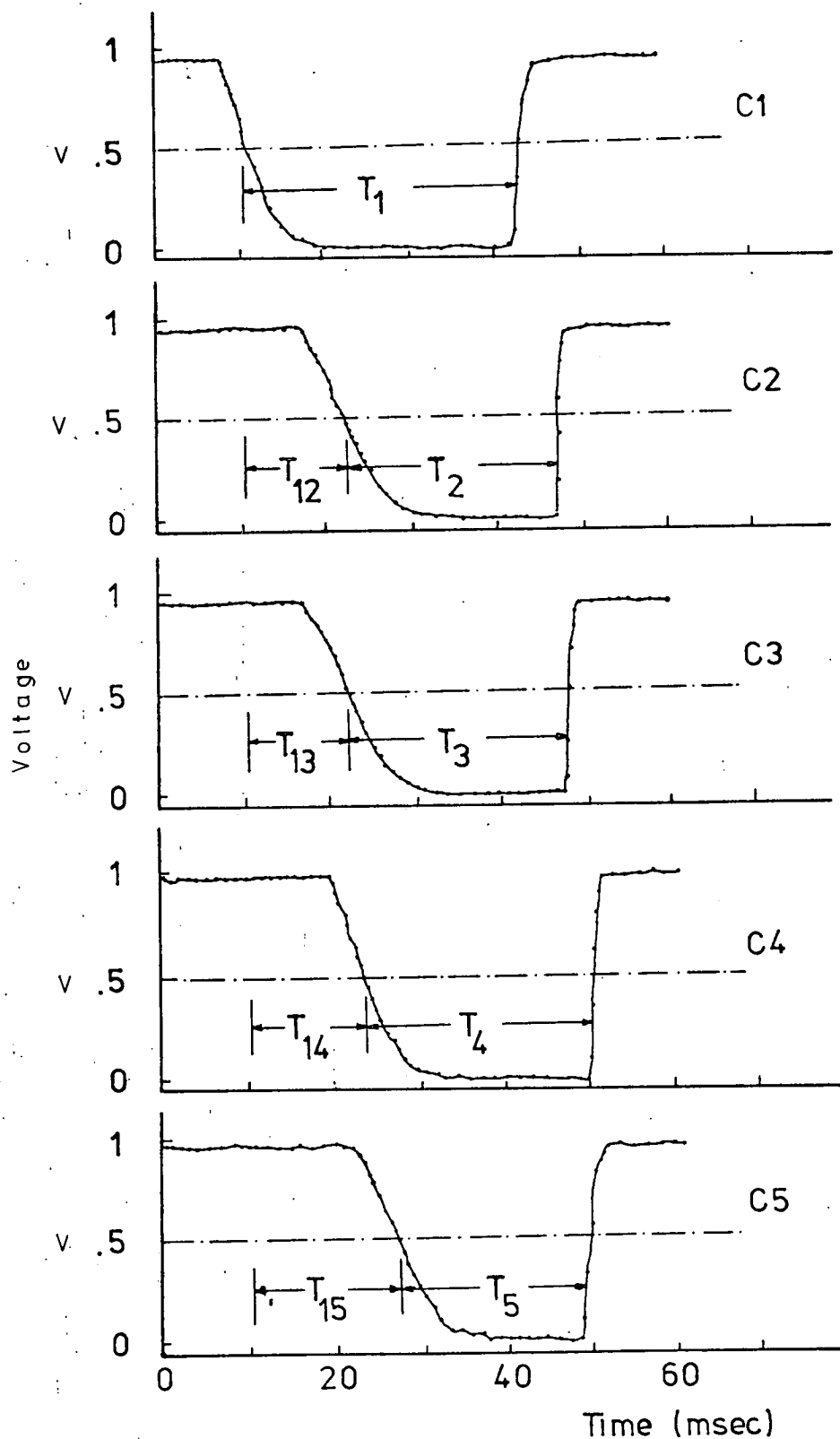


Figure 4.7

Ideal pulse sequence generated by the conductivity contacts when struck by a single, vertically rising bubble.

definition of the pulse duration time rather arbitrary due to the trapezoidal shape of the pulse. The above is clearly shown in figure 4.8 where a pulse experimentally obtained is presented, and one can verify that the time that it takes the leading edge of pulse to drop from 1V to 0V is approximately 10 milliseconds.

A correct pulse duration, assumed to be the pulse duration at $V = 1V$, can simply be obtained if the probe output voltage is recorded in the computer memory and for a successful bubble hit, the computer examines the curve backwards from c through linearisation over small time intervals in order to calculate point slopes of the curve. When the slope becomes zero a total correction time is computed and is added to the pulse duration time obtained at the threshold voltage, V_0 .

$$T_i = T_i^* + T_c$$

The above correction was applied to all channels and only to the bubble leading edge, as the rise time of the voltage is very small when the probe re-enters the dense phase and no correction will be made for the trailing edge of the pulse.

Analytical channel

In 4.5, the arrangement of the analytical channel in relation to the five contacts of the bubble parameter detector was described, as well as the discrimination necessary to obtain gas composition within the bubble and in its near environment.

As the frequency of response of the conductivity contacts are very high the accuracy of synchronization between the pulse mode of the three upper channels and the CTHWB output will depend only on the rise time of the analytical sensor.

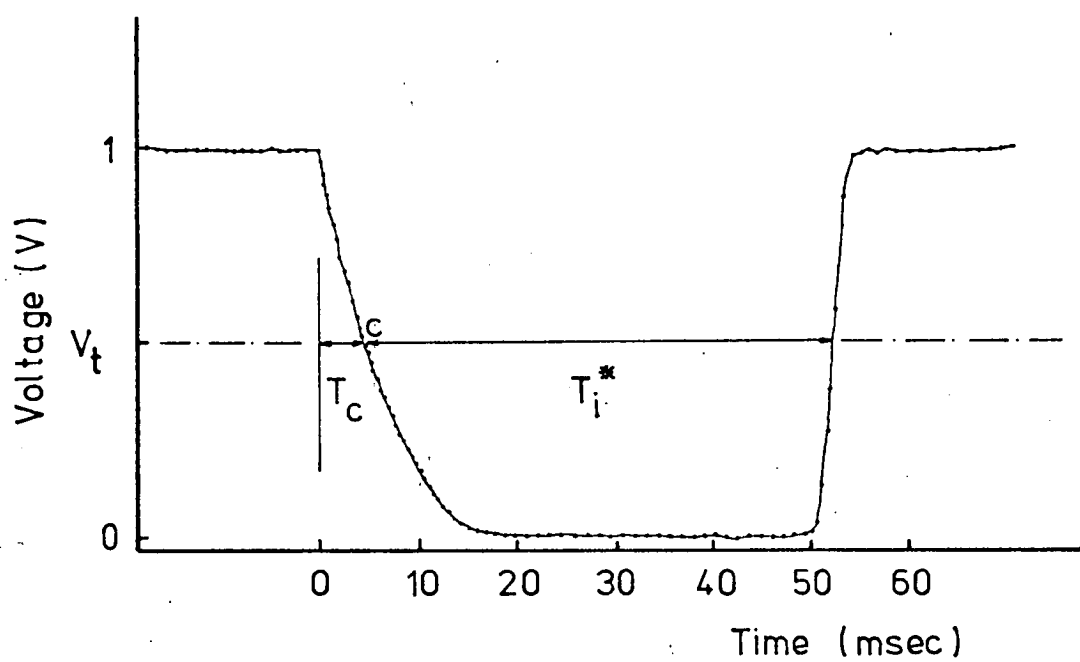


Figure 4.8

Experimentally obtained bubble pulse, showing rise time correction to obtain time of pulse duration.

Since the frequency of response of the CTHWB is greater than 10 KHz any time lag between the pulsing mode of the conductivity contacts and the CTHWB will be due to the flow of gas and the time necessary for the gas to travel between the sintered plate and the Hot-wire, which depend on the gas sampling rate. The gas sampling rate should be maintained to the minimum possible in order to avoid bubble volume variation during the analysis time, and also to avoid mixing of gas.

It would be ideal to sample gas with a flow rate which would give a superficial gas velocity, at the sintered plate level and based on the sensor head external diameter, equal to the gas superficial velocity at incipient fluidisation when the probe is in dense phase mode and equal to the bubble velocity when the probe is engulfed by a bubble. Such an ideal situation is impracticable and it was decided to sample gas at a rate that will give a superficial gas velocity, based on the diameter of the sensor head, intermediate between the gas superficial velocity at incipient fluidisation and the velocity of a bubble having an equivalent spherical diameter of 2 cm, $U_b = 31.47 \text{ cm sec}^{-1}$. The analytical gas sampling rate was then $3 \text{ cm}^3 \text{ sec}^{-1}$, superficial velocity, based on the sensor diameter 16.6 cm sec^{-1} . For the above bubble the total amount of gas sampled from the bubble when in contact with the probe will be 4.55% of its total volume, if assumed spherical, and only 2.27% if assumed hemispherical. In a fluidised bed of 15.4 cm diameter and a superficial gas velocity at incipient fluidisation of $0.575 \text{ cm sec}^{-1}$ the amount of gas sampled from the dense phase is only 2.86%.

A test to measure the rise time of the CTHWB at the above gas sampling rate was carried out, and is described in Appendix B. It

was verified to be 3.2 ± 0.5 milliseconds, which is similar to the rise time of the conductivity contacts, as shown in figure 4.8.

From the above it is concluded that a sampling rate of $3 \text{ cm}^3 \text{ sec}^{-1}$ does not produce much mixing of the gas, sampled from around the probe. This is only a small fraction of the gas in the dense phase and a small fraction of the total bubble volume, even for a bubble with an equivalent spherical diameter of 2 cm. Finally, the rise time of the CTHWB is adequate for a complete synchronization between its output and the pulsing mode of the three upper channels and hence with the bubble position.

The sensitivity of the helium concentration detected depends ultimately on the pre-set gain of the off-set computer line matching circuit and on the conversion of the analogue voltages to their discrete digital form. For the PDP-8e a 0V signal corresponds to a digital binary representation 000 000 000 and 1V corresponds to 111 111 111, which is 512 in decimal. Thus, if the output of the line matching circuit is made to be 0V for pure air (0% He) and the gain adjusted to give an output of 1V for pure helium (100% He), the sensitivity of the analogue-digital conversion will be 0.195%.

4.7 COMPUTER SOFTWARE

All the above mentioned examination logic of the probe output was translated into assembler language PAL-8 in order to be implemented in the PDP-8e computer.

The final program logic was able to examine all six channels at an approximate rate of 2 KH_2 , the sampling loop taking $500 \pm 1 \text{ } \mu\text{sec}$. Thus, we can obtain an accurate description of bubble velocity, and

size, as well as point composition of gas.

A complete description and timing methods of the program may be found in Appendix C. The program also contains the necessary instructions to operate automatically the fluidised bed and the experimental runs. Thus, the zeroing and calibration of the CTHWB and the bubble injection were done automatically by the computer.

CHAPTER 5

EXPERIMENTAL PROGRAMME

RESULTS

5.1 EQUIPMENT

The aims of the experimental programme were to determine instantaneous bubble size, rise velocity, shape, surface area, gas concentration in both phases and, hence, mass transfer rates between bubbles and dense phase.

The method employed to perform all the above measurements was described in the previous chapter and requires the injection of bubbles containing helium as a tracer in an air fluidised bed of electrically conducting particles.

A schematic diagram of the fluidised bed and ancillary equipment is shown in figure 5.1.

The fluidised bed used in these investigations was contained in a vessel of QVF glass pipe, 150 cm long and 15.4 cm diameter. To the lower end of the tube was fitted a mount which holds a sintered plate gas distributor of brass 0.635 cm thick of 25 μ m pore size. Two pressure tappings were inserted in the PVC mount, one above the sintered plate and one below, thus permitting the measurement of pressure drops due to the fluidised bed and due to both the fluidised bed and sintered plate. Figure 5.2 shows details of the mount, care being taken to ensure perfect sealing at the plate edge.

A calming section was fixed to the column below the PVC mount, which is also shown in figure 5.2.

The probe at the end of a supporting tube of brass, diameter 1.25 cm, was aligned vertically by means of a supporting bracket at the top of the QVF tube, as shown in figure 5.1. The bracket was made of two strips of Dural sheet and separated vertically by two spacers, 15 cm long, at their tips. A slot, 1.25 cm wide, ran

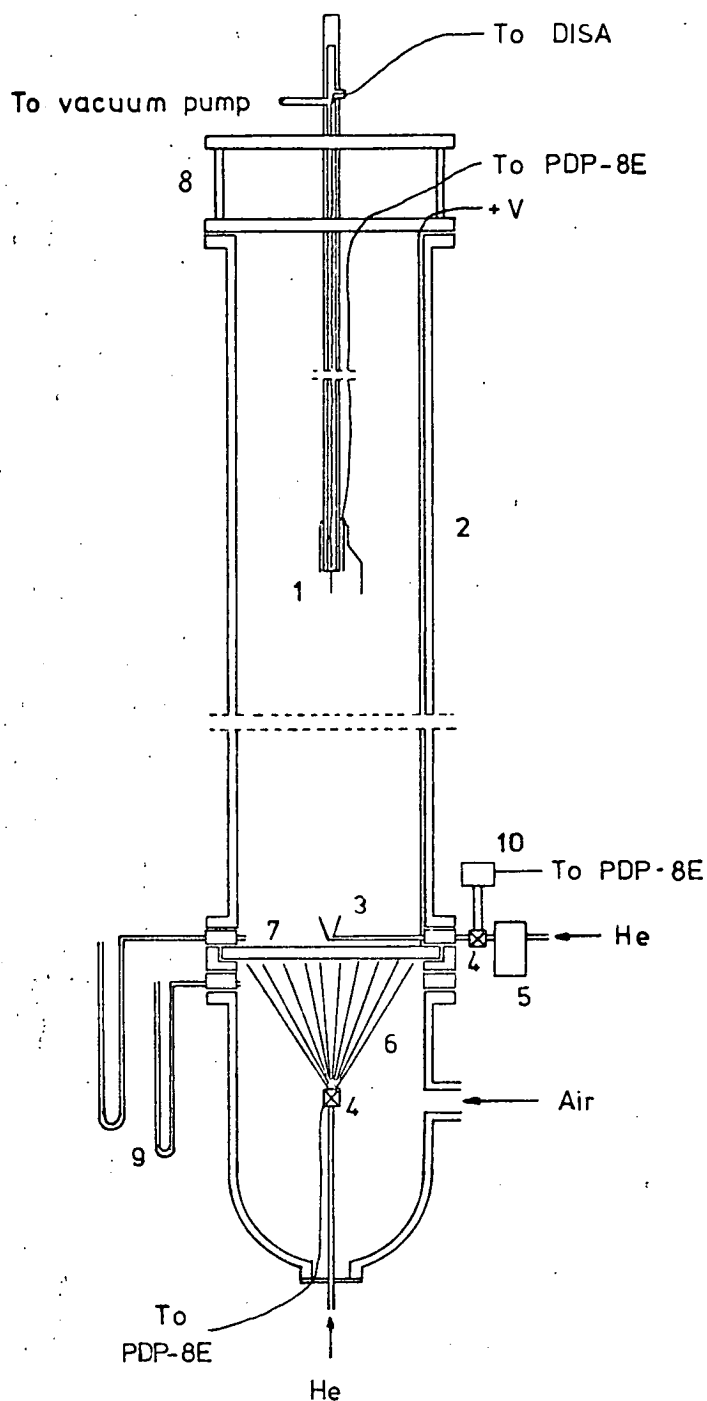


Figure 5.1

Fluidised bed equipment

1. Probe and supporting tube. 2. Column.
3. Expansion nozzle. 4. Solenoid valve.
5. Pressurized tank. 6. Hypodermic tube distributor.
7. Sintered plate distributor.
8. Supporting bracket. 9. Manometers.
10. Variable time delay switch.

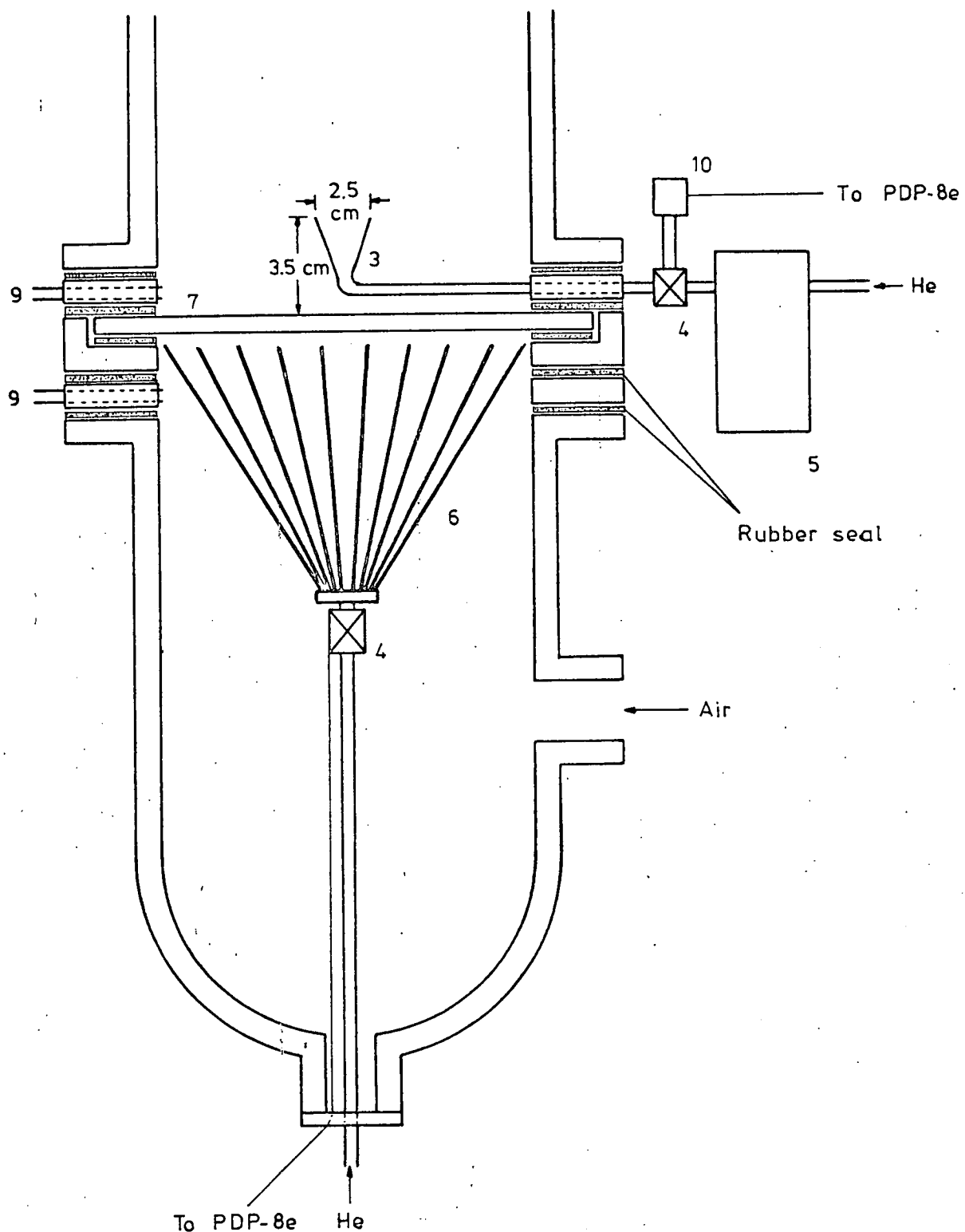


Figure 5.2 The gas distributor and bubble injector.
(For details see figure 5.1)

through each strip over the whole length. The probe supporting tube passed through these two slots and was then fitted in place. Because the Dural strips were placed over the diameter of the QVF tube, this arrangement permits the probe and supporting tube to be readily moved both vertically and radially.

The fluidised bed was made of petroleum coke which had been partially graphitised. The particles were screened and the fraction with size in the range $80\text{ }\mu\text{m}$ - $105\text{ }\mu\text{m}$ used to form the fluidised bed. To form a 90 cm high fluidised bed at incipient fluidisation 11500 g were utilized.

The specific gravity of the coke particles was determined experimentally and was found to be 1.58 g cm^{-3} .

5.2 INCIPIENT FLUIDISATION

Superficial gas velocity and bed voidage

The minimum fluidisation velocity was measured by the pressure drop technique, being determined by the intersection of the graphical asymptotes of the two sections of the $\Delta P/U_s$ curve (pressure drop across the bed versus superficial gas velocity), as shown in figure 5.3. Figure 5.4 shows bed height versus superficial gas flow rate.

From the above two figures it is found that the superficial gas velocity at incipient fluidisation is 0.575 cm sec^{-1} and the bed height is 90 cm.

The bed voidage at incipient fluidisation was calculated from the coke weight, its specific gravity and bed height at incipient fluidisation. It was found to be 0.557.

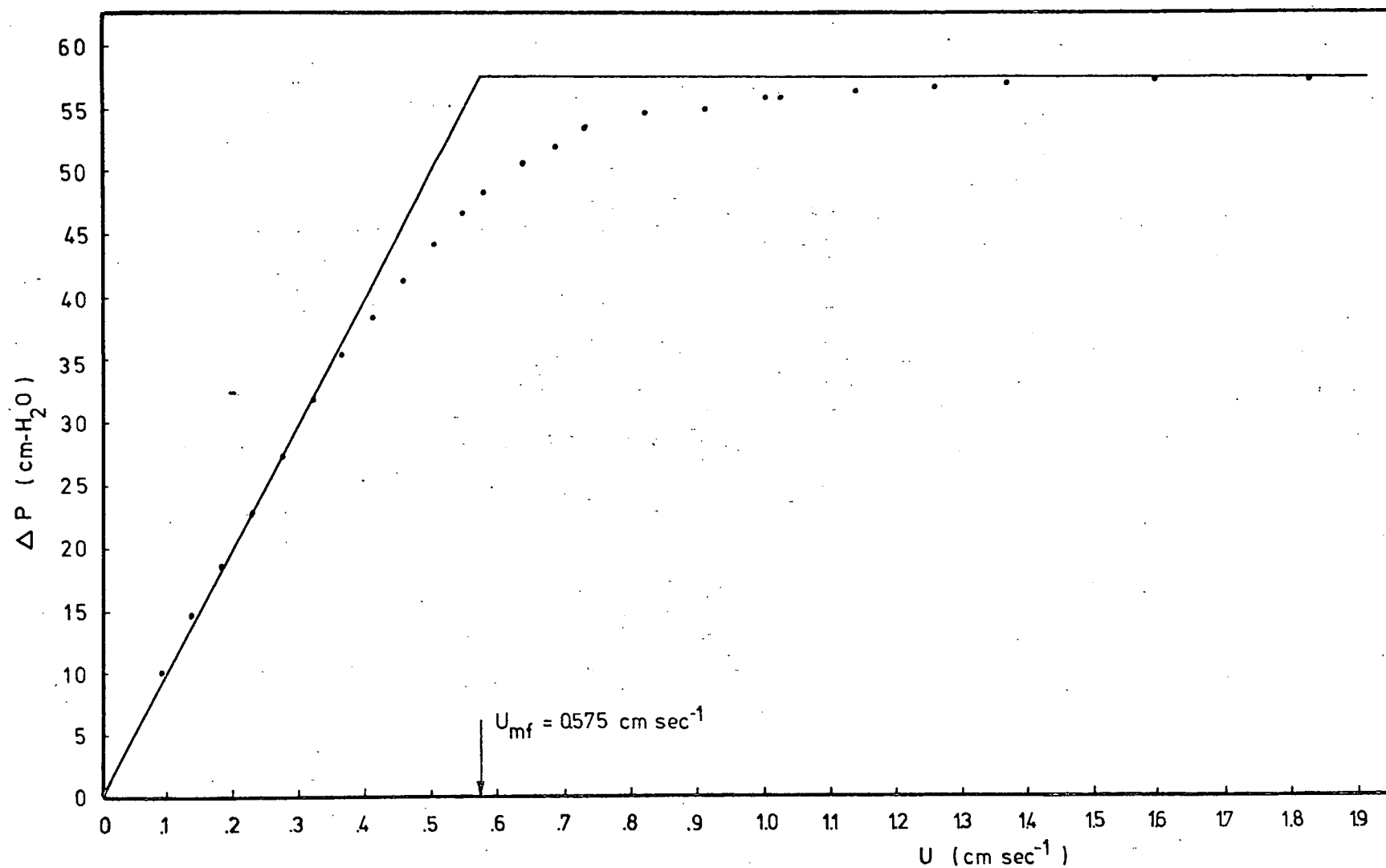


Figure 5.3

Determination of minimum fluidisation conditions

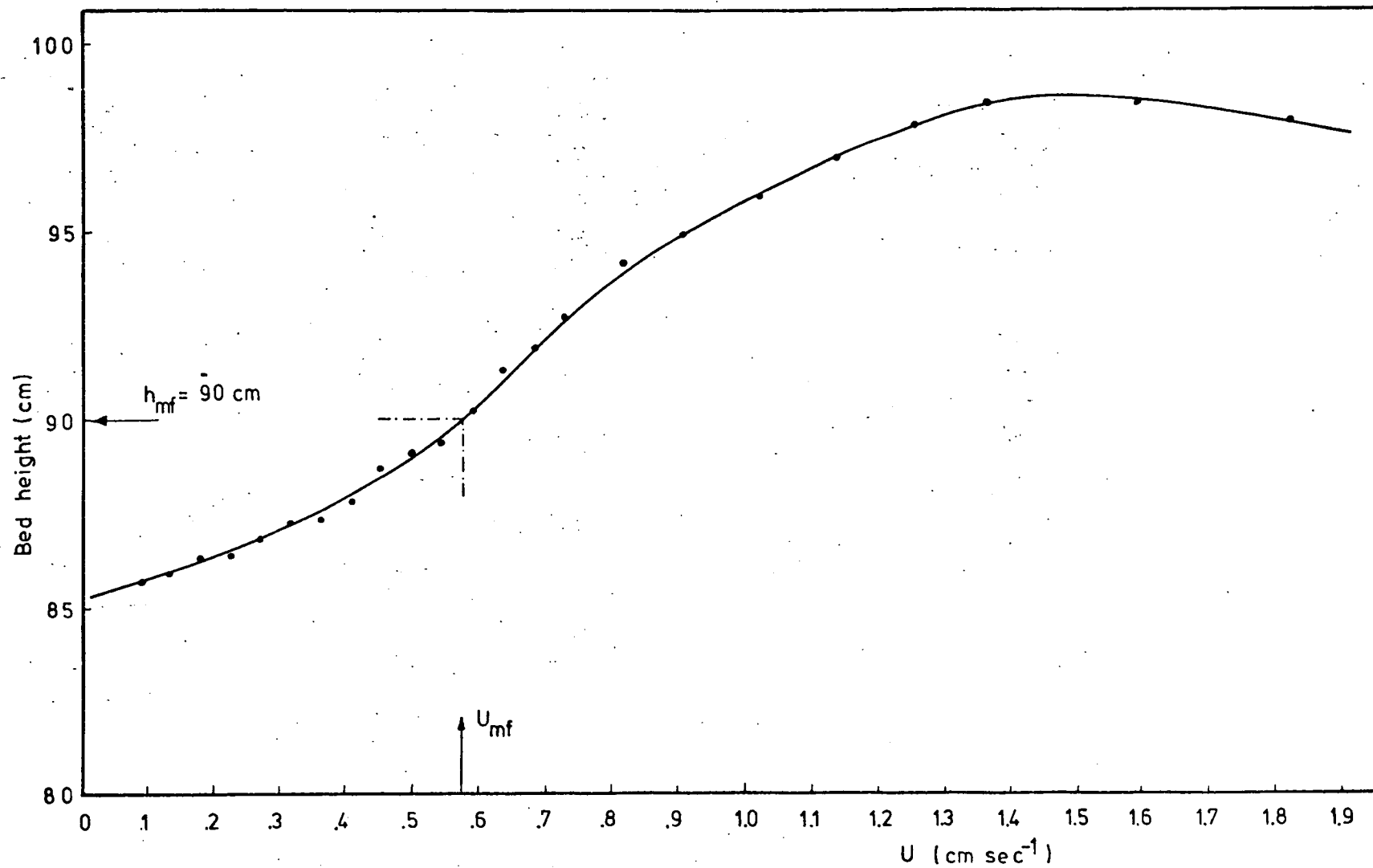


Figure 5.4 Determination of bed height at minimum fluidisation

5.3 BUBBLE INJECTION SYSTEM

Bubbles were injected into the bed at the column axis through an expansion nozzle situated immediately above the sintered plate distributor, as shown in figures 5.1 and 5.2.

The expansion nozzle at the column axis was connected, by a copper pipe (I.D. -0.635 cm), to a solenoid valve outside the bed. The valve was operated by a variable time delay switch and relay which could actuate the valve at various frequencies and made to stay open for 0.10 - 0.13 seconds.

Helium tracer was supplied from a pressurized tank connected to the valve. In order to obtain bubbles of different sizes the pressure in the tank was regulated in the range .2 - 2.5 kg cm⁻² gauge and two different sizes of tank were used , 590 cm³ and 139 cm³, in order to obtain bubbles of volume ranging from 30 to 900 cm³.

In each experimental run the time delay switch and relay was automatically energised by the computer and the bubble injection initiated by the pre-set frequency in the time delay switch, and ranging from 20 to 40 bubbles per minute.

5.4 CTHWB CALIBRATION

As shown in Appendix B, a linear relationship was found between the helium concentration and the CTHWB output, which is made compatible with the analogue digital characteristics of the PDP-8e computer through an off-set computer line-matching circuit with variable gain, which is described in figure 4.6.

From Appendix B it is also apparent that the CTHWB output is

dependent on the gas temperature. In order to compensate any possible drift of the bridge and computer matching circuit, the zero (corresponding to a helium concentration of 0%) and the proportionality constant between helium concentration and CTHWB output were determined automatically by the computer before any experimental run.

The zeroing of the CTHWB is trivial since it is the output corresponding to pure air in the bed.

In order to determine the above proportionality constant the air flow through the bed was metered by two calibrated rotameters, one of them fitted with a solenoid valve which was normally open. A third rotameter, also fitted with a solenoid valve, was used for metering helium and was set for the same flow rate as the rotameter metering air.

To determine the proportionality constant, the computer, after zeroing the CTHWB output, closed the magnetic valve of the air metering rotameter and opened the valve of the helium rotameter, thus permitting the introduction of a known mixture of helium-air in the bed without change in the flow through the bed. The calibrating mixture contained, invariably, 50% He and 50% air. The helium being introduced in the bed through a distributor of hypodermic tubes situated immediately below the sintered plate distribution, as shown in figure 5.1.

The computer analysed the CTHWB output until the gas composition in the bed reached the steady state, when the CTHWA output was recorded and the proportionality constant determined. In order to detect the approach to the steady state in the bed the computer analysed the CTHWB at constant time intervals of 500 msec and the bed was considered under steady state when the

ratio between two consecutive voltage outputs of the CTHWB was greater than 0.996.

After recording the steady state output voltage the flow of helium to the bed was stopped and the magnetic valve of the air rotameter switched on.

A period of 30 minutes was then allowed before the start of the experimental run, in order to eliminate any trace of helium.

5.5 EXPERIMENTAL RUN

The experimental run consisted of the injection of tracer bubbles at a given frequency and the simultaneous measurement of all the relevant bubble parameters and gas composition within the bubble and its surroundings.

The whole experimental run was carried out automatically by the computer, only the frequency of bubble injection being pre-set in the time-delay time switch.

In each experimental run the probe was positioned at a given level in the bed at its axis and then the time delay-relay circuit was automatically switched on and bubbles introduced into the bed. In order to reach steady state, sixty bubbles were introduced before the computer began the analysis of the probe output. Then, for each bubble encounter pulse duration and delay times between contacts were stored in computer memory, as was the tracer concentration in the dense phase just ahead, behind or within the bubble, as it passed a given level in the bed. Since the main sample loop of the program implemented in the computer takes $500 \pm 1 \mu\text{sec}$ to analyse all the channels, the gas composition was analysed and stored in computer

memory at the approximate rate of 2 KHz.

For a given bubble passing the probe, if the discrimination function described in chapter 4 was satisfied and the bubble passed the probe assembly "centrally", the data was kept in computer memory. For a successful bubble encounter the CTHWA output was analysed, after the bubble had passed the probe, for a further time interval corresponding to $\frac{3}{4}T$, where $T = \frac{1}{N}$ and N is the bubble injection frequency. The rate of sampling in this case was lowered to 0.1 KHz, i.e. one gas composition every 10 m sec.

If a given bubble encounter was unsuccessful all the memory addresses were reset and the memory storage overwritten by the probe output corresponding to another following bubble.

In each run after six successful bubble encounters, the computer analysis was stopped and the bubble injection was discontinued. All the data stored in computer memory was then punched out on paper tape for further computer processing on the Edinburgh Regional Centre IBM-4-75 computer. In each run it was found necessary to inject between 100 to 200 bubbles, depending on bubble size, in order to obtain the six successful bubble encounters. The probability of a successful encounter increases with the bubble size.

The process explained above in 5.4 and 5.5 was automatically repeated by the computer, the probe level being normally changed during the print-out of the data.

In all the experimental runs bubbles were introduced into the bed under minimum fluidisation conditions.

5.6 EXPERIMENTAL RESULTS

BUBBLE PARAMETERS

5.6.1 Bubble rise velocity

The measurement of bubble rise velocity is of primary importance since all the other parameters are derived from it.

Owing to the fine limits set by the velocity discriminator (typically in our experiments 0.95), and due to the rapid sampling rate of the computer permitting an accurate measurement of delay times, the error involved in the determination of the rise velocity of a bubble, using a probe with a separation distance of 2.48 cm between leading and upper contacts, will be less than 3%.

Figure 5.5 shows the mean rise velocity of single bubbles of different sizes. For each size the deviation from the mean value obtained in the various experiments is also shown. No systematic variation of bubble velocity with distance above the distributor was detected.

In figure 5.5 the bubble velocity is plotted versus the equivalent spherical diameter, given by

$$D_e = \sqrt[3]{\frac{6V_b}{\pi}} \quad (5.1)$$

The bubble volume, V_b was also deduced from the probe output, as will be shown below.

The results are compared with equation

$$U_b = 0.711 \sqrt{g D_e} \quad (5.2)$$

for the rising velocity of bubbles, and the equation

$$U_s = 0.35 \sqrt{g D} \quad (5.3)$$

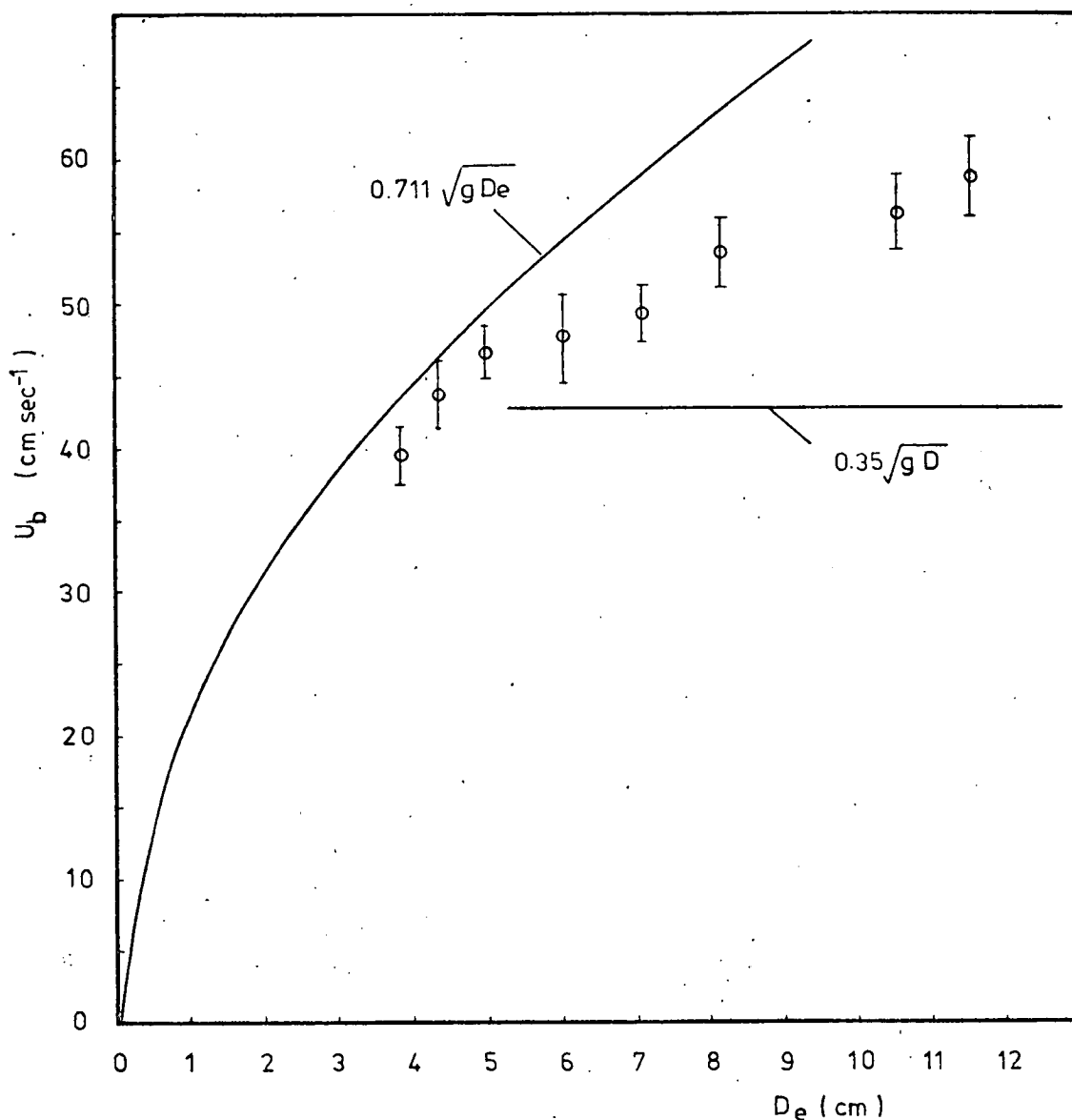


Figure 5.5

Bubble velocities versus the equivalent spherical diameter of the bubble. Each point represents the average of three sets of experimental runs made between 10 and 65 cm above the bubble injector at 5 cm intervals. For each one of the bubble sizes the standard deviation of the velocity is also presented. Its low value (less than 3 cm.sec $^{-1}$) shows that bubble velocity remains constant with height.

for the rise velocity of slugs.

The results clearly reflect an influence of the column walls since the bubbles examined follow a relationship intermediate between the two above.

Reference to the figure also shows that the reproducibility of the probe output is good, with single bubble rise velocity deviations of less than 6% from the mean.

5.6.2 Bubble shapes

Bubble geometric characteristics were derived from the probe output which permits the immediate determination of bubble central length, L_b , vertical distance to bubble interface from bubble nose at a radial distance, S_r , from the bubble centre line, L_d , and an off-centre bubble height, L_o , at the same radial position as L_d .

Figure 5.6 reports the mean bubble shape for each one of the different bubble volumes injected in the bed in terms of the sizes reported by the probe output (L_b , L_d and L_o) and following the process of shape analysis and fitting, as presented in Appendix A. The fluidised bed inner boundaries (column wall) are also represented in figure 5.6 in order to compare the bubbles sizes with the bed diameter.

The results shown in figure 5.6 show that a transition from a spherical cap shape to a paraboloid shape occurs as the bubble volume increases. Such transition is of similar character to the bubble shape variation in two dimensional fluidised beds obtained by Goldsmith and Rowe (27). Using photography they verified that a transition from a circular bubble, with an indented

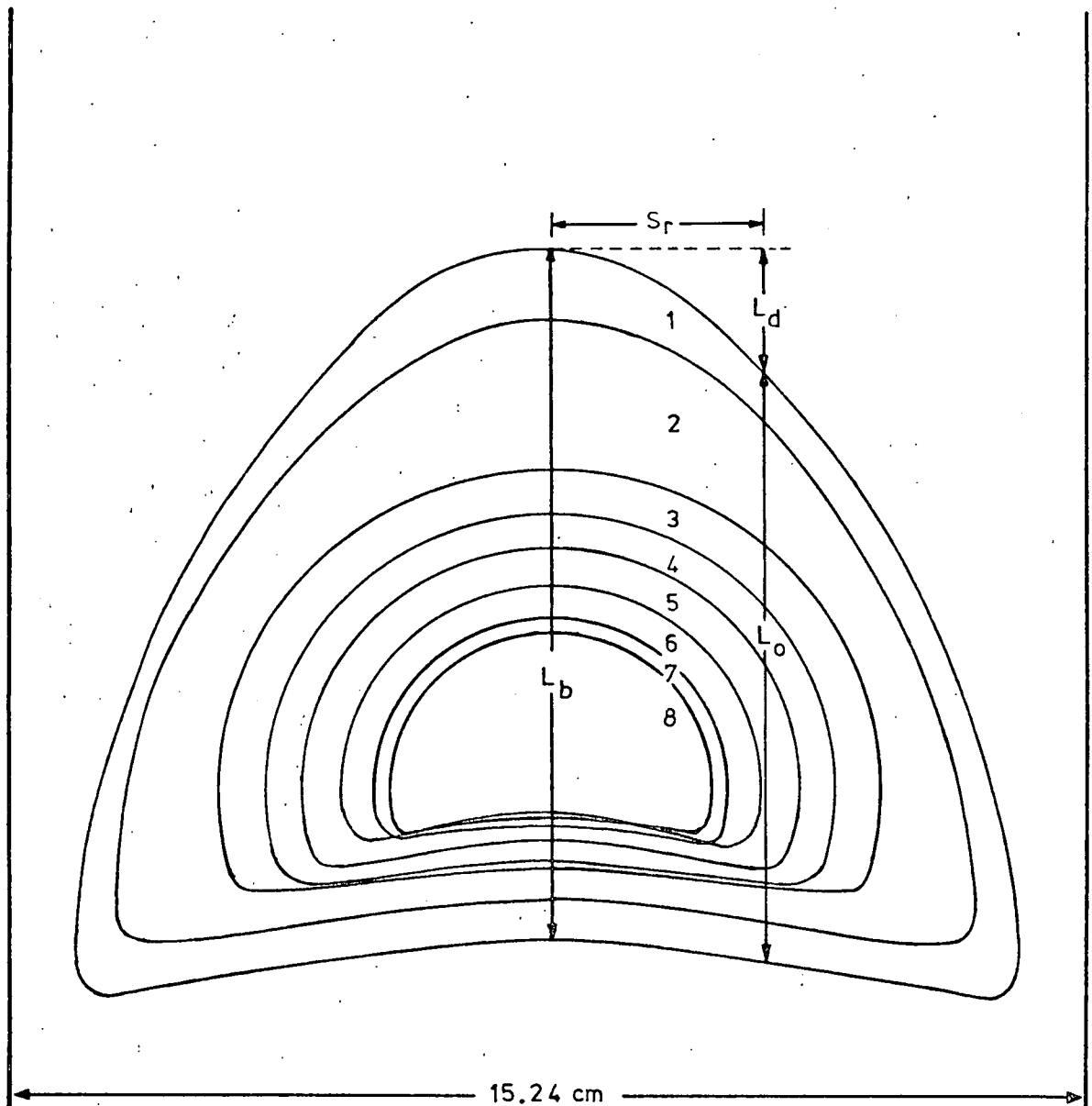


Figure 5.6

Bubble shapes obtained for different bubble volumes injected in the bed and referred to in Table 5.1 (pag. 79).

rear surface, to an elongated shape occurs as the bubble maximum width increases. The elongation is presumably due to a wall effect. This is probably also the reason for the transition in our case. It is interesting to note that Goldsmith and Rowe (27) found that the transition occurs for values of bubble width in the range of 8 to 10 centimeters. In our case, as is shown in figure 5.6, it was found that bubbles with a maximum width above 10 centimeters are paraboloid shaped and below 10 centimeters are spherical caps.

The spherical cap-shaped bubbles reported in figure 5.6 have similar geometry to that noted by Rowe and Partridge (69). Examination of their data for three dimensional bubbles reveals that for a particle size of around 90 μm , as used here, the wake angle θ_w should be almost 150° , with the mean wake fraction, f_w , approaching 0.35, which are independent of bubble size. We found in our experiments that bubbles have an indented rear surface, as $L_b < L_o + L_d$, with mean wake angle and wake fraction of 156° and 0.35 respectively, thus showing again some accord with the alternative technique of x-ray photography used by Rowe and Partridge (69). The scatter of f_w and θ_w shown in table 5.1 is similar to that of (69), who found considerable deviation from the mean values with wake angles up to 180° and wake fraction up to 0.5.

All the bubble parameters are reported in Table 5.1 for each one of the bubble sizes.

5.6.3 Bubble volume and surface area

Both the volume and surface area of the bubble were calculated using the equations for each of the solids of revolution corresponding

TABLE 5.1

Exp	Injection Frequency (min ⁻¹)	U_b (cm sec ⁻¹)	L_b (cm)	L_d (cm)	L_o (cm)	S_r (cm)
1	20	59.1±2.9	10.12±0.38	1.74±0.10	8.51±0.42	3.04
		59.2±2.6	9.93±0.43	1.86±0.14	8.42±0.48	
		58.4±2.7	10.04±0.38	1.77±0.09	8.51±0.51	
2	20	56.9±2.4	8.61±0.38	1.68±0.11	7.16±0.38	3.04
		55.4±2.6	8.42±0.34	1.73±0.09	7.21±0.41	
		56.6±2.6	8.47±0.35	1.75±0.12	7.17±0.37	
3	20	52.9±2.4	5.82±0.31	1.06±0.09	4.89±0.23	3.04
		54.0±2.5	5.71±0.29	1.21±0.11	4.73±0.28	
		53.8±2.4	5.75±0.26	1.09±0.08	4.87±0.27	
4	20	49.1±1.7	4.98±0.30	0.62±0.05	4.60±0.26	2.11
		51.2±2.0	5.08±0.24	0.57±0.07	4.55±1.27	
		47.9±2.1	5.12±0.31	0.61±0.06	4.59±0.21	
5	20	46.1±3.2	4.18±0.23	0.72±0.06	3.77±0.21	2.11
		48.2±2.8	4.07±0.19	0.69±0.06	3.76±0.17	
		48.8±2.9	4.14±0.26	0.69±0.05	3.82±0.19	
6	20	47.1±1.8	3.43±0.19	0.84±0.04	2.71±0.15	2.11
		46.3±1.7	3.41±0.17	0.91±0.05	2.63±0.11	
		46.1±1.7	3.27±0.25	0.92±0.04	2.61±0.12	
7	30	44.1±2.5	2.91±0.24	0.47±0.04	2.49±0.11	1.46
		44.2±2.5	3.16±0.22	0.51±0.04	2.41±0.13	
		43.1±2.3	3.08±0.25	0.46±0.03	2.51±0.11	
8	40	40.2±2.2	2.74±0.21	0.52±0.03	2.26±0.12	1.46
		39.2±1.9	2.64±0.19	0.54±0.03	2.26±0.12	
		38.2±2.2	2.60±0.24	0.56±0.04	2.22±0.13	

Bubble parameters obtained directly from the probe output.

The above reported data are the average of sets of experimental runs made between 10 and 65 cm, above the injector, at 5 cm intervals. The standard deviation is also presented and its low value, for all the parameters, permits the conclusion that there is no noticeable variation of the bubble size and shape with height.

TABLE 5.2

Exp	R_D (cm)	Shape	f_w	θ_w	Bubble volume (cm ³)	Bubble surface area frontal (cm ²)	total (cm ²)	Equivalent spherical diameter (cm)
1	3.47	PAR	-	-	810.5	904.0	1065.6	11.6
2	3.55	PAR	-	-	612.6	723.5	867.7	10.5
3	4.67	LEC	0.329	153.2	287.0	172.0	237.0	8.2
4	4.01	LEC	0.307	149.4	187.0	130.2	177.2	7.1
5	3.51	LEC	0.370	159.7	114.0	91.9	129.4	6.0
6	2.95	LEC	0.394	163.4	65.3	63.0	89.8	5.0
7	2.47	LEC	0.325	152.5	42.8	48.4	66.5	4.3
8	2.25	LEC	0.365	158.9	30.3	38.0	53.4	3.9

Bubble derived parameters. The above parameters were obtained from the mean values of the probe output parameters reported in table 5.1

to the fitted shape. In Table 5.2 the bubble volume and frontal and total surface areas of the mean bubble are given, where the mean bubble is the bubble resulting from the mean values of R_c , L_b , L_o and L_d . For each one of the bubble sizes the equivalent spherical bubble diameter (D_e) is also reported.

5.7 EXPERIMENTAL RESULTS

TRACER CONCENTRATION

The analytical technique previously described permits the measurement of tracer concentration both within the bubble and in the dense phase, not only in the immediate vicinity of the bubble but also during the interval between two consecutive bubbles. All experimental determinations of tracer concentration were realised under the sampling conditions specified earlier.

5.7.1 Tracer concentration profiles in the dense phase

Radial profiles

In each one of the experiments referred to in Tables 5.1 and 5.2 tracer concentration in the dense phase was measured at five different radial positions at levels between 20 and 60 cm, above the injector, at 10 cm intervals. The gas composition was obtained from the probe output at the mean time between the passage of two consecutive bubbles through a given level, thus the gas composition of the dense phase was not affected by the presence of any bubble.

Figure 5.7 shows the radial profiles obtained for experiment 5

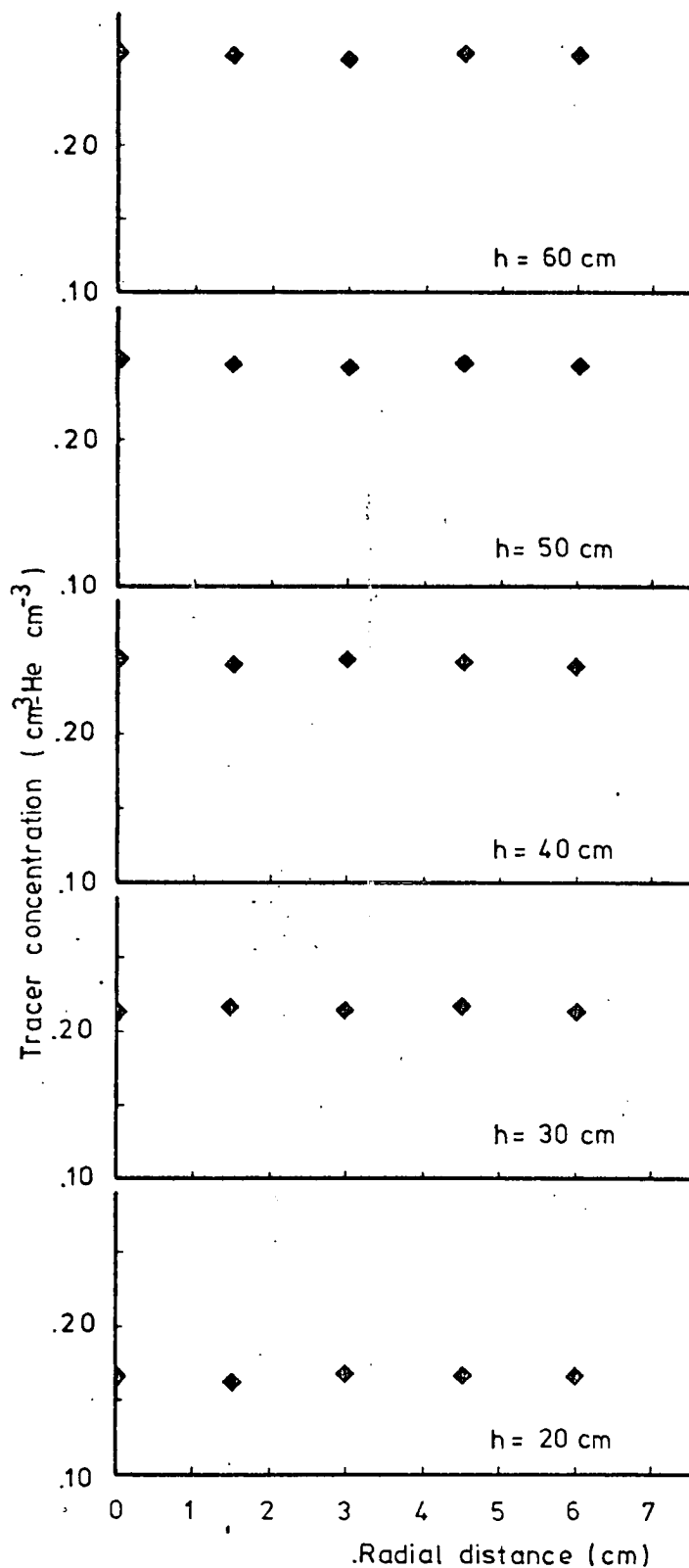


Figure 5.7 Radial profiles of tracer concentrations in the dense phase.

Tracer concentrations were obtained from the probe output and each point represents the average of three sets of six measurements, the standard deviation being always less than $0.012 \text{ cm}^3 \text{ He cm}^{-3}$.

(see Table 5.1 and 5.2) in which the bubble frequency of injection was 20 bubbles per minute and the bubble volume 114 cm^3 . Similar results were obtained for all the other experiments mentioned in Tables 5.1 and 5.2.

The above results show that a flat profile of concentration in the dense phase exists for all the levels above 20 cm. They show also that the concentration at each level increases with height.

Similar results were obtained by Davies and Richardson (20) who injected bubbles of carbon dioxide in an incipient fluidised bed of particles with a similar size to ours, and also by Stephens et al (72), again for a similar particle size.

The existence of flat radial concentration profiles is certainly due to the essential plug flow nature of the flow when the gas superficial velocity is low, and also due to the high diffusivity of helium in air.

Axial profiles

Axial profiles of concentration in the dense phase were obtained from the tracer concentration measurement at the axis of the fluidised bed, as above, at distances from the bubble injector between 10 and 65 cm at 5 cm intervals. The axial profiles are reported in figure 5.8 for all the experiments realised.

5.7.2 Tracer concentration within and around the bubble

Tracer concentration within and around a bubble was recorded as previously described in chapter 4.

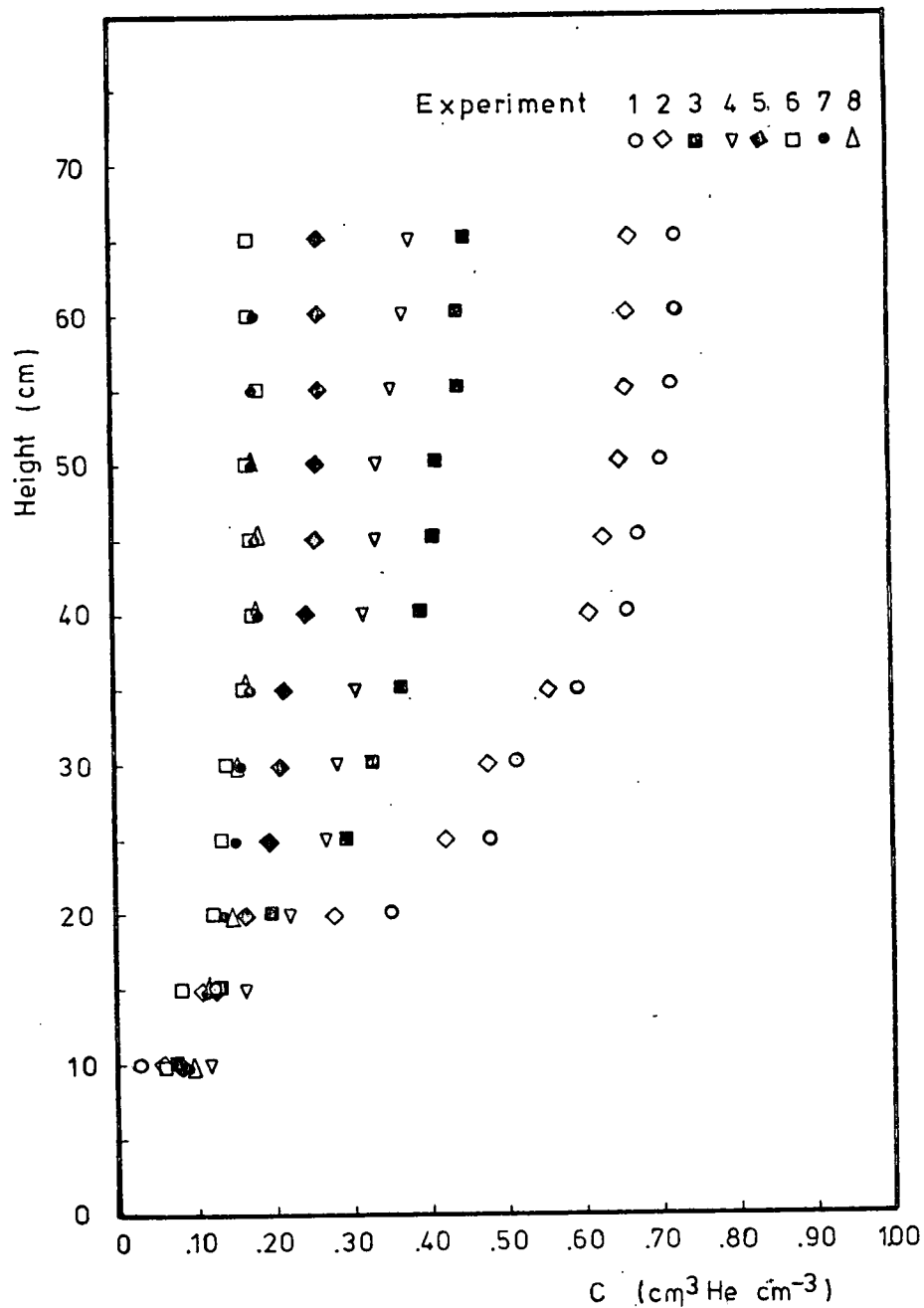


Figure 5.8 Axial profile of tracer concentration in the dense phase. For each experiment and each level the plotted data represents the average of three runs of six bubbles each, the standard deviation being always less than 0.010 cm³ He cm⁻³.

As bubbles passing the probe coaxially are the only ones accepted by the computer logic, a vertical profile of tracer concentration at the bubble axis was then obtained from the CTHWB output. Also, tracer concentration in the dense phase ahead and behind the bubble at the bubble axis was obtained since recording of the CTHWA began immediately after the leading contact enters into the bubble phase pulse mode, and was continued after the bubble had passed the probe array.

Figures 5.9 and 5.10 show the tracer profiles of concentration for bubbles with the same central length at four different levels in the bed and corresponding to bubbles of experiment 2 (figure 5.9) and experiment 5 (figure 5.10). In both cases the tracer concentration is plotted versus time and also versus the relative distance between bubble nose and the three upper channels, as the probe is stationary and the bubble moves vertically with a rise velocity, U_b .

Similar results were obtained in the other experiments mentioned.

The profiles obtained in both cases are similar and they distinguish three regions.

- A - The region ahead of the bubble corresponding to the sampling carried out between the pulse appearance on the leading contact and its appearance on the upper contacts.
- B - The bubble compositions corresponding to the pulse duration on the three upper contacts.
- C - The region behind the bubble analysed after the three upper contacts returned to the dense phase mode.

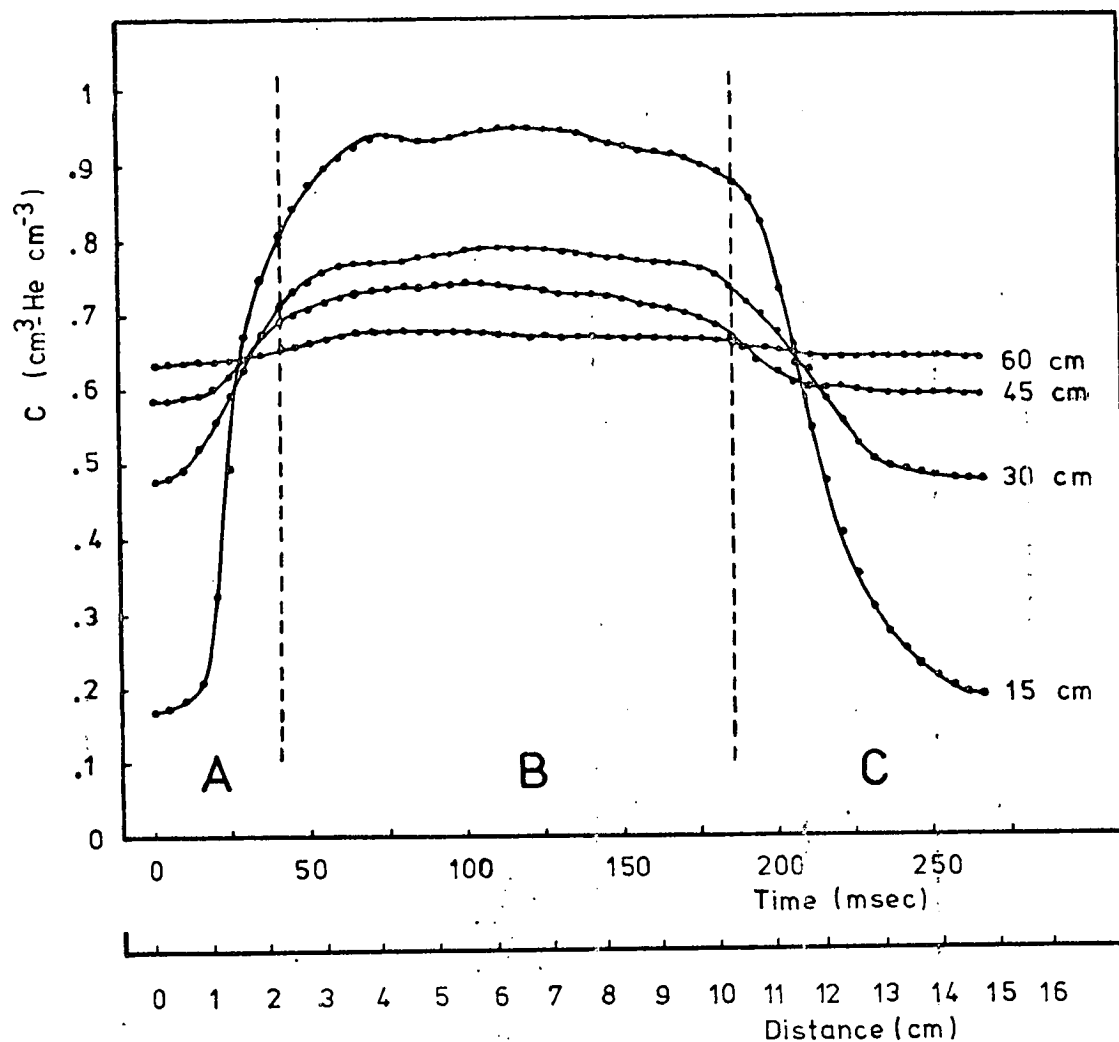


Figure 5.9

Tracer concentration within the bubble and in its surroundings, for a bubble 8.50 cm long, at four different levels in the bed.

The above profiles were obtained from the data yielded by four randomly sampled bubbles from the respective experimental runs.

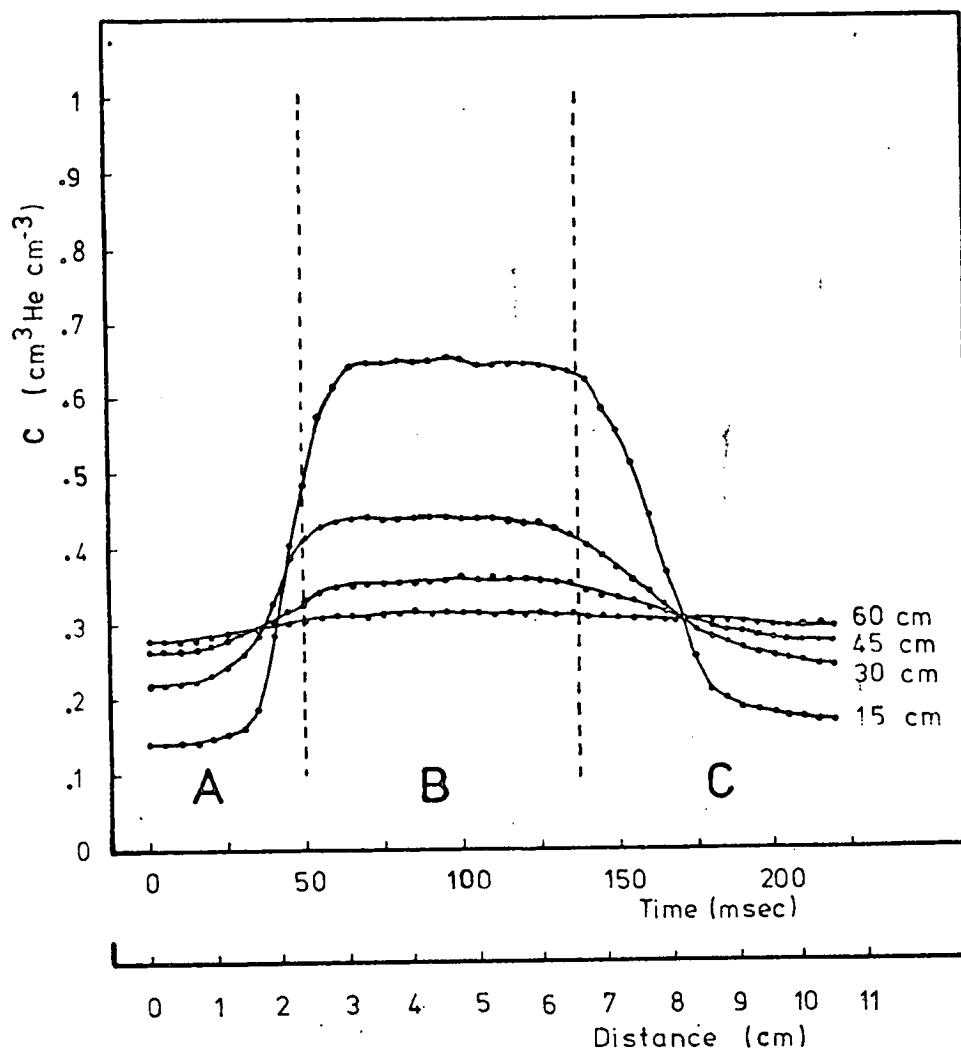


Figure 5.10 Tracer concentration within the bubble and in its surroundings, for a bubble 4.13 cm long, at four different levels in the bed.

The above profiles were obtained from the data yielded by four randomly sampled bubbles from the respective experimental runs.

The above profiles show a sharp transition of tracer concentration from region A to region B. The tracer concentration rises very sharply in the dense phase ahead of the bubble nose from its dense phase value to the bubble tracer concentration over a distance under one centimeter. The gas composition within the bubble shows an almost flat profile, the tracer concentration decreasing slightly to the rear of the bubble, as is apparent in figure 5.9 for the case of longer bubbles, and is certainly due to the fact that the bubble undergoes gas transfer with the dense phase as it rises.

In region C the tracer concentration profile shows a slower decrease in concentration when compared with region A, and is tailing over a longer distance. Figure 5.9 shows that the tracer concentration in the dense phase behind the bubble is always higher than the tracer concentration in the dense phase ahead of the bubble at $t = 0$.

The above profiles show an interesting agreement with the findings of Rowe and collaborators (70) who injected bubbles of NO_2 in two-dimensional fluidised beds. Their work, summarized by Rowe (65), gives the main quantitative evidence of the details of gas motion around bubbles and cloud and wake formation. From the above reported work in "Fluidization" edited by J.F. Davidson and D. Harrison (18), Rowe shows in figures 4.43 and 4.45 two photographs that corroborate our experimentally determined tracer concentration profiles. The comparison, although very important, has limitations due to the fact that in Rowe's case the ratio of bubble velocity to the gas velocity at minimum fluidisation is in the range 2.4 to 10 and in our case in the range 68 to 100. The cloud boundary coincides, in our case, with the bubble boundary and the profile shown in

region A of figures 5.9 and 5.10 are not necessarily the gas profile in the cloud.

The profiles described in figures 5.9 and 5.10 were systematically obtained throughout the experiments. Few exceptions were found and they only affected the region behind the bubble. Figure 5.11 shows one of the typical exceptions. It shows a sharp decrease in tracer concentration in the rear of the bubble, and a small peak of tracer concentration in the dense phase behind the bubble. Such might be due to cloud or wake shedding which occurred just before the bubble hit the probe. Rowe et al (65) (70), using their sophisticated x-ray technique, showed that cloud and wake shed at fairly regular intervals and this may explain our unsystematic findings that require shedding to occur 10 to 20 milliseconds before the bubble hits the probe in order to be detected.

5.7.3 Axial profiles of tracer concentration in the bubble phase

Due to the fact that bubbles move and the probe is static, an axial profile of tracer concentration in the bubble phase can only be obtained from the detection of tracer concentration within different bubbles at different distances from the gas distributor.

As explained before, in each run we recorded the bubble properties and tracer concentration profiles of six bubbles. Mean values of bubble properties and tracer concentration within the bubble were obtained for each run.

Since the tracer concentration within the bubble shows a flat profile, the gas concentrations of the six bubbles were averaged from their values measured in the centre of the bubble.

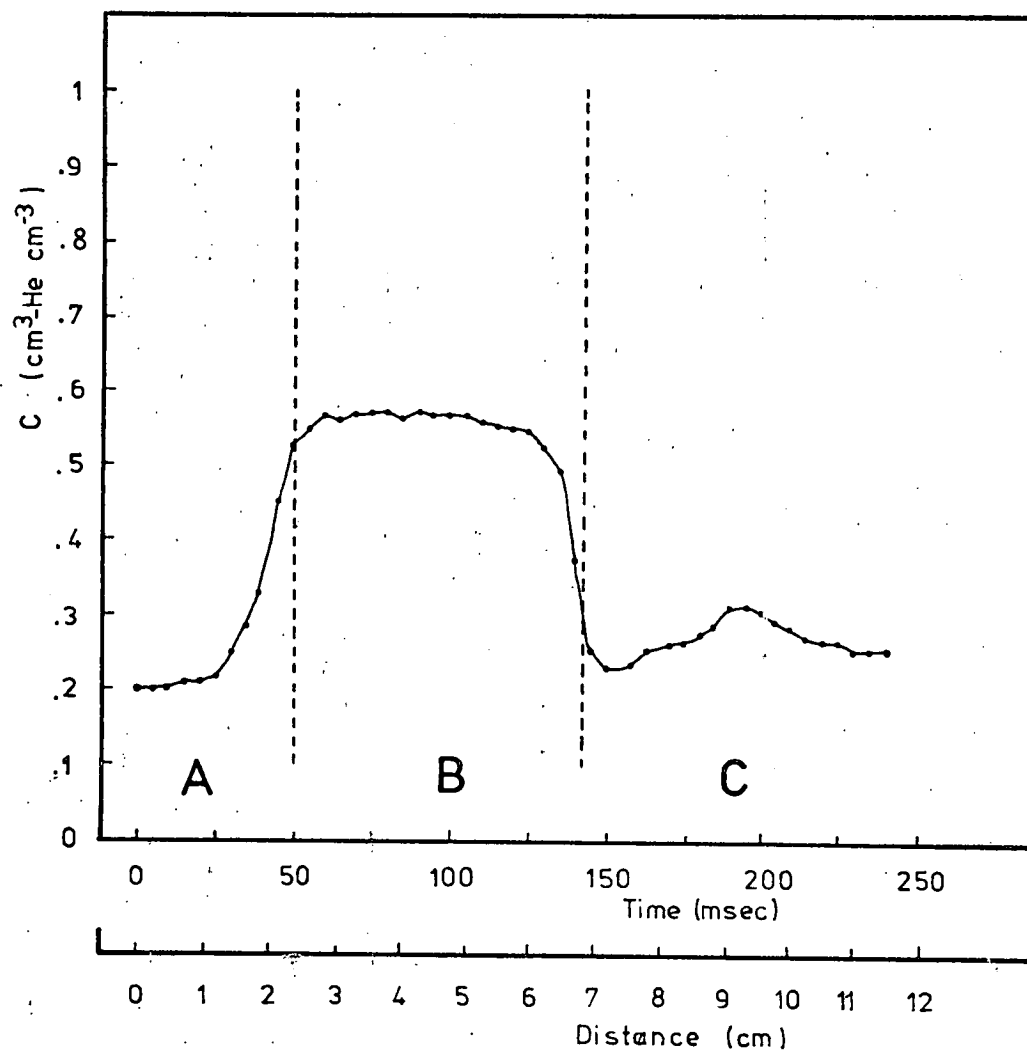


Figure 5.11 Tracer concentration within the bubble and its surroundings showing the effect of cloud shedding.

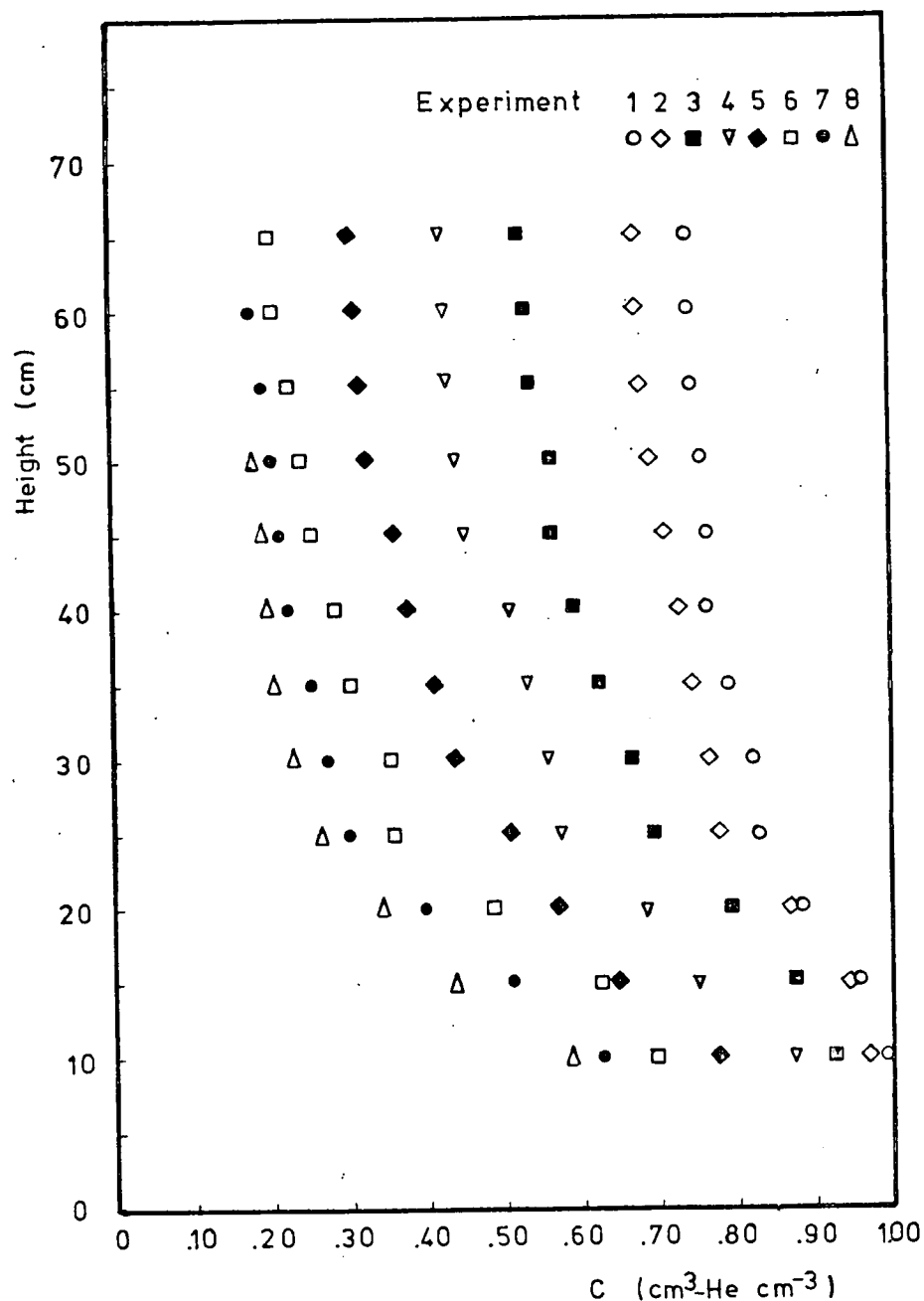


Figure 5.12 Axial profile of tracer concentration in the bubble. For each experiment and each level the plotted data represents the average of three runs, the standard deviation being always less than $0.013 \text{ cm}^3 \text{ He-cm}^{-3}$.

Figure 5.12 presents the axial profiles of tracer concentration obtained as above for heights between 10 and 65 cm, above the injector, at 5 cm intervals for all the experiments realised.

5.8 MASS TRANSFER BETWEEN BUBBLES AND DENSE PHASE

5.8.1 Determination of gas exchange coefficients

From the results reported in the previous sections the following assumptions can be made

- a) The bubble velocity and size remains constant with height.
- b) The bubble is well mixed.
- c) The gas in the dense phase is essentially in plug flow.
- d) The regular and rapid injection of bubbles will be conducive to the bed being at steady state.

The rate of gas transfer between a single rising bubble and the dense phase gives the equation

$$-\frac{d(V_b C_b)}{dt} = -V_b \cdot U_b \cdot \frac{dC_b}{dh} = K_o (C_b - C_p) \quad (5.4)$$

where V_b is the bubble volume, U_b its velocity, C_b and C_p the tracer concentration within the bubble and in the dense phase as measured by the probe at different levels in the bed, K_o is an overall gas exchange coefficient.

As bubbles were injected in the bed at a given frequency, N , a mass balance between the bubble phase, considered a continuum, and the dense phase gives

$$N V_b \frac{dC_b}{dh} + S U_{mf} \frac{dC_p}{dh} = 0 \quad (5.5)$$

where S is the cross sectional area of the bed.

From the above equations (5.4) and (5.5) it follows that

$$\frac{d \ln (C_b - C_p)}{dh} = - K_o \frac{S \cdot U_{mf} + N V_b}{S \cdot U_{mf} \cdot V_b \cdot U_b} \quad (5.6)$$

which integrates to give

$$2.303 \log \frac{C_b - C_p}{(C_b - C_p)_o} = - K_o \frac{S \cdot U_{mf} + N V_b}{S \cdot U_{mf} \cdot V_b \cdot U_b} (h - h_o) \quad (5.7)$$

Equation (5.7) permits the determination of the exchange coefficient, K_o , from the representation of $\log (C_b - C_p)$ against h . In equation 5.7 $(C_b - C_p)_o$ and h_o are integration constants.

Tracer concentration within the bubble and in the dense phase was determined at different levels, as explained in 5.7, and the values of $C_b - C_p$ at each level were then obtained from figures 5.8 and 5.12 and plotted against height, on semi-logarithmic co-ordinates. The results obtained are shown in figure 5.13 for all the experiments shown in table 5.1. The slope was determined by the least squares method, from which K_o was found for each one of the bubble sizes.

The above method of computation of K_o does not require the definition of the integration constants h_o and $(C_b - C_p)_o$ and the values of h presented in figure 5.13 refers to the height above the injector.

The mathematical formulation presented permits the calculation of the exchange coefficient during the steady rise of the bubble, thus ignoring any initial effects of mixing during the bubble injection.

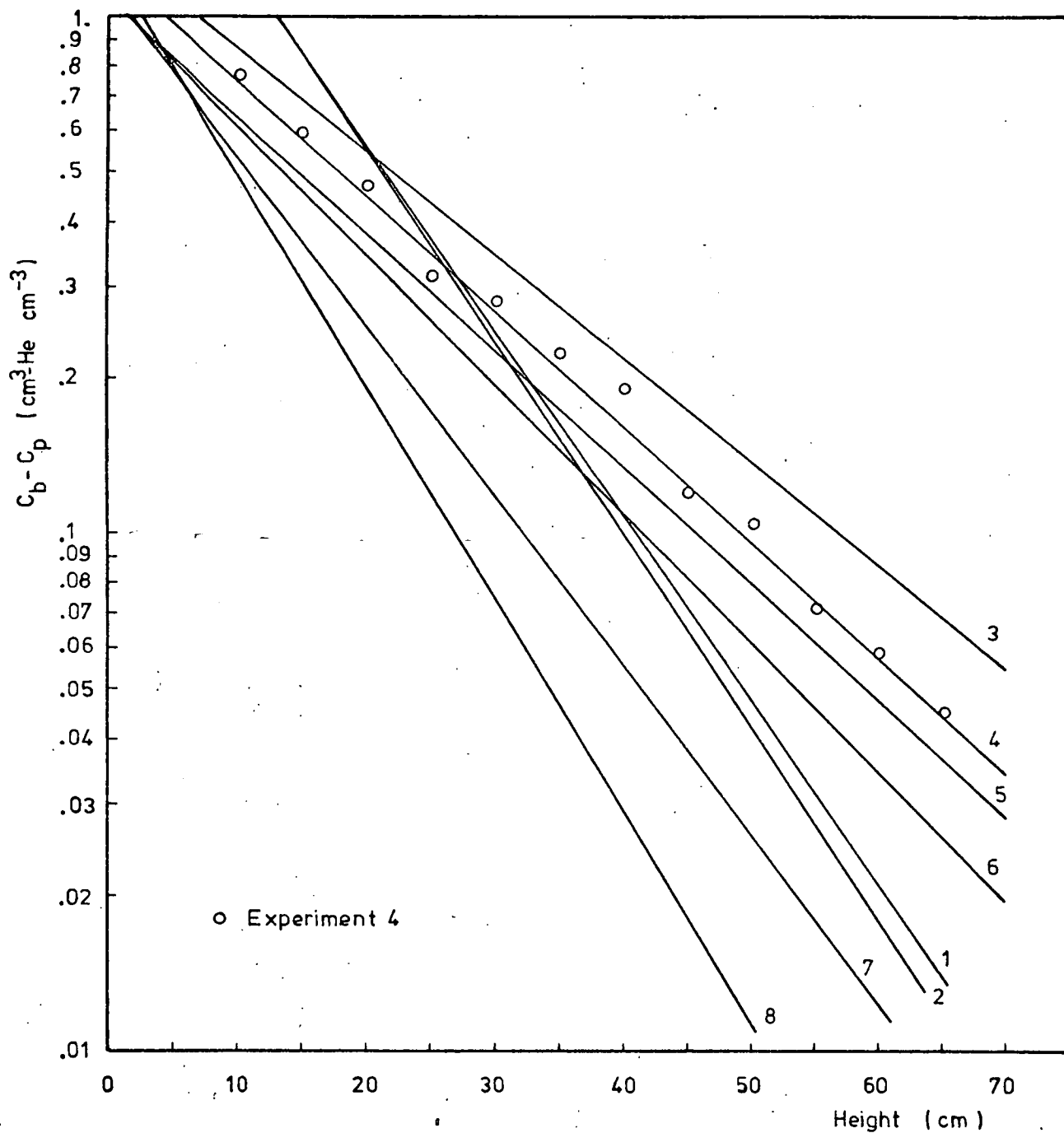


Figure 5.13 $C_b - C_p$ versus height. Reference height ($h = 0$) set at injector level.

5.8.2 Gas exchange during bubble injection

The above method of calculation of the exchange coefficient, although very simple, does not give information on gas mixing during bubble injection, as found by Szekely (74) and by Rowe et al (66) (this later work being realised in a two-dimensional fluidised bed).

The systematic work of Rowe et al (66) shows that the degree of mixing during bubble formation depends on the bubble size, being negligible in bubbles having a diameter greater than 10 cm and almost complete in those half this size. The similar findings of Szekely (74) in a three dimensional bed show a much more extensive initial mixing, but his experiments were realised in a shallow bed, thus presenting severe end effects since the bubble travels only a short distance between its injection and its bursting at the fluidised bed surface.

On the other hand, Davies and Richardson (20) found that bubbles need to travel a certain distance after injection before gas transfer may commence. They found that the above distance decreases with decreasing particle size and for each particle size is fairly independent of bubble volume. For bubbles with equivalent spherical diameter in the range 5.3 to 7.8 cm they found the above distance to be in the range 10 to 13 cm, when using particles with size 80 μm .

In our work it is found, from figure 5.13, that bubbles need to travel some distance, h_o , above the injector before gas interchange commences. Table 5.3 reports these values for each one of the bubble sizes, together with the bubble central length, L_b , and equivalent spherical diameter, D_e . It is seen that, in our case, h_o is of the same order of magnitude as the bubble central height, but it is interesting to note that for experiments 1 to 3 (large bubbles $D_e > 8 \text{ cm}$) h_o is greater than L_b and for experiments 4 to 8

TABLE 5.3

Experiment	D_e (cm)	L_b (cm)	h_o (cm)	$h_o - L_b$ (cm)
1	11.60	10.03	12.98	2.95
2	10.53	8.50	13.17	4.67
3	8.18	5.76	7.04	1.28
4	7.10	5.06	4.78	-0.28
5	6.01	4.12	1.82	-2.30
6	4.99	3.37	1.73	-1.64
7	4.33	3.06	1.52	-1.53
8	3.86	2.66	2.25	-0.41

($D_e < 8$ cm) h_o is smaller than L_b .

In the above analysis the reference height ($h = 0$) was set at the injector level although, if one considers that bubbles begin to move only after being totally formed, the reference height should be set at the bubble nose and the above values of h_o should be off-set by L_b , the bubble central length.

In table 5.2 the off-set values of h_o are also reported for all the experiments. It is seen they are positive for $D_e > 8$ cm and negative for $D_e < 6$ cm. This means that small bubbles ($D_e < 6$ cm) suffer some degree of exchange during formation and bubbles with equivalent spherical diameter greater than 6 cm are unaffected. In figure 5.14 the values of $C_b - C_p$ are plotted against the off-set height ($h - L_b$). From this representation, which was obtained by shifting the straight lines, of figure 5.13, L_b centimeters for each experiment, parallel to the h axis, it is seen that bubbles with $D_e < 6$ cm suffer a tracer gas dilution at the instant of formation. The bubble resulting from the injection of pure tracer has a tracer concentration that is only 90% of its original value. Such a degree of dilution seems, in our case, to be independent of bubble size; and lower in magnitude to that found by Rowe et al (66), although, in their case, bubbles are injected into a two-dimensional bed and perpendicularly to its direction of vertical motion.

The results of Davies and Richardson (20), mentioned earlier, are also explained by the above as their bubbles have equivalent spherical diameters ranging from 5.3 to 7.8 cm. In their case the initial mixing should be negligible and if the reference height is off-set by the bubble diameter their results are quite comparable to ours.

The above supports the initial mixing explanation as given

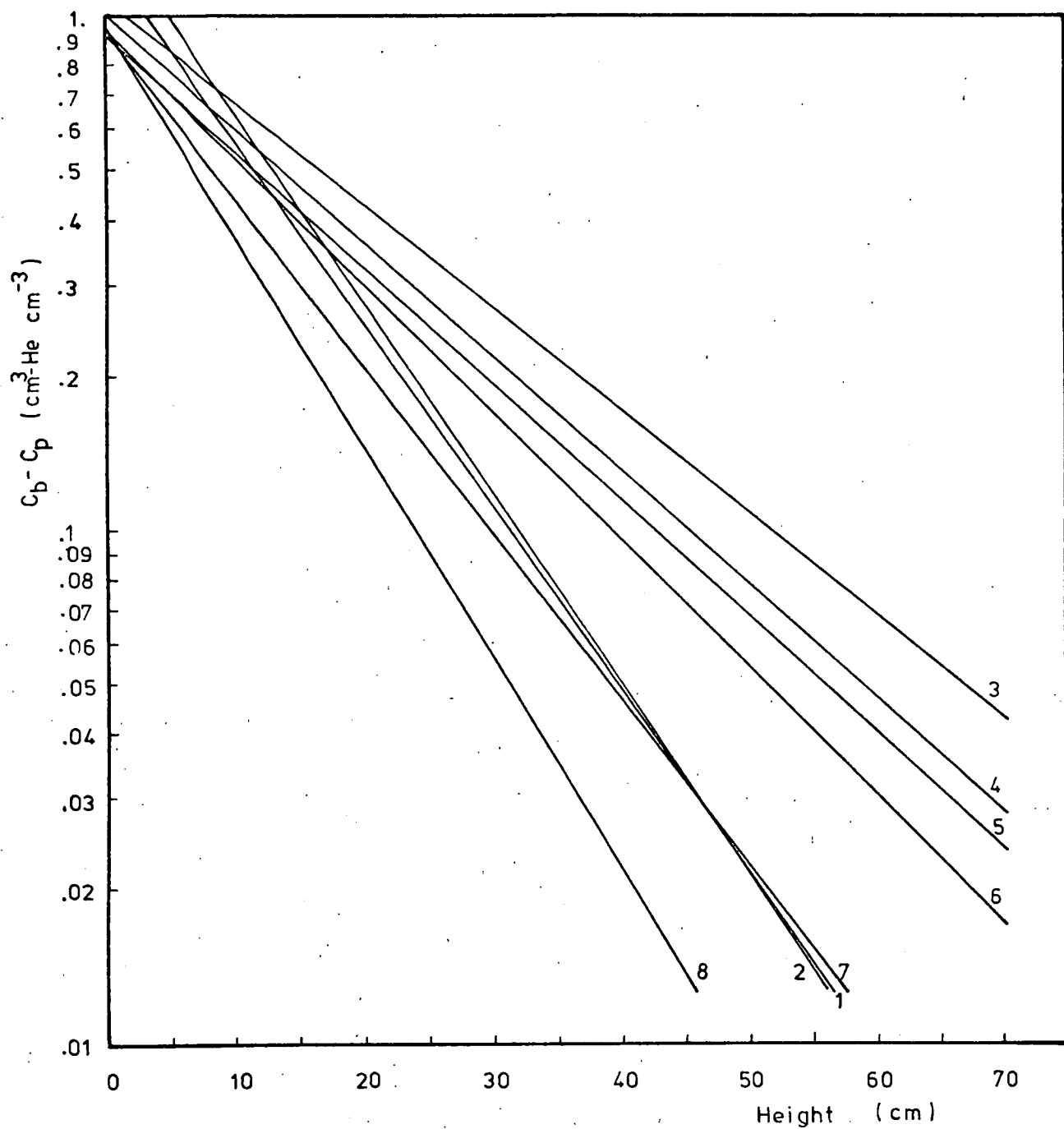


Figure 5.14

$C_b - C_p$ versus height. Reference height ($h = 0$) set at bubble nose after its complete formation.

by Rowe et al. (66) and its dependence on the initial bubble size.

5.8.3 Gas exchange coefficients

The overall mass transfer coefficients, K_o , were calculated from the slopes of the straight lines of the representation of $C_b - C_p$ versus h in semi-logarithmic scale.

The results obtained are reported in table 5.4. In order to compare our results with the different models proposed in the literature, and presented in chapter III, values of $K_v = K_o / V_b$ were calculated and are reported also in table 5.4. In figure 5.15 the above values of K_v , the exchange coefficient per unit of bubble volume, are plotted versus D_b and compared with the different existing models for gas exchange between bubbles and dense phase. It should be pointed out that the experimentally determined gas exchange coefficients are independent of bubble shape and the predictions from the different models assume bubbles to be spherical or spherical caps with indented rear surfaces.

Our experimentally determined gas exchange coefficients show agreement with the prediction of the different models as the mass transfer coefficient decreases with increasing bubble diameter; which is contrary to the findings of Rietema and collaborators (24) (62), although our technique is certainly more accurate than their techniques discussed in chapter III. The only exception to the above is shown by experiments 1 and 2 where bubbles are parabolic shape rather than a spherical cap or spherical shape as used by the different authors when presenting their models. The above anomaly, due to change in bubble shape, is discussed in the next section.

TABLE 5.4

Experiment	Bubble equivalent spherical diameter (cm)	K_o (cm ³ sec ⁻¹)	K_v (sec ⁻¹)
1	11.56	1094.0	1.35
2	10.53	997.6	1.63
3	8.18	371.6	1.30
4	7.10	298.4	1.60
5	6.01	208.3	1.83
6	4.99	145.1	2.23
7	4.33	115.4	2.75
8	3.86	93.7	3.09

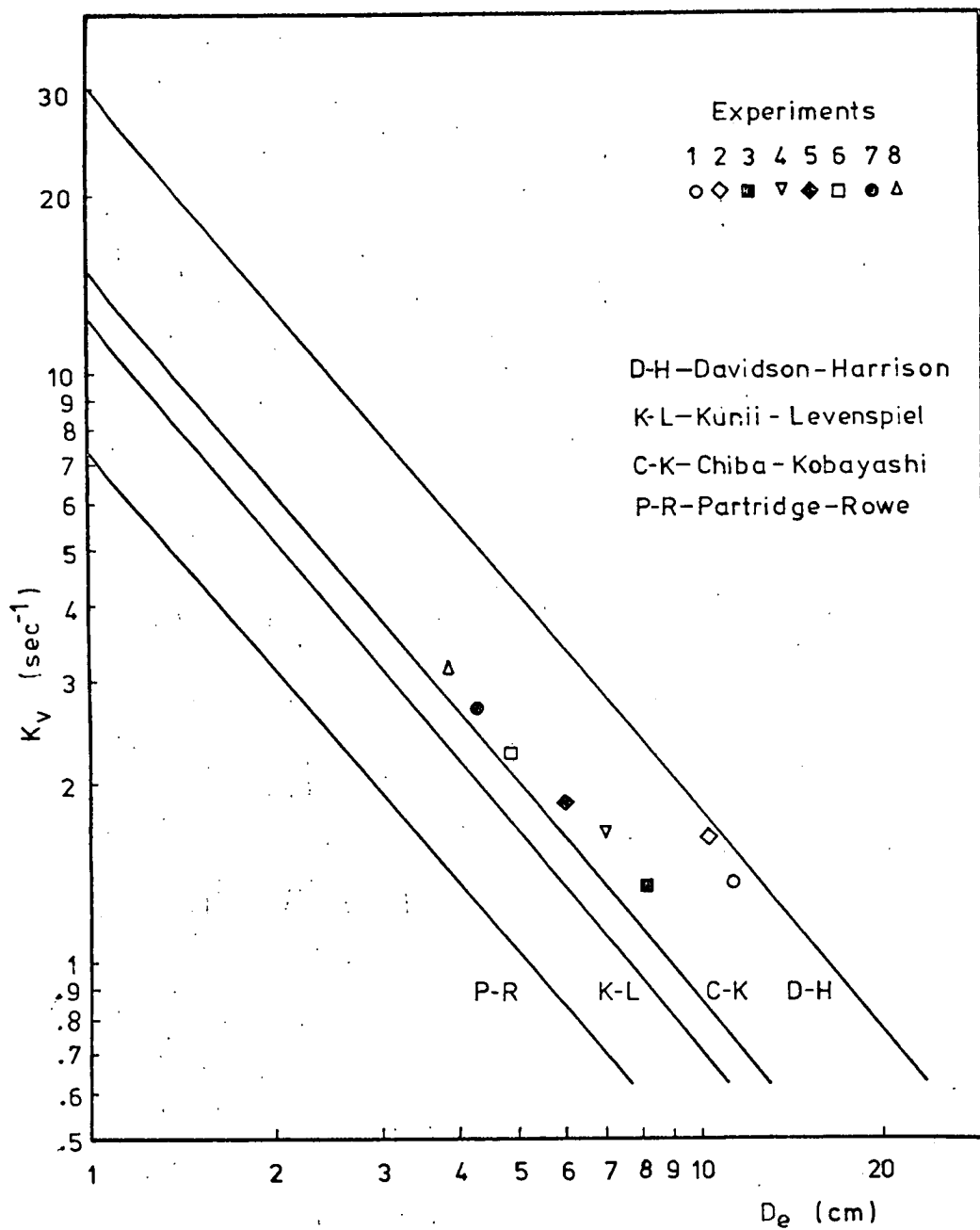


Figure 5.15

Overall mass transfer coefficients.

5.8.4 Mass transfer coefficients

From the literature discussed in chapter 3 it follows that several processes play a role in the total gas transfer between single bubbles and dense phase.

- Convective gas transfer due to gas circulation through the bubble.
- Diffusion through the cloud-bubble boundary.
- Diffusion through the cloud-dense phase boundary.
- Gas transfer due to cloud shedding.
- Gas transfer to solids in the cloud-solids raining through the bubble (in case they are porous or adsorbing).

The first three processes are the most well treated in the literature. Cloud shedding is a phenomena still not very well understood despite the preliminary work of Rowe et al. (70). Gas transfer due to adsorption on solids will make the study of the gas transfer very difficult indeed, both due to the fact of the unsystematic characteristics of particle raining through the bubble, not quantified until now, and due to the characteristics of gas adsorption processes.

Convective gas transfer due to gas circulation through the bubble was first proposed by Zenz (84), who suggested that the volumetric flow rate through the bubble was

$$Q_t = \frac{1}{2} \pi D_e^2 U_{mf} \quad (5.8)$$

Later, Davidson and Harrison (17) analysed the percolation of a void with diameter D_e by viscous fluid in a fixed bed of particles under

the influence of a uniform pressure gradient acting throughout the bed and proposed

$$Q_t = \frac{3}{4} \pi D_e^2 U_{mf} \quad (5.9)$$

The above equation 5.9 represents the flow of gas through the bubble for stationary bubbles. For very fast moving bubbles ($U_b/V_o \gg 1$, $V_o = U_{mf}/\epsilon_{mf}$) the above equation 5.9 represents the gas circulation between the bubbles and the cloud, as was seen in chapter II, figure 2.4. So, for very fast moving bubbles where circulation of gas appears, the theoretical foundations for a convective gas transfer contribution given by the above equation needs to be strengthened. Rowe and Partridge (68) and Stewart (73) suggest from their experiments that there is mixing within the main part of each bubble but streamline flow out of the bubble and into the particulate phase occurs. Suppose we assume that the gas in the cloud is in streamline flow until it reaches the wake region behind the bubble, where complete mixing takes place with the gas in the particulate phase at the same level as the wake, due to the fast moving particles in this region.

Then equation 5.9 represents the convective component of gas transfer between bubbles and dense phase, as the recirculating gas re-enters the bubble with the same composition as the dense phase. The tracer concentration profiles shown in figure 5.9 and 5.10 suggest that the above assumptions are acceptable as tracer concentration in the region behind the bubble is always lower than within the bubble and some mixing should then occur, although it is not possible to define accurately, in our case, the wake boundary.

Gas transfer due to diffusion was analysed theoretically by

various authors, as seen in chapter 3. Davidson and Harrison (17) estimated a mass transfer coefficient

$$K_g = 0.975 D_g^{\frac{1}{2}} D_e^{-\frac{1}{4}} g^{\frac{1}{4}} \quad (5.10)$$

based on an analogy with diffusion across the frontal surface of large bubbles rising in liquids and considering the bubble with a spherical cap shape having a flat base and a 50° angle between its rim and polar axis.

As the above bubble shape is far from the real bubble shape, Chiba and Kobayashi (13) proposed a similar estimate of the gas transfer coefficient based on Rowe's (59) bubble shape and using Murray's (52) stream function. Other models are discussed in chapter 3 and all of them require estimated properties of the bubbles.

In the present work bubble height and velocity were determined experimentally, and the ratio U_b/V_o was found to be always greater than 40. Thus, the cloud thickness is always less than $0.025 \cdot D_e$, as obtained from the Davidson and Harrison (17) bubble model. The height measured then can be considered as the height of the bubble-cloud.

If we now assume that the cloud has the same composition as the bubble, which is well mixed, gas is then exchanged between bubble and dense phase at the cloud boundary by diffusion. The exposure time, t_c , of all the elements of the interface moving from the top to the bottom of the bubble is the same

$$\text{i.e.:} \quad t_c = \frac{L_b}{U_b} \quad (5.11)$$

as $U_{mf} \ll U_b$.

The process of gas diffusion should be well represented by the Higbie (31) penetration model which is similar to liquid-film controlled diffusion from a rising bubble.

Thus, following Higbie (31)

$$K_g = \sqrt{\frac{4 D_g}{\pi t_c}} \quad (5.12)$$

using equation 5.11 above

$$K_g = \sqrt{\frac{4 D_g U_b}{\pi L_b}} \quad (5.13)$$

D_g being the diffusion coefficient, in our case that of helium in air, $D_g = 0.705 \text{ cm}^2 \text{ sec}^{-1}$.

The above equation gives the gas exchange due to diffusion over the frontal surface of the bubble, considered to be the unimpeded area, S_a

$$S_a = S_b \cdot \epsilon_{mf} \quad (5.14)$$

where S_b is the frontal surface of the bubble calculated as a solid or revolution, and is reported in table 5.1 for all the experiments.

Calderbank et al (9), studying the instantaneous rates of solution of gas bubbles in liquids, found that the contribution due to the wake is only a small percentage of the overall process. We consider then, in our case, that the gas transport between bubble and dense phase is a two-component process.

The overall gas transfer coefficient will be given by

$$Q_t = K_g \cdot S_a \quad (5.15)$$

Q_t , K_g and S_a are obtained from equations 5.9, 5.13 and 5.14

TABLE 5.5

Experiment	K_o (cm ³ sec ⁻¹)	D_e (cm)	Q_t (cm ³ sec ⁻¹)	t_c (sec)	K_g (cm sec ⁻¹)	S_a (cm ²)	$K_g.S_a$ (cm ³ sec ⁻¹)	$Q_t + K_g.S_a$ (cm ³ sec ⁻¹)
1	1094.0	11.60	182.3	0.1705	2.295	503.53	1155.3	1337.6
2	997.6	10.53	150.2	0.1510	2.438	403.14	982.9	1133.1
3	371.6	8.18	90.6	0.1075	2.890	95.80	276.9	367.5
4	298.4	7.10	68.3	0.1024	2.960	72.53	214.7	283.0
5	208.3	6.01	48.9	0.0866	3.219	51.19	164.8	213.7
6	145.1	4.99	33.7	0.0725	3.519	35.09	123.5	157.2
7	115.4	4.33	25.4	0.0698	3.586	26.97	96.7	122.1
8	93.7	3.86	20.2	0.0675	3.646	21.19	77.3	97.5

respectively.

In table 5.5 the experimentally determined gas exchange coefficient K_o obtained from the slopes of the straight lines in figure 5.13 and compared with the predictions from equation 5.15

The results show that the above model predicts the overall gas exchange coefficient within 10% for bubbles of spherical cap shape. For experiments 1 and 2 the bubbles were elongated in shape and the predicted results differ from the experimental ones by 22.2 per cent, in experiment 1, and 13.6 per cent in experiment 2. The deviation thus increases with bubble volume. In both cases it is interesting to note that the difference between the predicted overall gas exchange coefficient and the experimental one is of the same order of magnitude as the convective component of gas transfer. One possible explanation is due to the fact that in these two experiments bubbles are approaching the slug regime and in this case the convective component for gas transfer is

$$Q_t = \frac{1}{4} \pi D^2 U_{mf} \quad (5.16)$$

as proposed by Hovmand and Davidson (32); D being the diameter of the bed.

In our case, $Q_t = 104.8 \text{ cm}^3 \text{ sec}^{-1}$ and overall mass transfer coefficients become

Experiment	$Q_t + K_g \cdot S_a$
1	1260.1
2	1087.7

The deviation between K_o and $Q_t + K_g \cdot S_a$ being only 15.2 per cent in experiment 1 and 9.0 per cent in experiment 2.

The agreement seems now to be quite reasonable and the simple model of surface renewal at the cloud interface over the frontal surface of the bubble, accounts correctly for changes in observed mass-transfer due to changes in the frontal area of bubbles as well as their contact times.

The experimentally determined gas transfer coefficients

$$K_g = \frac{K_o - Q_t}{S_a} \quad (5.17)$$

are compared in figure 5.16 with the prediction from

$$K_g = \sqrt{\frac{4 D_g}{\pi t_c}} \quad (5.18)$$

Finally, the above results show the importance of gas transfer due to diffusion in the case of single bubbles rising in a fluidised bed of particles with low gas superficial velocity at incipient fluidisation. It is concluded from table 5.5 that the contribution of gas transfer due to diffusion to the overall process of gas transport between bubble and dense phase is always greater than 75 per cent, that is

$$\frac{K_g \cdot S_a}{K_o} > 0.75$$

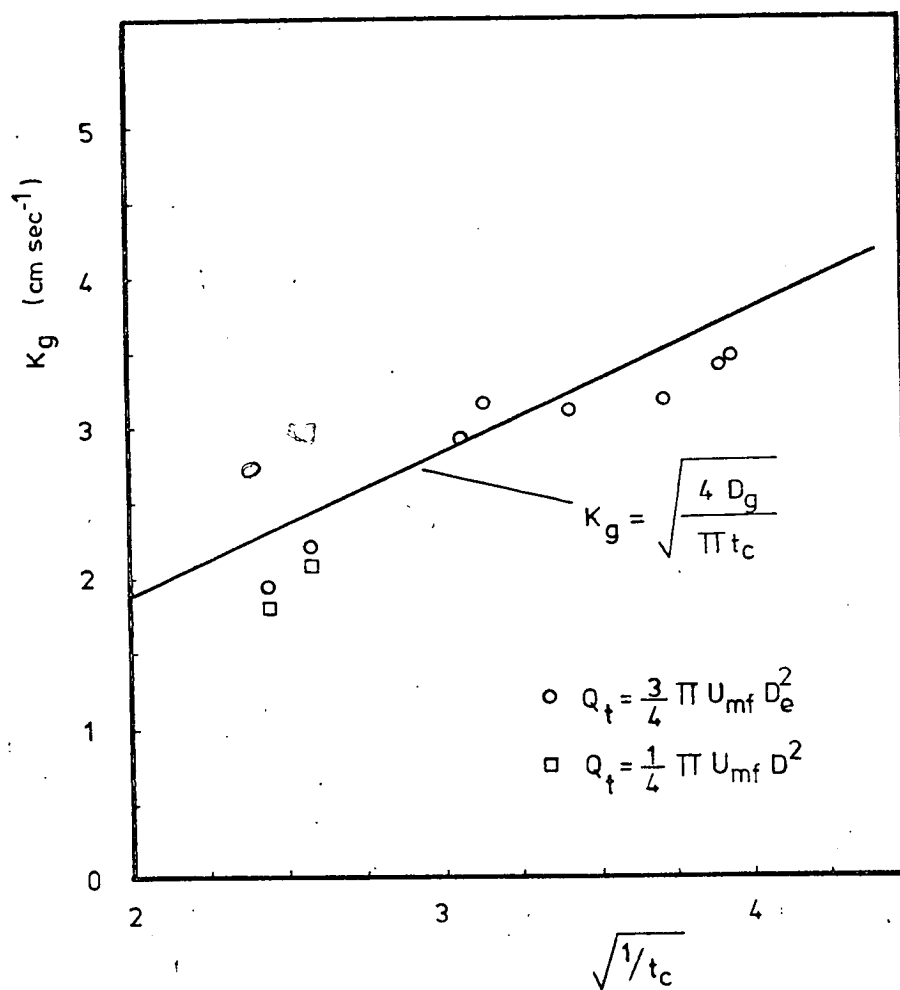


Figure 5.16

Mass transfer coefficients.

CHAPTER 6
COALESCENCE OF BUBBLES
GAS EXCHANGE BETWEEN BUBBLES
AND DENSE PHASE

6.1 INTRODUCTION

The phenomenon of coalescence has been studied, both theoretically and experimentally, by several researchers; Toor and Calderbank (11), Partridge and Rowe (58), Harrison and Leung (30), Matsumo and Rowe (48), Clift and Grace (14), Werther (79), Burgess and Calderbank (8) and Nguyen et al. (53).

The importance of the frequency and size of gas bubbles in determining the behaviour of aggregatively fluidised beds is well known and until now models that take into account the behaviour of bubbles and also consider coalescence, have been proposed by Partridge and Rowe (58), Toor and Calderbank (11) and Kato and Wen (36).

In the published works referred to above (36) (58) (11) only the interaction between bubbles concerned with growth or frequency has been studied and no attempt has been made to measure gas transfer between bubble and dense phase when bubbles coalesce.

This chapter describes an extension of the technique discussed in the previous chapters in order to study gas exchange between bubbles and dense phase when coalescence of two bubble occurs.

6.2 EXPERIMENTAL

As the objective of the work described in this chapter is the study of the effect of coalescence between two bubbles upon the gas transfer between the bubbles and dense phase, the injection system used in the study of gas transfer between a single rising bubble and dense phase was modified in order to inject bubble pairs.

Figure 6.1 shows the system used to introduce bubble pairs, consisting of two bubbles with equal volumes which were injected simultaneously in the bed. To achieve this, the injection system described in section 5.3 was duplicated and two expansion nozzles were inserted in the bed, in two radially opposed positions. One of the nozzles was slightly above the other, approximately 3.5 cm apart. The two expansion nozzles were connected to two solenoid valves and a helium tracer was supplied from two identical tanks, (volume 251 cm^3). The automatic injection of the bubble pairs was controlled by the computer, as before, and the injection frequency was set at 20 bubble pairs per minute for all the experiments.

The fluidised bed and all the other equipment and techniques were used without any modification.

Due to the small diameter of our column, (15.4 cm), which limits the range of volume of each bubble of the doublet, the pressure in the supply tank was controlled in order to obtain bubbles with a volume always less than 110 cm^3 . Also, in order to avoid coalescence of the two bubbles during injection, the nozzles were not only radially separated but were vertically separated as well.

6.3 INJECTION AND COALESCENCE

A visual observation of the movement of the injected bubbles was carried out. Pairs of air bubbles, of similar volume to those used in the gas transfer studies, were injected into the bed with different heights of fluidised particles. Thus by injecting bubble pairs and observing them breaking through the top of the bed with different heights, it was possible to detect a coalescence region

of the two bubbles. Bed height was increased in steps of 5 cm from a minimum bed height, the top of which was 10 cm above the top injector nozzle.

It was found that bubbles rise vertically or probably with a small movement towards the centre of the bed for distances, above the top injector, between 10 and 30 cm; it was also noticed that two bubbles break through the bed surface radially separated. For distances between 35 and 45 cm it appears that the breaking of the two bubbles was sometimes independent and sometimes appeared to be that of a single bubble. For distances greater than 50 cm it was found that only a single bubble resulting from the coalescence of the bubble pair was breaking through the bed surface.

6.4 BUBBLE VOLUME AND SHAPE

As the two bubbles were injected near the wall, no attempt was made to measure the volume and shape of the two individual bubbles. This was only done for the resulting bubble after coalescence and only at 65 cm above the injector. Only three different bubble sizes were used

TABLE 6.1

Experiment	Volume, (cm ³)	Shape	Equivalent spherical diameter (cm)
1	89.8	LEC	5.55
2	138.2	LEC	6.41
3	218.3	LEC	7.47

The volume of the individual bubbles is assumed to be half that of the resulting bubble.

TABLE 6.2

Experiment	Volume (cm ³)	Equivalent spherical diameter (cm)
1	44.9	4.40
2	69.1	5.09
3	109.1	5.92

6.5 EXPERIMENTAL RESULTS

TRACER CONCENTRATION

Tracer concentration in the dense phase and within the bubble was measured at different levels in the bed, between 20 and 65 cm above the gas distributor at 5 cm intervals.

For distances between 10 and 30 cm above the top injector nozzle the probe was vertically aligned with one or other of the injectors and the results averaged. For distances above 50 cm the probe was placed at the axis of the bed; and, finally, for distances between 30 and 50 cm the probe was positioned at various radial distances between the axis of the bed and the radial distance of the injectors.

In figures 6.2, 6.3 and 6.4 are presented the results obtained for each one of the experiments realized, where values of $C_b - C_p$ are plotted against height, on semi-logarithmic co-ordinates.

Tracer concentrations within the bubble show similar profiles

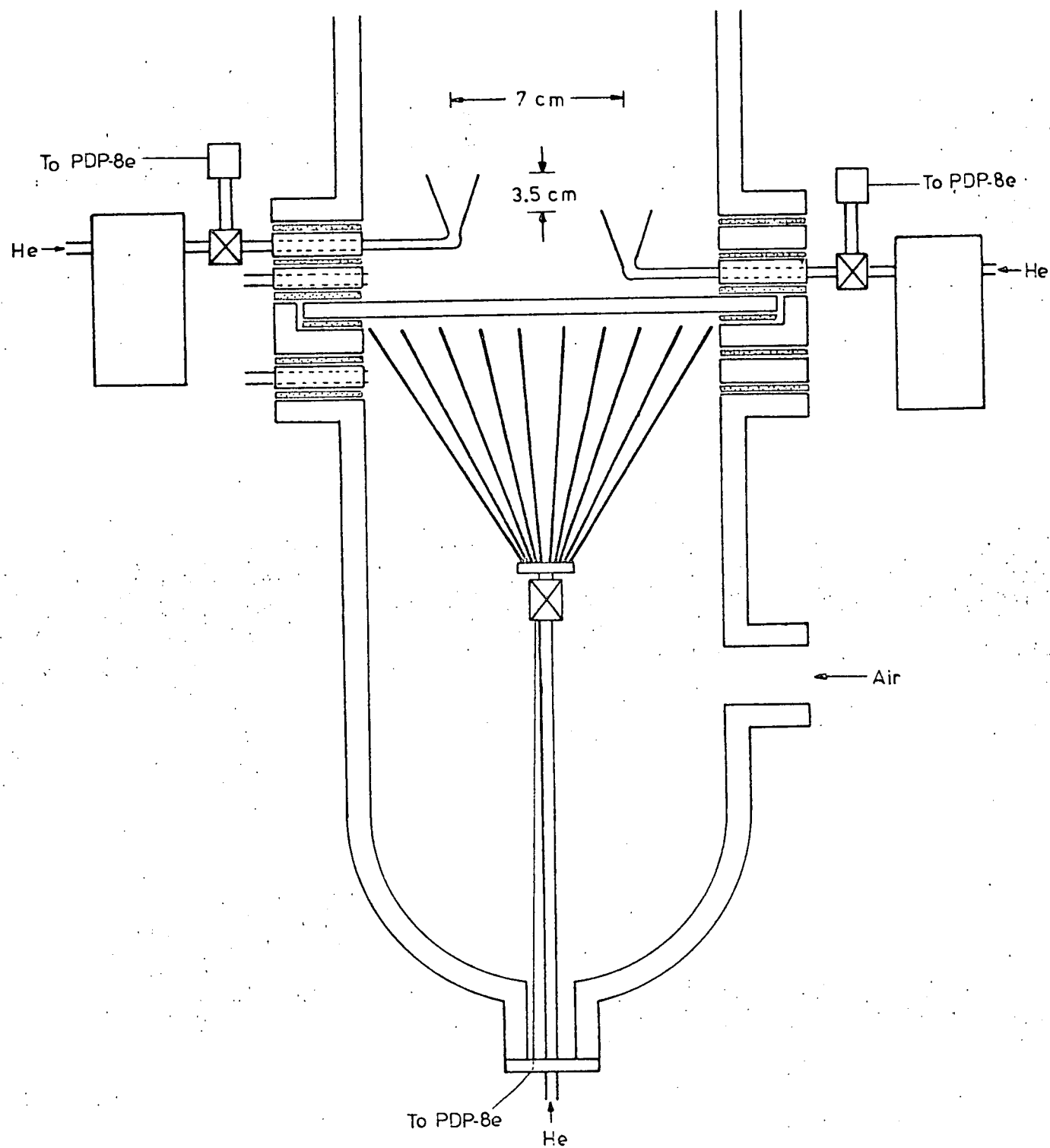


Figure 6.1 Experimental equipment for the introduction of bubble pairs into the fluidised bed.

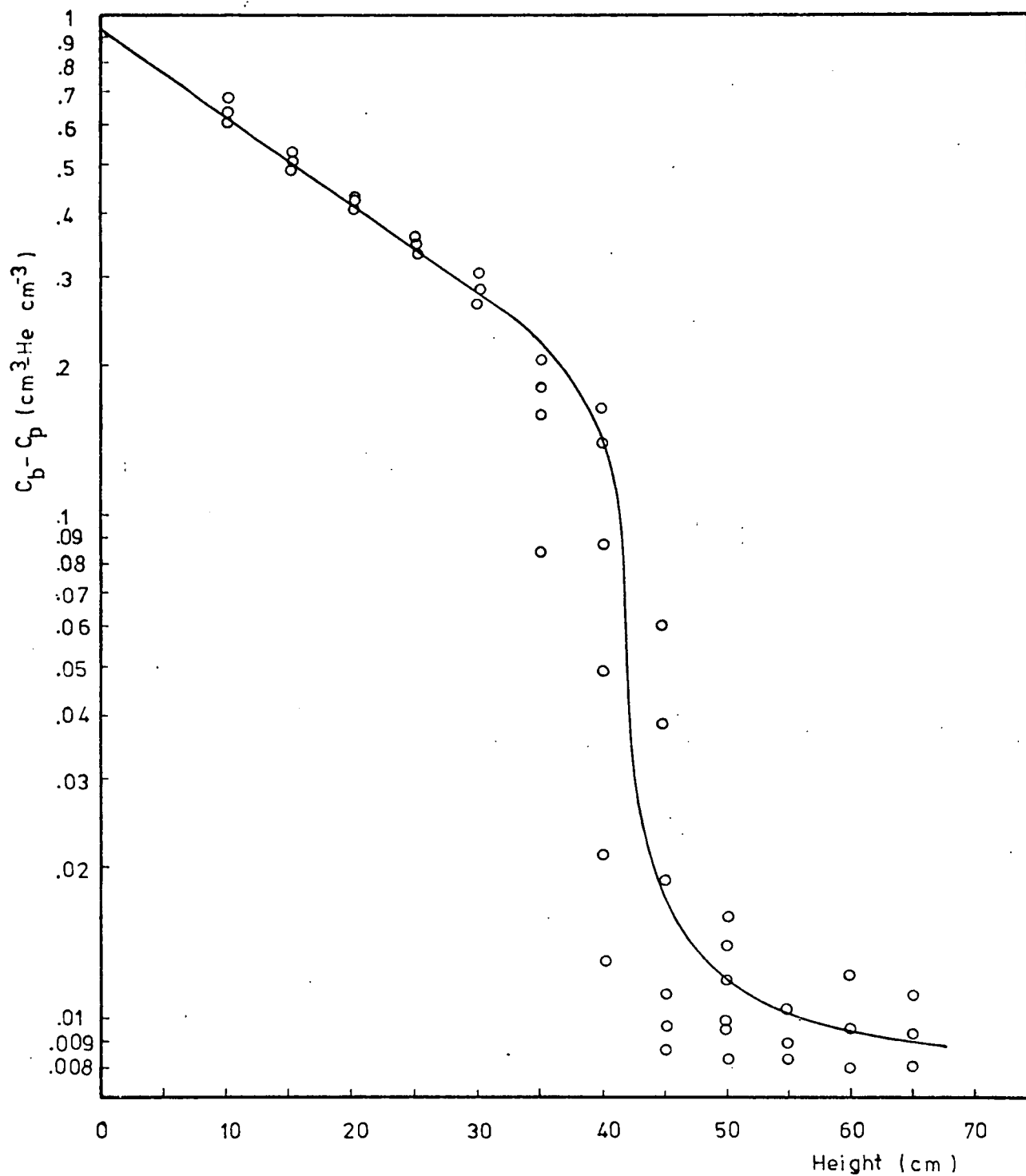


Figure 6.2

$C_b - C_p$ versus height. Rising and coalescence of two identical bubbles of equivalent spherical diameter

$$D_e = 5.55 \text{ cm.}$$

The above presents the data collected in all the runs.

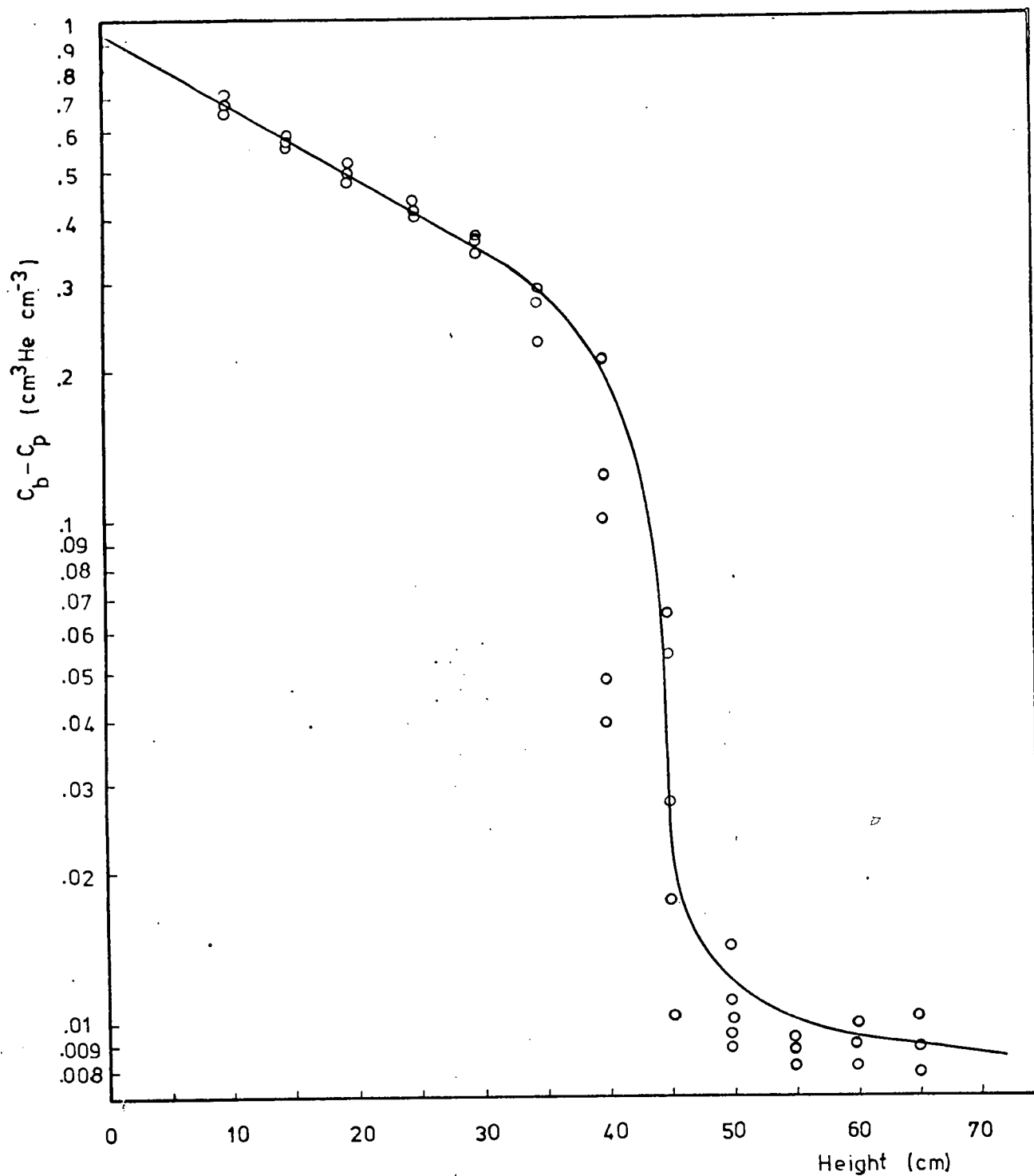


Figure 6.3

$C_b - C_p$ versus height. Rising and coalescence of two identical bubbles of equivalent spherical diameter

$$D_e = 6.41 \text{ cm.}$$

The above presents the data collected in all the runs.

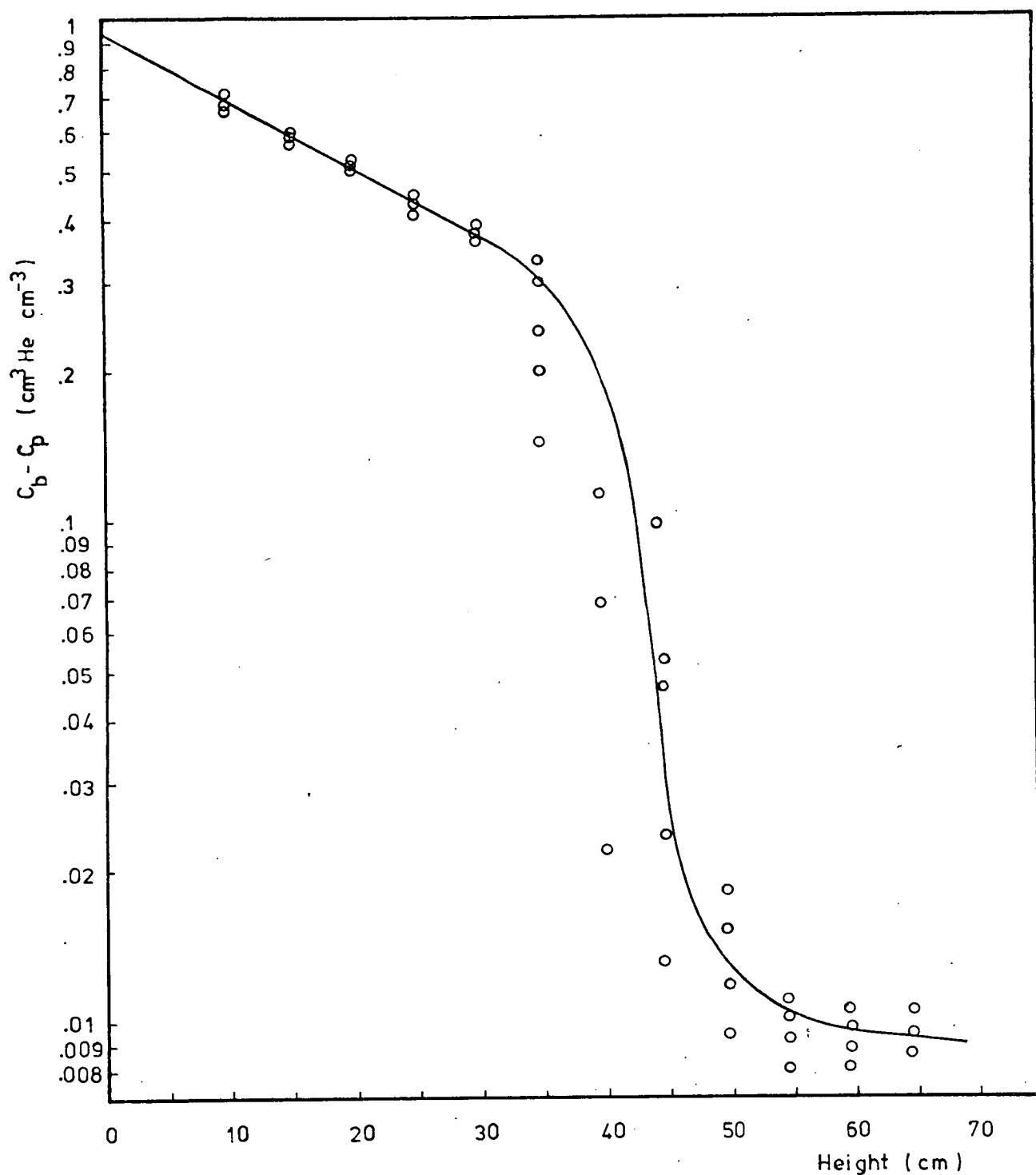


Figure 6.4

$C_b - C_p$ versus height. Rising and coalescence of two identical bubbles of equivalent spherical diameter

$$D_e = 7.47 \text{ cm.}$$

The above presents the data collected in all the runs.

to those obtained in the case of single bubbles and presented in figures 5.9 and 5.10. Also, radial profiles of concentration were similar to those shown in figure 5.7.

6.6 MASS TRANSFER BETWEEN BUBBLES AND DENSE PHASE

Figures 6.2, 6.3 and 6.4 show three distinguishable regions. The first between 0 and 30 cm corresponding to the rise of the isolated bubbles, the second between 40 and 50 cm where a scattering of $C_b - C_p$ values appears to exist ranging from the value for a single rising bubble to that of a bubble having practically the same composition as the dense phase, and finally a third region for heights above 50 cm where the concentration difference between the bubble and the dense phase seems to be non-existent as $C_b - C_p$ is always less than $0.01 \text{ cm}^3\text{-He cm}^{-3}$, which is the minimum value of $C_b - C_p$ that can be resolved by our detector from background noise in the CTHWB.

Gas exchange coefficients

Injected bubble

The overall gas exchange coefficient, K_o , was calculated from the slope of the straight lines of the representation of $C_b - C_p$ versus, h , on a semi-logarithmic scale, for the region between 0 and 30 cm above the top injector using equation (6.1)

$$\frac{d \ln (C_b - C_p)}{dh} = -K_o \frac{S \cdot U_{mf} + N \cdot 2 \cdot V_b}{S \cdot U_{mf} \cdot 2 \cdot V_b \cdot U_b} \quad (6.1)$$

The above equation is similar to equation (5.6), the only difference

being the multiplying factor 2 which is due to the fact that two bubbles are now injected simultaneously.

In table 6.3 below are presented the results obtained.

TABLE 6.3

Experiment	D_e (cm)	K_o (cm ³ sec ⁻¹)	K_v (sec ⁻¹)
1	4.40	128.6	2.86
2	5.09	158.9	2.29
3	5.92	212.7	1.95

The above results show that the overall mass transfer coefficients are slightly higher than predicted from interpolation of the results obtained when single bubbles were injected in the bed.

Coalescence

From figures 6.2, 6.3 and 6.4 it is apparent that there is complete mixing between bubble and dense phase when coalescence of the two bubbles, with equal volume, occurs.

Conclusion

The above results show clearly that the effect of coalescence of two identical bubbles does not lead solely to a bubble with a volume double that of the two coalescing ones. The resulting bubble is a new identity, totally independent of the coalescing bubbles as far as composition is concerned. There is complete mixing of gas between the bubble and the dense phase during the process of

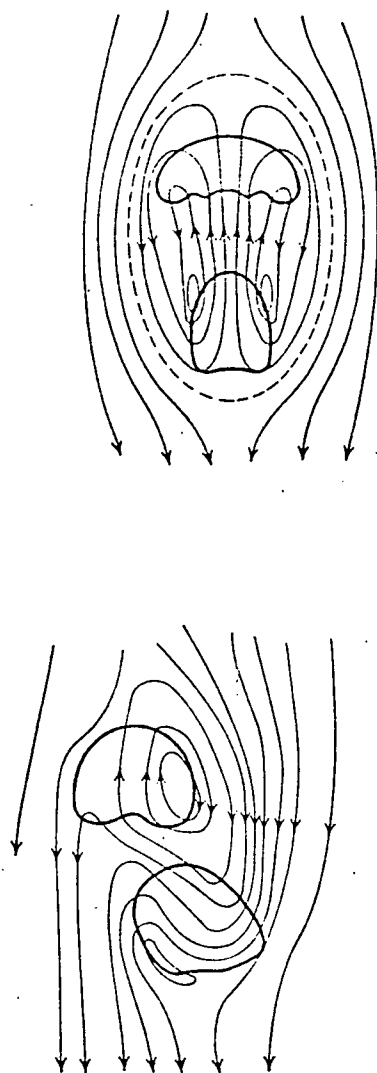


Figure 6.5

Gas streamlines relative to bubbles during
coalescence. (after Muchi et al. (51))

coalescence.

Muchi et al. (51) examined the streamlines of gas in the vicinity of two large coalescing bubbles and found that a single gas cloud can encompass both bubbles. So gas circulation exists between the two bubbles and through the dense phase between them, as shown schematically in figure 6.5. Coalescence thus promotes convective gas exchange between bubbles and dense phase.

The gas exchange coefficients between the "undisturbed" bubbles and the dense phase, obtained in the present set of experiments, are slightly higher than predicted from interpolation of the results obtained when single bubbles were injected in the bed. The difference, less than 10%, can be attributed to gas transfer enhancement during the earliest stage of the coalescence process, with the consequent increase in the convective component of the process of gas transfer and also an increase in the surface area of the trailing bubble due to its elongation prior to coalescence, which increases the diffusional transport of gas between the bubble and dense phase.

6.7 GAS EXCHANGE BETWEEN BUBBLES AND DENSE PHASE IN A FREELY BUBBLING BED

The results obtained in chapter 5 show that the transfer of gas between a single rising bubble and dense phase is a two component process. Gas exchange occurs by both convection and diffusion.

In a freely bubbling bed, where bubbles rise "undisturbed" and also coalesce, a third component for gas transfer seems to exist. The results presented in 6.6 show that there is complete mixing

between the gas in two coalescing bubbles and in the adjacent dense phase.

Calderbank and Toor (11) observed bubble frequencies at various flow rates and bed heights in a freely bubbling bed and found that bubble frequencies decrease rapidly with bed height, due to coalescence between rising bubbles.

The following relationship was then proposed

$$n = \alpha e^{-\beta h} \quad (6.2)$$

which relates the bubble frequency, n , with the bed height. α and β are two constants which were determined experimentally.

If we now assume that all gas in excess of that required to incipiently fluidise the bed, passes through the bed as bubbles which are uniform in size and equally distributed at each height in the bed, then

$$U - U_{mf} = N_v V_b U_b = n V_b \quad (6.3)$$

where V_b and U_b are the bubble volume and velocity and N_v the number of bubbles per unit of bed volume.

From equations 6.2 and 6.3, it results that the rate of bubble coalescence per unit of bed volume is

$$-\frac{d N_v}{d t} = \beta \cdot N_v \cdot U_b \quad (6.4)$$

The flow rate of gas between two coalescing bubbles and dense phase is then

$$Q_c = -2 V_b \frac{1}{N_v} \frac{d N_v}{d t} \quad (6.5)$$

and finally

$$Q_c = 2 \beta U_b V_b \quad (6.6)$$

In a freely bubbling bed the total flow rate of gas between the two phases is therefore

$$Q = Q_t + Q_d + Q_c \quad (6.7)$$

where

$$Q_t = \frac{3}{4} \pi U_{mf} D_e^2 \quad (6.8)$$

$$Q_d = K_g \cdot S_a \quad (6.9)$$

and

$$Q_c = 2 \beta U_b V_b \quad (6.10)$$

In the above equation (6.7) the first two terms account for gas transfer between the single rising bubble and the dense phase, and the final term accounts for gas transfer as a result of the coalescence of two identical bubbles.

In chapter 5 it was concluded that for single rising bubbles the contribution of gas transfer due to diffusion to the overall process of gas transfer between the bubble and the dense phase is always greater than 75 per cent. If we consider now that two bubbles, with the same characteristics as those used in experiments 3 and 8 and described in Tables 5.1 and 5.2, rise and coalesce in a fluidised bed as used in our experiments, the following values for each one of the three components of the overall process of gas transfer are obtained

TABLE 6.4

Experiment	D_e (cm)	Q_t (cm ³ sec ⁻¹)	Q_d (cm ³ sec ⁻¹)	Q_c (cm ³ sec ⁻¹)
3	8.18	90.6	276.9	392.4
8	3.86	20.2	77.3	30.5

and their relative contributions to the overall process are

TABLE 6.5

Experiment	Q_t (%)	Q_d (%)	Q_c (%)
3	11.9	36.4	51.7
8	15.7	60.4	23.8

In the above computation of Q_c the mean value of β ($\beta = 1.275$ cm⁻¹) reported by Calderbank and Toor (11) was used.

The above tables show clearly the significant contribution of coalescence to the overall mass transfer rate between bubbles and dense phase which should be included in the models for chemical reaction in fluidised beds of catalyst, as proposed by Calderbank and Toor (11).

CHAPTER 7

SUMMARY AND CONCLUSIONS

SUMMARY AND CONCLUSIONS

The development of a compound probe which comprises a five element conductivity sensor to detect bubble parameters and a constant temperature hot wire sensor to measure the concentration of a tracer, both in the dense phase and within the bubble, has provided a rigorous method for the simultaneous measurement of bubble velocity, shape and size, as well as mass transfer rates between bubble and dense phase in a gas fluidised bed of electrically conducting particles.

The analysis of the compound probe output involves a sequence of decision criteria which must be made in real time. The probe is therefore coupled to a high-speed digital computer which is capable of performing the required analysis fast enough to identify, and follow, those bubbles which strike the probe centrally.

The probe has measured the local bubble interface contour with high accuracy and this has provided a convenient method of accepting only those bubbles whose axis are coincident with the probe axis.

The technique was applied to the cases where single bubbles, or bubble doublets, were injected into an air fluidised bed of carbon particles, with size in the range 80 - 105 μm , which was maintained at incipient fluidisation, the bubble rising velocity being much greater than the dense phase interstitial gas velocity.

Bubbles of various sizes were injected into the bed and their velocities, shapes and sizes obtained from the pulse duration and delay times on the conductivity channels. Bubble velocities were measured at different heights in the bed and no variation with

height was found for a given bubble size. Bubble velocities, experimentally determined, were found to lie within the limits of the predictions for single rising bubbles and single rising slugs. These results, which are similar to those found by Calderbank et al. (9), reflect the wall proximity effects and the approach to slug flow conditions for sufficiently large bubbles.

The bubble shapes measured by the probe were found to be low eccentricity spherical caps for bubbles with equivalent spherical diameter less than 8.18 cm ($D_e/D \leq 0.53$), and bubbles with equivalent spherical diameter greater than 8.18 cm were found to have a paraboloid shape. Such transition is of similar character to the bubble shape variation in two dimensional beds obtained by Goldsmith and Rowe (27), which is due to a wall effect. In the first case, bubbles were found to have a shape similar to that found by Rowe and Partridge (69) using the x-ray photography of three dimensional bubbles in similar beds.

The low standard deviation of the measured bubble parameters at different bed heights permitted the conclusion that there is no noticeable bubble size variation with height.

Tracer concentration was obtained from a CTHWB output which was stored in computer memory. The helium concentration associated with the bubble and its near environment is given by the CTHWB at particular times determined by the signal of the three upper contacts of the conductivity sensor. As the frequency of response of the conductivity contacts is very high, the accuracy of synchronization between the signal of the three upper contacts and the CTHWB output will depend only on the response time of the analytical sensor. A test was carried out to measure the latter, which proved to be of similar order of magnitude to that of the conductivity sensor. Thus,

the CTHWB output was seen to be adequately synchronized with the signal of the three upper contacts and hence with the bubble position.

Tracer concentration was detected near to and within a rising bubble as well as, in the dense phase, at the mean time between the passage of two consecutive bubbles through a given level in the bed. Radial profiles of tracer concentration in the dense phase, recorded at various bed heights for all experiments in which bubble volume and injection frequency was varied over a wide range, were in all cases flat. This is certainly due to the essential plug flow nature of the flow when the gas superficial velocity is low, and also due to the high diffusivity of helium in air. It was also found that tracer concentration increases with height.

Tracer concentration near to and within the bubble was obtained from the CTHWB output which was stored in computer memory according to the output of the three upper contacts of the conductivity sensor. The tracer concentration within the bubbles shows an almost flat profile, the tracer concentration decreasing slightly to the rear of the bubble which is certainly due to the fact that the bubble undergoes mass transfer with the dense phase as it rises.

The absence of significant concentration gradients within the bubble shows that the principal resistance to mass transfer lies at the bubble-dense phase interface.

The concentration recorded near to and within a rising bubble showed that the principal concentration difference resided close to the frontal interface of the bubble (the non-wake region). The concentration in the bubble wake does not greatly differ from that in the adjoining dense phase so that this region is well mixed, but mass transfer between the rear surface of the bubble and the wake is slow since it is not apparently necessary to make allowance for it in the

mass-transfer theoretical model discussed below.

From the axial profiles of tracer concentration in the bubble and in the dense phase, gas transfer rates between the two phases were calculated.

The overall steady-state rates of mass transfer obtained experimentally compare well with that predicted from a simple surface renewal model over the frontal interface in which the unobscured area for transfer was calculated from the dense phase porosity.

Although steady state mass transfer rates from the rear surface of the bubble appeared to be small, it was not possible using the present technique to identify, with certainty, periodic wake-shedding phenomena such as are known to exist.

Further experiments were conducted in which two helium bubbles were caused to coalesce as they rose through the air fluidised bed. It was shown that the resulting single bubble almost instantaneously assumed a concentration very close to that of the adjacent dense phase. This observation makes it clear that bubble coalescence is responsible for a significant contribution to the overall mass-transfer rate. It is shown how models for chemical reaction in fluidised beds of catalyst may be modified accordingly.

NOMENCLATURE

a	=	radius of bubble (cm)
A	=	radius of cloud (cm)
C	=	concentration of tracer ($\text{cm}^3 \text{ cm}^{-3}$)
C_b	=	concentration of tracer in the bubble ($\text{cm}^3 \text{ cm}^{-3}$)
C_p	=	concentration of tracer in the dense phase ($\text{cm}^3 \text{ cm}^{-3}$)
D	=	diameter of fluidised bed (cm)
D_b	=	diameter of bubble (cm)
D_c	=	diameter of cloud (cm)
D_e	=	equivalent spherical diameter of the bubble (cm)
D_g	=	diffusivity of tracer in fluidising gas ($\text{cm}^2 \text{ sec}^{-1}$)
f_w	=	wake fraction
F_1	=	fluidising gas flow rate ($\text{cm}^3 \text{ sec}^{-1}$)
F_2	=	flow rate of gas in bubbles ($\text{cm}^3 \text{ sec}^{-1}$)
g	=	gravitational acceleration (cm sec^{-2})
h	=	height in the fluidised bed (cm)
h'	=	mass transfer coefficient (cm sec^{-1})
h_d	=	gas transfer coefficient (cm sec^{-1})
h_o	=	reference height (cm)
k	=	thermal conductivity of gas ($\text{cal/sec cm}^2 (^\circ\text{C/cm})$)
K	=	constant equation (2.1)
K'	=	constant equation (2.3)
K^*	=	mass transfer coefficient (cm sec^{-1})
K_c	=	mass transfer coefficient (cm sec^{-1})
K_{bc}	=	gas exchange coefficient between bubble and cloud (cm sec^{-1})
K_{cd}	=	gas exchange coefficient between cloud and dense phase (cm sec^{-1})
K_{bd}	=	overall mass transfer coefficient between bubble and dense phase (cm sec^{-1})
K_g	=	mass transfer coefficient (cm sec^{-1})
K_o	=	overall gas exchange coefficient ($\text{cm}^3 \text{ sec}^{-1}$)
K_p	=	permeability constant equation (2.11)
K_v	=	gas exchange coefficient per unit of bubble volume (sec^{-1})
L_b	=	bubble vertical axis length (cm).
L_d	=	distance between bubble nose and interface at radial distance S_r (cm)
L_o	=	off-centre bubble height (cm)
n	=	bubble frequency in a freely bubbling bed (sec^{-1})
N	=	bubble injection frequency (sec^{-1})
Nu	=	Nusselt number
N_v	=	number of bubbles per unit of bed volume (cm^{-3})

P	=	pressure (cm-H ₂ O)
Q	=	overall mass transfer coefficient (cm ³ sec ⁻¹)
Q_c	=	gas exchange coefficient for coalescence (cm ³ sec ⁻¹)
Q_d	=	gas exchange coefficient for diffusion (cm ³ sec ⁻¹)
Q_t	=	gas exchange coefficient for through flow (cm ³ sec ⁻¹)
r	=	polar co-ordinate
r_b	=	bubble frontal nose (cm)
R_b	=	bubble nose radius (cm)
Re	=	Reynolds number
S	=	cross sectional area of the bed (cm ²)
S_a	=	bubble frontal unimpeded area (cm ²)
S_b	=	bubble frontal area (cm ²)
Sc	=	Schmidt number
S_d	=	vertical separation distance between leading and upper contacts (cm)
Sh	=	Sherwood number
S_r	=	radial separation of contact five (cm)
S_t	=	total surface area of bubble (cm ²)
t	=	time (sec)
t_c	=	exposure time (sec)
T	=	bubble injection period (sec)
T_c	=	pulse rise time (sec)
T_i	=	pulse duration on contact i (sec)
$T_{l,j}$	=	pulse delay time between leading and contact j (sec)
U_r, U_o	=	components of the velocity of fluidising gas in polar co-ordinates
U_b	=	bubble velocity (cm sec ⁻¹)
U_{mf}	=	superficial velocity of gas at minimum fluidisation (cm sec ⁻¹)
U_r	=	relative velocity between bubble and interstitial gas (cm sec ⁻¹)
U_s	=	slug velocity (cm sec ⁻¹)
v_x	=	gas velocity through the curved surface of the bubble (cm sec ⁻¹)
V	=	voltage (V)
V_b	=	bubble volume (cm ³)
V_i	=	voltage (V)
V_o	=	interstitial velocity of fluidising gas (cm sec ⁻¹)
V_t	=	threshold voltage (V)
z	=	bed height (cm)
α	=	constant equation (6.2)
β	=	constant equation (6.2)
ϵ_{mf}	=	bed porosity at incipient fluidisation

θ	=	polar co-ordinate
θ_w	=	wake angle
μ	=	viscosity of gas ($\text{gr cm}^{-1} \text{sec}^{-1}$)
ρ	=	specific gravity (gr cm^{-3})
ϕ	=	velocity potential
γ	=	stream function

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APPENDICES

APPENDIX AA.1 BUBBLE SHAPE SELECTION

The constraints imposed by the real-time computer speed restrict the shape measurement facilities to only one of the probe channels of information. The probe array therefore offers only a relatively small amount of information upon which to base the bubble volume and area computation. It is in this frame of reference that the following fitting procedure was developed.

Firstly, consider the information reported by the probe after a successful hit. Figure A.1 shows the variables reported, where:

L_b is the central bubble vertical length

L_d is the distance between bubble pole and bubble interface at the radial distance S_p

L_o is the bubble vertical chord length at the radial distance S_r

and

R_b is the bubble nose radius

$$R_b = \frac{S_r^2 + L_d^2}{2 \cdot L_d} \quad (A.1)$$

In order to process the data, the following assumptions are made

1. The central length L_b is the true central length on the bubble axis as the probe discrimination level was always set at 0.95.
2. The successful bubble hit is generated by a bubble which

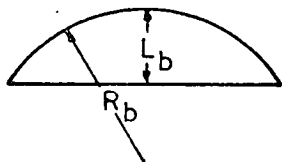
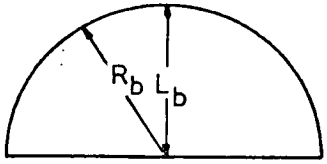
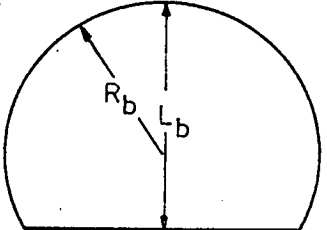
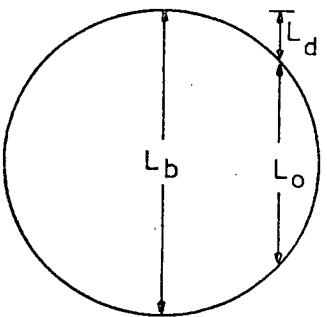
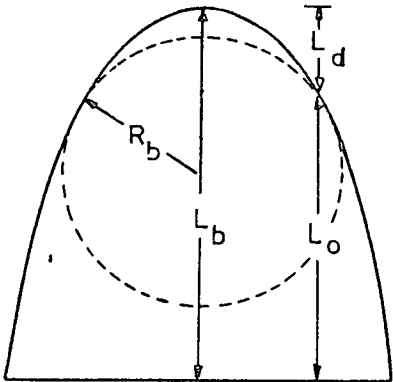
CONDITION	REPRESENTATION	SHAPE
$L_b < R_b$		SPHERICAL CAP
$L_b = R_b$		HEMISPHERE
$L_b > R_b$ and $L_b < 2.R_b$		LOW ECCENTRICITY CAP,
$L_b = 2.R_b$ $L_d + L_o < 2.R_b$		SPHERE
$L_b \geq 2.R_b$ and $L_d + L_o \geq 2.R_b$		PARABOLOID

Figure A.1 Bubble shape selection criteria.

is moving vertically and thus parallel to the probe axis.

3. The bubble is symmetrical in shape.

The problem now becomes one of selecting an appropriate bubble shape to fit the bubble parameters L_b , L_d , L_o and R_b for the fixed separation distance, S_r , between the central contact (C1) and the off-centre contact (C5).

As outlined in chapter 2, x-ray examination of single bubbles in three dimensional beds (69) indicates that bubbles are of spherical cap shape with an indented rear surface. Also, using photography (27) it was found that in a given fluidised bed when bubble size increases in such a way that wall effects begin to be important, a transition from spherical shaped bubbles to elongated bubbles, in a vertical direction, is obtained.

The shape selection criteria for a given bubble passing the probe coaxially will then be simplified to the simple comparison of the probe output and the directly derived bubble nose radius, with the properties of the solids of revolution representing the experimentally determined bubble shapes, as obtained by Rowe and collaborators (69) (27).

Figure A.1 gives the selection criteria in tabular form, and a discussion for each of the criteria follows.

1. $L_b < R_b$

The bubble central height is less than the bubble nose radius. The shape assumed for the bubble is a
SPHERICAL CAP.

2. $L_b = R_b$

The bubble fitted to the data under these conditions is

HEMISPHERICAL. This shape is based on the fact that the single bubbles measured by Rowe and Partridge (69) using x-rays are very close to hemispherical for the same types of fluidised particulate material.

3. $L_b < 2 R_b$

This condition accounts for those bubbles which are of the shape detected by Rowe (143) using x-rays where the wake fraction is less than $\frac{1}{2}$ the total sphere volume. The shape fitted here is given by a LOW ECCENTRICITY CAP.

4. $L_b = 2 R_b$ and $L_d + L_o < 2.R_b$

This condition corresponds to the limit of the case presented in 3 above. It corresponds to a SPHERICAL BUBBLE in which the wake fraction is zero.

5. $L_b \geq 2.R_b$ and $L_d + L_o \geq 2.R_b$

This condition reflects the fact that the bubble is elongated due to wall effects, as detected by Goldsmith and Rowe (27) and the most convenient shape to fit the probe output data is the PARABOLOID of revolution.

A.2 BUBBLE VOLUME AND SURFACE AREAS

The following represent the geometric relationships used to compute the volume (V_b) and surface areas, frontal (S_b) and total (S_t), for each one of the solids of revolution reported by the probe.

1. SPHERICAL CAP

$$V_b = \pi \cdot L_b \cdot \frac{3 \cdot R_r^2 + L_b^2}{6} \quad (A.2)$$

$$S_b = 2 \cdot \pi \cdot R_b \cdot L_b \quad (A.3)$$

$$S_t = S_b + \pi R_r^2 \quad (A.4)$$

where R_r is the radius of the base of the bubble

$$R_r = \sqrt{2 \cdot L_b \cdot R_b - L_b^2} \quad (A.5)$$

2. HEMISPHERE

$$V_b = \frac{2}{3} \pi R_b^3 \quad (A.6)$$

$$S_b = 2 \cdot \pi \cdot R_b^2 \quad (A.7)$$

$$S_t = 3 \cdot \pi \cdot R_b^2 \quad (A.8)$$

3. LOW ECCENTRICITY CAP

$$V_b = \frac{4}{3} \cdot \pi \cdot R_b^3 - \frac{\pi}{6} \cdot L_w \cdot (3 R_r^2 + L_w^2) \quad (A.9)$$

$$S_b = 4 \cdot \pi \cdot R_b^2 - 2 \cdot \pi \cdot R_b \cdot L_w \quad (A.10)$$

$$S_t = S_b + \pi \cdot R_r^2 \quad (A.11)$$

L_w is defined as

$$L_w = 2 \cdot R_b - L_b \quad (A.12)$$

and R_r is obtained from equation A.5.

4. SPHERE

$$V_b = \frac{4}{3} \pi R_b^3 \quad (A.13)$$

$$S_b = S_t = 4 \cdot \pi \cdot R_b^2 \quad (A.14)$$

5. PARABOLOID

The general equation of a parabola is of the form

$$Y = a X^2 + b \quad (A.15)$$

Applying the boundary conditions

$$Y = L_b \quad X = 0$$

$$Y = L_o \quad X = S_r$$

then

$$b = L_b \quad (A.16)$$

and

$$a = - \frac{L_b - L_o}{S_r^2} \quad (A.17)$$

From geometry

$$L_b = L_o + L_d \quad (A.18)$$

then

$$Y = - \frac{L_d}{S_r^2} X^2 + L_b \quad (A.19)$$

The volume and surface areas of the paraboloid of revolution will be

$$V_b = \frac{\pi}{2} \cdot R_r^2 \cdot L_b \quad (\text{A.20})$$

$$S_b = 4 \cdot \pi \cdot L_b \cdot R_r \quad (\text{A.21})$$

$$S_t = S_b + \pi \cdot R_r^2 \quad (\text{A.22})$$

where R_r , the radius of the base of the paraboloid is

$$R_r = \sqrt{\frac{S_r^2 \cdot L_b}{L_d}} \quad (\text{A.23})$$

APPENDIX B

B.1 CTHWB - Principle and calibration

The hot-wire anemometre was first used by Koepsel (38) in the measurement of very small air velocities and was brought to the general knowledge by King (37).

In principle the constant temperature anemometer consists of a Wheatstone Bridge and a servo amplifier. A schematic representation is shown in figure B.1. When the bridge is in balance no voltage difference exists between the inputs of the servo amplifier. Any change in the flow, or fluid properties acting on the probe, will affect the temperature of the sensor, causing it to grow cooler or hotter as the case may be. The resultant resistance change brings about a voltage difference at the servo amplifier inputs. The output voltage of the servo amplifier is applied to the bridge so that the original sensor temperature is restored by increasing or reducing the bridge operating voltage.

Let us now consider a thin wire stretched between two needles which serve as current leads. When a voltage, V , is applied across the wire, it will reach a given temperature, T , which can be considered constant over its length. If gas at constant temperature, T_0 , flows steadily past the wire, heat transfer by forced convection between the wire and the gas occurs.

Under steady state

$$RI^2 = h A (T - T_0) \quad (B.1)$$

where R is the wire resistance, A its area, h the heat transfer

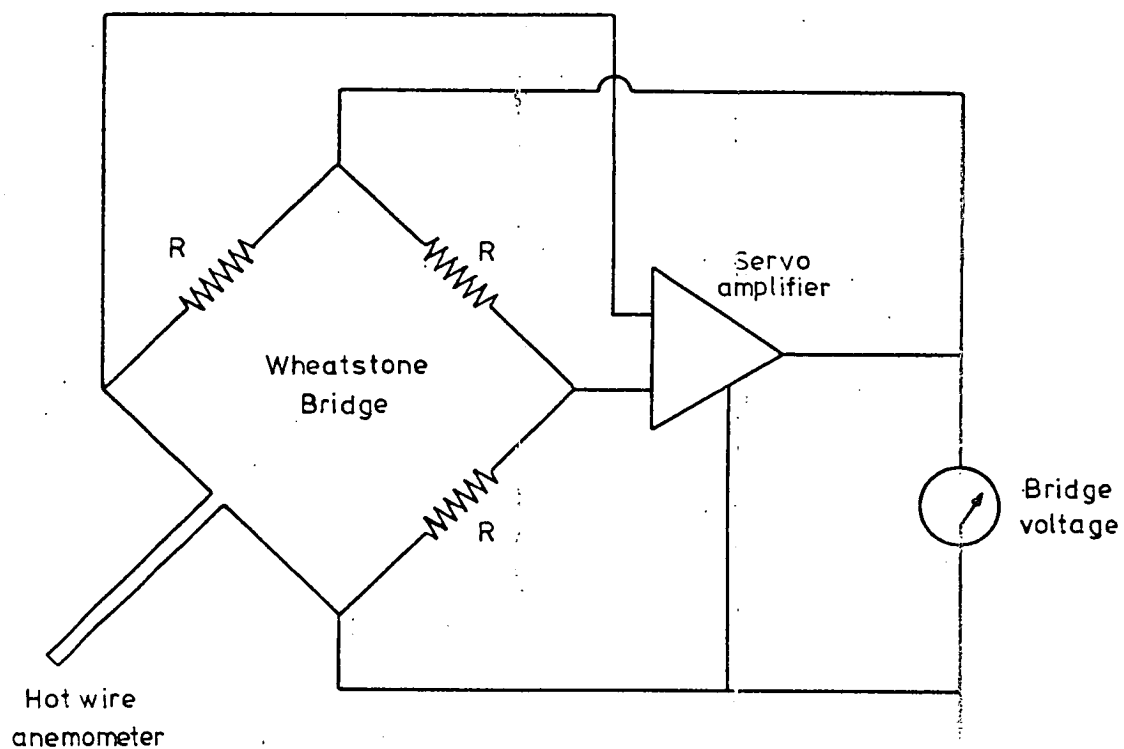


Figure B.1 Constant temperature hot wire anemometer principle.

coefficient, and I the electric current.

For forced heat transfer between the wire and the flowing gas, the heat transfer coefficient, h , depends on the gas velocity, U , diameter of the wire, d , gas physical properties (specific gravity, ρ , viscosity, μ , thermal conductivity, k , and specific heat, c_p).

A dimensionless analysis will obtain the following relations between the Nusselt, Reynolds and Prandtl dimensionless numbers

$$Nu = C. Re^{\alpha} Pr^{\beta} \quad (B.2)$$

where C is a constant and

$$Nu = \frac{h d}{k} \quad (B.3)$$

$$Re = \frac{U \rho d}{\mu} \quad (B.4)$$

$$Pr = \frac{c_p \mu}{k} \quad (B.5)$$

From the above equations (C.1) and (C.2)

$$\frac{R I^2}{A (T - T_o)} \frac{d}{k} = C. Re^{\alpha} Pr^{\beta} \quad (B.6)$$

or as a function of the voltage, V

$$\frac{V^2}{R A (T - T_o)} \frac{d}{k} = C. Re^{\alpha} Pr^{\beta} \quad (B.7)$$

and finally

$$V^2 = C \frac{R A (T - T_o) k}{d} \left[\frac{U \rho d}{\mu} \right]^\alpha \left[\frac{C_p \mu}{k} \right]^\beta \quad (B.8)$$

The above equation (B.8) shows clearly that the voltage necessary to keep the wire at a given temperature will depend on changes in the flow of the gas, as well as on the gas properties.

Ulsamer (77) found the following values of C, α and β

Re	C	α	β
0.1 - 50	0.385	0.31	0.91
50 - 10,000	0.50	0.31	0.60

and the above equation (B.8) thus permits an accurate calibration of the bridge if all the variables are known.

The above is a simple process in the case of pure gases. However, if gas mixtures are studied, properties such as viscosity, thermal conductivity, and specific heat are complex functions of the composition, (Jacob (34)). The ordinary mixing rule cannot be applied and the best approach is certainly the experimental calibration.

In our application, the gas flow is kept constant, thus the bridge voltage will depend on the gas composition, a mixture of air and helium. The bridge used in our experiments was a constant temperature hot wire bridge, Disa mod. 55A01 with linearised output. Its calibration was effected by placing our probe in a tube, into which a mixture of helium and air was introduced. The gas sampling rate was made to be the same as used in the experimental work, 3 cm³/sec. The relationship between V and the per cent of helium in the gas mixture is shown in figure B.2. It was found that a linear relationship exists between the bridge output and helium percentage in the gas mixture.

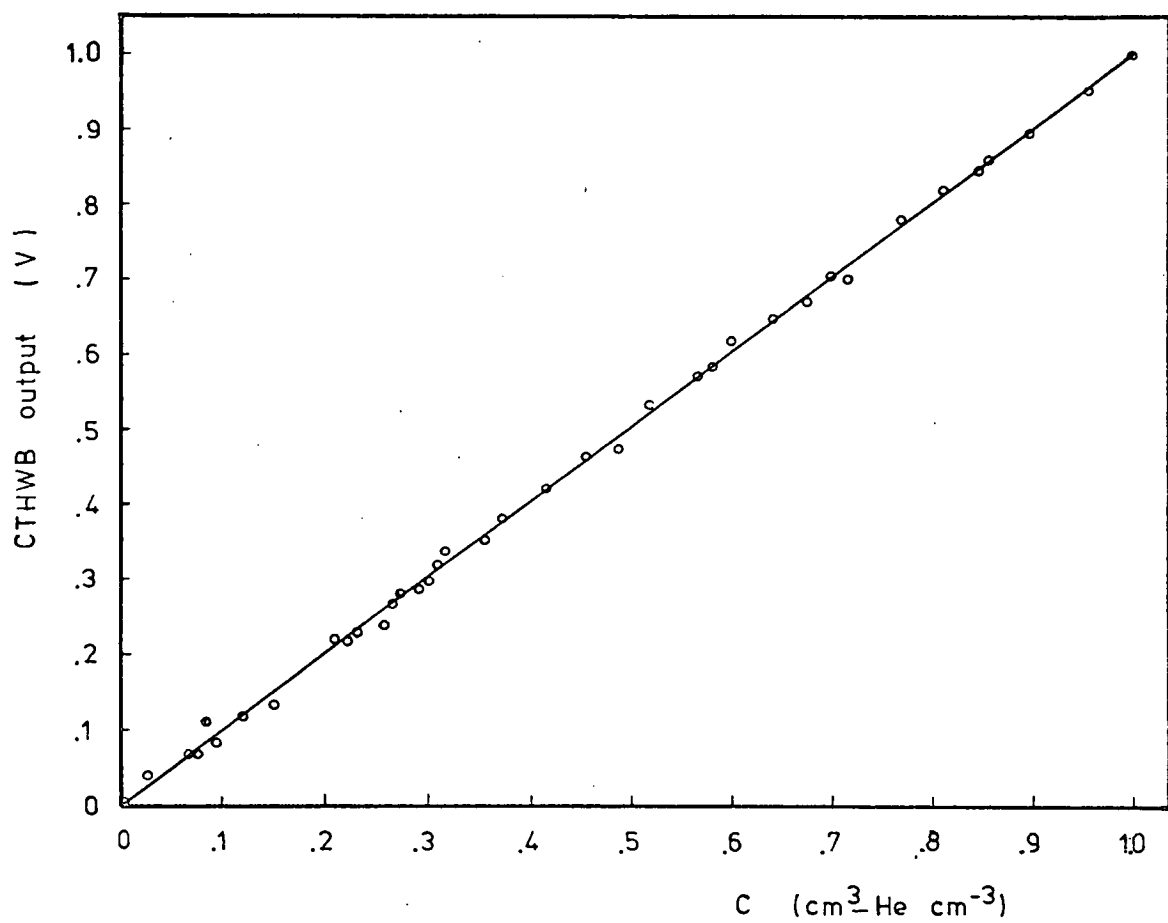


Figure B.2 Calibration of the CTHWB.

B.2 CTHWB - Testing

The helium concentration associated with the bubble and its near environment is given by the CTHWB output at positions determined by the pulse mode of the three upper contacts. These were placed coaxially and at the same level as the sintered plate, at the end of the plastic tube containing the hot wire sensor.

As the frequency of response of the conductivity contacts is very high, thus permitting an accurate determination of bubble position, the output of the CTHWB can only be associated with bubble location if the rise time of the tracer sensor head is very short. The frequency of response of the CTHWB being greater than 10 KHz, any time lag in the determination of gas composition will be due to the flow of gas between the sintered plate and the hot wire sensor.

The hot wire is contained in a chamber with a diameter of 0.22 cm, and 0.05 to 0.1 cm distant from the sintered plate.

As the rate of gas sampling is $3 \text{ cm}^3 \text{ sec}^{-1}$, the gas superficial velocity within the chamber is 79 cm sec^{-1} and the delay time, Δt , between gas sampling and composition detection will be

$$\frac{0.05}{79} < \Delta t < \frac{0.1}{79}$$

that is, the delay time is between 0.63 and 1.26 m sec.

A test was carried out to measure the rise time of the hot wire detector at the gas sampling rate used in practice, $3 \text{ cm}^3 \text{ sec}^{-1}$.

A step change in composition from air to helium was effected by puncturing balloons of helium with the probe and using a microphone actuated time switch connected to the computer. The

output of the CTHWB being recorded in computer memory during the transient state.

Figure B.3 shows the experimental arrangement used in order to obtain a step change in gas composition without disturbing the sampling rate. A rubber balloon containing helium was placed in a metal box, diameter 30 cm, height 45 cm. A small hole, diameter 2 cm, was opened in the top end of the box, through which the probe could pass. To the probe, and at the same level of the sintered plate, was fixed a sharp needle. Thus, by introducing the probe into the box, through the small hole, the balloon was punctured and a step change in the gas sampled by the probe, from air to helium, was affected. The other end of the metal box was left open in order to eliminate, or minimize, the existence of a shock wave, due to the puncturing of the balloon, which could affect the gas sampling rate.

The instant of puncturing the balloon was detected by a microphone, fixed to the box wall, which was interfaced with the PDP-8e computer. This arrangement permitted an accurate determination of the instant at which the step change was created. The CTHWB output was then recorded in computer memory at 100 μ sec intervals during 10 m sec, after which the data stored in the memory was printed out.

The flow chart of the program implemented in the computer is presented in figure B.4.

Figure B.4 presents the CTHWB output obtained and recorded as above. It shows that the probe has a rise time of 3.2 m sec, that is the time that it takes to reach 95% of the steady state voltage. Let us note the possibility of some gas dilution, due to the open end of the box, which will affect the final voltage of the

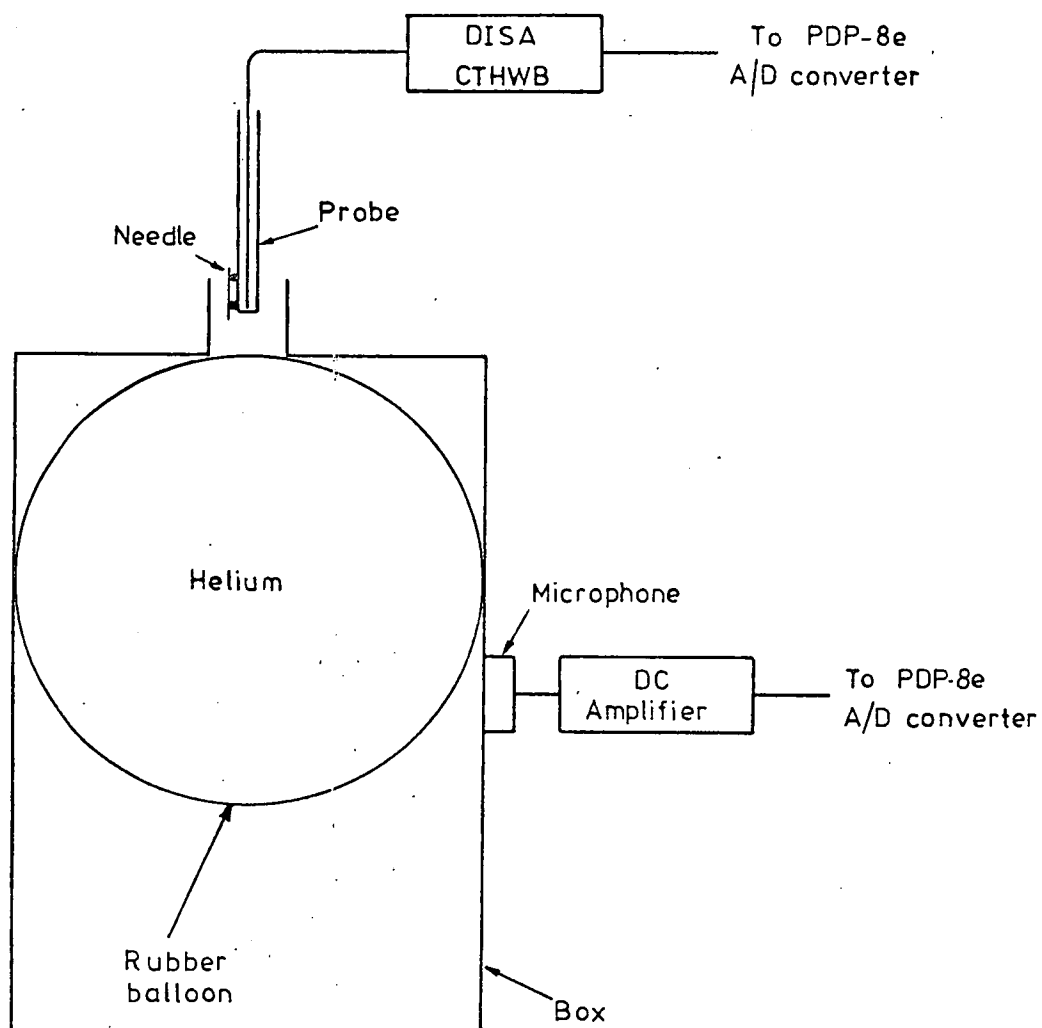


Figure B.3

Determination of the frequency of response of the hot wire detector. Schematic diagram of the assembly.

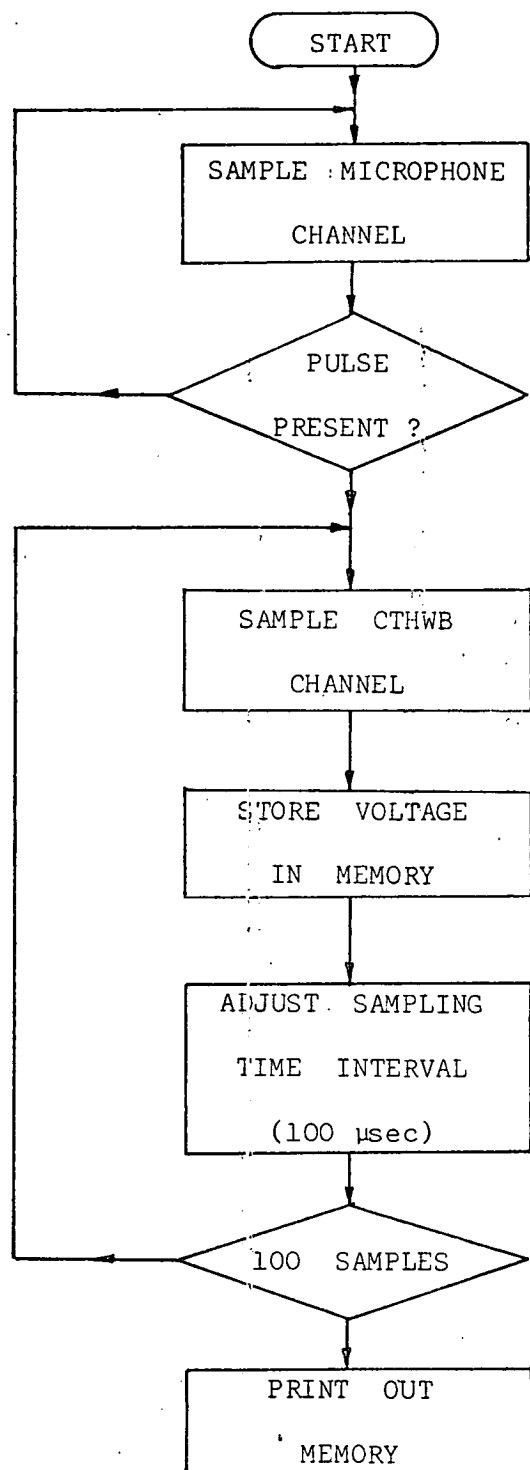


Figure B.4

Flow chart of the program implemented in the PDP-8e computer to determine the rise time of the hot wire detector.

CTHWB output, decreasing it.

Figure B.5 presents the result of an experiment that obtained the average rise time. Other experiments presented values between 2.7 and 3.6 m sec.

The experimentally determined rise time of the CTHWB detector being

$$3.2 \pm 0.5 \text{ m sec}$$

which is slightly greater than the predicted delay time, due to gas flow.

The experimentally determined rise time of the detector head shows that the CTHWB output is adequately synchronized with the pulsing mode of the three upper conductivity contacts of the compound probe and, hence, with the bubble position.

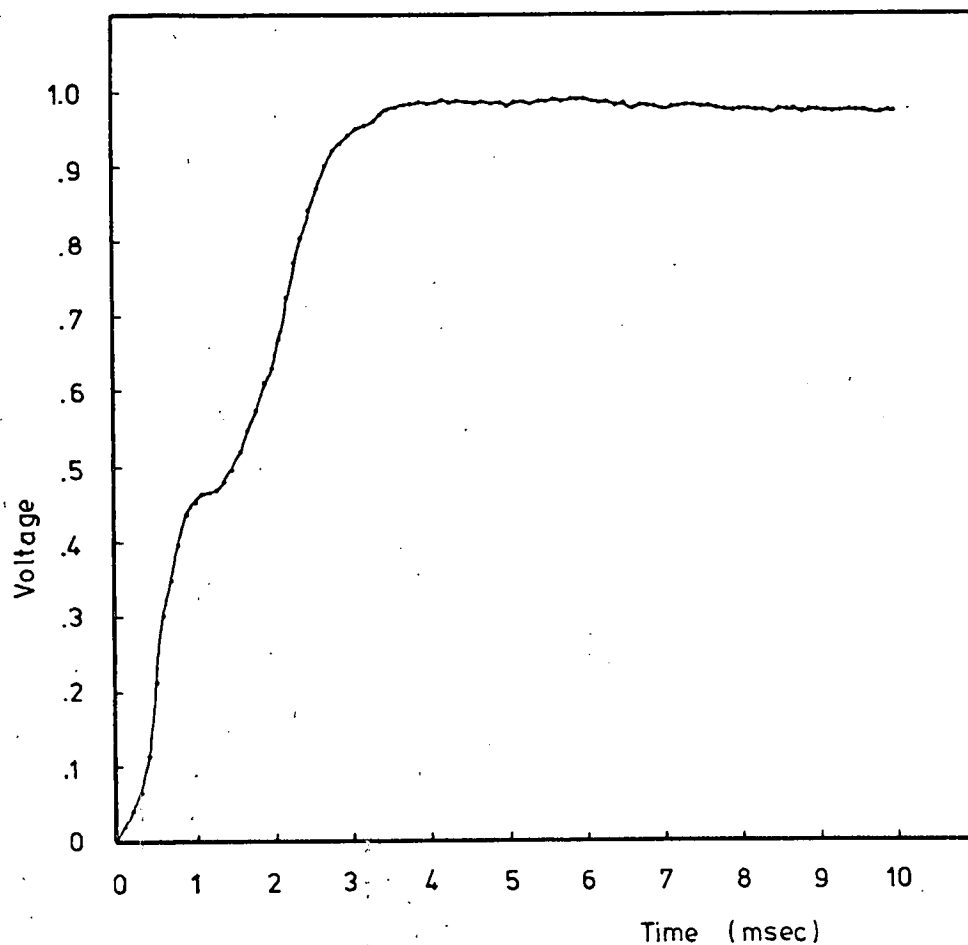


Figure B.5 Response of the CTHWB to a step change in
the gas composition.

APPENDIX CC.1 Real time data processing methods

The real time sampling logic programs were written in Program Assembler Language, PAL-8, for a Digital Equipment Corporation PDP - 8E digital computer. Detailed language description and computer characteristics may be found elsewhere (61).

The PAL-8 language consists of symbolic micro-instruction elements, found within the permanent symbol table in the software assembler supplied by Digital. The following are the micro-instructions used in the programs herein, their execution times are also indicated:

Instruction	Operation	Time (μsec)
TAD	Two's complement add to AC	2.6
DCA	Deposit and clear AC	2.6
ISZ	Increment and skip if zero	2.6
JMS	Jump to subroutine	2.6
JMP	Jump	1.2
NOP	No operation	1.2
IAC	Increment AC	1.2
RAL	Rotate AC and Link one place left	1.2
RTL	Rotate AC and Link two places left	1.2
RAR	Rotate AC and Link one place right	1.2
RTR	Rotate AC and Link two places right	1.2
CML	Complement Link	1.2
CMA	Complement AC	1.2
CIA	Complement and increment AC	2.4

Instruction	Operation	Time (μsec)
CLL	Clear Link	1.2
CLA	Clear AC	1.2
HLT	Halt	-
SKP	Skip unconditional	1.2
SNL	Skip on non-zero link	1.2
SZL	Skip on zero link	1.2
SZA	Skip on zero AC	1.2
SNA	Skip on non-zero AC	1.2
SMA	Skip on minus AC	1.2
SPA	Skip on zero or positive AC	1.2
LAS	Load AC with Switch Register	2.4
CDF	Change data field	1.2

The above execution times for instructions TAD, DCA, ISZ, JMS and JMP are for direct address mode. An indirect address will require an additional time of 1.2 μsec for its execution.

The instructions for the analogue to digital converter are:

ADLM	Load Multiplexer
ADST	Start conversion
ADSK	Skip on A/D done flag
ADRB	Read A/D Buffer

The analogue to digital converter operated with a combined conversion time of 26.2 μsec per channel.

The converted voltage (between 0 and +1 volt) was stored in memory or examined in binary form, with the following conversion

relationship:

Voltage (Volt)	Binary
0.0	000 000 000
+0.5	011 111 111
+1.0	111 111 111

The accuracy of voltage measurement was therefore restricted to ± 2 mV.

C.2. Program

A complete program listing, which includes routines for CTHWA zeroing and calibration, automatic control of the bubble injector and the real time analysis of the conductivity probe pulses and gas composition detection within the bubble and in its immediate surroundings, is presented in the following pages. It is written in PAL-8 assembler language and an explanation of the logical meaning of the different instruction is given in each one of the main routines.

The basic routine for pulse sequence examination of the conductivity sensors and the CTHWB output analysis consisted of two general sections; the real time pulse sequence logic and storage in memory of the analytical channel and the subsequent floating point computation for the purpose of goodness of fit examination. The process was therefore one of intermittent sampling of the continuous analogue voltages.

/INJECTED BUBBLES

PAL8-V7

PAGE 1

			/INJECTED BUBBLES
1			
2			
3		5000	FIXMRI FGET=5000
4		7000	FIXMRI FVOR=7000
5		6000	FIXMRI FPUT=6000
6		0000	FIXMRI FEKT=0000
7		1000	FIXMRI FADD=1000
8		2000	FIXMRI FSUB=2000
9		3000	FIXMRI FMPY=3000
10		4000	FIXMRI FDIV=4000
11		6531	FIXMRI ADLM=6531
12		6532	FIXMRI ADST=6532
13		6533	FIXMRI ADRB=6533
14		6534	FIXMRI ADSK=6534
15		6506	FIXMRI DR50=6506
16		6505	FIXMRI DBC0=6505
17			
18		0005	*0005
19	00005	7400	7400
20	00006	7200	7200
21	00007	5600	5600
22			
23		0100	*100
24	00100	0000	CRLF,0
25	00101	7200	CLA
26	00102	1107	TAD K215
27	00103	4111	JMS TYPE
28	00104	1110	TAD K212
29	00105	4111	JMS TYPE
30	00106	5500	JMP I CRLF
31	00107	0215	K215,215
32	00110	0012	K212,12
33			
34	00111	0000	TYPE,0
35	00112	6041	TSF
36	00113	5112	JMP --1
37	00114	6046	TLS
38	00115	7200	CLA
39	00116	5511	JMP I TYPE
40			
41	00117	0000	SPACE,0
42	00120	7200	CLA
43	00121	1124	TAD K240
44	00122	4111	JMS TYPE
45	00123	5517	JMP I SPACE
46	00124	0240	K240,240
47			
48	00125	0000	SLA,0
49	00126	7200	CLA
50	00127	1132	TAD SL
51	00130	4111	JMS TYPE
52	00131	5525	JMP I SLA
53	00132	0257	SL,257
54			
55	00133	0000	BLANKS,0

56	00134	7200	CLA	
57	00135	1143	TAD NURLAN	
58	00136	3144	DCA NURLOU	
59	00137	4111	JMS TYPE	
60	00140	2144	ISZ NURLOU	
61	00141	5137	JMP .-2	
62	00142	5533	JMP I BLANKS	
63	00143	7300	NURLAN,-500	
64	00144	0000	NURLOU,0	
65				
66		0200	*200	
67	00200	7200	CLA	
68	00201	6342	TCF	
69	00202	6046	TLS	
70	00203	7200	CLA	
71	00204	7604	LAS	
72	00205	3733	DCA I DPZ	/THRESHOLD VOLTAGE
73	00206	7604	LAS	
74	00207	7041	CIA	
75	00210	3734	DCA I DMZ	
76	00211	7402	HLT	
77	00212	7604	LAS	
78	00213	3735	DCA I RATVEL	/COINCIDENCE
79				/DISCRIMINATOR RATIO.
80	00214	7402	HLT	
81	00215	7604	LAS	
82	00216	3336	DCA RATPL	/VERTICAL RISE
83				/DISCRIMINATOR RATIO.
84	00217	7402	HLT	
85	00220	7604	LAS	
86	00221	3337	DCA BPLOOP	/NUMBER OF BUBBLES
87				/PER RUN.
88	00222	7402	HLT	
89	00223	7604	LAS	
90	00224	3752	DCA I STCON2	/HELIUM CONC. FOR
91				/CTHWA CALIBRATION.
92	00225	7402	HLT	
93	00226	7000	L210,NOP	
94	00227	4747	JMS I DRIDET	/CTHWA ZEROING.
95	00230	7200	CLA	
96	00231	1353	TAD RCH1	
97	00232	6506	DBSO	/OPERATE VALVES FOR
98				/CTHWA CALIBRATION.
99	00233	7200	CLA	
100	00234	1327	TAD INUM1	
101	00235	3330	DCA DLNU1	
102	00236	4745	L901,JMS I BSDELY	/DELAY TIME.
103	00237	2330	ISZ DLNU1	
104	00240	5236	JMP L901	
105	00241	7200	CLA	
106	00242	4750	JMS I STVLDT	/CTHWA CALIBRATION.
107	00243	7200	CLA	/CALIBRATION FINISHED.
108	00244	1353	TAD RCH1	/STOP HELIUM FLOW.
109	00245	6505	DBCO	
110	00246	7200	CLA	

111	00247	4751	JMS I STVLST	
112	00250	4100	JMS Z CRLF	
113	00251	1331	TAD DNUM	
114	00252	3332	DCA DLNU	
115	00253	4745	L902,JMS I BSDELY	/DELAY
116	00254	2332	ISZ DLNU	
117	00255	5253	JMP L902	
118	00256	7200	CLA	
119	00257	4747	JMS I DRIDET	/CHECK ZERO.
120	00260	7200	CLA	/START BUBBLE
121	00261	1354	TAD RCH2	/INJECTION
122	00262	1355	TAD RCH3	
123	00263	6506	DBSO	
124				
125	00264	7200	CLA	
126	00265	1325	TAD DNUM2	
127	00266	3326	DCA DLNU2	
128	00267	4745	L903,JMS I BSDELY	/INTRODUCE 60 BUBBLES
129	00270	2326	ISZ DLNU2	/BEFORE STARTING
130	00271	5267	JMP L903	/BUBBLE ANALYSIS
131	00272	7200	CLA	
132				
133	00273	4744	L291,JMS I MAINSA	/START BUBBLE ANALYSIS.
134				
135	00274	7200	CLA	/PRESET NUMBER OF
136	00275	1354	TAD RCH2	/BUBBLES OBTAINED.
137	00276	1355	TAD RCH3	/STOP BUBBLE INJECTION.
138	00277	6505	DBCO	
139	00300	7200	CLA	
140	00301	1337	TAD BPLOOP	
141	00302	7041	CIA	
142	00303	3340	DCA NBTTO	
143				
144	00304	4100	L285,JMS Z CRLF	/PRINTOUT DATA.
145	00305	4746	JMS I PCARBU	/BUBBLE PARAMETERS.
146	00306	4100	JMS Z CRLF	
147	00307	4100	JMS Z CRLF	
148				/GAS COMPOSITION.
149	00310	4741	JMS I PCONDF	/AHEAD OF BUBBLE.
150	00311	4100	JMS Z CRLF	
151	00312	4100	JMS Z CRLF	
152	00313	4742	JMS I PCONBU	/WITHIN BUBBLE.
153	00314	4100	JMS Z CRLF	
154	00315	4100	JMS Z CRLF	
155	00316	4743	JMS I PCONWK	/BEHIND THE BUBBLE.
156	00317	4100	JMS Z CRLF	
157	00320	4100	JMS Z CRLF	
158	00321	2340	ISZ NBTTO	
159	00322	5304	JMP L285	
160	00323	7402	LHLT,HLT	
161	00324	5226	JMP L210	/REPEAT THE EXPERIMENT.
162				
163	00325	7772	DNUM2,-6	
164	00326	0000	DLNU2,0	
165	00327	7766	DNUM1,-12	

166	00330	0000	DLNU1,0
167	00331	7750	DNUM,-30
168	00332	0000	DLNU,0
169	00333	2330	DPZ,DP
170	00334	2331	DMZ,DM
171	00335	2567	RATVEL,RATDD
172	00336	2570	RATPL,RATPD
173	00337	0000	RPL00P,0
174	00340	0000	NBTTO,0
175	00341	3400	PCONDF,PCDF
176	00342	3443	PCONBU,PCBF
177	00343	3506	PCONWK,PCWF
178	00344	1400	MAINSA,SAMPL
179	00345	0356	BSDELY,SBDELY
180	00346	3672	PCARRU,PR11
181	00347	0400	DRIDET,DETDR1
182	00350	0600	STVLDT,DTSTST
183	00351	1061	STVLST,STDRDV
184	00352	3634	STCON2,STDC2
185	00353	1000	RCH1,1000
186	00354	2000	RCH2,2000
187	00355	4000	RCH3,4000
188			
189	00356	0000	SRDELY,0
190	00357	7220	CLA
191	00360	1374	TAD DEL99
192	00361	3375	DCA ESP199
193	00362	7000	L6079,NOP
194	00363	1374	TAD DEL99
195	00364	3376	DCA ESP299
196	00365	7000	L6089,NOP
197	00366	7000	NOP
198	00367	2376	ISZ ESP299
199	00370	5365	JMP L6089
200	00371	2375	ISZ ESP199
201	00372	5362	JMP L6079
202	00373	5756	JMP I SBDELY
203			
204	00374	5413	DEL99,-2365
205	00375	0000	ESP199,0
206	00376	0000	ESP299,0
207			
208		0400	*400
209	00400	0000	DETDRI,
210	00401	7200	CLA
211	00402	3254	DCA DLS
212	00403	3255	DCA DHS
213	00404	1257	TAD DRIVOL
214	00405	7041	CIA
215	00406	3260	DCA CT442
216	00407	7200	L443,CLA
217	00410	4662	JMS I DRDDEL
218	00411	7200	CLA
219	00412	6531	ADLM
220	00413	6532	ADST

/ROUTINE FOR ZEROING
/THE CTHWA.

221	00414	6534	ADSK
222	00415	5214	JMP .-1
223	00416	6533	ADRR
224	00417	1254	TAD DLS
225	00420	7500	SMA
226	00421	5231	JMP L452
227	00422	1253	TAD OV4000
228	00423	3256	DCA DOVERF
229	00424	1255	TAD DHS
230	00425	1261	TAD PL1
231	00426	3255	DCA DHS
232	00427	1256	TAD DOVERF
233	00430	5246	JMP L457
234	00431	7000	L452,NOP
235	00432	7000	NOP
236	00433	7000	NOP
237	00434	7000	NOP
238	00435	7000	NOP
239	00436	7000	NOP
240	00437	7000	NOP
241	00440	7000	NOP
242	00441	7000	NOP
243	00442	7000	NOP
244	00443	7000	NOP
245	00444	7000	NOP
246	00445	7000	NOP
247	00446	3254	L457,DCA DLS
248	00447	2260	ISZ CT442
249	00450	5227	JMP L443
250	00451	4263	JMS BFDRFC
251	00452	5620	JMP I DETDRI
252	00453	4000	OV4000,4000
253	00454	0000	DLS,0
254	00455	0000	DHS,0
255	00456	0000	DOVERF,0
256	00457	1750	DRIVOL,1750
257	00460	0000	CT442,0
258	00461	0001	PL1,1
259	00462	1100	DRDDEL,DELDDR
260			
261	00463	0000	BFDRFC,0
262	00464	7200	CLA
263	00465	1257	TAD DRIVOL
264	00466	3325	DCA FLT11+1
265	00467	1323	TAD D3777
266	00470	3330	DCA FLT12+1
267	00471	4407	JMS I 7
268	00472	5324	FGET FLT11
269	00473	7000	FNOR
270	00474	6332	FPUT NDRVOL
271	00475	5327	FGET FLT12
272	00476	7000	FNOR
273	00477	6335	FPUT ND3777
274	00500	0000	FEXT
275	00501	7200	CLA

276	00502	1254	TAD DLS
277	00503	3325	DCA FLT11+1
278	00504	1255	TAD DHS
279	00505	3330	DCA FLT12+1
280	00506	4407	JMS I 7
281	00507	5324	FGET FLT11
282	00510	7000	FNOR
283	00511	6340	FPUT FLT13
284	00512	5327	FGET FLT12
285	00513	7000	FNOR
286	00514	3335	FMPY ND3777
287	00515	1340	FADD FLT13
288	00516	4332	FDIV NDRVOL
289	00517	6343	FPUT DRIFT1
290	00520	0000	FEXT
291	00521	7200	CLA
292	00522	5663	JMP I BFDRFC
293			
294	00523	3777	D3777, 3777
295	00524	0013	FLT11, 13
296	00525	0000	0
297	00526	0000	0
298	00527	0013	FLT12, 13
299	00530	0000	0
300	00531	0000	0
301	00532	0000	NDRVOL, 0
302	00533	0000	0
303	00534	0000	0
304	00535	0000	ND3777, 0
305	00536	0000	0
306	00537	0000	0
307	00540	0000	FLT13, 0
308	00541	0000	0
309	00542	0000	0
310	00543	0000	DRIFT1, 0
311	00544	0000	DRIFT2, 0
312	00545	0000	DRIFT3, 0
313			
314		0600	*600
315	00600	0000	DTSTST, 0
316	00601	7200	CLA
317	00602	1740	TAD I ZDRI1
318	00603	3335	DCA ZDRIFT
319	00604	1741	TAD I ZDRI2
320	00605	3336	DCA ZDRIFT+1
321	00606	1742	TAD I ZDRI3
322	00607	3337	DCA ZDRIFT+2
323	00610	7200	CLA
324	00611	1316	TAD TESTS
325	00612	3315	DCA THRTIM
326	00613	7000	L873, NOP
327	00614	4773	JMS I STSAM
328	00615	1717	TAD I MMMLS
329	00616	3322	DCA VT1LS+1
330	00617	1720	TAD I MMMHS

/ROUTINE FOR CTHWA
/CALIBRATION.

331	00620	3325	DCA VT1HS+1
332	00621	4772	JMS I SSTDEL
333	00622	4773	JMS I STSAM
334	00623	1717	TAD I MMMLS
335	00624	3330	DCA VT2LS+1
336	00625	1720	TAD I MMMHS
337	00626	3333	DCA VT2HS+1
338	00627	4407	JMS I 7
339	00630	5354	FGET N1000
340	00631	7000	FNOR
341	00632	6354	FPUT N1000
342	00633	5343	FGET ZZ3777
343	00634	7000	FNOR
344	00635	6343	FPUT ZZ3777
345	00636	5332	FGET VT2HS
346	00637	7000	FNOR
347	00640	3343	FMPY ZZ3777
348	00641	6346	FPUT FLT222
349	00642	5327	FGET VT2LS
350	00643	7000	FNOR
351	00644	1346	FADD FLT222
352	00645	4354	FDIV N1000
353	00646	2335	FSUB ZDRIFT
354	00647	6367	FPUT MAX1
355	00650	5324	FGET VT1HS
356	00651	7000	FNOR
357	00652	3343	FMPY ZZ3777
358	00653	6351	FPUT FLT333
359	00654	5321	FGET VT1LS
360	00655	7000	FNOR
361	00656	1351	FADD FLT333
362	00657	4354	FDIV N1000
363	00660	2335	FSUB ZDRIFT
364	00661	4367	FDIV MAX1
365	00662	3354	FMPY N1000
366	00663	6357	FPUT NRATIO
367	00664	0000	FEXT
368	00665	7200	CLA
369	00666	1362	TAD MM13
370	00667	1357	TAD NRATIO
370	00670	3363	DCA TP11
372	00671	1360	TAD NRATIO+1
373	00672	7110	L1524, CLL RAR
374	00673	2363	ISZ TP11
375	00674	5272	JMP L1524
376	00675	3366	DCA BRATIO
377	00676	1366	TAD BRATHO
378	00677	1364	TAD CC996
379	00700	7510	SPA
380	00701	5213	JMP L873
381	00702	7200	CLA
382	00703	1366	TAD BRATIO
383	00704	7041	CIA
384	00705	1365	TOD CC1000
385	00706	7500	SMA

386	00707	5311	JMP L1525
387	00710	5213	JMP L873
388	00711	7000	L1525,NOP
389	00712	2315	ISZ THRTM
390	00713	5213	JMP L873
391	00714	5600	JMP H DTSTST
392			
393	00715	0000	THRTM,0
394	00716	7775	TESTS,-3
395	00717	1052	MMMLS,MLS
396	00720	1053	MMMHS,MHS
397	00721	0013	VT1LS,13
398	00722	0000	0
399	00723	0000	0
400	00724	0013	VT1HS,13
401	00725	0000	0
402	00726	0000	0
403	00727	0013	VT2LS,13
404	00730	0000	0
405	00731	0000	0
406	00732	0013	VT2HS,13
407	00733	0000	0
408	00734	0000	0
409	00735	0000	ZDRIFT,0
410	00736	0000	0
411	00737	0000	0
412	00740	0543	ZDRI1,DRIFT1
413	00741	0544	ZDRI2,DRIFT2
414	00742	0545	ZDRI3,DRIFT3
415	00743	0013	ZZ3777,13
416	00744	3777	3777
417	00745	0000	0
418	00746	0000	FLT222,0
419	00747	0000	0
420	00750	0000	0
421	00751	0000	FLT333,0
422	00752	0000	0
423	00753	0000	0
424	00754	0013	N1000,13
425	00755	1750	1750
426	00756	0000	0
427	00757	0000	NRATIO,0
428	00760	0000	0
429	00761	0000	0
430	00762	7765	MM13,-13
431	00763	0000	TP11,0
432	00764	6033	CC996,-1745
433	00765	1753	CC1000,1753
434	00766	0000	BRATIO,0
435	00767	0000	MAX1,0
436	00770	0000	MAX2,0
437	00771	0000	MAX3,0
438	00772	1113	SSTDEL,DELSST
439	00773	1000	STSAM,SAMST
440			

441		1000	*1000	
442	01000	0000	SAMST,0	/SAMPLING OF ANALYTICAL
443	01001	7200	CLA	/CHANNEL.
444	01002	3252	DCA MLS	/1000 POINTS PER SEC.
445	01003	3253	DCA MHS	
446	01004	1255	TAD MAXVOL	
447	01005	7041	CIA	
448	01006	3256	DCA CT4429	
449	01007	7200	L4489,CLA	
450	01010	4660	JMS I MADDEL	
451	01011	6531	ADLM	
452	01012	6532	ADST	
453	01013	6534	ADSK	
454	01014	5213	JMP .-1	
455	01015	6533	ADRR	
456	01016	1252	TAD MLS	
457	01017	7500	SMA	
458	01020	5230	JMP L4529	
459	01021	1251	TAD CV4000	
460	01022	3254	DCA MOVERF	
461	01023	1253	TAD MHS	
462	01024	1257	TAD PL1M	
463	01025	3253	DCA MHS	
464	01026	1254	TAD MOVERF	
465	01027	5245	JMP L4579	
466	01030	7000	L4529,NOP	
467	01031	7000	NOP	
468	01032	7000	NOP	
469	01033	7000	NOP	
470	01034	7000	NOP	
471	01035	7000	NOP	
472	01036	7000	NOP	
473	01037	7000	NOP	
474	01040	7000	NOP	
475	01041	7000	NOP	
476	01042	7000	NOP	
477	01043	7000	NOP	
478	01044	7000	NOP	
479	01045	3252	L4579,DCA MLS	
480	01046	2256	ISZ CT4429	
481	01047	5207	JMP L4489	
482	01050	5600	JMP I SAMST	
483				
484	01051	4000	CV4000,4000	
485	01052	0000	MLS,0	
486	01053	0000	MHS,0	
487	01054	0000	MOVERF,0	
488	01055	1750	MAXVOL,1750	
489	01056	0000	CT4429,0	
490	01057	0001	PL1M,1	
491	01060	1100	MADDEL,DELDDR	
492				
493	01061	0000	STDRDV,0	
494	01062	7200	CLA	
495	01063	1672	TAD I SMX10	

496	01064	3675	DCA I NSTV1
497	01065	1673	TAD I SMX20
498	01066	3676	DCA I NSTV2
499	01067	1674	TAD I SMX30
500	01070	3677	DCA I NSTV3
501	01071	5661	JMP I STDRDV
502	01072	0767	SMX10,MAX1
503	01073	0770	SMX20,MAX2
504	01074	0771	SMX30,MAX3
505	01075	3644	NSTV1,NSTDV1
506	01076	3645	NSTV2,NSTDV2
507	01077	3646	NSTV3,NSTDV3
508			
509	01100	0000	DELDDR,0
510	01101	7200	CLA
511	01102	1311	TAD DEL931
512	01103	7041	CIA
513	01104	3312	DCA ESP931
514	01105	7000	L931,NOP
515	01106	2312	ISZ ESP931
516	01107	5305	JMP L931
517	01110	5700	JMP I DELDDR
518			
519	01111	0272	DEL931,272
520	01112	0000	ESP931,0
521			
522	01113	0000	DELSST,0
523	01114	7200	CLA
524	01115	1332	TAD DEL88
525	01116	7041	CIA
526	01117	3334	DCA ESP18
527	01120	7000	L6078,NOP
528	01121	1333	TAD DEL8
529	01122	7041	CIA
530	01123	3335	DCA ESP28
531	01124	7000	L6088,NOP
532	01125	2335	ISZ ESP23
533	01126	5324	JMP L6088
534	01127	2334	ISZ ESP18
535	01130	5320	JMP L6078
536	01131	5713	JMP I DELSST
537			
538	01132	0130	DEL88,130
539	01133	1750	DEL8,1750
540	01134	0000	ESP18,0
541	01135	0000	ESP28,0
542			
543		1200	*1200
544			
545			/ANALYTICAL CHANNEL
546			/EXAMINATION AND GAS
547			/COMPOSITION STORAGE
548			/IN COMPUTER MEMORY.
549	01200	0000	DENCON,0
550	01201	7200	CLA

/PROBE IN DENSE PHASE
/AHEAD OF THE BUBBLE.

551	01202	4712	JMS I ADC4	
552	01203	6211	CDF 10	
553	01204	3620	DCA I DFPOI	
554	01205	6201	CDF 00	
555	01206	2220	ISZ DFPOI	
556	01207	7000	NOP	
557	01210	1220	TAD DFPOI	
558	01211	1221	TAD DFPOIM	
559	01212	7440	SZA	
560	01213	5216	JMP L794	
561	01214	7200	CLA	
562	01215	5711	JMP I STOP	
563	01216	7200	L794, CLA	
564	01217	5600	JMP I DENCON	
565				
566	01220	0000	DFPOI, 0	
567	01221	2010	DFPOIM, -5770	
568				
569	01222	0000	BUBCON, 0	/PROBE WITHIN THE BUBBLE.
570	01223	7200	CLA	
571	01224	4712	JMS I ADC4	
572	01225	6221	CDF 20	
573	01226	3642	DCA I BBPOI	
574	01227	6201	CDF 00	
575	01230	2242	ISZ BBPOI	
576	01231	7000	NOP	
577	01232	1242	TAD BBPOI	
578	01233	1243	TAD BBPOIM	
579	01234	7440	SZA	
580	01235	5240	JMP L795	
581	01236	7200	CLA	
582	01237	5711	JMP I STOP	
583	01240	7200	L795, CLA	
584	01241	5622	JMP I BUBCON	
585				
586	01242	0000	BBPOI, 0	
587	01243	0010	BBPOIM, -7770	
588				
589	01244	0000	WAKCON, 0	/PROBE IN THE DENSE
590	01245	7200	CLA	/PHASE BEHIND THE BUBBLE.
591	01246	4712	JMS I ADC4	
592	01247	6231	CDF 30	
593	01250	3664	DCA I WFPOI	
594	01251	6201	CDF 00	
595	01252	2264	ISZ WFPOI	
596	01253	7000	NOP	
597	01254	1264	TAD WFPOI	
598	01255	1265	TAD WFPOIM	
599	01256	7440	SZA	
600	01257	5262	JMP L796	
601	01260	7200	CLA	
602	01261	5711	JMP I STOP	
603	01262	7200	L796, CLA	
604	01263	5644	JMP I WAKCON	
605				

606	01264	0000	WFPOI,0	
607	01265	0010	WFPOIM,-7770	
608				
609	01266	0000	EXWKC,0	/BUBBLE PASSED THE PROBE.
610	01267	7200	CLA	/CONTINUE SAMPLING
611	01270	1305	TAD WCONCE	/PROBE OUTPUT ONCE
612	01271	3306	DCA WCOUNT	/EVERY 5 MSEC.
613	01272	7200	LW,CLA	
614	01273	4704	JMS I COWAKE	
615	01274	7200	CLA	
616	01275	1307	TAD WDEL	
617	01276	3310	DCA WCOU	
618	01277	2310	ISZ WCOU	
619	01300	5277	JMP .-1	
620	01301	2306	ISZ WCOUNT	
621	01302	5272	JMP LW	
622	01303	5666	JMP I EXWKC	
623				
624	01304	1244	COWAKE,WAKCON	
625	01305	7470	WCONCE,-310	
626	01306	0000	WCOUNT,0	
627	01307	5356	WDEL,-2422	
628	01310	0000	WCOU,0	
629				
630	01311	0323	STOP,LHLT	
631	01312	2333	ADC4,ADC	
632				
633	01313	0000	BUBPOI,0	/MEMORY ADDRESS POINTERS
634	01314	7200	CLA	/WHERE DATA STORAGE FOR
635	01315	2340	ISZ DF	/EACH BUBBLE BEGINS.
636	01316	2341	ISZ BF	
637	01317	2342	ISZ WF	
638	01320	1220	TAD DFPOI	
639	01321	6211	CDF 10	
640	01322	3740	DCA I DF	
641	01323	6201	CDF 00	
642	01324	7200	CLA	
643	01325	1242	TAD BBPOI	
644	01326	6211	CDF 10	
645	01327	3741	DCA I BF	
646	01330	6201	CDF 00	
647	01331	7200	CLA	
648	01332	1264	TAD WFPOI	
649	01333	6211	CDF 10	
650	01334	3742	DCA I WF	
651	01335	6201	CDF 00	
652	01336	7200	CLA	
653	01337	5713	JMP I BUBPOI	
654	01340	0000	DF,0	
655	01341	0000	BF,0	
656	01342	0000	WF,0	
657				
658	01343	7200	RESET,CLA	/UNSUCCESSFUL HIT.
659	01344	1767	TAD I BFLAG1	/RESET STARTING ADDRESSES
660	01345	7450	SNA	/FOR DATA STORAGE OF

661	01346	5765	JMP I L5AI	/NEXT BUBBLE.
662	01347	7200	CLA	
663	01350	6211	CDF 10	
664	01351	1740	TAD I DF	
665	01352	6201	CDF 00	
666	01353	3220	DCA DFPOI	
667	01354	6211	CDF 10	
668	01355	1741	TAD I BF	
669	01356	6201	CDF 00	
670	01357	3242	DCA RBPOI	
671	01360	6211	CDF 10	
672	01361	1742	TAD I WF	
673	01362	6201	CDF 00	
674	01363	3264	DCA WFPOI	
675	01364	5766	JMP I L5BI	
676				
677	01365	1407	L5AI,L5A	
678	01366	1410	L5BI,L5	
679	01367	1472	BFLAG1,BFLAG	
680				
681		1400	*1400	
682	01400	0000	SAMPL,0	/MAIN RUN ROUTINE
683	01401	7200	CLA	/EXAMINATION OF ALL
684	01402	3272	DCA BFLAG	/CHANNEL .
685	01403	1675	TAD I BLOOP	/LOGICAL DECISIONS
686	01404	7041	CIA	/AND DATA STOTAGE.
687	01405	3276	DCA BUBNUM	
688	01406	4714	JMS I SET10	/SET POINTERS
689	01407	4715	L5A,JMS I SET20	/AND
690	01410	4702	L5,JMS I CLR	/CLEAR COUNTS.
691	01411	7200	L10,CLA	
692	01412	4704	JMS I INPUTA	/EXAMINE CH1 FOR
693	01413	1713	TAD I SAM11	/PULSE LEADING EDGE
694	01414	7450	SNA	/CORRECTION.
695	01415	5211	JMP L10	
696	01416	7200	L11,CLA	
697	01417	4704	JMS I INPUTA	
698	01420	4677	JMS I STR	
699	01421	7200	CLA	
700	01422	1713	TAD I SAM11	/BUBBLE APPROACHING THE
701	01423	7440	SZA	/PROBE ?
702	01424	5216	JMP L11	/NO. RETURN.
703	01425	4703	L6,JMS I TK	/YES. EXAMINE ALL
704				/CHANNELS.
705	01426	7200	CLA	
706	01427	1266	TAD ADJTIM	
707	01430	7041	CIA	
708	01431	3267	DCA COUTIM	
709	01432	2267	ISZ COUTIM	
710	01433	5232	JMP --1	
711				
712	01434	7200	CLA	/ARE ALL UPPER CONTACTS
713	01435	1710	TAD I P333	/ALSO IN PULSE MODE ?
714	01436	1711	TAD I P444	
715	01437	1712	TAD I P555	

716	01440	1270	TAD MIN3	
717	01441	7450	SNA	
718	01442	5250	JMP L602	/YES.CONTINUE IN L602
719				/BELOW.
720	01443	7200	CLA	/NO.
721	01444	1713	TAD I SAM11	/HAS PULSE ON LEADING
722	01445	7440	SZA	/CONTACT FINISHED?
723	01446	5665	JMP I RESET1	/YES.STOP SAMPLING.
724				/RESET ADDRESSES.
725	01447	5225	JMP L6	/NO.CONTINUE.
726				
727	01450	7200	L602,CLA	/LEADING AND UPPER
728				/CONTACTS IN PULSE MODE.
729	01451	4705	JMS I OKK	/CONTINUE EXAMINING ALL
730				/CHANNELS IN ROUTINE OK.
731	01452	4707	JMS I CEXWK	/BUBBLE PASSED THE PROBE.
732				/CONTINUE EXAMINING
733				/ANALATYCAL CHANNEL.
734	01453	4736	JMS I ADD	/SAMPLING FINISHED.
735				/TAKE ALL PULSE
736				/DURATION AND DELAY TIME.
737	01454	4701	JMS I TESTER	/CHECK DISCRIMINATORS.
738	01455	4700	JMS I EXM	/DISCRIMINATORS SATISFIED.
739				/EXAM LEADING EDGE OF PULSE
740				/AND CORRECT ITS DURATION.
741	01456	4673	JMS I BCSTR	/BUBBLE PASSED THE PROBRE
742	01457	4674	JMS I CPOIM	/SQUARELY.STORE BUBBLE
743				/PARAMETERS AND BUBBLE
744				/POINTERS.
745	01460	1271	TAD RFL	
746	01461	3272	DCA BFLAG	
747	01462	2276	ISZ BUBNUM	/RUN SHOULD BE FINISHED ?
748	01463	5210	JMP L5	/NO.WAIT FOR NEXT BUBBLE.
749	01464	5600	JMP I SAMPL	/YES.EXIT MAIN ROUTINE
750				/PROCEED TO STOP BUBBLE
751				/INJECTION AND TO PRINT OUT
752				/ALL THE DATA OF THIS RUN.
753				
754	01465	1343	RESET1,RESET	
755	01466	0012	ADJTIM,12	
756	01467	0000	COUTIM,0	
757	01470	7775	MIN3,-3	
758	01471	0001	BFL,1	
759	01472	0000	BFLAG,0	
760	01473	3112	BCSTR,BSTORE	
761	01474	1313	CPOIM,BUBPOI	
762	01475	0337	BLOOP,BPLOOP	
763	01476	0000	BUBNUM,0	
764	01477	3065	STR,STORE	
765	01500	3200	EXM,EXAM	
766	01501	2471	TESTER,TEST	
767	01502	3000	CLR,CLEAR	
768	01503	1600	TK,TAKE	
769	01504	2342	INPUTA,INA	
770	01505	2000	OKK,OK	

771	01506	1516	ADD, ADDN	
772	01507	1266	CEXW, EXWKC	
773	01510	1733	P333, P3	
774	01511	1734	P444, P4	
775	01512	1735	P555, P5	
776	01513	1747	SAM11, SAM1	
777	01514	4131	SET1, SET1	
778	01515	4112	SET2, SET2	
779				
780	01516	0000	ADDN, 0	
781	01517	7200	CLA	
782	01520	1743	TAD I CPL1	
783	01521	1752	TAD I CP1I	
784	01522	3743	DCA I CPL1	
785	01523	1744	TAD I CPL2	
786	01524	1753	TAD I CP2I	
787	01525	3744	DCA I CPL2	
788	01526	1745	TAD I CPL3	
789	01527	1754	TAD I CP3I	
790	01530	3745	DCA I CPL3	
791	01531	1746	TAD I CPL4	
792	01532	1755	TAD I CP4I	
793	01533	3746	DCA I CPL4	
794	01534	1747	TAD I CPL5	
795				
796	01535	1756	TAD I CP5I	
797	01536	3747	DCA I CPL5	
798	01537	1751	TAD I CD2I	
799	01540	1750	TAD I CDL2	
800	01541	3751	DCA I CD2I	
801	01542	5716	JMP I ADDN	
802				
803	01543	1742	CPL1, CPUL1	
804	01544	1743	CPL2, CPUL2	
805	01545	1744	CPL3, CPUL3	
806	01546	1745	CPL4, CPUL4	
807	01547	1746	CPL5, CPUL5	
808	01550	2153	CDL2, CSD2	
809	01551	1736	CD2I, CDEL2	
810	01552	2146	CP1I, CP1	
811	01553	2147	CP2I, CP2	
812	01554	2150	CP3I, CP3	
813	01555	2151	CP4I, CP4	
814	01556	2152	CP5I, CP5	
815				
816		1600	*1'600	
817	01600	0000	TAXE, 0	
818	01601	7200	CLA	/CH1 IN GAS MODE,
819	01602	2342	ISZ CPUL1	/A PULSE IS PRESENT ON CH1.
820				
821	01603	4722	JMS I INPUT	/EXAMINE OTHER CHANNELS.
822				
823	01604	7200	CLA	
824	01605	1350	TAD SAM2	/FIRSTLY CHANNEL 2 -
825				/ - SHAPE CHANNEL:

826	01606	7450	SNA	
827	01607	5217	JMP L12	
828	01610	7200	CLA	/CH2 PULSE IN LIQUID MODE.
829				
830	01611	1332	TAD P2	/HAS CH2 BEEN IN GAS
831				/MODE BEFORE DURING
832	01612	7440	SZA	/THIS SEQUENCE?
833	01613	5224	JMP L1	/YES, PULSE OVER ON CH2.
834				
835	01614	2336	ISZ CDEL2	/NO, INCREMENT CH2 DELAY
836	01615	1354	TAD ZERO	/COUNTER.
837	01616	5227	JMP L13	
838				
839	01617	2343	L12, ISZ CPUL2	/CH2 IN GAS MODE,
840				/INCREMENT DURATION
841	01620	7200	CLA	/COUNTER.
842	01621	1323	TAD C1	
843	01622	3332	DCA P2	/SET CH2 PULSE PRESENT
844	01623	5227	JMP L13	/FLAG.
845				
846	01624	7200	L1, CLA	
847	01625	1323	TAD C1	/CH2 PULSE OVER.
848	01626	3326	DCA P20	/SET CH2 PULSE OVER FLAG.
849				
850	01627	7200	L13, CLA	/EXAMINE CH3 WITH THE SAME
851	01630	1351	TAD SAM3	/LOGIC.
852	01631	7450	SNA	
853	01632	5242	JMP L14	
854	01633	7200	CLA	
855	01634	1333	TAD P3	
856	01635	7440	SZA	
857	01636	5247	JMP L2	
858	01637	2337	ISZ CDEL3	
859	01640	1354	TAD ZERO	
860	01641	5252	JMP L15	
861				
862	01642	2344	L14, ISZ CPUL3	
863	01643	7200	CLA	
864	01644	1323	TAD C1	
865	01645	3333	DCA P3	
866	01646	5252	JMP L15	
867				
868	01647	7200	L2, CLA	
869	01650	1323	TAD C1	
870	01651	3327	DCA P30	
871				
872	01652	7200	L15, CLA	/EXAMINE CH4 WITH THE SAME
873	01653	1352	TAD SAM4	/LOGIC.
874	01654	7450	SNA	
875	01655	5265	JMP L16	
876	01656	7200	CLA	
877	01657	1334	TAD P4	
878	01660	7440	SZA	
879	01661	5272	JMP L3	
880	01662	2340	ISZ CDEL4	

881	01663	1354	TAD ZERO	
882	01664	5275	JMP L4	
883				
884	01665	2345	L16, ISZ CPUL4	
885	01666	7200	CLA	
886	01667	1323	TAD C1	
887	01670	3334	DCA P4	
888	01671	5275	JMP L4	
889				
890	01672	7200	L3, CLA	
891	01673	1323	TAD C1	
892	01674	3330	DCA P40	
893				
894	01675	7200	L4, CLA	/EXAMINE CH5 WITH THE SAME
895	01676	1353	TAD SAM5	/LOGIC.
896	01677	7450	SNA	
897	01700	5310	JMP L101	
898	01701	7200	CLA	
899	01702	1335	TAD P5	
900	01703	7440	SZA	
901	01704	5315	JMP L102	
902	01705	2341	ISZ CDEL5	
903	01706	1354	TAD ZERO	
904	01707	5320	JMP L103	
905				
906	01710	2346	L101, ISZ CPUL5	
907	01711	7200	CLA	
908	01712	1323	TAD C1	
909	01713	3335	DCA P5	
910	01714	5320	JMP L103	
911				
912	01715	7200	L102, CLA	
913	01716	1323	TAD C1	
914	01717	3331	DCA P52	
915				
916	01720	4724	L103, JMS I DETCON	/EXAMINE PULSE MODE OF
917				/ALL 3 UPPER CONTACTS.
918				/SAMPLE ANALATYCAL CHANNEL
919				/AND STORE DATA ACCORDING
920				/TO THEIR PULSE MODE
921				
922	01721	5600	JMP I TAKE	
923				
924	01722	2200	INPUT, IN	
925	01723	00F1	C1, 1	
926	01724	2400	DETCN, GETCON	
927	01725	0000	P10, 0	
928	01726	0000	P20, 0	
929	01727	0000	P30, 0	
930	01730	0000	P40, 0	
931	01731	0000	P50, 0	
932	01732	0000	P2, 0	
933	01733	0000	P3, 0	
934	01734	0000	P4, 0	
935	01735	0000	P5, 0	

936	01736	0000	CDEL2,0	
937	01737	0000	CDEL3,0	
938	01740	0000	CDEL4,0	
939	01741	0000	CDEL5,0	
940	01742	0000	CPUL1,0	
941	01743	0000	CPUL2,0	
942	01744	0000	CPUL3,0	
943	01745	0000	CPUL4,0	
944	01746	0000	CPUL5,0	
945	01747	0000	SAM1,0	
946	01750	0000	SAM2,0	
947	01751	0000	SAM3,0	
948	01752	0000	SAM4,0	
949	01753	0000	SAM5,0	
950	01754	0000	ZERO,0	
951				
952		2000	*2000	
953	02000	0000	OK,0	
954				
955				/THERE ARE NOW PULSES ON
956				/ALL CHANNELS IN THE
957				/CORRECT SEQUENCE.
958				/EXAMINE A/D INPUT UNTIL
959				/ALL PULSES ARE FINISHED.
960				
961	02001	7200	L627,CLA	
962	02002	4731	JMS I INPUT2	/GET THE SAMPLES.
963				
964	02003	7200	CLA	
965	02004	1734	TAD I P10I	/CHECK CH1 FIRST.
966	02005	7440	SZA	
967	02006	5214	JMP L620	/CH1 PULSE ALREADY OVER,
968				/CHECK CH2.
969				
970				
971	02007	1741	TAD I SM1	/CH1 PULSE NOT FINISHED YET,
972				/CHECK FOR PRESENT LOOP.
973				
974	02010	7440	SZA	/IS CH1 IN PULSE MODE STILL?
975	02011	5214	JMP L620	
976	02012	2346	ISZ CP1	/YES INCREMENT COUNTER.
977	02013	5217	JMP L619	
978				
979	02014	7200	L620,CLA	/NO, SET PULSE OVER FLAG=1
980	02015	1332	TAD C10	
981	02016	3734	DCA I P10I	
982				
983	02017	7200	L619,CLA	/CHECK CH2. HAS CH2 PULSE
984	02020	1733	TAD I P2I	/COMMENCED?
985	02021	7440	SZA	
986	02022	5234	JMP L6019	/YES, PROCEED.
987				
988	02023	1742	TAD I SM2	/NO, HAS IT ARRIVED
989	02024	7450	SNA	/THIS LOOP?
990	02025	5230	JMP L6002	

991	02026	2353	ISZ CSD2	/NO, INCREMENT COUNTER.
992	02027	5250	JMP L621	
993				
994	02030	2347	L6002, ISZ CP2	/YES PULSE HAS COMMENCED,
995				/INCREMENT DURATION COUNTER.
996	02031	1332	TAD C10	
997	02032	3733	DCA I P21	/SET PULSE ARRIVED FLAG=1
998	02033	5250	JMP L621	
999				
1000	02034	7200	L6019, CLA	/PULSE HAS ARRIVED ALREADY.
1001				/HAS IT FINISHED THIS LOOP?
1002	02035	1735	TAD I P20I	
1003	02036	7440	SZA	
1004	02037	5250	JMP L621	/FINISHED ALREADY.
1005	02040	1742	TAD I SM2	/CHECK FOR THIS LOOP.
1006	02041	7440	SZA	
1007	02042	5245	JMP L622	
1008	02043	2347	ISZ CP2	/INCREMENT DURATION COUNTER.
1009	02044	5250	JMP L621	
1010				
1011	02045	7200	L622, CLA	/SET PULSE OVER FLAG=1.
1012	02046	1332	TAD C10	
1013	02047	3735	DCA I P20I	
1014				
1015	02050	7200	L621, CLA	/CHECK CH3 WITH CH1 LOGIC.
1016	02051	1736	TAD I P30I	
1017	02052	7440	SZA	
1018	02053	5261	JMP L624	
1019	02054	1743	TAD I SM3	
1020	02055	7440	SZA	
1021	02056	5261	JMP L624	
1022	02057	2350	ISZ CP3	
1023	02060	5264	JMP L623	
1024				
1025	02061	7200	L624, CLA	
1026	02062	1332	TAD C10	
1027	02063	3736	DCA I P30I	
1028				
1029	02064	7200	L623, CLA	/CHECK CH4 WITH CH1 LOGIC.
1030	02065	1737	TAD I P40I	
1031	02066	7440	SZA	
1032	02067	5275	JMP L626	
1033	02070	1744	TAD I SM4	
1034	02071	7440	SZA	
1035	02072	5275	JMP L626	
1036	02073	2351	ISZ CP4	
1037	02074	5300	JMP L625	
1038				
1039	02075	7200	L626, CLA	
1040	02076	1332	TAD C10	
1041	02077	3737	DCA I P40I	
1042				
1043	02100	7200	L625, CLA	
1044	02101	1740	TAD I P50I	/CHECK CH5 WITH CH1 LOGIC.
1045	02102	7440	SZA	

1046	02103	5311	JMP L730	
1047	02104	1745	TAD I SM5	
1048	02105	7440	SZA	
1049	02106	5311	JMP L730	
1050	02107	2352	ISZ CP5	
1051	02110	5314	JMP L727	
1052				
1053	02111	7200	L730,CLA	
1054	02112	1332	TAD C10	
1055	02113	3740	DCA I P50I	
1056				
1057	02114	4730	L727,JMS I DTCCON	/EXAMINE PULSE MODE OF
1058				/ALL 3 UPPER CONTACTS.
1059				/SAMPLE ANALATYCAL CHANNEL
1060				/AND STORE DATA ACCORDING
1061				/TO THEIR PULSE MODE.
1062				
1063				
1064	02115	1324	TAD ATIME	
1065	02116	3325	DCA CTIME	
1066	02117	2325	ISZ CTIME	
1067	02120	5317	JMP .-1	
1068				
1069	02121	4726	JMS I CHECK	
1070	02122	4727	JMS I FINI	
1071	02123	5600	JMP I OK	
1072				
1073	02124	7774	ATIME,-4	
1074	02125	0000	CTIME,0	
1075	02126	2154	CHECK,CK	
1076	02127	2446	FINI,FINISH	
1077	02130	2400	DTCCON,GETCON	
1078	02131	2200	INPUT2,IV	
1079	02132	0001	C10,1	
1080	02133	1732	P2I,P2	
1081	02134	1725	P10I,P10	
1082	02135	1726	P20I,P20	
1083	02136	1727	P30I,P30	
1084	02137	1730	P40I,P40	
1085	02140	1731	P50I,P50	
1086	02141	1747	SM1,SAM1	
1087	02142	1750	SM2,SAM2	
1088	02143	1751	SM3,SAM3	
1089	02144	1752	SM4,SAM4	
1090	02145	1753	SM5,SAM5	
1091	02146	0000	CP1,0	
1092	02147	0000	CP2,0	
1093	02150	0000	CP3,0	
1094	02151	0000	CP4,0	
1095	02152	0000	CP5,0	
1096	02153	0000	CSD2,0	
1097				
1098				
1099	02154	0000	CK,0	
1100	02155	7200	CLA	

1101	02156	1736	TAD I P301	
1102	02157	1737	TAD I P401	
1103	02160	1740	TAD I P501	
1104	02161	1374	TAD MI3	
1105	02162	7440	SZA	
1106	02163	5372	JMP L2423	
1107	02164	7200	CLA	
1108	02165	1733	TAD I P21	
1109	02166	7440	SZA	
1110	02167	5372	JMP L2423	
1111	02170	7200	CLA	
1112	02171	5775	JMP I RESI	
1113				
1114	02172	7200	L2423, CLA	
1115	02173	5754	JMP I CK	
1116				
1117	02174	7775	MI3, -3	
1118	02175	1343	RESI, RESET	
1119				
1120		2200	*2200	
1121	02200	0000	IN, 0	/INPUT ROUTINE FROM
1122	02201	7200	CLA	/CONDUCTIVITY CONTACTS.
1123				/SETTING OF PULSE MODE.
1124				
1125				
1126	02202	1316	TAD C0	
1127	02203	4333	JMS ADC	
1128	02204	3311	DCA S1	
1129				
1130	02205	1317	TAD C2	
1131	02206	4333	JMS ADC	
1132	02207	3312	DCA S2	
1133				
1134	02210	1320	TAD C3	
1135	02211	4333	JMS ADC	
1136	02212	3313	DCA S3	
1137				
1138	02213	1321	TAD C4	
1139	02214	4333	JMS ADC	
1140	02215	3314	DCA S4	
1141				
1142	02216	1322	TAD C5	
1143	02217	4333	JMS ADC	
1144	02220	3315	DCA SS555	
1145				
1146	02221	1311	L1630, TAD S1	
1147	02222	1331	TAD DM	
1148	02223	7510	SPA	
1149	02224	5231	JMP L1640	
1150	02225	7200	CLA	
1151	02226	1330	TAD DP	
1152	02227	3723	DCA I SS1	
1153	02230	5234	JMP L1641	
1154				
1155	02231	1332	L1640, TAD CORRE9	

1156	02232	7200	CLA
1157	02233	3723	DCA I SS1
1158			
1159	02234	1312	L1641, TAD S2
1160	02235	1331	TAD DM
1161	02236	7510	SPA
1162	02237	5244	JMP L1642
1163	02240	7200	CLA
1164	02241	1330	TAD DP
1165	02242	3724	DCA I SS2
1166	02243	5247	JMP L1643
1167			
1168	02244	1332	L1642, TAD CORRE9
1169	02245	7200	CLA
1170	02246	3724	DCA I SS2
1171			
1172	02247	1313	L1643, TAD S3
1173	02250	1331	TAD DM
1174	02251	7510	SPA
1175	02252	5257	JMP L1644
1176	02253	7200	CLA
1177	02254	1330	TAD DP
1178	02255	3725	DCA I SS3
1179	02256	5262	JMP L1645
1180			
1181	02257	1332	L1644, TAD CORRE9
1182	02260	7200	CLA
1183	02261	3725	DCA I SS3
1184			
1185	02262	1314	L1645, TAD S4
1186	02263	1331	TAD DM
1187	02264	7510	SPA
1188	02265	5272	JMP L1646
1189	02266	7200	CLA
1190	02267	1330	TAD DP
1191	02270	3726	DCA I SS4
1192	02271	5275	JMP L1647
1193			
1194	02272	1332	L1646, TAD CORRE9
1195	02273	7200	CLA
1196	02274	3726	DCA I SS4
1197			
1198	02275	1315	L1647, TAD SS555
1199	02276	1331	TAD DM
1200	02277	7510	SPA
1201	02300	5305	JMP L1648
1202	02301	7200	CLA
1203	02302	1330	TAD DP
1204	02303	3727	DCA I SS5
1205	02304	5310	JMP L1649
1206			
1207	02305	1332	L1648, TAD CORRE9
1208	02306	7200	CLA
1209	02307	3727	DCA I SS5
1210			

1211	02310	5600	L1649, JMP I IN	
1212				
1213	02311	0000	S1, 0	
1214	02312	0000	S2, 0	
1215	02313	0000	S3, 0	
1216	02314	0000	S4, 0	
1217	02315	0000	SS555, 0	
1218	02316	0001	CC, 1	
1219	02317	0002	C2, 2	
1220	02320	0003	C3, 3	
1221	02321	0004	C4, 4	
1222	02322	0005	C5, 5	
1223	02323	1747	SS1, SAM1	
1224	02324	1750	SS2, SAM2	
1225	02325	1751	SS3, SAM3	
1226	02326	1752	SS4, SAM4	
1227	02327	1753	SS5, SAM5	
1228	02330	0000	DP, 0	
1229	02331	0000	DM, 0	
1230	02332	0000	CORRE9, 0	
1231				
1232	02333	0000	ADC, 0	/ A/D CONVERSION ROUTINE.
1233	02334	6531	ADLM	
1234	02335	6532	ADST	
1235	02336	6534	ADSK	
1236	02337	5336	JMP .-1	
1237	02340	6533	ADRB	
1238	02341	5733	JMP I ADC	
1239				
1240	02342	0000	INA, 0	/EXAMINATION OF LEADING
1241	02343	7200	CLA	/CONTACT FOR BUBBLE
1242	02344	1371	TAD CCH1	/LEADING EDGE CORRECTION.
1243	02345	4333	JMS ADC	/SAMPLING OF CH1 FOUR
1244	02346	3372	DCA SA1	/TIMES FASTER THAN
1245	02347	1372	TAD SA1	/MAIN LOOP.
1246	02350	1774	TAD I DMA	
1247	02351	7510	SPA	
1248	02352	5357	JMP L26A	
1249	02353	7200	CLA	
1250	02354	1775	TAD I DPA	
1251	02355	3773	DCA I SA2	
1252	02356	5364	JMP L26B	
1253				
1254	02357	7200	L26A, CLA	
1255	02360	7000	NOP	
1256	02361	7000	NOP	
1257	02362	7000	NOP	
1258	02363	3773	DCA I SA2	
1259				
1260	02364	1376	L26B, TAD SRA	
1261	02365	3377	DCA SR	
1262	02366	2377	ISZ SR	
1263	02367	5366	JMP .-1	
1264	02370	5742	JMP I INA	
1265				

1266	02371	0001	CCH1,1
1267	02372	0000	SA1,0
1268	02373	1747	SA2,SAM1
1269	02374	2331	DMA,DM
1270	02375	2330	DPA,DP
1271	02376	7776	SRA,-2
1272	02377	0000	SR,0
1273			
1274		2400	*2400
1275	02400	0000	GETCON,0
1276			
1277	02401	7200	CLA
1278	02402	1635	TAD I P33
1279	02403	1636	TAD I P44
1280	02404	1637	TAD I P55
1281	02405	1240	TAD MINUS3
1282	02406	7500	SMA
1283	02407	5217	JMP L3001
1284			
1285	02410	7200	CLA
1286	02411	1241	TAD CORR
1287	02412	3242	DCA COUN
1288	02413	2242	ISZ COUN
1289	02414	5213	JMP --1
1290	02415	4643	JMS I CODEPH
1291	02416	5600	JMP I GETCON
1292			
1293	02417	7200	L3001,CLA
1294	02420	1664	TAD I PA3
1295	02421	1665	TAD I PA4
1296	02422	1666	TAD I PA5
1297	02423	1240	TAD MINUS3
1298	02424	7500	SMA
1299	02425	5232	JMP L3002
1300			
1301	02426	7000	NOP
1302	02427	7200	CLA
1303	02430	4644	JMS I COBUPH
1304	02431	5600	JMP I GETCON
1305			
1306	02432	7200	L3002,CLA
1307	02433	4645	JMS I COWKPH
1308	02434	5600	JMP I GETCON
1309	02435	1733	P33,P3
1310	02436	1734	P44,P4
1311	02437	1735	P55,P5
1312	02440	7775	MINUS3,-3
1313	02441	7775	CORR,-3
1314	02442	0000	COUN,0
1315	02443	1200	CODEPH,DENCON
1316	02444	1222	COBUPH,BUBCON
1317	02445	1244	COWKPH,WAKCON
1318			
1319			
1320			

/EXAMINATION OF PULSE MODE
/OF THREE UPPER CONTACTS.

1321	02446	0000	FINISH,0	
1322	02447	7200	CLA	
1323	02450	1662	TAD I PA1	
1324	02451	1663	TAD I PA2	
1325	02452	1664	TAD I PA3	
1326	02453	1665	TAD I PA4	
1327	02454	1666	TAD I PA5	
1328	02455	1270	TAD MINUS5	
1329	02456	7440	SZA	
1330	02457	5667	JMP I L6271	
1331				
1332	02460	7200	CLA	
1333	02461	5646	JMP I FINISH	
1334				
1335	02462	1725	PA1,P10	
1336	02463	1726	PA2,P20	
1337	02464	1727	PA3,P30	
1338	02465	1730	PA4,P40	
1339	02466	1731	PA5,P50	
1340	02467	2001	L6271,L627	
1341	02470	7773	MINUS5,-5	
1342				
1343				
1344	02471	0000	TEST,0	/TEST THE PULSE SEQUENCE
1345				/FOR THE DESIRED ANGLE
1346				/OF APPROACH OF THE
1347				/BUBBLE INTERFACE.
1348				
1349	02472	7200	CLA	/PRELIMINARY TEST FOR
1350	02473	1751	TAD I CDEL2I	/ZERO DELAY COUNTERS
1351	02474	7450	SNA	/ON ANY CHANNEL.
1352				
1353	02475	5755	JMP I L5I	/CH2 ZERO, ->START.
1354	02476	7200	CLA	
1355	02477	1752	TAD I CDEL3I	
1356	02500	7450	SNA	
1357	02501	5755	JMP I L5I	
1358	02502	7200	CLA	
1359	02503	1753	TAD I CDEL4I	
1360	02504	7450	SNA	
1361	02505	5755	JMP I L5I	
1362	02506	7200	CLA	
1363	02507	1754	TAD I CDEL5I	
1364	02510	7450	SNA	
1365	02511	5755	JMP I L5I	
1366				
1367	02512	7200	CLA	/ALL OUTER CHANNELS HAVE
1368				/DELAYS.
1369				
1370	02513	1752	TAD I CDEL3I	/SECONDARY TEST FOR DELAY
1371	02514	3756	DCA I D1I	/VARIATION BETWEEN OUTER
1372	02515	1753	TAD I CDEL4I	/CHANNELS.
1373	02516	3757	DCA I D2I	
1374	02517	1754	TAD I CDEL5I	
1375	02520	3760	DCA I D3I	

1376				
1377	02521	4761	JMS I MAXM	/GET MAXIMUM AND MINIMUM.
1378	02522	4762	JMS I RATIO	/CALCULATE RATIO.
1379	02523	7200	CLA	/COMPARE WITH DESIRED.
1380	02524	1763	TAD I RAT1	
1381	02525	7041	CIA	
1382	02526	1367	TAD RATDD	
1383	02527	7500	SMA	
1384	02530	5755	JMP I L5I	/PULSES NOT COINCIDENT,
1385				/RETURN TO START.
1386				
1387	02531	7200	CLA	/PULSES COINCIDENT,
1388				/PROCEED.
1389				
1390	02532	1764	TAD I CPUL3I	/TEST FOR PULSE DURATION
1391	02533	3756	DCA I D1I	/VARIATION.
1392	02534	1765	TAD I CPUL4I	
1393	02535	3757	DCA I D2I	
1394	02536	1766	TAD I CPUL5I	
1395	02537	3760	DCA I D3I	
1396	02540	4761	JMS I MAXM	
1397	02541	4762	JMS I RATIO	
1398	02542	7200	CLA	
1399	02543	1763	TAD I RAT1	
1400	02544	7041	CIA	
1401	02545	1370	TAD RATPD	
1402	02546	7500	SMA	
1403	02547	5755	JMP I L5I	/PULSES OF DIFFERENT
1404				/LENGTH, RETURN TO START.
1405				
1406	02550	5671	JMP I TEST	/RETURN.
1407				
1408	02551	1736	CDEL2I, CDEL2	
1409	02552	1737	CDEL3I, CDEL3	
1410	02553	1740	CDEL4I, CDEL4	
1411	02554	1741	CDEL5I, CDEL5	
1412				
1413	02555	1343	L5I, RESET	
1414	02556	2704	D1I, D1	
1415	02557	2705	D2I, D2	
1416	02560	2706	D3I, D3	
1417	02561	2600	MAXM, MAXMIN	
1418	02562	2707	RATIO, RATC	
1419	02563	2767	RAT1, RAT01	
1420	02564	1744	CPUL3I, CPUL3	
1421	02565	1745	CPUL4I, CPUL4	
1422	02566	1746	CPUL5I, CPUL5	
1423				
1424	02567	0000	RATDD, 0	
1425	02570	0000	RATPD, 0	
1426				
1427		2600	*2600	
1428	02600	0000	MAXMIN, 0	
1429				/CALCULATE THE MAXIMUM
1430				/AND MINIMUM OF THREE

/BINARY NUMBERS WHICH
/REPRESENT TIME.

1431			
1432			
1433	02601	7200	CLA
1434	02602	1301	TAD C3777
1435	02603	3303	DCA MIN
1436	02604	3302	DCA MAX
1437	02605	1302	TAD MAX
1438	02606	7041	CIA
1439	02607	1304	TAD D1
1440	02610	7540	SMA SZA
1441	02611	5227	JMP L1430
1442			
1443	02612	7200	L1433, CLA
1444	02613	1302	TAD MAX
1445	02614	7041	CIA
1446	02615	1305	TAD D2
1447	02616	7540	SMA SZA
1448	02617	5233	JMP L1431
1449			
1450	02620	7200	L1434, CLA
1451	02621	1302	TAD MAX
1452	02622	7041	CIA
1453	02623	1306	TAD D3
1454	02624	7540	SMA SZA
1455	02625	5237	JMP L1432
1456	02626	5242	JMP L1435
1457			
1458	02627	7200	L1430, CLA
1459	02630	1304	TAD D1
1460	02631	3302	DCA MAX
1461	02632	5212	JMP L1433
1462			
1463	02633	7200	L1431, CLA
1464	02634	1305	TAD D2
1465	02635	3302	DCA MAX
1466	02636	5220	JMP L1434
1467			
1468	02637	7200	L1432, CLA
1469	02640	1306	TAD D3
1470	02641	3302	DCA MAX
1471			
1472	02642	7200	L1435, CLA
1473	02643	1303	TAD MIN
1474	02644	7041	CIA
1475	02645	1304	TAD D1
1476	02646	7510	SPA
1477	02647	5265	JMP L1439
1478			
1479	02650	7200	L1437, CLA
1480	02651	1303	TAD MIN
1481	02652	7041	CIA
1482	02653	1305	TAD D2
1483	02654	7510	SPA
1484	02655	5271	JMP L1438
1485			

1486	02656	7200	L1452, CLA
1487	02657	1303	TAD MIN
1488	02660	7041	CIA
1489	02661	1306	TAD D3
1490	02662	7510	SPA
1491	02663	5275	JMP L1451
1492	02664	5600	JMP I MAXMIN
1493			
1494	02665	7200	L1439, CLA
1495	02666	1304	TAD D1
1496	02667	3303	DCA MIN
1497	02670	5250	JMP L1437
1498			
1499	02671	7200	L1438, CLA
1500	02672	1305	TAD D2
1501	02673	3303	DCA MIN
1502	02674	5256	JMP L1452
1503			
1504	02675	7200	L1451, CLA
1505	02676	1306	TAD D3
1506	02677	3303	DCA MIN
1507	02700	5600	JMP I MAXMIN
1508			
1509	02701	3777	C3777, 3777
1510	02702	0000	MAX, 0
1511	02703	0000	MIN, 0
1512	02704	0000	D1, 0
1513	02705	0000	D2, 0
1514	02706	0000	D3, 0
1515			
1516			
1517			
1518	02707	0000	RATC, 0
1519			
1520			
1521			
1522			
1523	02710	7200	CLA
1524	02711	1302	TAD MAX
1525	02712	3335	DCA FLT1+1
1526	02713	1303	TAD MIN
1527	02714	3340	DCA FLT2+1
1528	02715	4407	JMS I 7
1529	02716	5350	FGET CF100
1530	02717	7000	FNOR
1531	02720	6350	FPUT CF100
1532	02721	5334	FGET FLT1
1533	02722	7000	FNOR
1534	02723	6342	FPUT FLT3
1535	02724	5337	FGET FLT2
1536	02725	7000	FNOR
1537	02726	4342	FDIV FLT3
1538	02727	3350	FMPY CF100
1539	02730	6345	FPUT RATO
1540	02731	0000	FEXT

/CALCULATE THE RATIO
/OF MINIMUM TO MAXIMUM
/TIME IN FLOATING POINT
/MODE.

1541	02732	4353	JMS OCTC
1542	02733	5707	JMP I RATO
1543	02734	0013	FLT1,13
1544	02735	0000	0
1545	02736	0000	0
1546	02737	0013	FLT2,13
1547	02740	0000	0
1548	02741	0000	0
1549	02742	0000	FLT3,0
1550	02743	0000	0
1551	02744	0000	0
1552	02745	0000	RATO,0
1553	02746	0000	0
1554	02747	0000	0
1555	02750	0013	CF100,13
1556	02751	0144	144
1557	02752	0000	0
1558			
1559			
1560			
1561			
1562	02753	0000	OCTC,0
1563			
1564			
1565	02754	7200	CLA
1566	02755	1366	TAD M13
1567	02756	1345	TAD RATO
1568	02757	3370	DCA TP1
1569	02760	1346	TAD RATO+1
1570	02761	7110	L151,CLL RAR
1571	02762	2370	ISZ TP1
1572	02763	5361	JMP L151
1573			
1574	02764	3367	DCA RATO1
1575	02765	5753	JMP I OCTC
1576			
1577	02766	7765	M13,-13
1578	02767	0000	RATO1,0
1579	02770	0000	TP1,0
1580			
1581		3000	*3000
1582	03000	0000	CLEAR,0
1583	03001	7200	CLA
1584	03002	3636	DCA I DEL2
1585	03003	3637	DCA I DEL3
1586	03004	3640	DCA I DEL4
1587	03005	3641	DCA I DEL5
1588	03006	3642	DCA I PUL1
1589	03007	3643	DCA I PUL2
1590	03010	3644	DCA I PUL3
1591	03011	3645	DCA I PUL4
1592	03012	3646	DCA I PUL5
1593	03013	3647	DCA I PE2
1594	03014	3650	DCA I PE3
1595	03015	3651	DCA I PE4

/CONVERT FLOATING POINT
/NUMBER TO 12 BIT BINARY.

1596	03016	3652	DCA I PE5
1597	03017	3653	DCA I PE10
1598	03020	3654	DCA I PE20
1599	03021	3655	DCA I PE30
1600	03022	3656	DCA I PE40
1601	03023	3657	DCA I PE50
1602	03024	3635	DCA I CSD2I
1603			
1604	03025	3660	DCA I CP1Z
1605	03026	3661	DCA I CP2Z
1606	03027	3662	DCA I CP3Z
1607	03030	3663	DCA I CP4Z
1608	03031	3664	DCA I CP5Z
1609			
1610	03032	1306	TAD POINT3
1611	03033	3307	DCA PT3
1612			
1613	03034	5600	JMP I CLEAR
1614			
1615	03035	2153	CSD2I, CSD2
1616	03036	1736	DEL2, CDEL2
1617	03037	1737	DEL3, CDEL3
1618	03040	1740	DEL4, CDEL4
1619	03041	1741	DEL5, CDEL5
1620	03042	1742	PUL1, CPUL1
1621	03043	1743	PUL2, CPUL2
1622	03044	1744	PUL3, CPUL3
1623	03045	1745	PUL4, CPUL4
1624	03046	1746	PUL5, CPUL5
1625	03047	1732	PE2, P2
1626	03050	1733	PE3, P3
1627	03051	1734	PE4, P4
1628	03052	1735	PE5, P5
1629	03053	1725	PE10, P10
1630	03054	1726	PE20, P20
1631	03055	1727	PE30, P30
1632	03056	1730	PE40, P40
1633	03057	1731	PE50, P50
1634	03060	2146	CP1Z, CP1
1635	03061	2147	CP2Z, CP2
1636	03062	2150	CP3Z, CP3
1637	03063	2151	CP4Z, CP4
1638	03064	2152	CP5Z, CP5
1639			
1640			
1641	03065	0000	STORE, 0
1642	03066	7200	CLA
1643	03067	1327	TAD PT3
1644	03070	7041	CIA
1645	03071	1306	TAD POINT3
1646	03072	1710	TAD I SNUM
1647	03073	7510	SPA
1648	03074	5302	JMP LST1
1649	03075	7200	CLO
1650	03076	1711	TAD I SMP1

/STORE THE CH1 PROBE
/SIGNAL WHILE WAITING FOR
/THE CH1 PULSE.

1651	03077	3707	DCA I PT3
1652	03100	2307	ISZ PT3
1653	03101	5665	JMP I STORE
1654			
1655	03102	7200	LST1, CLA
1656	03103	1306	TAD POINT3
1657	03104	3307	DCA PT3
1658	03105	5665	JMP I STORE
1659			
1660	03106	4200	POINT3, 4200
1661	03107	0000	PT3, 0
1662	03110	3325	SNUM, SN
1663	03111	2372	SMP1, SA1
1664			
1665			
1666	03112	0000	BSTORE, 0
1667	03113	7200	CLA
1668	03114	1760	TAD I CD311
1669	03115	6211	CDF 10
1670	03116	3766	DCA I DCBS
1671	03117	6201	CDF 00
1672	03120	2366	ISZ DCBS
1673			
1674	03121	7200	CLA
1675	03122	1761	TAD I CD411
1676	03123	6211	CDF 10
1677	03124	3766	DCA I DCBS
1678	03125	6201	CDF 00
1679	03126	2366	ISZ DCBS
1680			
1681	03127	7200	CLA
1682	03130	1762	TAD I CD511
1683	03131	6211	CDF 10
1684	03132	3766	DCA I DCBS
1685	03133	6201	CDF 00
1686	03134	2366	ISZ DCBS
1687			
1688	03135	7200	CLA
1689	03136	1763	TAD I CP1111
1690	03137	6211	CDF 10
1691	03140	3766	DCA I DCBS
1692	03141	6201	CDF 00
1693	03142	2366	ISZ DCBS
1694			
1695	03143	7200	CLA
1696	03144	1764	TAD I CD2111
1697	03145	6211	CDF 10
1698	03146	3766	DCA I DCBS
1699	03147	6201	CDF 00
1700	03150	2366	ISZ DCBS
1701			
1702	03151	7200	CLA
1703	03152	1765	TAD I CP2111
1704	03153	6211	CDF 10
1705	03154	3766	DCA I DCBS

1706	03155	6201	CDF 00	
1707	03156	2366	ISZ DCBS	
1708				
1709	03157	5712	JMP I BSTORE	
1710	03160	1737	CD311, CDEL3	
1711	03161	1740	CD411, CDEL4	
1712	03162	1741	CD511, CDEL5	
1713	03163	1742	CP1111, CPUL1	
1714	03164	1736	CD2111, CDEL2	
1715	03165	1743	CP2111, CPUL2	
1716	03166	0000	DCBS, 0	
1717				
1718				
1719		3200	*3200	
1720	03200	0000	EXAM, 0	/EXAMINE THE PREVIOUSLY
1721	03201	7200	CLA	/STORED RISE OF THE CH1
1722	03202	1330	TAD M1	/PULSE FOR THE MINIMUM
1723	03203	1720	TAD I PT3I	/SPECIFIED RISE SLOPE
1724	03204	3321	DCA PT4	/AND CORRECT CH1 PULSE
1725	03205	1323	TAD C24	/DURATION.
1726	03206	7041	CIA	
1727	03207	1321	TAD PT4	/SET UP STORE COUNTERS.
1728	03210	3322	DCA PT5	
1729	03211	3334	DCA CCOR	/ZERO TIME COUNTER.
1730	03212	1322	TAD PT5	
1731	03213	7041	CIA	
1732	03214	1724	TAD I POINT2	
1733	03215	7540	SMA SZA	
1734	03216	5733	JMP I BACK	/STORAGE COUNTER <
1735				/MINIMUM STORE LOCATION,
1736				/TRY AGAIN.
1737	03217	7200	L222, CLA	
1738	03220	1322	TAD PT5	
1739	03221	7041	CIA	
1740	03222	1724	TAD I POINT2	/TEST STORE COUNTERS
1741	03223	7540	SMA SZA	/FOR TOP END OF STORE.
1742	03224	5277	JMP L223	
1743	03225	7200	CLA	
1744	03226	1321	TAD PT4	
1745	03227	7041	CIA	
1746	03230	1724	TAD I POINT2	
1747	03231	7540	SMA SZA	
1748	03232	5272	JMP L221	
1749	03233	7200	CLA	
1750	03234	1322	TAD PT5	
1751	03235	1720	TAD I PT3I	
1752	03236	7450	SNA	
1753	03237	5733	JMP I BACK	/COMPLETE STORE EXAMINED
1754				/WITHOUT SUCCESS, TRY
1755				/AGAIN.
1756				
1757				
1758				
1759	03240	7200	L224, CLA	/EXAMINE PULSE RISE SLOPE.
1760				

1761	03241	2334	ISZ CCOR	/INCREMENT TIME COUNTER
1762	03242	1721	TAD I PT4	
1763	03243	3326	DCA A	
1764	03244	1722	TAD I PT5	
1765	03245	3327	DCA B	
1766	03246	1327	TAD B	
1767	03247	7041	CIA	
1768	03250	1326	TAD A	
1769	03251	1335	TAD SLOPE	
1770	03252	7500	SMA	
1771	03253	5304	JMP L225	/SLOPE < DESIRED,
1772				/STOP EXAMINING.
1773	03254	7200	CLA	
1774	03255	1326	TAD A	
1775	03256	7041	CIA	
1776	03257	1327	TAD B	
1777	03260	7510	SPA	
1778	03261	5304	JMP L225	/SLOPE CHANGED DIRECTION,
1779				/STOP EXAMINING
1780	03262	7200	CLA	
1781	03263	1330	TAD M1	
1782	03264	1321	TAD PT4	
1783	03265	3321	DCA PT4	
1784	03266	1330	TAD M1	
1785	03267	1322	TAD PT5	
1786	03270	3322	DCA PT5	
1787	03271	5217	JMP L222	/SLOPE STILL TOO STEEP,
1788				/CONTINUE EXAMINING.
1789	03272	7200	L221, CLA	
1790	03273	1724	TAD I POINT2	
1791	03274	1325	TAD SN	
1792	03275	3321	DCA PT4	
1793	03276	5240	JMP L224	
1794				
1795	03277	7200	L223, CLA	
1796	03300	1325	TAD SN	
1797	03301	1724	TAD I POINT2	
1798	03302	3322	DCA PT5	
1799	03303	5240	JMP L224	
1800				
1801	03304	7200	L225, CLA	/ADJUST TIME COUNTER
1802	03305	1334	TAD CCOR	/FOR FASTER SAMPLING
1803	03306	7110	CLL RAR	/(DIV. BY FOUR)
1804	03307	7110	CLL RAR	
1805	03310	3334	DCA CCOR	
1806	03311	1731	TAD I CPC1	/ADD CORRECTOR TO CH1
1807	03312	1334	TAD CCOR	/PULSE DURATION.
1808	03313	3731	DCA I CPC1	
1809	03314	1732	TAD I CPC2	/ADD CORRECTOR TO CH2
1810	03315	1334	TAD CCOR	/PULSE DURATION.
1811	03316	3732	DCA I CPC2	
1812	03317	5600	JMP I EXAM	
1813				
1814	03320	3107	PT3I, PT3	
1815	03321	0000	PT4, 0	

1816	03322	0000	PT5,0
1817	03323	0024	C24,24
1818	03324	3106	POINT2,POINT3
1819	03325	0377	SN,377
1820	03326	0000	A,0
1821	03327	0000	R,0
1822	03330	7777	M1,-1
1823	03331	1742	CPC1,CPUL1
1824	03332	1743	CPC2,CPUL2
1825	03333	1343	BACK,RESET
1826	03334	0000	CCOR,0
1827	03335	0000	SLOPE,0
1828			
1829			
1830		3400	*3400
1831	03400	0000	PCDF,0
1832	03401	7200	CLA
1833	03402	1352	TAD PPLINE
1834	03403	3353	DCA PPL
1835	03404	6211	CDF 10
1836	03405	1640	TAD I DFMX
1837	03406	6201	CDF 00
1838	03407	7041	CIA
1839	03410	3241	DCA DFMX
1840	03411	7200	L4410,CLA
1841	03412	6211	CDF 10
1842	03413	1642	TAD I CCDF
1843	03414	6201	CDF 00
1844	03415	3751	DCA I OUT2
1845	03416	4754	JMS I PRC0
1846	03417	2353	ISZ PPL
1847	03420	5225	JMP L4415
1848	03421	4100	JMS Z CRLF
1849	03422	1352	TAD PPLINE
1850	03423	3353	DCA PPL
1851	03424	7410	SKP
1852	03425	4117	L4415,JMS Z SPACE
1853	03426	7200	CLA
1854	03427	2242	ISZ CCDF
1855	03430	7200	CLA
1856	03431	1242	TAD CCDF
1857	03432	1241	TAD DFMX
1858	03433	7440	SZA
1859	03434	5211	JMP L4410
1860	03435	2240	ISZ DFMX
1861	03436	7000	NOP
1862	03437	5600	L4430,JMP I PCDF
1863	03440	0000	DFMX,0
1864	03441	0000	DFMX,0
1865	03442	0000	CCDF,0
1866			
1867	03443	0000	PCBF,0
1868	03444	7200	CLA
1869	03445	1352	TAD PPLINE
1870	03446	3353	DCA PPL

1871	03447	6211	CDF 10
1872	03450	1703	TAD I BFMX
1873	03451	6201	CDF 00
1874	03452	7041	CIA
1875	03453	3304	DCA BFMX
1876	03454	7200	L5510, CLA
1877	03455	6221	CDF 20
1878	03456	1705	TAD I CCBF
1879	03457	6201	CDF 00
1880	03460	3751	DCA I OUT2
1881	03461	4754	JMS I PRCO
1882	03462	2353	ISZ PPL
1883	03463	5270	JMP L5515
1884	03464	4100	JMS Z CRLF
1885	03465	1352	TAD PPLINE
1886	03466	3353	DCA PPL
1887	03467	7410	SKP
1888	03470	4117	L5515, JMS Z SPACE
1889	03471	7200	CLA
1890	03472	2305	ISZ CCBF
1891	03473	7200	CLA
1892	03474	1305	TAD CCBF
1893	03475	1304	TAD BFMX
1894	03476	7440	SZA
1895	03477	5254	JMP L5510
1896	03500	2303	ISZ BFMX
1897	03501	7000	NOP
1898	03502	5643	L5530, JMP I PCBF
1899	03503	0000	BFMX, 0
1900	03504	0000	BFMX, 0
1901	03505	0000	CCBF, 0
1902			
1903	03506	0000	PCWF, 0
1904	03507	7200	CLA
1905	03510	1352	TAD PPLINE
1906	03511	3353	DCA PPL
1907	03512	6211	CDF 10
1908	03513	1746	TAD I WFMX
1909	03514	6201	CDF 00
1910	03515	7041	CIA
1911	03516	3347	DCA WFMX
1912	03517	7200	L6610, CLA
1913	03520	6231	CDF 30
1914	03521	1750	TAD I CCWF
1915	03522	6201	CDF 00
1916	03523	3751	DCA I OUT2
1917	03524	4754	JMS I PRCO
1918	03525	2353	ISZ PPL
1919	03526	5333	JMP L6615
1920	03527	4100	JMS Z CRLF
1921	03530	1352	TAD PPLINE
1922	03531	3353	DCA PPL
1923	03532	7410	SKP
1924	03533	4117	L6615, JMS Z SPACE
1925	03534	7200	CLA

1926	03535	2350	ISZ CCWF
1927	03536	7200	CLA
1928	03537	1350	TAD CCWF
1929	03540	1347	TAD WFMX
1930	03541	7440	SZA
1931	03542	5317	JMP L6610
1932	03543	2346	ISZ WFMX
1933	03544	7000	NOP
1934	03545	5706	L6630, JMP I PCWF
1935	03546	0000	WFMX, 0
1936	03547	0000	WFMX, 0
1937	03550	0000	CCWF, 0
1938			
1939	03551	3653	OUT2, OUTC2
1940	03552	7766	PPLINE, -12
1941	03553	0000	PPL, 0
1942	03554	3600	PRCO, PRICON
1943			
1944		3600	*3600
1945	03600	0000	PRICON, 0
1946	03601	7200	CLA
1947	03602	1625	TAD I DRI1
1948	03603	3241	DCA NDRIFT
1949	03604	1626	TAD I DRI2
1950	03605	3242	DCA NDRIFT+1
1951	03606	1627	TAD I DRI3
1952	03607	3243	DCA NDRIFT+2
1953	03610	4407	JMS I 7
1954	03611	5233	FGET STDC1
1955	03612	7000	FNOR
1956	03613	6236	FPUT SLT14
1957	03614	5252	FGET OUTC1
1958	03615	7000	FNOR
1959	03616	2241	FSUB NDRIFT
1960	03617	3236	FMPY SLT14
1961	03620	4244	FDIV NSTDV1
1962	03621	6247	FPUT FCOUT1
1963	03622	0000	FEXT
1964	03623	4256	JMS FBCONV
1965	03624	5600	JMP I PRICON
1966	03625	0543	DRI1, DRIFT1
1967	03626	0544	DRI2, DRIFT2
1968	03627	0545	DRI3, DRIFT3
1969	03630	0767	MX1, MAX1
1970	03631	0770	MX2, MAX2
1971	03632	0771	MX3, MAX3
1972	03633	0013	STDC1, 13
1973	03634	0000	STDC2, 0
1974	03635	0000	STDC3, 0
1975	03636	0000	SLT14, 0
1976	03637	0000	0
1977	03640	0000	0
1978	03641	0000	NDRIFT, 0
1979	03642	0000	0
1980	03643	0000	0

1981	03644	0000	NSTDV1,0
1982	03645	0000	NSTDV2,0
1983	03646	0000	NSTDV3,0
1984	03647	0000	FCOUT1,0
1985	03650	0000	0
1986	03651	0000	0
1987	03652	0013	OUTC1,13
1988	03653	0000	OUTC2,0
1989	03654	0000	OUTC3,0
1990			
1991	03655	4000	PRIN,PRINT4
1992			
1993	03656	0000	FBCONV,0
1994	03657	1270	TAD ZZ13
1995	03660	1247	TAD FCOUT1
1996	03661	3271	DCA FB1
1997	03662	1250	TAD FCOUT1+1
1998	03663	7110	L4921,CLL RAR
1999	03664	2271	ISZ FB1
2000	03665	5263	JMP L4921
2001	03666	4655	JMS I PRIN
2002	03667	5656	JMP I FBCONV
2003	03670	7765	ZZ13,-13
2004	03671	0000	FB1,0
2005			
2006			
2007			
2008	03672	0000	PRT1,0
2009	03673	4100	JMS Z CRLF
2010			
2011	03674	7200	CLA
2012	03675	6211	CDF 10
2013	03676	1777	TAD I AA7000
2014	03677	6201	CDF 00
2015	03700	3371	DCA CD31
2016	03701	2377	ISZ AA7000
2017			
2018	03702	7200	CLA
2019	03703	6211	CDF 10
2020	03704	1777	TAD I AA7000
2021	03705	6201	CDF 00
2022	03706	3372	DCA CD41
2023	03707	2377	ISZ AA7000
2024			
2025	03710	7200	CLA
2026	03711	6211	CDF 10
2027	03712	1777	TAD I AA7000
2028	03713	6201	CDF 00
2029	03714	3373	DCA CD51
2030	03715	2377	ISZ AA7000
2031			
2032	03716	7200	CLA
2033	03717	1371	TAD CD31
2034	03720	4655	JMS I PRIN
2035	03721	4100	JMS Z CRLF

2036			
2037	03722	7200	CLA
2038	03723	1372	TAD CD41
2039	03724	4655	JMS I PRIN
2040	03725	4100	JMS Z CRLF
2041			
2042	03726	7200	CLA
2043	03727	1373	TAD CD51
2044	03730	4655	JMS I PRIN
2045	03731	4117	JMS Z SPACE
2046			
2047	03732	7200	CLA
2048	03733	6211	CDF 10
2049	03734	1777	TAD I AA7000
2050	03735	6201	CDF 00
2051	03736	3374	DCA CP111
2052	03737	2377	ISZ AA7000
2053			
2054	03740	7200	CLA
2055	03741	1374	TAD CP111
2056	03742	4655	JMS I PRIN
2057	03743	4117	JMS Z SPACE
2058			
2059	03744	7200	CLA
2060	03745	6211	CDF 10
2061	03746	1777	TAD I AA7000
2062	03747	6201	CDF 00
2063	03750	3375	DCA CD211
2064	03751	2377	ISZ AA7000
2065			
2066	03752	7200	CLA
2067	03753	1375	TAD CD211
2068	03754	4655	JMS I PRIN
2069	03755	4117	JMS Z SPACE
2070			
2071	03756	7200	CLA
2072	03757	6211	CDF 10
2073	03760	1777	TAD I AA7000
2074	03761	6201	CDF 00
2075	03762	3376	DCA CP211
2076	03763	2377	ISZ AA7000
2077			
2078	03764	7200	CLA
2079	03765	1376	TAD CP211
2080	03766	4655	JMS I PRIN
2081	03767	4117	JMS Z SPACE
2082	03770	5672	JMP I PRT1
2083			
2084	03771	0000	CD31,0
2085	03772	0000	CD41,0
2086	03773	0000	CD51,0
2087	03774	0000	CP111,0
2088	03775	0000	CD211,0
2089	03776	0000	CP211,0
2090	03777	0000	AA7000,0

2091			
2092		4000	*4000
2093	04000	0000	PRINT4,0
2094	04001	7100	CLL
2095	04002	7510	SPA
2096	04003	7061	CML CMA IAC
2097	04004	3241	DCA VAL
2098	04005	3237	DCA BOX
2099	04006	1236	TAD CTR
2100	04007	3240	DCA CNT
2101	04010	1235	TAD ADDR
2102	04011	3217	DCA XYZ+1
2103	04012	1246	TAD C240
2104	04013	7430	SZL
2105	04014	1247	TAD C7615
2106	04015	4111	JMS Z TYPE
2107	04016	1241	XYZ, TAD VAL
2108	04017	1242	TAD CON
2109	04020	7510	SPA
2110	04021	5225	JMP .+4
2111	04022	2237	ISZ BOX
2112	04023	3241	DCA VAL
2113	04024	5216	JMP XYZ
2114	04025	7200	CLA
2115	04026	1237	TAD BOX
2116	04027	4250	JMS OUT
2117	04030	3237	DCA BOX
2118	04031	2217	ISZ XYZ+1
2119	04032	2240	ISZ CNT
2120	04033	5216	JMP XYZ
2121	04034	5600	JMP I PRINT4
2122	04035	1242	ADDR, TAD CON
2123	04036	7774	CTR, -4
2124	04037	0000	BOX, 0
2125	04040	0000	CNT, 0
2126	04041	0000	VAL, 0
2127	04042	6030	CON, 6030
2128	04043	7634	7634
2129	04044	7765	7765
2130	04045	7777	7777
2131	04046	0240	C240, 240
2132	04047	7615	C7615, 7615
2133			
2134	04050	0000	OUT, 0
2135	04051	3305	DCA V1
2136	04052	3304	DCA V2
2137	04053	7100	CLL
2138	04054	1306	TAD V3
2139	04055	3307	DCA V4
2140	04056	1305	TAD V1
2141	04057	7010	RAR
2142	04060	7430	SZL
2143	04061	2304	ISZ V2
2144	04062	7000	NOP
2145	04063	2307	ISZ V4

2146	04064	5257	JMP .-5
2147	04065	7300	CLA CLL
2148	04066	1304	TAD V2
2149	04067	7310	RAK
2150	04070	7430	S4L
2151	04071	5277	JMP ST1
2152	04072	7200	CLA
2153	04073	1305	TAD V1
2154	04074	1311	TAD C60
2155	04075	4111	JMS Z TYPE
2156	04076	5650	JMP I OUT
2157			
2158	04077	7300	ST1, CLA CLL
2159	04100	1305	TAD V1
2160	04101	1310	TAD C260
2161	04102	4111	JMS Z TYPE
2162	04103	5650	JMP I OUT
2163			
2164	04104	0000	V2, 0
2165	04105	0200	V1, 0
2166	04106	7774	V3, -4
2167	04107	0000	V4, 0
2168	04110	0260	C260, 260
2169	04111	0060	C60, 60
2170			
2171			
2172	04112	0000	SET2, 0
2173	04113	7200	CLA
2174	04114	3723	DCA I DFPOI1
2175	04115	3726	DCA I CCDF9
2176	04116	3724	DCA I BBPOI1
2177	04117	3727	DCA I CCBF9
2178	04120	3725	DCA I WFPOI1
2179	04121	3730	DCA I CCWF9
2180	04122	5712	JMP I SET2
2181	04123	1220	DFPOI1, DFPOI
2182	04124	1242	BBPOI1, BBPOI
2183	04125	1264	WFPOI1, WFPOI
2184	04126	3442	CCDF9, CCDF
2185	04127	3505	CCBF9, CCBF
2186	04130	3550	CCWF9, CCWF
2187			
2188	04131	0000	SET1, 0
2189	04132	7200	CLA
2190	04133	1371	TAD A5777
2191	04134	3766	DCA I DFI
2192	04135	1372	TAD A6077
2193	04136	3767	DCA I BFI
2194	04137	1373	TAD A6177
2195	04140	3770	DCA I WFI
2196	04141	1354	TAD A6000
2197	04142	3761	DCA I DFMAX9
2198	04143	1355	TAD A6100
2199	04144	3762	DCA I BFMAX9
2200	04145	1356	TAD A6200

2201	04146	3763	DCA I WFMAX9
2202	04147	1360	TAD A7000
2203	04150	3764	DCA I DCBSI
2204	04151	1360	TAD A7000
2205	04152	3765	DCA I AB7000
2206	04153	5731	JMP I SET1
2207	04154	6200	A6000,6000
2208	04155	6100	A6100,6100
2209	04156	6200	A6200,6200
2210	04157	6300	A6300,6300
2211	04160	7000	A7000,7000
2212	04161	3440	DFMAX9,DFMAX
2213	04162	3503	BFMAX9,BFMAX
2214	04163	3546	WFMAX9,WFMAX
2215	04164	3166	DCBSI,DCBS
2216	04165	3777	AB7000,AA7000
2217	04166	1340	DFI,DF
2218	04167	1341	BFI,BF
2219	04170	1342	WFI,WF
2220	04171	5777	A5777,5777
2221	04172	6077	AA6077,6077
2222	04173	6177	A6177,6177
2223			
2224			
2225			
2226			\$

C.3 Timing methods

The bubble velocity and size measurements were obtained by counting the number of times the computer followed a particular logic train whilst undergoing the real time examination of the compound probe output. For example, consider the detection and storage of the voltage on channel 0 corresponding to a gas composition within the bubble, BUBCON routine in page 2.1 of the program listing:

Symbolic micro-instructions	time (μ sec)
BUBCON,0	
CLA	1.2
JMS I ADC4	3.8 + 28.4
CDF 20	1.2
DCA I BBPOI	2.4
CDF 00	1.2
ISZ BBPOI	2.4
NOP	1.2
TAD BBPOI	2.6
TAD BBPOIM	2.6
SZA	1.2
JMP L795	1.2
CLA	1.2
JMP I STOP	1.2
L795, CLA	1.2
JMP I BUBCON	2.4

The above routine thus requiring 53.0 μ sec to be executed.

Let us note that the instructions

CLA

JMP I STOP

are not accounted for as they correspond to a trap which will halt the computer due to overwriting of the CTHWA output in memory, thus indicating that the experiment should be repeated.

The instruction

JMS I ADC4

calls indirectly the A/D conversion routine in page 3-4 of the program listing,

ADC , 0

ADLM

ADST

ADSK

JMP .-1

ADRB

JMP I ADC

which requires 28.4 μ sec for its execution.

The complete PAL-8 logic sequence times outlined in the program listing were timed in the above manner to yield a total loop time of 500 μ sec.

APPENDIX D

PUBLISHED WORK

D.1 MASS TRANSFER IN THE SPINNING CATALYST BASKET REACTOR

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Chemical Engineering Science, 30, 167-175 (1975)

MASS TRANSFER IN THE SPINNING CATALYST BASKET REACTOR

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Received 28 January 1974; accepted 17 June 1974)

Abstract—A mathematical model is derived for the evaluation of the relative velocity between the gas and the pellets in two configurations of the spinning catalyst basket reactor. This slip velocity is then used to calculate the particle Reynolds number and hence mass transfer coefficients from well known correlations for mass transfer in packed beds. Experimentally determined mass transfer coefficients were found to be in good agreement with the model. The effect on mass transfer rates of the position of the pellet and rotor dimensions for the cruciform spinning catalyst basket reactor is seen to be well predicted from the model. The performance of a typical spinning catalyst basket reactor is calculated when a catalytic oxidation reaction takes place at various jacket temperatures and rotational speeds.

INTRODUCTION

The spinning catalyst basket reactor (SCBR) has been used to measure the kinetics of catalytic reactions since it was developed for this purpose by Carberry *et al.*[1] and Brisk *et al.*[2]. A general review of different reactors of this type was presented by Choudary *et al.*[3] and Bennet *et al.*[4]. From the above references one concludes that near-ideal mixing is readily obtained, but much higher rotational speeds are needed to achieve the heat and mass transfer rates that will be necessary to ensure insignificant temperature and concentration gradients at the catalyst surface at the rates of reaction which may require measurement. Relations between the particle-gas mass transfer coefficients and rotating basket speed were reported by Brisk *et al.*[2, 5] and Choudary *et al.*[3]. They found that mass transfer rates were little affected by the position of the pellet in the basket.

In the present work mass transfer coefficients were experimentally determined and a mathematical model derived to calculate the gas velocity through the rotating bed in the two more common SCBR geometries, the cylindrical and the cruciform spinning baskets. The mathematical model permits one to calculate the mass transfer coefficients, and by analogy heat transfer coefficients, using published empirical relations for packed beds.

EXPERIMENTAL

A model of the SCBR was constructed in "Perspex" and comprised an unbaffled cylindrical vessel (14.2 cm i.d. and 6.2 cm high), containing a

rotor supported and driven by a shaft connected to a variable speed motor.

The rotor was formed from two "Perspex" discs (13.2 cm dia. and 0.635 cm thick) separated and connected through their centres by a "Perspex" rod (dia. 1.905 cm and 4.45 cm length). The discs were systematically drilled and rods of stainless steel (dia. 0.158 cm and 5.08 cm length) were passed through them to support the particles forming the simulated catalyst bed. Different patterns of holes were drilled to provide different packing geometries and many more large holes were drilled for gas circulation. Figure 1 shows the disc and the different configurations of SCBR used in this work are shown in Table 1.

In reactors T1, T2 and T3 the particles in the basket, supported by the stainless steel rods, were glass spheres of 0.635 cm diameter in a cubic arrangement ($\epsilon = 0.476$). In reactor T4 the particles in the basket were alundum spheres of 0.642 cm dia., the calculated porosity of the bed being 0.364.

In operation air entered beneath the centre of the bottom disc and left above the top disc, the temperature being measured to within 0.1°C by a thermometer inserted within the reactor through a sealed port.

The rate of sublimation of naphthalene spheres into the air stream at 25°C in the different rotating baskets described above was measured at steady state conditions.

Naphthalene spheres were cast (diameter 0.635 ± 0.009 cm) and for each run two spheres of naphthalene were inserted in the basket replacing

Table 1. Characteristics of the CBSR.

Reactor	Geometry	Thickness of the arm (cm)	Length of the arm (cm)	Packing spheres size (cm)	Bed porosity
T1	Cruciform	0.635	5.08	Glass—0.635	0.476
T2	Cruciform	1.905	5.08	Glass—0.635	0.476
T3	Cruciform	0.635	Variable	Glass—0.635	0.476
T4	Cylindrical	Diameter—12.00 cm		Alundun 0.642	0.364

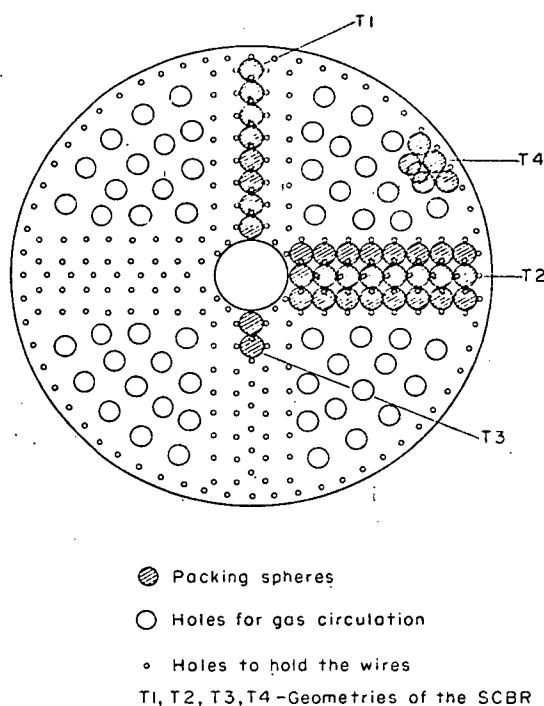


Fig. 1. Disc.

two inert spheres which otherwise constituted the packing.

The concentration of naphthalene in the gas leaving the reactor was determined with a spectrophotometer (Pye Unicam SP. 500) set at 2200 Å [6]. A special housing was constructed to hold an approximately 40 cm path length cell with quartz windows.

Under steady state conditions, in a well mixed CSTR

$$CF = K_g S (C_0 - C) \quad (1)$$

The experimental results show that the mass transfer coefficient increases with the rotating speed

as well as the radial position of the active pellet for all the reactor geometries tested.

In Figs. 2 and 3 the experimental mass transfer coefficient is plotted versus the rotary speed of the basket for different radial positions of the active pellet in reactors T1 and T4 respectively. In Fig. 4 the mass transfer coefficient is plotted against the rotary speed of the basket for reactor T2, at two different radial active pellet positions. The effect of active pellet circumferential position within the basket is also shown. Figure 5 shows the influence of the gap width between the arm and reactor wall on the mass transfer coefficient at different radial positions of the active pellet; in Fig. 5 the mass transfer coefficient is plotted vs. $1_a/L$ at 1705 rev/min. Similar results were obtained at 472, 1010 and 2630 rev/min.

The flow rate of air through the reactor was 500 cm³/min for all the experiments here reported, and no variation in K_g was detected when the flow rate was changed in the range of 250–1000 cm³/min.

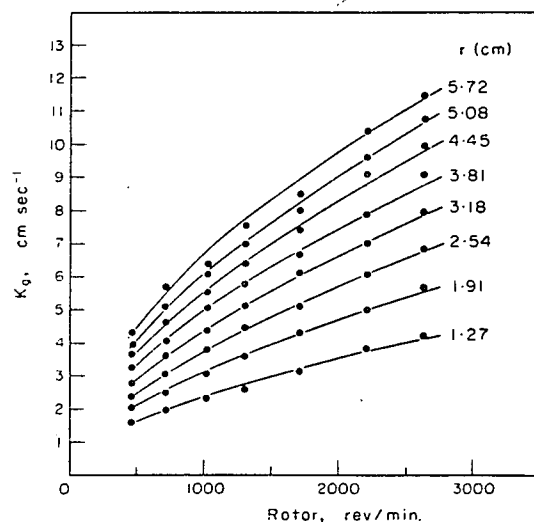


Fig. 2. Mass transfer coefficient dependence on radial position of the pellet and rotating speed in reactor T1 ($Sc = 2.53$, $T = 25^\circ C$).

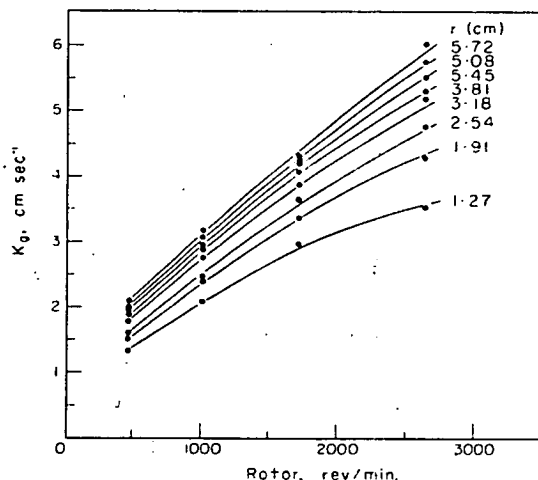


Fig. 3. Mass transfer coefficient dependence on radial position of the pellet and rotating speed in reactor T4 ($Sc = 2.53$, $T = 25^\circ C$).

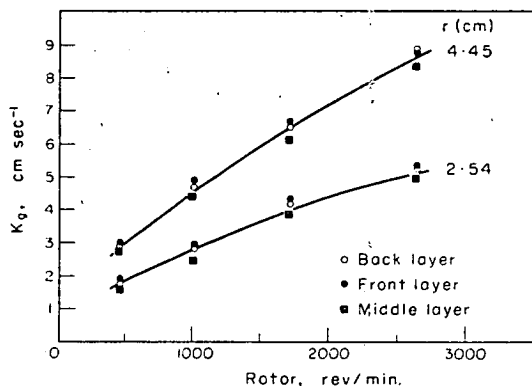


Fig. 4. Influence of the thickness of the arm, relative position of the active pellet and rotating speed upon the mass transfer coefficient ($Sc = 2.53$, $T = 25^\circ C$).

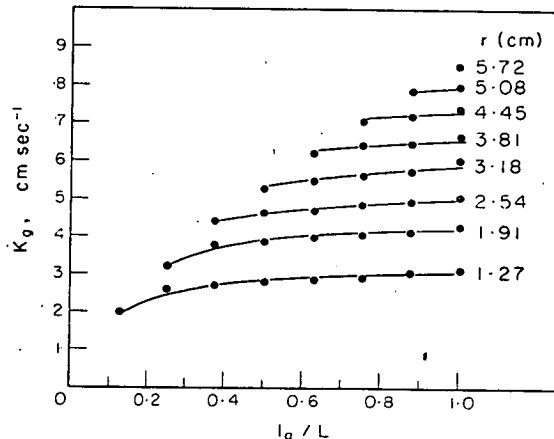


Fig. 5. Influence of the gap width between the arm and reactor wall on the mass transfer coefficient at different position of the active pellet at 1705 rev/min ($Sc = 2.53$, $T = 25^\circ C$).

All the experiments were carried out under ideal conditions of mixing, the reactor being tested by the step change method. Thus the gas to the reactor was changed from helium to nitrogen at $t = 0$, without variation in the flow rate, and the composition of the gas leaving the reactor analysed over a period of time by means of a katharometer.

MATHEMATICAL MODEL OF GAS FLOW IN SCBR

(a) Spinning basket with rotor arms

Consider a cylindrical CSTR in which a rotor with radial arms revolves. The gas between the arms will revolve and also move outwards. The movement of each element of gas can probably be best described by a logarithmic spiral, the equation for which is

$$r = r_0 e^{ct} = r_0 e^{c\omega t} \quad (2)$$

where c is a dimensionless constant, and r the radial position of the element of gas.

The acceleration impressed on the fluid will have two components, the radial acceleration, a_r , and the tangential acceleration, a_θ . The pressure-head will increase radially from r_1 to r_2 because of the radial component of the acceleration, and will decrease at a given radius from the front of one arm to the back of the next due to the tangential acceleration component.

If we consider an element of gas trapped between two consecutive arms, the mass of this element of gas is

$$dm = \rho \alpha r dr dz \quad (3)$$

where ρ is the density of the gas, and α the angle between the arms.

The tangential force applied to this element of gas producing the tangential acceleration, a_θ , is

$$F_\theta = a_\theta dm \quad (4)$$

and if P_θ is the pressure-head between the front of one arm and the back of the next at a given radial distance, i.e. the pressure-head across the element of fluid, F_θ is also given by,

$$F_\theta = P_\theta dr dz \quad (5)$$

and from Eqs (3)–(5)

$$P_\theta = \rho a_\theta r \alpha \quad (6)$$

The tangential acceleration can be calculated

from Eq. (2) and Eq. (7) below:

$$a_\theta = r \frac{d^2\theta}{dt^2} + 2 \frac{dr}{dt} \frac{d\theta}{dt} \quad (7)$$

which describes the curvilinear motion of a particle.

At a given rotating speed

$$\frac{d\theta}{dt} = \omega = \text{const.} \quad (8)$$

and

$$\frac{d^2\theta}{dt^2} = 0 \quad (9)$$

from which the pressure-head is given by

$$P_\theta = 2c\rho\alpha\omega^2 r^2 \quad (10)$$

If the rotor comprises n arms of thickness l_b , symmetrically disposed, the angle between two arms at a given radial distance r , is

$$\alpha = \frac{2\pi}{n} - 2 \arcsin \left(\frac{l_b}{2r} \right) \quad (11)$$

if

$$\frac{l_b}{2r} \ll 1 \quad (12)$$

then

$$\alpha = \frac{2\pi}{n} - \frac{l_b}{r} \quad (13)$$

and finally

$$P_\theta = 2c\rho\omega^2 r^2 \left(\frac{2\pi}{n} - \frac{l_b}{r} \right) \quad (14)$$

As P_θ is the pressure-head at a given radial distance between the front of one arm and the back of the next, P_θ will also be the pressure-head between the front and the back of the same arm at this radial position if the rotor is symmetrical.

We now consider the case where the rotor arms are porous, being comprised of catalyst particles confined in a packed configuration by a perforated basket as for example in a cruciform arrangement. Using the Ergun Eq. (7) for flow-through packed beds,

$$\frac{P_\theta}{l_b} = \frac{150}{D_p^3} \frac{(1-\epsilon)^2}{\epsilon^3} U + \frac{1.75\rho(1-\epsilon)}{D_p\epsilon^3} U^2 \quad (15)$$

From Eqs. (14) and (15),

$$A_1 U^2 + A_2 U - A_3 c \omega^2 r^2 = 0 \quad (16)$$

where

$$A_1 = \frac{1.75}{D_p} \rho \frac{1-\epsilon}{\epsilon^3} \quad (17)$$

$$A_2 = \frac{150\mu}{D_p^3} \frac{(1-\epsilon)^2}{\epsilon^3} \quad (18)$$

$$A_3 = 2\rho \left(\frac{2\pi}{n} - \frac{l_b}{r} \right) / l_b \quad (19)$$

In Eq. (16) only the constant c is unknown. This can be experimentally determined if the torque on the rotor is measured. If P_θ is the pressure-head across the arm at a given radial position,

$$T = \int_{r_1}^{r_2} P_\theta l_b r \, dr \quad (20)$$

assuming that P_θ does not change axially. The total torque on the basket is thus,

$$T_t = n \int_{r_1}^{r_2} P_\theta l_b r \, dr \quad (21)$$

where l_b is the axial width of the arm.

From Eqs. (14) and (22) and integrating, T_t becomes

$$T_t = 2n\rho\omega^2 l_b \left[\frac{\pi}{2n} (r_2^4 - r_1^4) - \frac{l_b}{3} (r_2^3 - r_1^3) \right] c \quad (22)$$

The experimental measurement of T_t will thus permit one to evaluate " c ".

(b) Cylindrical spinning basket

This can be represented as before but with an infinite number of rotor arms. Since the angle between two consecutive arms will now be $d\alpha$ the following equation equivalent to Eq. (10) obtains,

$$dP_\theta = 2c\rho\omega^2 r^2 d\alpha \quad (23)$$

The thickness of each blade will be

$$dl_b = r \, d\alpha \quad (24)$$

and from Eq. (24) we obtain

$$\frac{dP_\theta}{dl_b} = 2c\rho\omega^2 r \quad (25)$$

Equalising now this equation with the Ergun

equation in differential form,

$$A_1 U^2 + A_2 U - A_4 \omega^2 r c = 0 \quad (26)$$

where A_1 and A_2 are defined as before and

$$A_4 = 2\rho \quad (27)$$

The total torque on the basket will be

$$T_t = \int_0^{2\pi} \int_{r_1}^{r_2} dP_{1,r} r dr \quad (28)$$

replacing dP_θ in Eq. (28), and integrating, T_t will become

$$T_t = 4\pi\rho\omega^2 l [(r_2^4 - r_1^4)/4] c \quad (29)$$

from which the constant c can be determined for the cylindrical spinning basket.

EXPERIMENTAL DETERMINATION OF "c"

The constant c (Eqs 16 and 26) was determined by measuring the torque applied to the basket at various rotating speeds for reactors T1 and T4 using a conventional dynamometer arrangement.

Average values of c of 0.510 and 1.733 for the reactor T1 and T4 respectively were found.

When the bed of spheres in reactor T1 was replaced by a solid blade, the decrease in c was found to be less than 10 per cent.

J FACTOR FOR MASS TRANSFER

If the concentration of transferred component is small, the J_d factor for mass transfer may be defined [8] as,

$$J_d = \frac{K_g}{U} S_c^{1/3} \quad (30)$$

Several empirical relationships between J_d and various modified Reynolds numbers for packed beds have been found. Thus Hogen [9] most recently proposed

$$J_d = 0.84 Re^{*-0.51} \quad 0.01 < Re^* < 50 \quad (31)$$

and

$$J_d = 0.57 Re^{*-0.41} \quad 50 < Re^* < 1000 \quad (32)$$

where

$$Re^* = \frac{U\rho}{a\mu} \quad (33)$$

For a bed of uniform spheres with dia D and porosity ϵ

$$Re^* = \frac{U\rho D}{6(1-\epsilon)\mu} \quad (34)$$

In Fig. 6 the J_d coefficients in reactors T1 and T4,

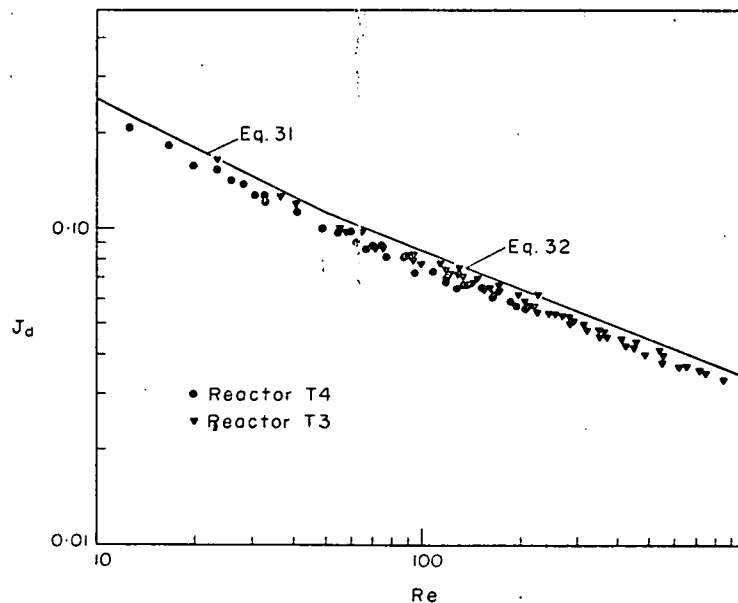


Fig. 6. J_d coefficient vs Re^* .

calculated from the experimental values of K_c and slip velocity from Eqs. (16, 26) are plotted against the above Reynolds number.

PREDICTION OF MASS TRANSFER COEFFICIENTS

Effect of pellet size and bed porosity

The predicted effect of the diameter of the spheres upon the mass transfer coefficients for reactor T1 is shown in Fig. 7, where the mass transfer coefficient is plotted against the rotating speed for different values of the sphere dia and two radial positions of the active pellet. It is assumed that the experimental constant c is independent of the size of the pellets and a cubic arrangement of the pellets in the arms obtains ($\epsilon = 0.476$) for all sizes of pellets. The experimental results obtained in reactor T1 are also shown.

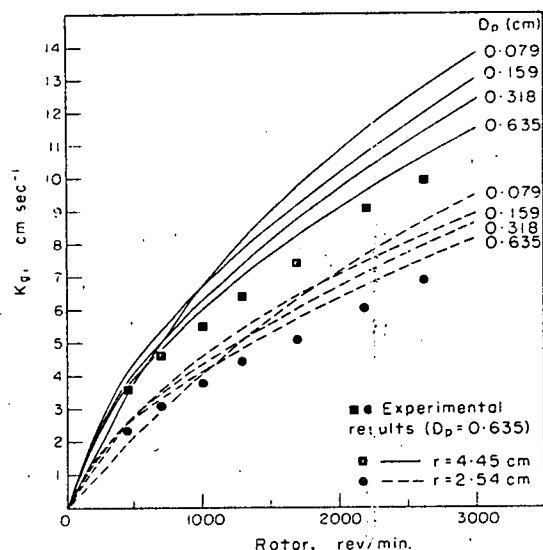


Fig. 7. Mass transfer coefficient predicted from the model. Effect of diameter of the pellets and rotating speed of reactor T1 ($Sc = 2.53$, $T = 25^\circ C$).

The effect of the porosity of the bed upon the mass transfer coefficients in reactor T1, is predicted from the model, and in Fig. 8 the mass transfer coefficient is plotted against the rotary speed in reactor T1 for beds packed with spheres of 0.318 cm dia in four different systematic arrangements.

From the above results it can be seen that the mass transfer coefficient generally increases, at a given radial distance with decrease in the diameter of the pellet at constant porosity. On the other hand the mass transfer coefficient decreases with decrease in the porosity, for beds with the same size pellets.

Brisk *et al.* [2, 5] have found in their spinning basket reactor that the mass transfer coefficient

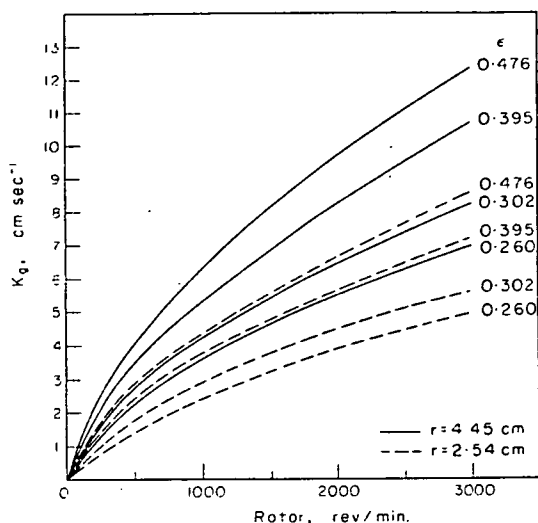


Fig. 8. Mass transfer coefficient predicted from the model. Effect of rotating speed and porosity of the bed in reactor T1 ($Sc = 2.53$, $T = 25^\circ C$).

decreases with pellet size. However no reference was made to the resulting changes in the porosity of the bed.

According to Leva [10] the porosity in cylindrical packed beds depends systematically on the ratio of the diameter of the pellet to the diameter of the reactor. Assuming that these results can be applied to our SCBR it was found that if the spinning basket is first randomly filled with spheres of 0.318 cm dia and porosity 0.45 and secondly with spheres of 0.079 cm dia, the porosity in the latter case falls to 0.36. In Fig. 9 the mass transfer coefficient in reactor

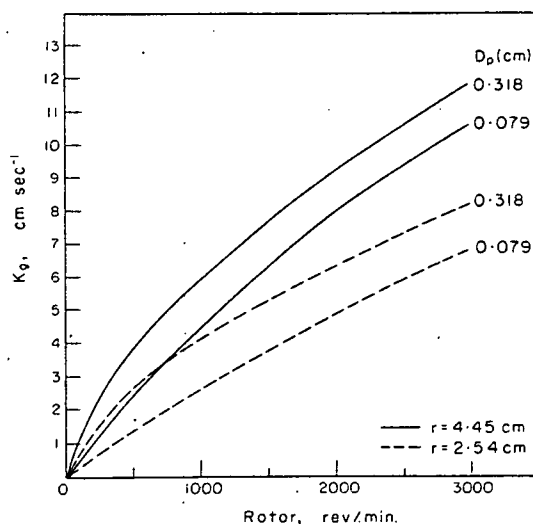


Fig. 9. Simultaneous effect of the size of the pellets and porosity of the bed on mass transfer coefficients in reactor T1 ($Sc = 2.53$, $T = 25^\circ C$).

T1 is plotted against rotary speed, for these two cases, at two different radial positions of the active pellet.

It is seen that the foregoing results of Brisk *et al.* are predicted as due to simultaneous changes in porosity and diameter rather than to changes in the pellet size per se. The other data given by Brisk *et al.* show the same trends as found here but the transfer coefficients were somewhat lower than we have reported.

DISCUSSION AND CONCLUSIONS

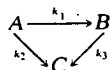
Two spinning catalyst basket configurations have been studied. The experimental results show that the mass transfer coefficient increases with the rotating speed as well as the radial position of the active pellet. In the cruciform spinning basket, it was found that the mass transfer coefficient increases when the gap width between the radial arm and reactor wall is decreased. This is due to the fact that for smaller gap widths the gas is forced to pass through the bed rather than to circulate in the space between the reactor wall and the edge of the arm. The experimental results show that the mass transfer coefficient decreases when the thickness of the arm in the cruciform basket increases. This is because, for a greater thickness of arm, the resistance to gas flow through the bed is higher.

In reactor T2 it was found that the mass transfer coefficient was higher for particles in the front and back layer than for particles in the middle of the basket. This result is not explicable in terms of our simple mathematical model. However secondary rotational flow of the gas trapped between the arms can be assumed by analogy with the flow of liquids in similar situations[8], and this counter-rotation will increase the mass transfer rate from the spheres in the front and back layers of the arm.

The experimental mass transfer coefficients and the J_d factors calculated therefrom agree reasonably with prediction from the model using empirical published relationships for mass transfer.

Application of the foregoing to a typical catalytic oxidation reaction

The catalytic oxidation of *o*-xylene with air over a commercial vanadia catalyst may be adequately represented by the following scheme of first-order reactions where 'A' is *o*-xylene, 'B' is phthalic anhydride and 'C' represents the products of combustion[11].



where

$$\frac{k_1}{k_1 + k_2} = 0.75$$

$$k_1 + k_2 = \begin{cases} \exp\left(18.6 - \frac{27,120}{RT}\right) & T \leq 713^\circ\text{K} \\ \exp\left(2.64 - \frac{4,476}{RT}\right) & T > 713^\circ\text{K} \end{cases}$$

$$k_1 \propto \exp\left(15.4 - \frac{30,500}{RT}\right) \text{ all } T.$$

in the units (gmol/ghr atm).

These data were obtained in a S.C.B.R. with a very dilute *o*-xylene feed.

Conlin[12] has numerically computed the temperature differences between catalyst pellets and jacket for a S.C.B.R. in which the above exothermic reactions are taking place under the following conditions which include high xylene feed concentrations. A cylindrical basket contained 210, 0.5 cm dia spherical catalyst particles weighing 32 g in a chamber of 60 cm³ volume. The packed bed had a voidage of 0.4 and a radius equal to that of 3 particles. Feed concentrations of (1–2) mol/cm³ [up to 1 per cent by volume] *o*-xylene in air with a flow rate of (12–18) cm³/sec were employed in the calculations representing upper limits of explosion and imperfect mixing respectively. Basket speeds were varied from (1–8), 000 rev/min.

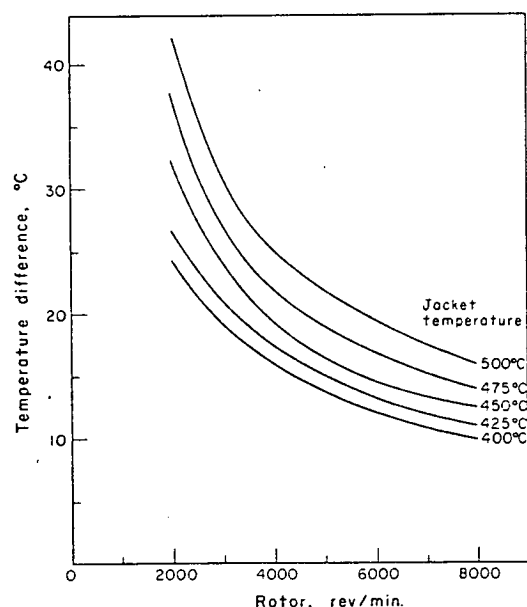


Fig. 10. Difference between pellet and jacket temperatures at various jacket temperatures and rotational speeds.

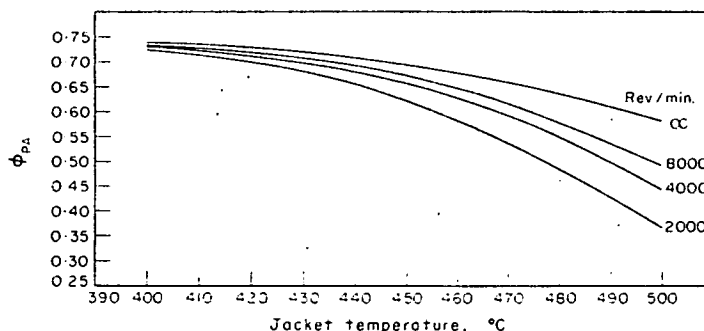


Fig. 11. Selectivities to phthalic anhydride with and without transport resistance compared at various jacket temperatures and rotational speeds.

Heat generation and removal rates at the catalyst particles were calculated at various jacket temperatures taking the foregoing kinetics which had been observed at low feed concentrations and high basket speeds. The physical and chemical rate resistances were summed up in the usual way taking $J_h = 1.07 J_d$ where J_d was calculated from Eqs (30) and (26). The intercepts of the heat removal and generation lines were determined graphically to give the steady-state particle temperatures. These varied by <20 per cent over the basket radius and were averaged. Figure 10 shows a typical set of results at the high flow rate and in Fig. 11 the consequential selectivities to phthalic anhydride are compared with those which are obtained in the absence of any physical transport resistance ($R = \infty$). Selectivities for the above reaction scheme are given by

$$\phi_{P.A.} = \frac{\frac{k_1}{k_1 + k_2}}{1 + \frac{k_3}{k_1 + k_2} \left(\frac{X}{1 - X} \right)} \quad (35)$$

The non-isothermality of the S.C.B.R. in this example is very apparent and could lead to a lack of appreciation of the proper selectivities achievable with the catalyst. It is apparent that the feed concentration of *o*-xylene would have to be largely reduced to obtain reliable kinetic data at the basket speeds conveniently obtainable. A reduction in flow rate can also be advantageous in that the conversion is increased and the rate of reaction consequently reduced.

Acknowledgement—One of us, J.R.P., wishes to thank Instituto de Alta Cultura, Lisbon, for financial assistance which made this research possible.

NOTATION

A_1, A_2, A_3, A_4 constants defined by Eqs (15–17) and (24)

- a surface area for mass transfer per unit volume of bed, cm^2/cm^3
- C concentration of naphthalene in air, mol cm^{-3}
- C_0 saturation concentration of naphthalene in air, mol cm^{-3}
- c dimensionless constant [Eq. (1)]
- D_p dia of the spheres, cm
- F flow rate, $\text{cm}^3 \text{sec}^{-1}$
- F_θ tangential force, dyne
- J_d dimensionless J factor for mass transfer
- J_h dimensionless J factor for heat transfer
- K_g mass transfer coefficient, cm sec^{-1}
- $k_{1,2,3}$ first order reaction velocity constants, gmol/ghr atm
- L maximum length of the arm, cm
- l_a length of the arm, cm
- l_b thickness of the arm, cm
- l_z axial width of the arm, cm
- n number of arms
- P_θ tangential pressure-head, dyne cm^{-2}
- Re^* modified Reynolds number
- R revolutions per minute
- r radial distance of the active pellet, cm
- r_1, r_2 inner and outer radius of the arm, cm
- S surface area for mass transfer, cm^2
- Sc schmidt number
- T torque on each arm of the rotating basket, dyne-cm
- T_t total torque on the rotating basket, dyne-cm
- t time, sec
- U superficial gas velocity, cm sec^{-1}
- X fractional conversion
- α angle between two consecutive arms

ϵ porosity of the bed
 ρ density, g cm^{-3}
 θ cylindrical co-ordinate
 μ viscosity, $\text{g cm}^{-1} \text{sec}^{-1}$
 ω angular velocity
 ϕ_{PA} selectivity to phthalic anhydride

$$= \left(\frac{\text{P.A. produced}}{\text{Xylene consumed}} \right)$$

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D.2 THE PREDICTION OF DISTILLATION PLATE EFFICIENCIES
FROM FROTH PROPERTIES

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THE PREDICTION OF DISTILLATION PLATE EFFICIENCIES FROM FROTH PROPERTIES

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SUMMARY:

A compound optical probe is described which is used to determine the distribution of bubble sizes and velocities when liquids are distilled on sieve trays in the froth regime. It is shown how the plate efficiencies may be thereby evaluated by thus taking into proper account the derived distribution of vapour residence times and mass transfer coefficients. The results are seen to agree with published data. Froth properties for several boiling liquids on two sieve plates are reported.

INTRODUCTION:

Geddes (1) first expounded a mass-transfer model of the distillation plate, estimating bubble sizes and velocities. Attempts have since been made to measure bubble properties in froths to refine the model (2) and (3) but the techniques used have been laborious and have sometimes truncated the size distributions leading to neglect of the relatively small number of large bubbles which however convey much of the vapour because of their high velocity. Burgess and Calderbank (4) developed a compound conductivity probe which greatly improved bubble measurement in water. An optical sensor is now described which permits bubble properties to be similarly measured in boiling organic liquids on distillation trays. The use of fibre optics has the added practical advantage of being flame-proof.

THE OPTICAL PROBE:

Optical probes (dip-sticks) have been previously described (5); they have been used to sense changes in refractive index at their extremity. In this paper we describe a very simple and miniature version of the optical 'dip-stick' which we have incorporated into a system of compound-probe and logic circuitry to measure bubble sizes and velocities when organic liquids are distilled on a sieve-tray. A set of five optical probes sense the passage of bubbles and select for measurement only those which strike the array /

array coaxially to give unambiguous measurement of the maximum bubble height and bubble velocity by means of an external discrimination circuit and accurate time base. The system of measurement has been previously described when electrical conductivity probes were used (4) but these suffered from the disadvantages that conducting liquids only could be studied and that the finite rate of drainage of liquid from the probe-tip required the use of an on-line computer to measure pulse durations accurately. The present system uses hard-wired logic circuitry, and is here used to demonstrate its application to boiling froth on 6 in. diameter sieve plates.

An optical 'dip-stick' comprises a light guide which conveys light to its extremity whereupon the light is either internally reflected back up the guide and sensed, if the specially shaped terminal is in a medium of low refractive index (gas) or is lost into the medium if it is of high refractive index (liquid). We have found that these functions can be easily achieved by taking a rigid light guide of glass (typically 3 mm. in diameter) and drawing it out sharply to produce a much smaller rod (typically 0.2 mm. in diameter) which is then cut at any desired length to produce a tip which is then ground to a point. Light is then passed down the large diameter section through one leg of a Y-form fibre optics guide and provided the tip of the probe is always wet, but otherwise in gas, light is strongly reflected back and may be registered by a photo-detector at the end of the other leg of the Y-form guide. In liquids, no light is reflected back from the probe tip. (Rigid light guides are available commercially and comprise a glass rod of high R.I. having an outer skin of low R.I. Glass). We have found that the small diameter section of the probe may be very small indeed and in our work has been restricted only by convenience in handling. When the tip is suddenly removed from a liquid, it strongly reflects light for several minutes until the lenticular film of liquid evaporates so that in a bubbling liquid, the tip is always reflective on exposure to gas or vapour. Further, one obtains a step change in the reflected light on changing from gas to liquid or vice-versa so that the pulse duration on passage of a bubble is accurately defined.

The arrangement of light source probe and detector is shown in Fig. K¹ for a single channel. The detectors used were transistor photo diodes in which detector and amplifier are contained on a single chip. A standard comparator circuit was used to give an output of 4V on reflection and 0V in the dark. A LED gave visual indication of pulses and is very useful for trimming the comparator circuit.

Fig. /

Fig. 2 shows the direct analogue photocell output from a typical encounter between the probe and a bubble, first stored in computer memory and then displayed. The signal rise-time is of the order of 0.5 m.sec. in a signal duration of about 100 times this period. For comparison, the response of an electrical conductivity probe in water is shown.

THE COMPOUND PROBE:

This has been previously described in detail when conductivity transducers were used (4) and only the principal features are repeated here.

The compound probe is an array of five sensors in which a leading probe is symmetrically surmounted by three identical probes as shown in Fig. 3. A bubble is known to be coaxial with the leading probe when it is later coincidentally sensed by the three upper probes. The time delay between 'making' the leading and upper contacts and 'breaking' the lower contact then give the maximum bubble height and bubble velocity. The fifth probe (Fig. 3) is used to measure the position and height of a vertical chord at a known distance from the centre line, from which the bubble shape can be deduced, if this is required. In order to avoid noise from multiple bubble encounters with the array, the further following discriminatory functions are exercised by the data processing logic.

1. All channels are initially in the same ('break') mode.
2. The pulse sequence $[1, (2,3,4)]$ Make, $[1, 2,3,4)]$ Break, is obligatory so that the measurement field is initially clear of bubbles, the sampled bubble hits the array 'squarely' and leaves the measurement field in the same fashion, without interference from other bubbles. This constraint limits the smallest size of bubble which can be measured to the vertical probe separation distance.
3. Small bubbles may be assumed to be spherical but those which trigger the fifth, shape measuring, probe are required to do so in the make and break sequences. 1, (2,3,4), 5 to ensure that only one bubble is involved.
4. All channels return to the same ('break') mode.

At the end of the above sequences, the delay times $1_M - (2,3,4)_M$; $1_M - 1_B$;

$1_M - 5_M$; $5_M - 5_B$, where the suffixes M and B refer to 'make' and 'break' respectively, are stored, whereafter sampling and measurement is continued. The technique is one of sample, hold and store repeated until a statistically significant sample is obtained. The time to obtain, say 2000 samples, depends on the coincidence jitter (2,3,4) allowed; this was adjustable (2-4 m sec) and was such that a typical experimental run with a sieve-tray froth at one vapour flow rate, took about four hours.

DATA /

DATA PROCESSING:

Pulses from the five channels were fed to a transistor logic circuit* which performed the foregoing discriminatory functions using a series of 'gates'. The state of the five channels (make or break) was continuously monitored and the delay times measured and sent to a high speed paper punch only when the correct sequence and coincidence were obtained, the latter having an adjustable 'jitter', set to secure a convenient sampling rate.

The punched tape was interfaced off-line with the Edinburgh Regional Centre computer using a programme to calculate froth and mass transfer properties, as described below. In developing the system, the analogue signals from the five phototransistors were simultaneously sent to a PDP/8 computer, as previously used with the compound conductivity probe (4) and programmed with the same software. The results using the hard-wired logic and software were nearly identical. Small differences were attributable to the more accurate timing available with the hard-wired logic circuit which has the additional advantage of a shorter off-stream data output time which allows a smaller coincidence jitter to be used.

DISTILLATION COLUMN:

A single 6 in. diameter sieve tray was fed with vapour from a steam-heated reboiler and liquid was returned from a total condenser, the whole being built of industrial glass components with copper heating coils. Condensate could be run to a flask for measurement of the boil-up rate.

One sieve plate made of $\frac{3}{16}$ in. 'Dural' sheet, comprised 31 $\frac{1}{4}$ in. holes and had a free area of 5.38% being of a configuration widely used in the chemical industry and previously employed in this laboratory to study an air-water froth. A second sieve plate comprised 43 $\frac{3}{32}$ in. holes giving a free area of 1%. The froth level was controlled at 10 cms. by means of a weir in the external liquid overflow line to the reboiler and the vapour rate was variable over the range 30 - 65 cm/sec. by alteration to the steam pressure on the reboiler. The compound probe was normally positioned with its tip at the mid-point of the froth but its height could be adjusted using a series of spacers where the head was flanged onto the column. The liquids used were water, methanol, acetone, benzene and toluene.

COMPUTATION /

* The hard-wired logic circuit was devised and built by Dr. R. Mackie of the Wolfson Microelectronics Unit, University of Edinburgh.

COMPUTATION OF THE PHYSICAL AND MASS-TRANSFER PROPERTIES OF FROTH:

It was decided to treat the bubbles formed on distillation trays as spherical since it had previously been shown (4) that the shape evaluation with the fifth channel was not of great significance to the more important froth parameters and also because it was proposed later on to use an on-line microprocessor where memory had to be optimised. However, a shape-correction procedure was embodied so that the total computed bubble flow rate was made equal to the known vapour flow rate. The procedure was as follows:

Measure the total bubble encounter frequency (γ) on the leading probe and store.

Measure and store the bubble heights ($D = \frac{t_2}{t_1} \ell$) in 100 histogram slots, ascending in size from $D = \ell$.

Let the n^{th} slot contain J_n bubbles of mean height D_n .

Compute and store in each slot the mean bubble velocity, ($\bar{V}_n = \frac{\sum_{j=1}^{J_n} V_n}{J_n}$)

where $V_n = \frac{\ell}{t_1}$

For each slot compute and store the fraction of bubbles of size D_n and velocity $\bar{V}_n$ contained in the dispersion.

$$f_n = \frac{\frac{J_n}{D_n^2 \bar{V}_n}}{\sum_{n=1}^{100} \frac{J_n}{D_n^2 \bar{V}_n}} \dots\dots\dots (1)$$

This relationship allows for the varying probability of a central hit between bubbles of different sizes and the compound probe, and also for the effect of varying bubble velocity on the frequency of such a collision, as explained in detail previously (4).

The fraction of the total flow passing through the n^{th} slot is also computed as:

$$f_f = \frac{f_n D_n^3 \bar{V}_n}{\sum_{n=1}^{100} f_n D_n^3 \bar{V}_n} = \frac{J_n D_n}{\sum_{n=1}^{100} J_n D_n} \dots\dots\dots (2)$$

A shape factor is then derived as,

$$S = \frac{6V_s \sum_{n=1}^{100} f_n D_n^2 \bar{V}_n}{\gamma \sum_{n=1}^{100} f_n D_n^3 \bar{V}_n} = \frac{6V_s \sum_{n=1}^{100} J_n}{\gamma \sum_{n=1}^{100} J_n D_n} \dots\dots\dots (3)$$

which allows one to check on any departure from the assumption of bubble sphericity in equating the net vapour flow as spherical bubbles to the known superficial vapour velocity.

The /

The Sauter-mean bubble diameter for spheres can be calculated as,

$$\bar{D}_s = \frac{\sum_{1}^{100} f_n D_n^3}{\sum_{1}^{100} f_n D_n^2} = \frac{\sum_{1}^{100} \frac{J_n D_n}{\bar{V}_n}}{\sum_{1}^{100} \frac{J_n}{\bar{V}_n}} \dots\dots\dots (4)$$

and the volume fraction vapour hold-up as,

$$H_g = \frac{V_s}{V_t} = \frac{S_y}{6} \frac{\sum_{1}^{100} f_n D_n^3}{\sum_{1}^{100} f_n D_n^2 \bar{V}_n} = \frac{V_s \sum_{1}^{100} \frac{J_n D_n}{\bar{V}_n}}{\sum_{1}^{100} J_n D_n} \dots\dots\dots (5)$$

assuming a vertical flow of bubbles.

Again the vapour-liquid interfacial area per unit volume of dispersion may be derived as,

$$a = \frac{6H_g}{\bar{D}_s} = \frac{6V_s \sum_{1}^{100} \frac{J_n}{\bar{V}_n}}{\sum_{1}^{100} J_n D_n} \dots\dots\dots (6)$$

The above mean properties, by themselves, are of limited interest since they do not lead to the mean mass-transfer efficiency which is computed as follows,

We assume a vertical flow of bubbles through a well-mixed liquid and for each slot in the bubble-size distribution, write a mass-balance over a differential height 'dh' of the dispersion

Thus, for the n^{th} slot,

$$G_n dy = (K_L A)_n (x_o - x^*) \frac{\rho_L}{(M.W)} \frac{dh}{h} = (K_g A)_n (y^* - y) \frac{\rho_g}{(M.W)} \cdot \frac{dh}{h}$$

$$G_n dy = (K_{og} A)_n (mx_o - y) \frac{\rho_g}{(M.W)} \cdot \frac{dh}{h} \dots\dots\dots (7)$$

$$\text{where } K_{og} = \frac{K_g}{1 + m \frac{K_g \rho_g}{K_L \rho_L}} \quad \text{and} \quad \frac{A_n \rho_g}{(M.W) h G_n} = \left(\frac{a}{V_s n} \right)$$

If we now assume that the mass-transfer coefficients in both vapour and liquid phases are given by the surface renewal theory following from the potential flow behaviour of large bubbles,

$$K_{og} = \frac{2 \sqrt{\frac{D_g}{\pi t}}}{1 + m \frac{\rho_g}{\rho_L} \sqrt{\frac{D_g}{D_L}}} \dots\dots\dots (8)$$

$$\text{where } t = \frac{D_n}{V_n}$$

Integrating /

Integrating equation 7 over the height of dispersion (h), the fractional approach to equilibrium of the vapour with the liquid is given by,

$$(E_{og})_n = 1 - e^{-\left(\frac{K_{og} a}{V_s}\right)_n h}$$

so that for all the slots in the distribution,

$$E_{og} = \sum_{n=1}^{100} f_{f,n} \left(1 - e^{-\left(\frac{K_{og} a}{V_s}\right)_n h} \right) \dots \dots \dots (9)$$

and since, $a = \frac{6H_g}{D}$; $H_g = \frac{V_s}{V}$ and $K_g = 2\sqrt{\frac{D_g \bar{V}_n}{\pi D_n}}$,

it follows from equation (8) that,

$$\left(\frac{K_{o,g} a}{V_s}\right)_n h = \frac{12 \sqrt{\frac{D_g}{\pi}} h}{D_n^{3/2} \bar{V}_n^{1/2} \left(1 + m \frac{\rho_g}{\rho_L} \sqrt{\frac{D_g}{D_L}}\right)} \dots \dots \dots (10)$$

From equations (2), (9) and (10),

$$E_{og} = \frac{\sum_{n=1}^{100} J_n D_n \left(1 - e^{-\frac{K}{(D_n \bar{V}_n)^{1/2} D_n}} \right)}{\sum_{n=1}^{100} J_n D_n} \dots \dots \dots (11)$$

where the system constant K is given by,

$$K = \frac{12 \sqrt{\frac{D_g}{\pi}} h}{1 + \frac{m \rho_g}{\rho_L} \sqrt{\frac{D_g}{D_L}}} \dots \dots \dots (12)$$

Equation (11) gives the point plate efficiency by making simultaneous allowance for the distribution functions of vapour residence time, vapour-liquid interfacial area and mass transfer coefficients resulting from the variation of bubble sizes and velocities.

TEST /

TEST OF EQUATION (11):

Equation (11) was first applied to oxygen and ammonia dissolution in water, representing liquid and gas-phase mass-transfer control respectively. The appropriate system properties (D_m and m) were used and combined with those calculated from the bubble sizes and velocities recorded by the compound probe positioned at the mid-height of a froth formed by passing air through the 5% free-area sieve plate into cold water. Figures 4 and 5 show that the agreement with published efficiencies is satisfactory.

It may also be seen that the predictions agree with those of Burgess and Calderbank (4) who used a conductivity probe in 20 cm of froth on a similar sieve plate, despite the fact that the gas hold-up in the present work was considerably higher, due, no doubt, to the absence of a positive liquid flow over the plate as existed in the former case. It would appear that a higher gas hold-up is offset by a larger bubble size in its effect on the plate efficiency in this instance.

Comparison was then made with some distillation systems, the compound probe being operated in the boiling liquids as described earlier. It will be seen from figures 6 and 7 that agreement is again quite reasonable in view of the differences in distillation apparatus and uncertainty about the froth height in published work.

FROTH PROPERTIES IN BOILING LIQUIDS:

The variation of bubble size and related properties in boiling froths follows the pattern previously reported for the air-water system (4). Some results for boiling benzene are shown in Figures 8 and 9. It may be seen that bubbles undergo dispersion as they rise a short distance from the tray and then begin to coalesce as the free surface is approached. The vapour velocity has very little effect on the predicted plate efficiency but the fluid physical properties exert a substantial influence. This is shown in figure 10 where the system constant 'K' from equation (11) is plotted for the various boiling liquids such that an efficiency $E_{o,g}$ of 50% is always obtained. K is related to the height of a transfer unit with froth properties excluded and was calculated by trial and error from Equation (11) for each boiling liquid using its measured froth properties. Although the results are few, it appears, as might be expected, that a low surface tension is favourable to dispersion and rapid mass-transfer, as signified by a low value of $K_{50\%}$. It must not be forgotten that froth properties can be largely affected when two liquids undergo mass transfer (6) and we have so far examined these properties for pure liquids only. The /

The fact that we have been able to predict therefrom some binary distillation plate efficiencies, which vary principally with composition due to changes in 'm', is due to a relatively small change in froth properties on mixing in the examples quoted but this may not be the invariable rule.

The sieve plate with the 1% free-area gave a higher mass-transfer efficiency than that with 5% due to the generation of a smaller proportion of 'big' bubbles in the former case, as is shown in Fig. 11.

A microprocessor is being adapted for use with the system of measurement described prior to a further study of the effect of fluid properties and tray characteristics on plate efficiencies.

NOMENCLATURE:

A	total vapour-liquid interfacial area
a	vapour-liquid interfacial area per unit volume of dispersion
D	bubble height (diameter)
$\bar{D}_s$	Sauter-mean bubble diameter
D_g	diffusivity in vapour phase
D_L	diffusivity in liquid phase
E_{og}	point plate efficiency in terms of vapour compositions
f_n	fraction of the number of bubbles in n^{th} slot of the distribution
f_f	fraction of the vapour flow in n^{th} slot of the distribution
G	molar flow rate of vapour
H_g	volume fraction gas in dispersion
h	froth height
J_n	number of bubbles in the n^{th} slot of the bubble-size distribution
$K_{L,g}$	mass-transfer coefficient
K	system constant (defined in equation (12))
K_{og}	overall mass-transfer coefficient in terms of vapour compositions
ℓ	distance between leading and upper probes
m	slope of the equilibrium line ($y^* = mx^*$)
(M.W)	molecular weight
S	shape factor (defined in equation (3))
t	mass-transfer exposure time of fluid elements at a bubble surface
t_1	delay time between 'making' channels of leading and upper probes
t_2	delay time between 'making and breaking' leading probe channel
V	bubble velocity
$\bar{V}_n$	mean bubble velocity in n^{th} slot
V_s	superficial vapour velocity
x	mol fraction of key component in liquid
y	mol fraction of key component in vapour
ρ	density
γ	bubble frequency at a point.

SUFFIXES:

n	the n^{th} slot in the bubble size distribution
o	outlet
*	equilibrium
L	liquid
g	vapour or gas
o,g	overall, vapour.

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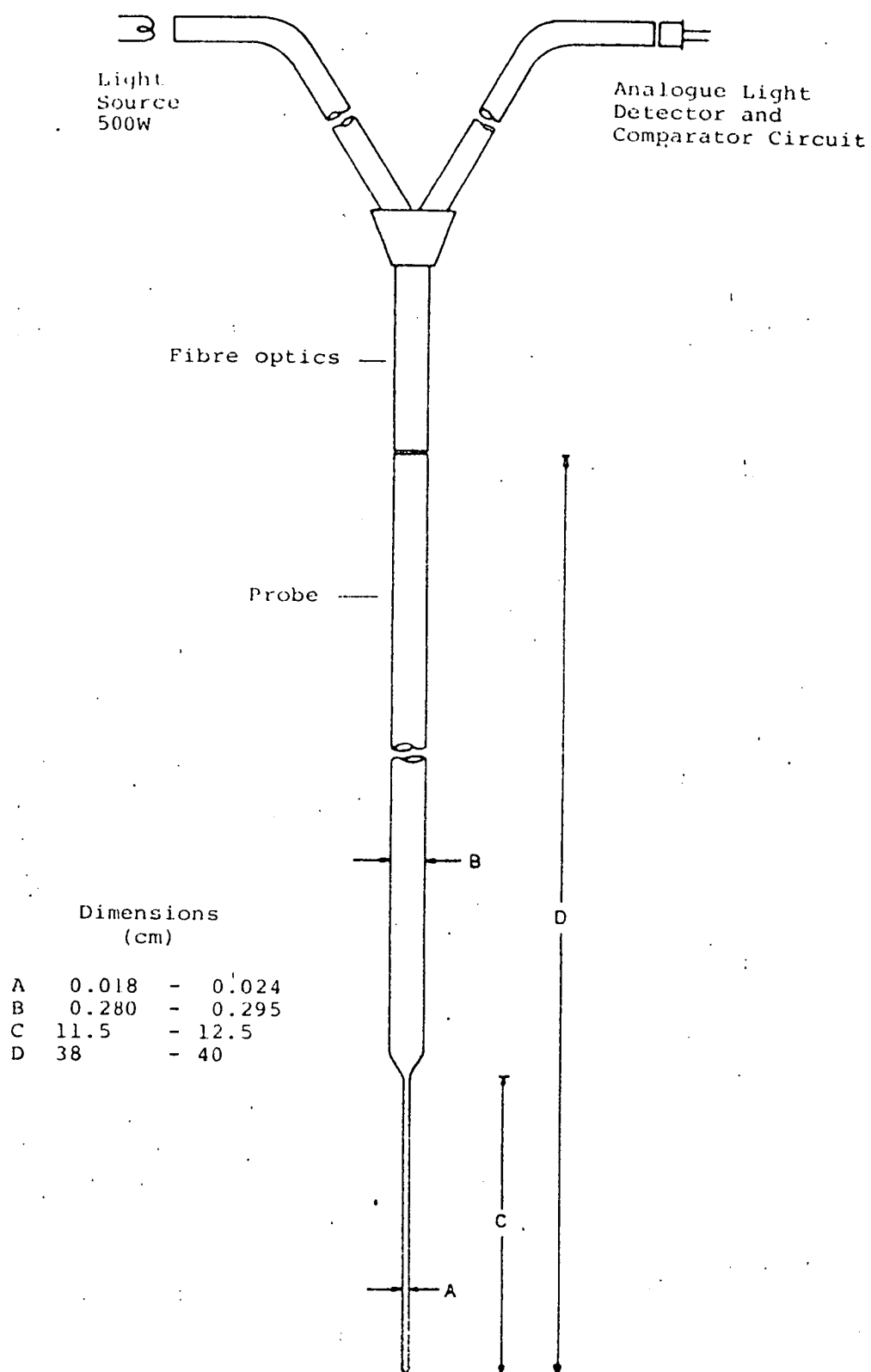
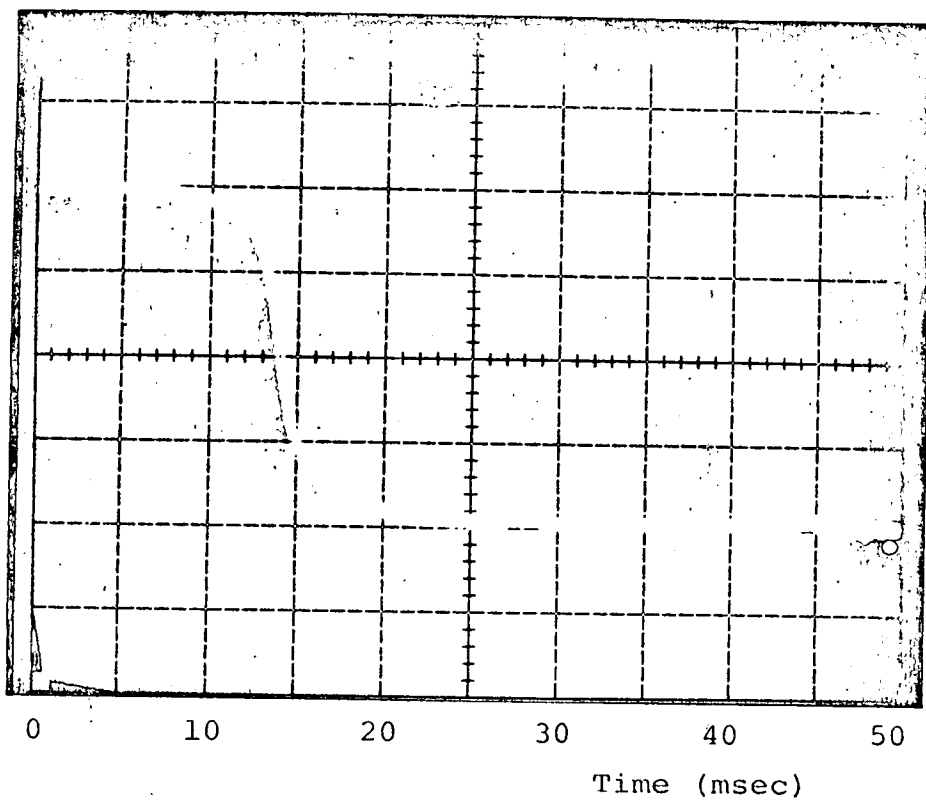
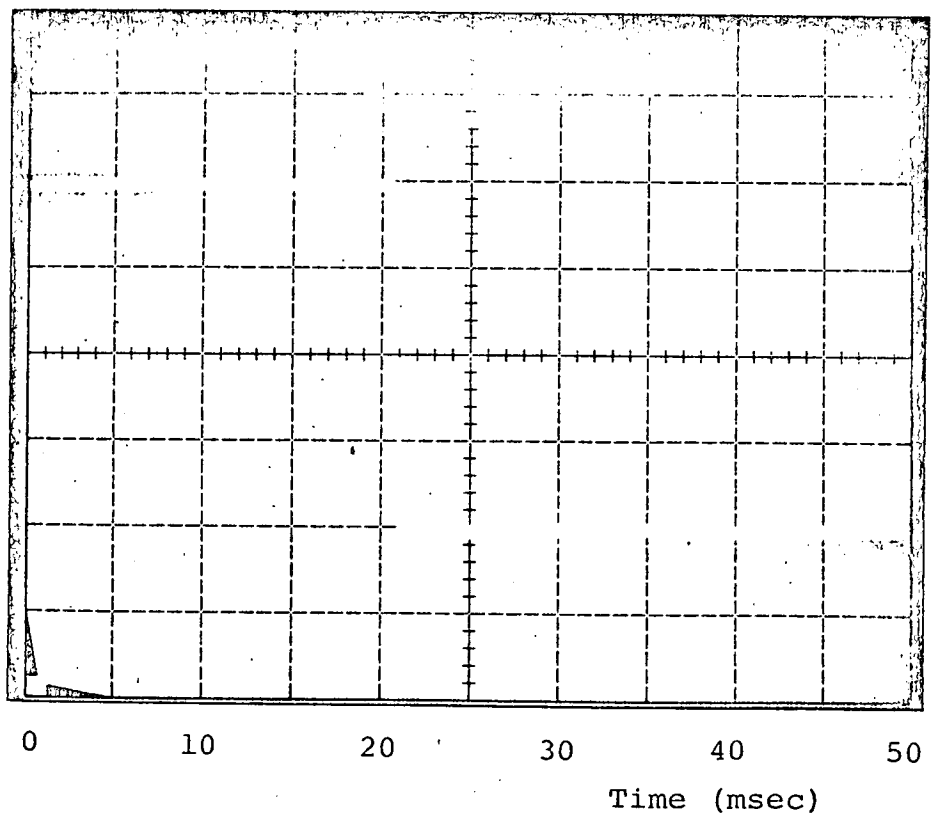


Fig. 1. A single probe.



(a)



(b)

Fig. 2. Typical analogue response to bubble encounter.

(a) Conductivity probe

(b) Optical probe.

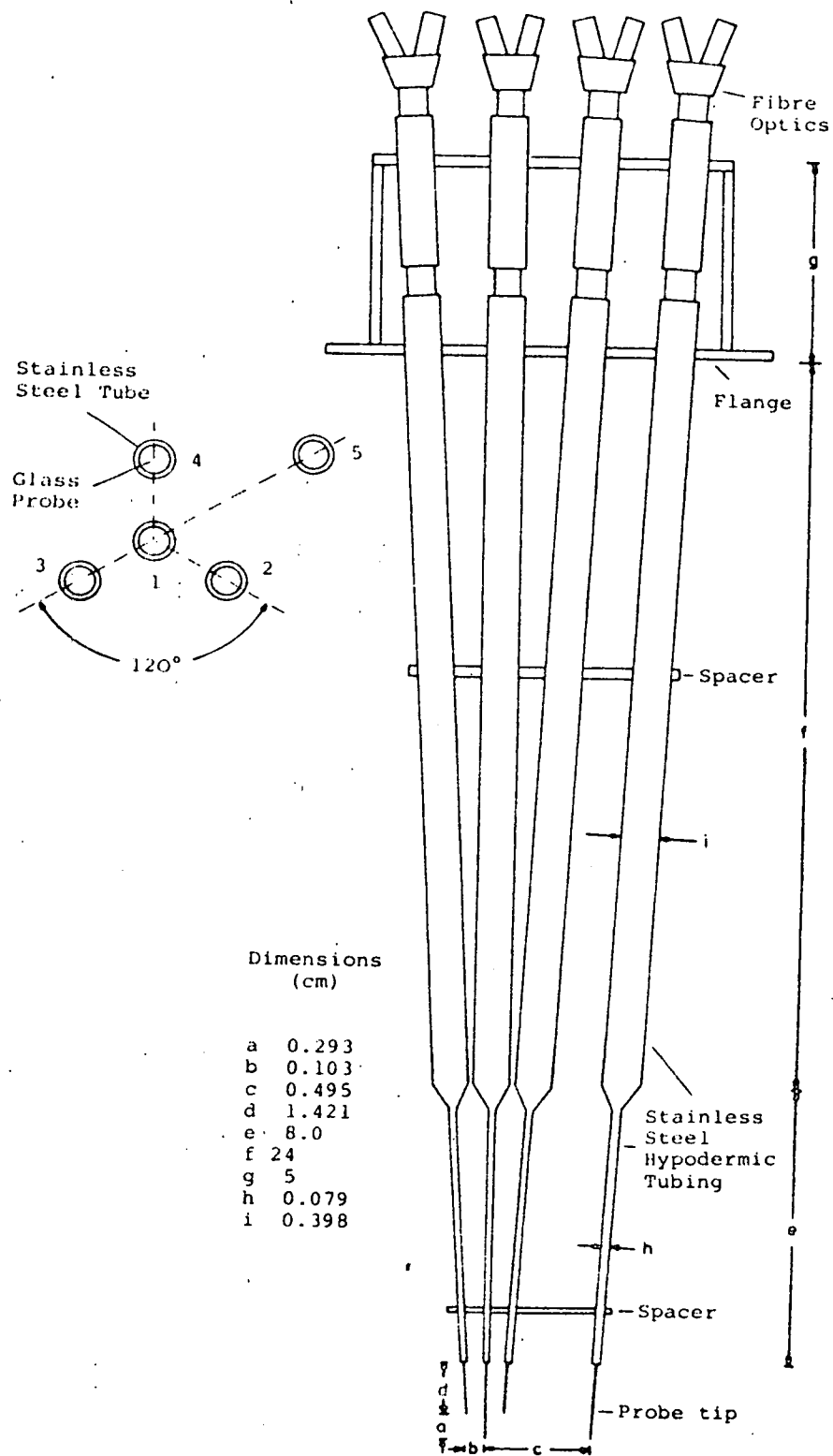


Fig. 3 The compound probe.

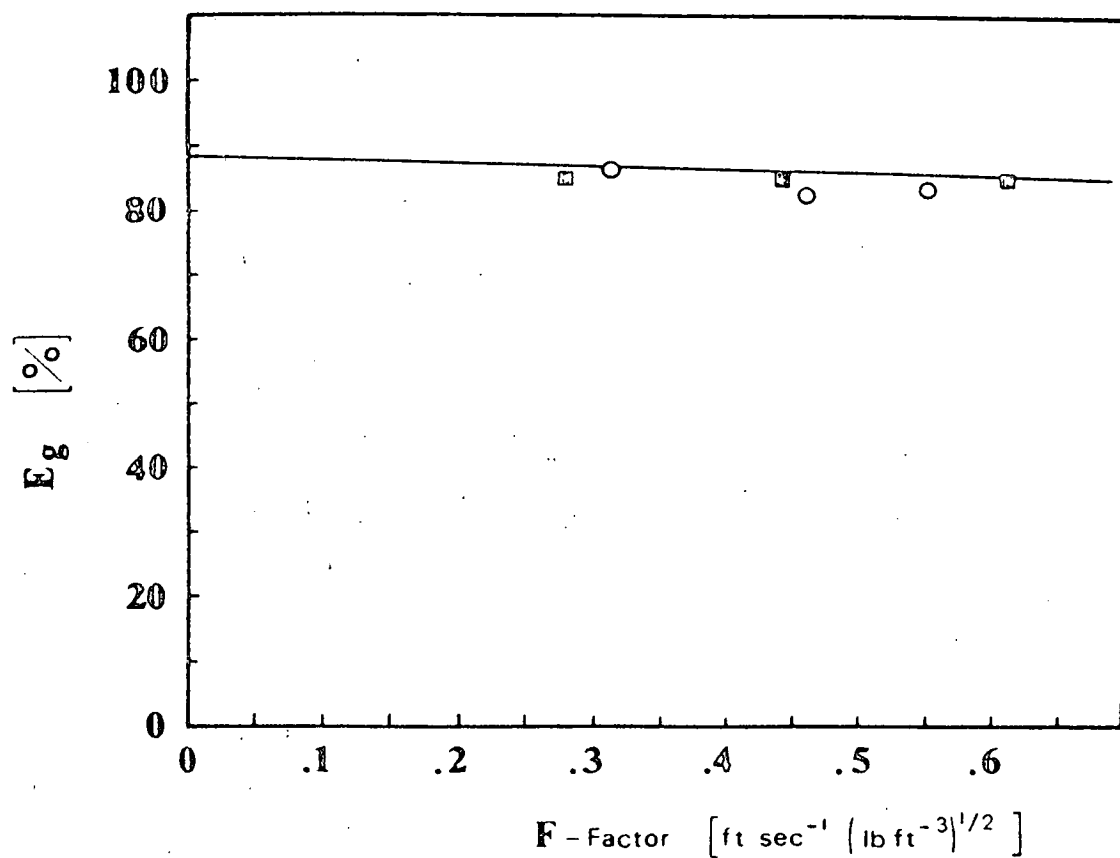


Fig. 5 Vapour phase plate efficiencies
 (Ammonia absorption at 20°C, $D_g = 0.22 \text{ cm}^2 \text{ sec}^{-1}$)
 ○ This work - 5.38% Free area
 ■ Burgess and Calderbank (4) - Tray with 4.8% Free area
 — A.I.Ch.E. design correlation (8)

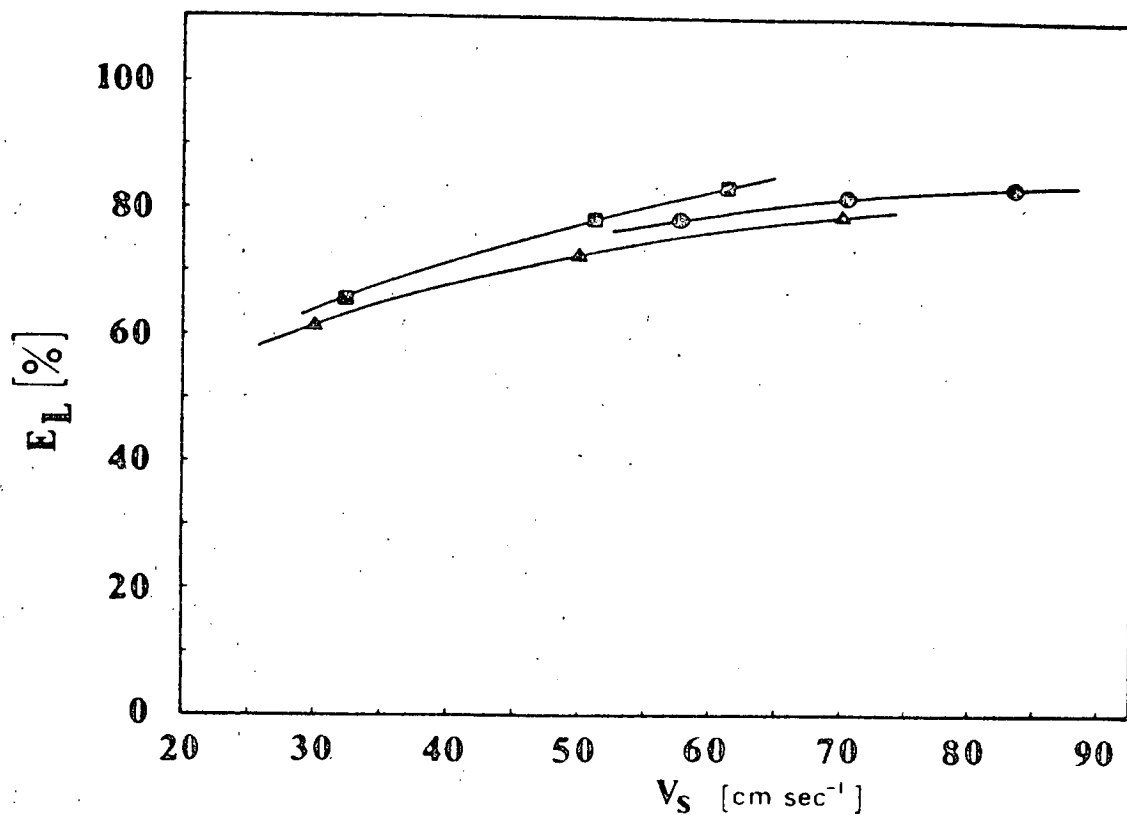


Fig. 4. Liquid phase sieve plate efficiencies
 (Oxygen - dissolution at 20°C, $D_L = 2.36 \cdot 10^{-5} \text{ cm}^2 \text{ sec}^{-1}$)

- This work - 5.38% free area
- ▲ Burgess and Calderbank (4) - 4.8% Free area
- Foss and Gerster (7) - Experimental data

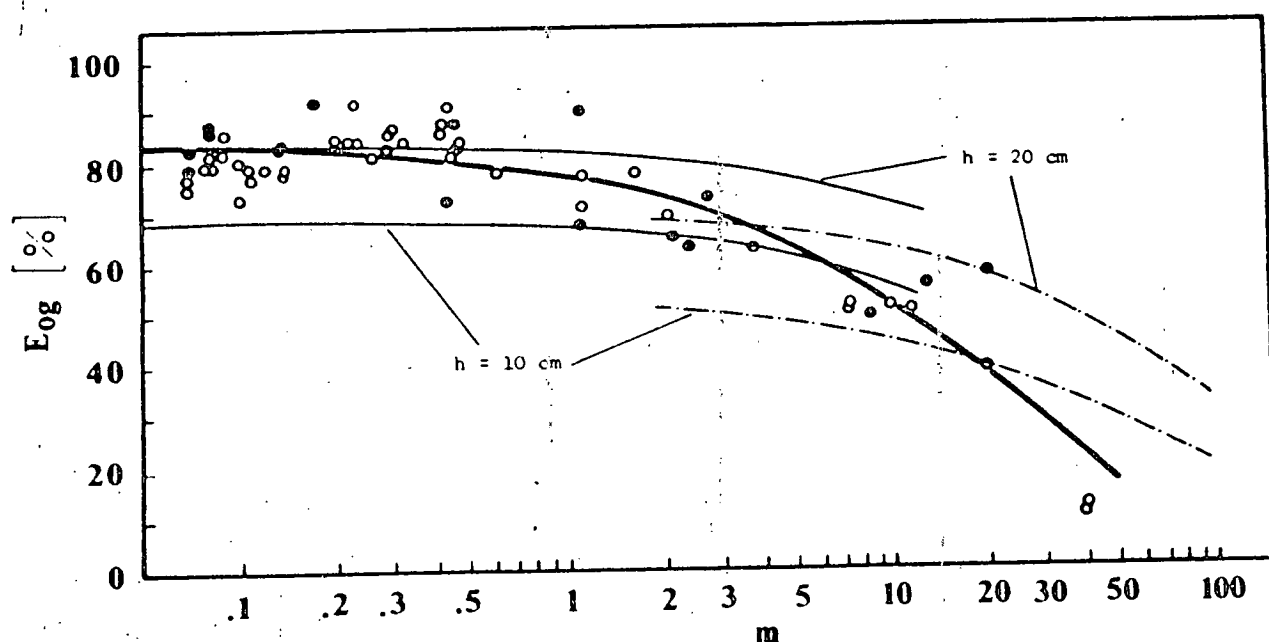


Fig. 7 Plate efficiencies for the acetone-water system

This work - 5.38% Free area

--- Water froth properties ($T = 100^{\circ}\text{C}$, $D_L = 5.02 \cdot 10^{-5} \text{ cm}^2 \text{ sec}^{-1}$,
 $D_g = 0.14 \text{ cm}^2 \text{ sec}^{-1}$)

— Acetone froth properties ($T = 58^{\circ}\text{C}$, $D_L = 1.31 \cdot 10^{-4} \text{ cm}^2 \text{ sec}^{-1}$,
 $D_g = 0.11 \text{ cm}^2 \text{ sec}^{-1}$)

● Bakowski, S. (11) - Experimental data

○ Schoenborn, E.M. et al (12) - Experimental data

— Hubner, W. et al. (10)

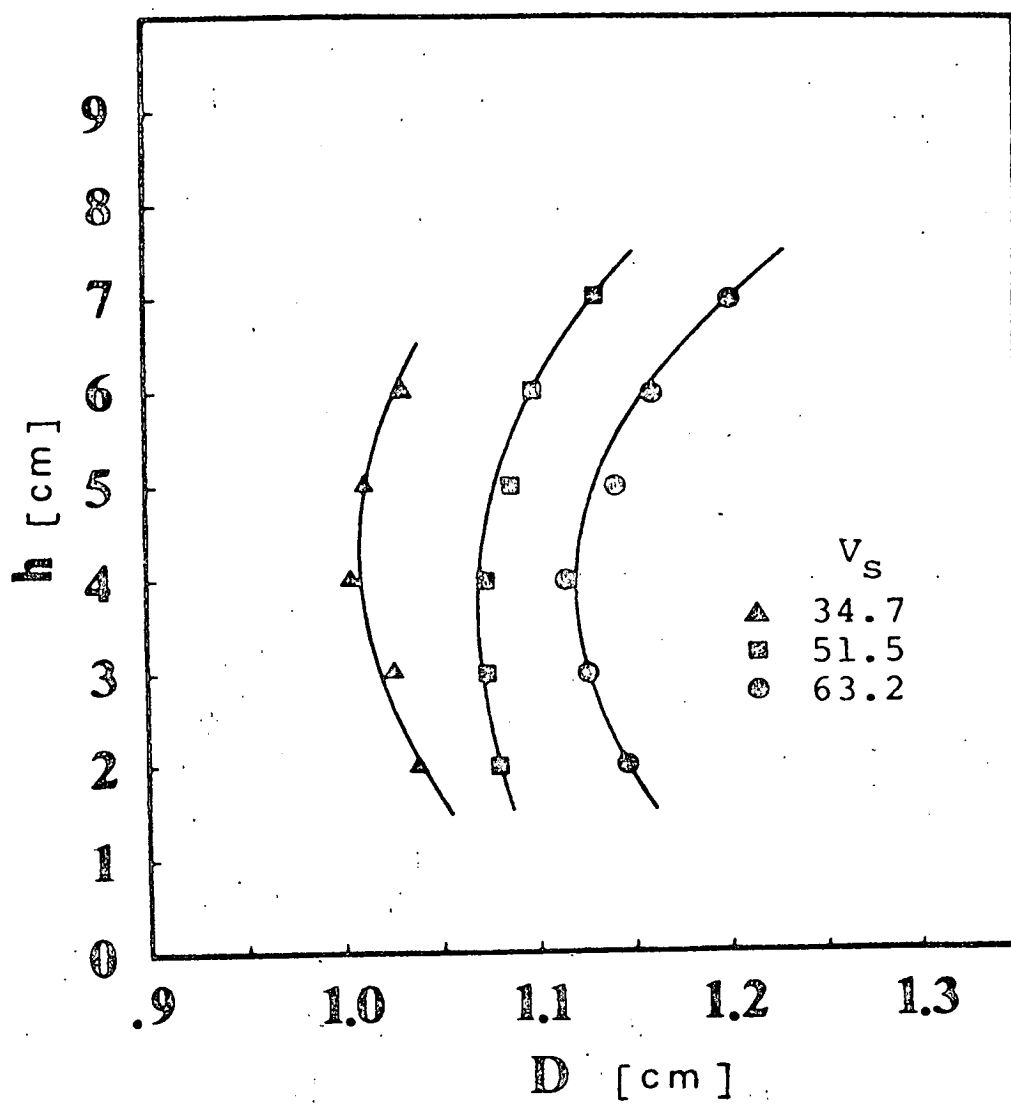


Fig. 8. Variation of bubble size with height in froth
(Boiling benzene - 5.38% free area)

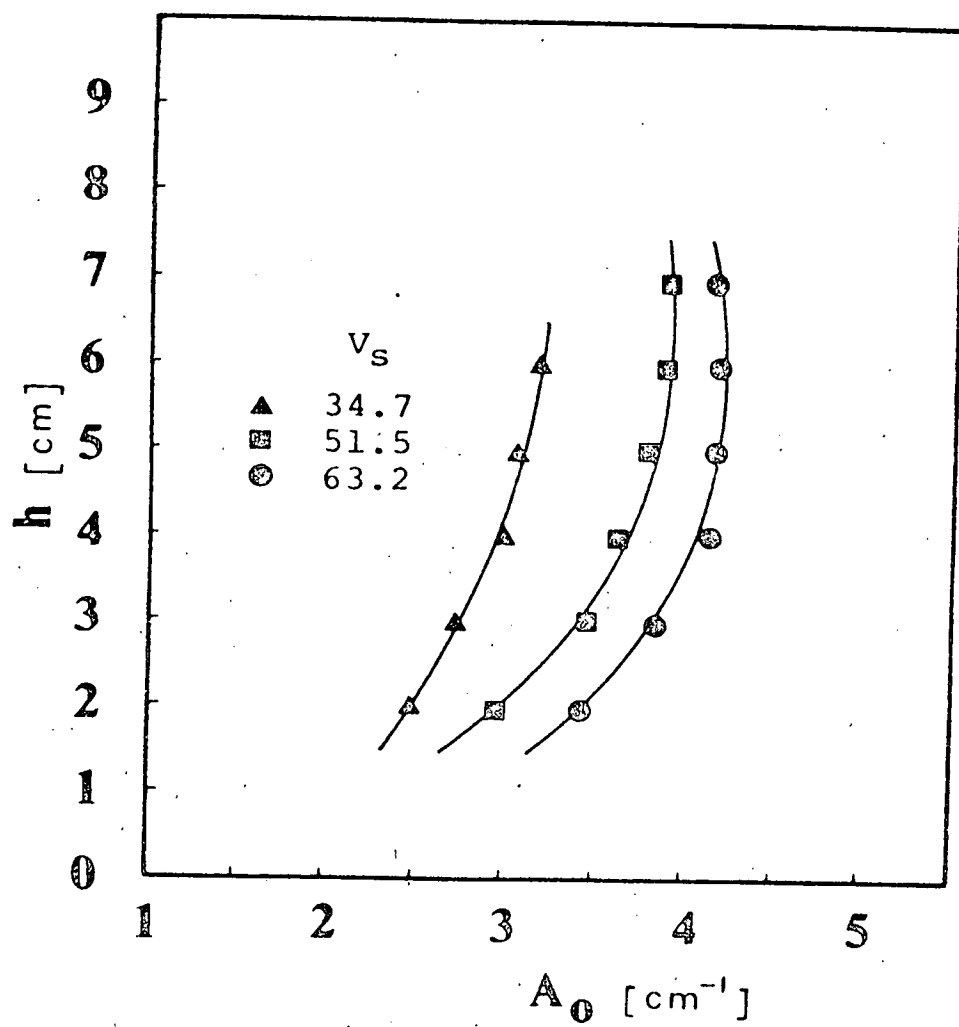


Fig. 9. Variation of vapour-liquid interfacial area with height in froth
(Boiling benzene - 5.38% free area)

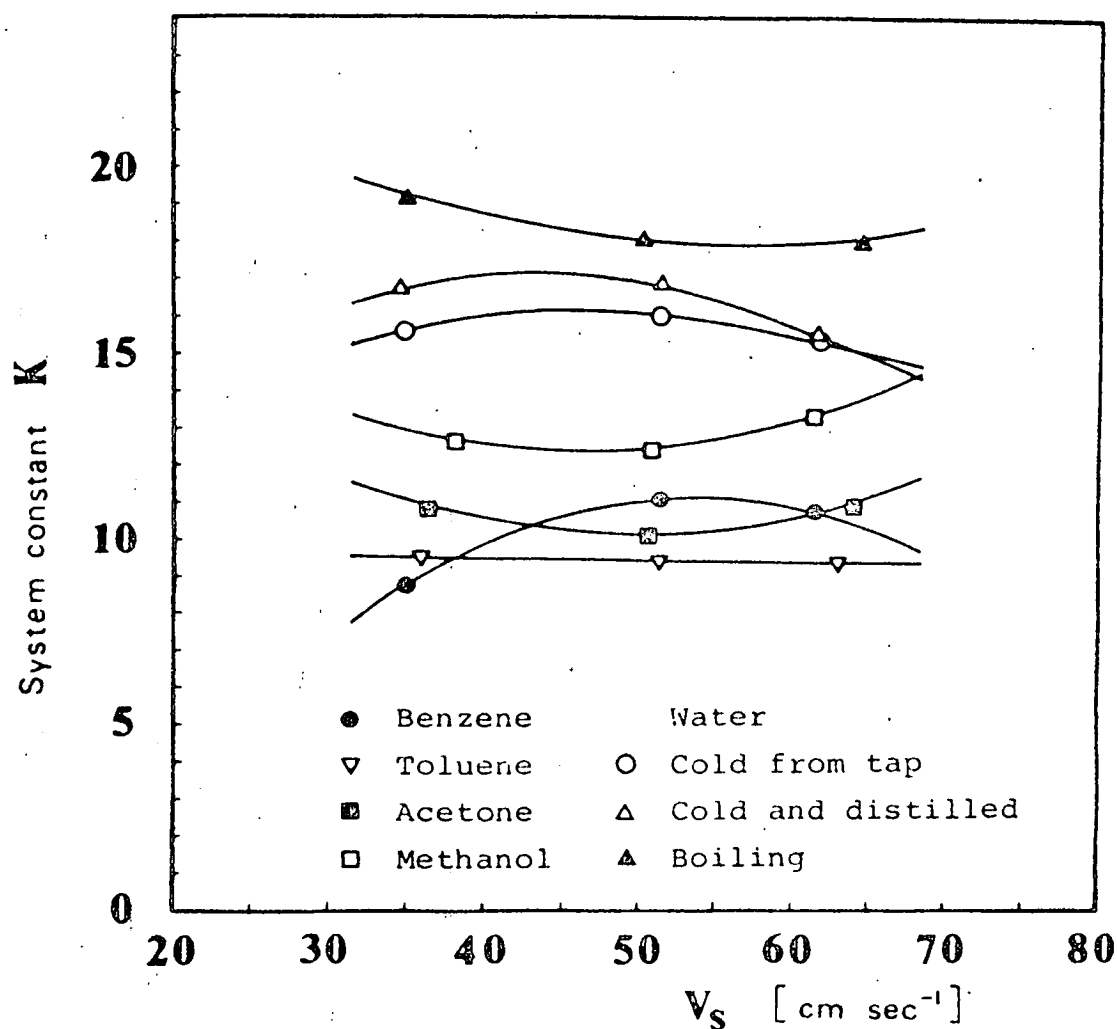


Fig. 10. The 'System Constant' (equation 12) for various boiling liquids on 5.3% free area tray, calculated for 50% plate efficiency.

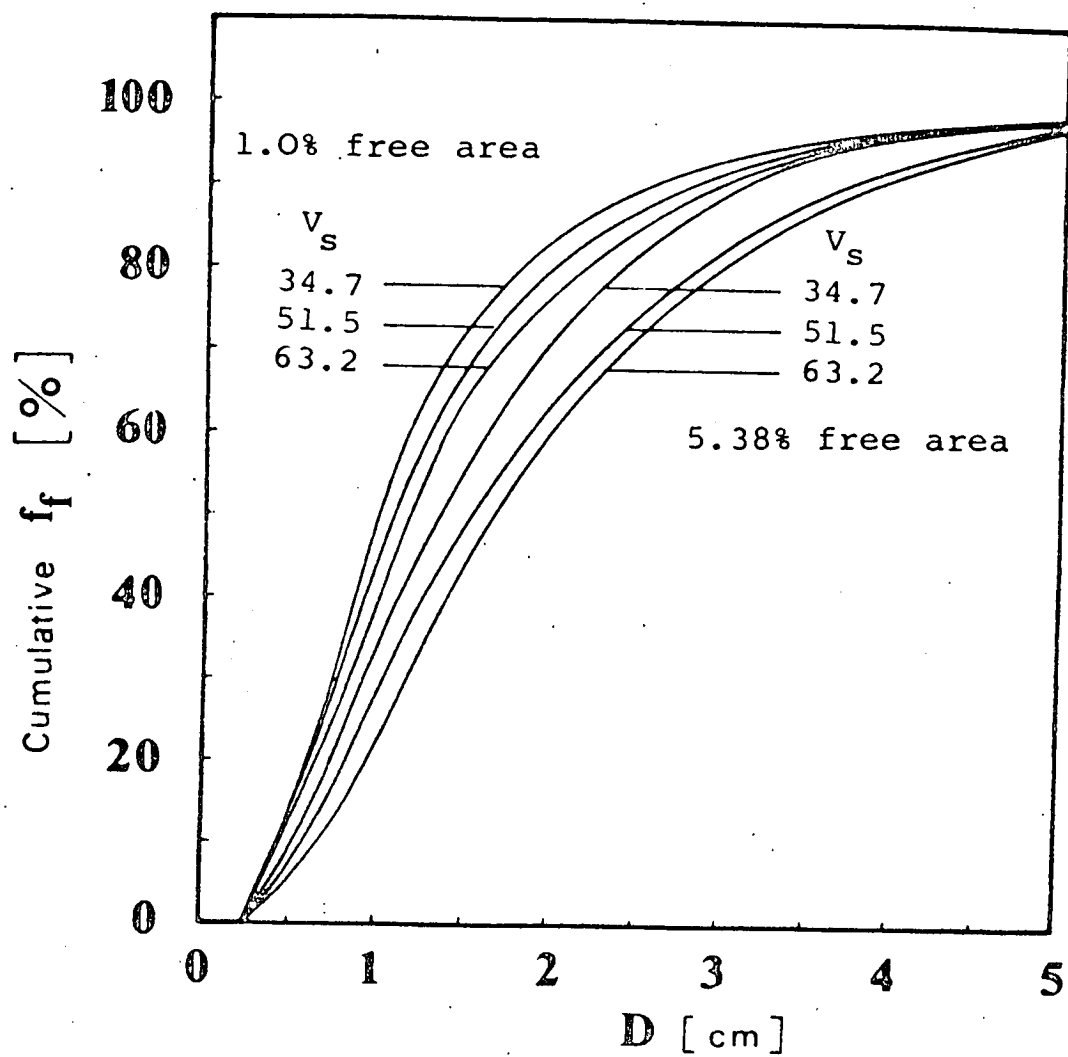


Fig. 11. The cumulative bubble flow distribution function for boiling benzene.
(5.38% and 1.0% free area sieve plates compared)

ERRATUM

The equation for frontal surface area, S_b , in the case of a paraboloidal - cap bubble (A. 21) should be

$$S_b = \frac{R_r}{6L_b^2} \left[(R_r^2 + 4L_b^2)^{\frac{3}{2}} - R_r^3 \right]$$

This affects the bubble frontal surface areas in experiments 1 and 2 (Table 52) which should be 348.6 cm^2 and 284.7 cm^2 respectively.

The validity of experimental results for K_g is not affected; the theoretical model given on pages 105 - 108 and on Fig. 5.16 should, however, be modified as far as experiments 1 and 2 only are concerned.