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DROPLET DEPOSITION AND INTERCHANGE IN ANNULAR TWO-PHASE FLOW

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Abstract—A mathematical description of droplet entrainment and deposition is presented. This enables both unidirectional deposition and interchange experiments to be analysed. This approach permits description of deposition by both the diffusion like mechanism conventionally used and by the direct impaction mechanism which has recently been identified. A criterion has been derived to differentiate between the two mechanisms of deposition based on the balance between the initial drop momentum and the drag force of a turbulent eddy. The transition criterion has been substantiated with experimental data derived from unidirectional deposition experiments. A preliminary examination of interchange data has been carried out.

INTRODUCTION

The experimental investigation of droplet motion in annular flow recently reported by James *et al.* (1980) shows that large droplets entrained by the gas are ejected from disturbance waves on the film and continue to travel in straight lines at about their initial velocity until they redeposit on the liquid film. This mechanism can be called direct impaction. However the visualisation technique employed by James *et al.* had a lower limit of resolution of 250 μm while the drop size measurements of Andreussi *et al.* (1978) and Azzopardi *et al.* (1980) indicate that a substantial volume fraction of the drops were of diameters below 250 μm. Therefore it still seems possible that a significant fraction of the entrained droplets deposit by a random diffusion-like mechanism caused by a succession of interactions with gas phase eddies.

The eddy diffusion mechanism of droplet motion appears to be amenable to a very simple mathematical representation based on the experimental finding that droplets whose transverse component of velocity is initially zero deposit according to an exponential decay law (Farmer *et al.* 1970), which implies that the rate of deposition in proportional to droplet concentration. Cousins & Hewitt (1968) measured the rate of deposition by removing the liquid film through a porous wall section and measuring the film flow rate at various points downstream. Care was taken to note whether the film flow rate exceeded the minimum value required for re-entrainment. Thus all the deposited liquid was contained in the film. The present paper re-examines these data which also show that, after about 20 tube diameters from the film removal section of the apparatus, droplet deposition is proportional to droplet concentration through a deposition coefficient which has been found, by these authors, to be almost independent of flow conditions and tube diameter.

When a significant portion of entrained droplets travels and deposits by direct impaction with a residence time much lower than the average residence time of other droplets, the simple model described above may be largely in error. In the present paper it is shown that available data on deposition and new measurements of the interchange of liquid between the film and the gas core confirm that the direct impaction and the

diffusion mechanisms are both important over a wide range of gas velocities. A new method for modelling droplet transport in a fully developed annular flow is then developed, based on the use of the cumulative distribution function of droplet residence length.

MATHEMATICAL MODELLING OF DROPLET DEPOSITION AND INTERCHANGE

(a) *The droplet residence length distribution function*

In a fully developed annular flow the rates of atomization and deposition may be assumed to be equal and constant along the tube. For a developed flow, let $E(t)$ be the volume fraction of droplets atomised at $t = 0$ still existing after a distance t. If R_A is the rate of atomization per unit area, $\pi D_t R_A E(x - z)\,dz$ is the liquid atomised between z and $z + dz$ which still exists in the form of droplets at x. For long tubes the entrained liquid flowrate at x is then given by

$$W_{LE} = \pi D_t R_A \int_{-\infty}^{x} E(x - z)\,dz \qquad [1]$$

where R_A is assumed constant. From [1] substituting $x - z$ by t

$$W_{LE} = W_{LE}(0) = \pi D_t R_A \int_{0}^{\infty} E(t)\,dt. \qquad [2]$$

When the liquid film is removed and droplets which deposit on the wall are not re-entrained, R_A may be assumed to be equal to zero for $x \geqslant 0$. In this case, from [1] W_{LE} is given by

$$W_{LE}(x) = \pi D_t R_A \int_{-\infty}^{0} E(x - z)\,dz \qquad [3]$$

or, substituting $x - z$ with t, and normalising with respect to $W_{LE}(0)$

$$W^{+}_{LE}(x) = \frac{W_{LE}(x)}{W_{LE}(0)} = \frac{\int_{x}^{\infty} E(t)\,dt}{\int_{-\infty}^{0} E(t)\,dt}. \qquad [4]$$

This equation may be adopted for the analysis of measurements of deposition and in particular, $E(t)$ may be determined by evaluating the derivative of $W_{LE}(x)$.

The rate of droplet transfer in a fully developed annular flow can also be evaluated by means of the tracer injection method first developed by Quandt (1965) and more recently critically reviewed by Andreussi & Zanelli (1976) and by Hewitt (1979). This method consists of injection of a tracer into the liquid film and measurement of the tracer concentration at a number of stations further downstream. If the tracer is injected at $x = 0$ and $C(x)$ is the tracer concentration in the liquid film, assumed to be uniform, the tracer carried by the liquid entrained between z and $z + dz$ is equal to

$$dW_S = \pi D_t R_A C(z)\,dz \qquad [5]$$

where W_S is the flow rate of tracer in the entrained liquid. The tracer flow rate along the tube is then given by

$$W_S(x) = \pi D_t R_A \int_{0}^{x} E(x - z)C(z)\,dz \qquad [6]$$

and the tracer balance between the injection point and any downstream point is

$$W_{LF}C(x) + \pi D_t R_A \int_0^x E(x-z)C(z)\,dz = W_{LF}C(0) \quad [7]$$

where W_{LF} is the liquid film flow rate.

By introducing the dimensionless concentration

$$q(x) = \frac{C(x) - C_\infty}{C(0) - C_\infty} \quad [8]$$

where C_∞, the tracer concentration for $x \to \infty$, is given by the overall balance

$$W_L C_\infty = W_{LF}C(0) \quad [9]$$

where W_L is the total film flow rate, and defining δ and η as

$$\delta = \frac{\pi D_t R_A}{W_{LF}}, \quad \eta = \frac{\pi D_t R_A}{W_{LE}}. \quad [10a, b]$$

Equation [7] becomes

$$q(x) + \delta \int_0^x E(x-z)q(z)\,dz + \eta \int_0^x E(x-z)\,dz = 1 \quad [11]$$

or, with $t = x - z$

$$q(x) + \delta \int_0^x E(t)q(x-t)\,dt + \eta \int_0^x E(t)\,dt = 1. \quad [12]$$

The derivative with respect to x of this equation is

$$\frac{dq}{dx}(x) + (\delta + \eta)E(x) + \int_0^x E(t)\frac{dq(x-t)}{dx}\,dt = 0. \quad [13]$$

Also in this case $E(t)$ may be evaluated from the first derivative of the experimental measurements of $q(x)$.

In the simple case of $E(t)$ given by an exponential decay function

$$E(t) = \exp(-\alpha t) \quad [14]$$

It is seen from [4] and [13] that $W_{LE}(x)$ and $q(x)$ have a similar form

$$W_{LE}(x) = \exp(-\alpha x) \quad [15]$$

$$q(x) = \exp[-(\delta + \eta)x] \quad [16]$$

Also, from [2] and [10b], we have

$$\alpha = \eta = \frac{1}{\int_0^\infty E(t)\,dt} \quad [17]$$

Farmer *et al.* (1970) showed that their measurements of the rate of deposition of single sized droplets injected at the tube axis could be correlated by means of [15].† The experimental investigation of the deposition of droplets originating from a wall layer reported by Cousins & Hewitt (1968), and also of the interchange determined by tracer injection reported by Cousins *et al.* (1965) and Jagota *et al.* (1973) show that at a sufficiently high gas velocity [15] and [16] give a good fit to experimental measurements. At lower gas velocities, appreciable deviations from the simple exponential decay law can be observed. We suggest that these deviations are due to the effect of the ejection velocity, which causes a rapid deposition of entrained droplets before their motion can be affected by turbulent eddies.

Available measurements of deposition and interchange are not accurate enough to allow a reliable determination of $E(t)$ by means of local values of the first derivative of $W_{LE}(x)$ and $q(x)$. In order to estimate, at least, the main features of this function, we assume that $E(t)$ can be represented as the sum of two distinct functions,

$$E(t) = e_D(t) + e_T(t) \qquad [18]$$

with

$$e_D(t) = E_D \exp(-\alpha t) \qquad [19]$$

and $e_T(t)$ representing the deviations from an exponential decay law due to the trajectory motion of larger droplets. As the lifetime of droplets depositing by direct impaction appears to be, from the shadowgraphic experiments, much lower than for the eddy diffusivity, it can be assumed that $e_T(t) \ll e_D(t)$ for t larger than some value b which has to be determined from available experiments. As $E(0) = 1$, we have $E_T = e_T(0) = 1 - E_D$. We can also introduce the parameter β defined as

$$\beta = \frac{E_T}{\int_0^\infty e_T(t)\,\mathrm{d}t} \qquad [20]$$

$1/\beta$ is proportional to the mean residence length of droplets represented by $e_T(t)$. The residence length of droplets entrained by turbulent eddies is proportional to $1/\alpha$, so we have that, in general, $1/\beta < 1/\alpha$.

Using [2] and [17]–[20], the entrained liquid flowrate, in a fully developed flow, can be expressed as

$$W_{LE} = \pi D_t R_A (E_D/\alpha + E_T/\beta). \qquad [21]$$

As can be seen from this equation, the simple expression used for $E(t)$ suggests that the entrained liquid can be divided into two streams with flowrates proportional to the fractions

$$F_D = \frac{E_D/\alpha}{E_D/\alpha + E_T/\beta} \qquad [22]$$

and

$$F_T = 1 - F_D. \qquad [23]$$

†Ganic & Mastanaiah (1981) have found their results to obey an equation of the form of [15]. However, it is not possible to compare their data with annular flow data as they did not measure drop sizes.

These streams can be assumed to represent droplets depositing by turbulent diffusion and direct impaction, respectively.

(b) *Interpretation of deposition measurements*

In the analysis of deposition experiments[4], using [18] and [19] for $E(t)$, can be written as

$$(W_{LE}^{+})_1 = \frac{(E_D/\alpha)\,e^{-\alpha x} + \int_x^{\infty} e_T(t)\,dt}{E_D/\alpha + E_T/\beta} \quad \text{for } x \leqslant b \qquad [24]$$

$$(W_{LE}^{+})_2 = \frac{(E_D/\alpha)\,e^{-\alpha x}}{E_D/\alpha + E_T/\beta} \quad \text{for } x > b \qquad [25]$$

Equation [25] indicates that measured values of $\ln(W_{LE}^{+})$ plotted versus x should lie on a straight line for $x > b$. This behaviour is confirmed by the experimental measurements of Cousins & Hewitt (1968). The intercept of this line at $x = 0$ is equal to F_D and the slope is α. These parameters of the $E(t)$ function can then by determined directly from droplet deposition experiments and for this, no assumptions are required about $e_T(t)$, besides that $e_T(t) \simeq 0$ for large values of t.

Diffusional deposition is usually described by the equation

$$R_{ED} = kC_{ED} \qquad [26]$$

where k is a mass transfer coefficient, R_{ED} is the rate of deposition of diffusing droplets and C_{ED} their concentration. In a fully developed flow C_{ED} is given by

$$C_{ED} = \frac{F_D W_{LE}}{W_G}\rho_G \qquad [27]$$

where W_G is the flow rate of the gas and ρ_G its density.

In deposition experiments, for $x > b$, deposition is only due to an eddy diffusivity mechanism, we can then assume that

$$\text{for } x > b \quad C_{ED}(x) = \frac{W_{LE}(x)\rho_G}{W_G} \qquad [28]$$

$$\left[\frac{dW_{LE}^{+}(x)}{dx}\right] = \frac{-\pi D_t R_{ED}(x)}{W_{LE}(0)} \qquad [29]$$

From [25], [26], [28] and [29] we obtain that α is related to k by

$$\alpha = \frac{\pi D_t}{W_G}\rho_G k. \qquad [30]$$

(c) *Interpretation of tracer transfer experiments*

In tracer transfer experiments a solution of known concentration of tracer is injected into the film. Samples are taken downstream of this point and the concentration of tracer in the film determined as a function of axial distance. In the analysis of these tracer injection experiments we can see from [13] that the initial slope of $\ln[q(x)]$ vs x is given

by

$$\phi = \delta + \eta = \pi D_t R_A \left[\frac{1}{W_{LE}} + \frac{1}{W_{LF}} \right]. \quad [31]$$

The value of ϕ is not affected by the actual form of the $E(t)$ function. Available measurements of the interchange and also the experiments presented in this paper show that the slope of $\ln [q(x)]$ is about constant for $0 < x < 20D_t$. This can be seen in figure 1 (Sabatini 1978). For larger values of the axial coordinate the slope tends to assume a lower value. This experimental result is confirmed by previous measurements of Cousins *et al.* (1965) and Jagota *et al.* (1973). In figure 1 we also represented the solution of [13] for $E(t)$ given by

$$E(t) = E_D \exp(-\alpha t) + E_T \exp(-\gamma t^N). \quad [32]$$

In this equation the values of E_D, α and γ have been chosen to give the best fit to experimental measurements ($E_D = 0.7$); the form of $e_T(t)$ and the value of the constant $N(N = 1.7)$ have been suggested by the analysis of shadowgraph experiments (Andreussi & Azzopardi 1982).

The rate of interchange can easily be determined from the initial slope of $\ln [q(x)]$ by means of [31]. However, when the effect of ejection velocity is appreciable, the parameter α, or k, which characterises the eddy diffusivity mechanism of deposition, cannot be directly evaluated from tracer interchange measurements. In fact, α is related to the rate of deposition R_D (equal to R_A) through [21]. Using [22], [21] can be written as

$$\alpha = \frac{E_D}{F_D} \frac{\pi D_t R_A}{W_{LE}} = \frac{E_D}{F_D} \eta. \quad [33]$$

From this equation it can be seen that the parameter η, which can be derived from the initial slope of $\ln [q(x)]$, is equal to the coefficient α for $E_D = F_D = 1$, otherwise η is always larger than α, because $F_D > E_D$ when $1/\beta < 1/\alpha$.

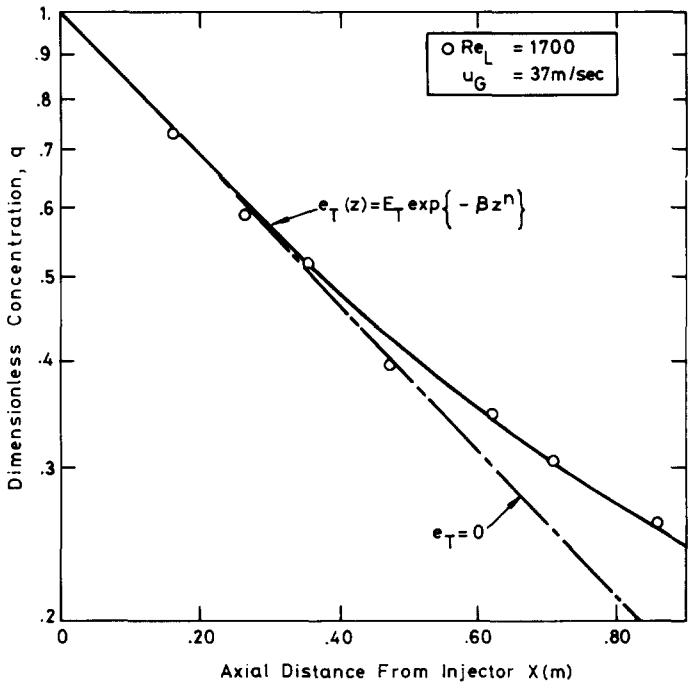

Figure 1. Dimensionless tracer concentration for different residence length distributions.

More information about $E(t)$ can be obtained from the results of tracer analysis by choosing a specific form for $e_T(t)$ and determining the value of various parameters by an optimization technique. This has been carried out for the experiment represented in figure 1, but due to experimental errors, the number of free parameters, and also to the uncertainty about the actual form of $e_T(t)$, this type of analysis does not seem to be reliable.

An approximate analytical solution of [13] can be obtained for $x > b$ by assuming that, in the calculation of the integral in this equation, $q(x)$ can be approximated by

$$q(x) = \exp\left[-(\delta + \eta)x\right] \qquad [34]$$

as it is suggested by the value of the slope the range $0 < x < b$. For $x > b$, we have $e_i(x) \simeq 0$, $E(x) \simeq e_D(x)$ and [13] becomes

$$\frac{dq}{dx} = -(\delta + \eta)\, e^{-(\delta+\eta)x}\left[E_D\, e^{(\delta+\eta-\alpha)x}\frac{\eta - \alpha}{\delta + \eta - \alpha} + \frac{\delta E_D}{\delta + \eta - \alpha} - \delta\int_0^\infty e_T(t)\, e^{(\delta+\eta)t}\, dt\right]. \qquad [35]$$

For small deviations from the eddy diffusivity mechanism of deposition, the term within brackets in [35] can be approximated by E_D, as $\eta \simeq \alpha$ and δ is in general small when the effect of ejection velocity is appreciable. Close to $x = b$, [35] can then be written as

$$\frac{dq}{dx} = -(\delta + \eta) q(x) E_D \qquad [36]$$

from which it can be seen that an approximate value for E_D can also be obtained from tracer experiments as the ratio between initial and final slopes of the curve $\ln[q(x)]$ vs x. The position at which the slope changes gives an estimate of the maximum deposition length of droplets depositing by direct impaction.

CRITERION FOR CLASSIFICATION OF DROPLET MOTION

The main features of the residence length distribution function $E(t)$ can be predicted for different flow regimes with the hypotheses that the initial transversal velocity and the direction of the velocity vector of entrained droplets do not depend on droplet size. These assumptions are confirmed by the measurements made by the shadowgraph technique and from this Andreussi & Azzopardi (1982) have suggested a simple relationship between droplet initial transverse velocity, v_i, and gas friction velocity, u^*,

$$v_i = 12\sqrt{\rho_G/\rho_L}\, u^* \qquad [37]$$

where ρ_L is the liquid density. This correlation is based on data from a number of velocities at two pressures. The only other data available for this parameter is that of Chang (1973). However, this worker varied neither densities nor velocities.

The motion of droplets of different size leaving the film at the same transverse velocity can be simulated by means of the model developed by James *et al.* (1980). This models the successive interactions between a drop and gas eddies. The results of this simulation are shown in figure 2 from which it is seen that droplets below 80 μm in diameter are noticeably affected by the gas turbulence, while droplets above 120μm in diameter are hardly affected by turbulent eddies. The criterion for this transition is arbitrary and can be based, for instance, on the fraction of droplets which deposit in the quadrant opposite the point of ejection.

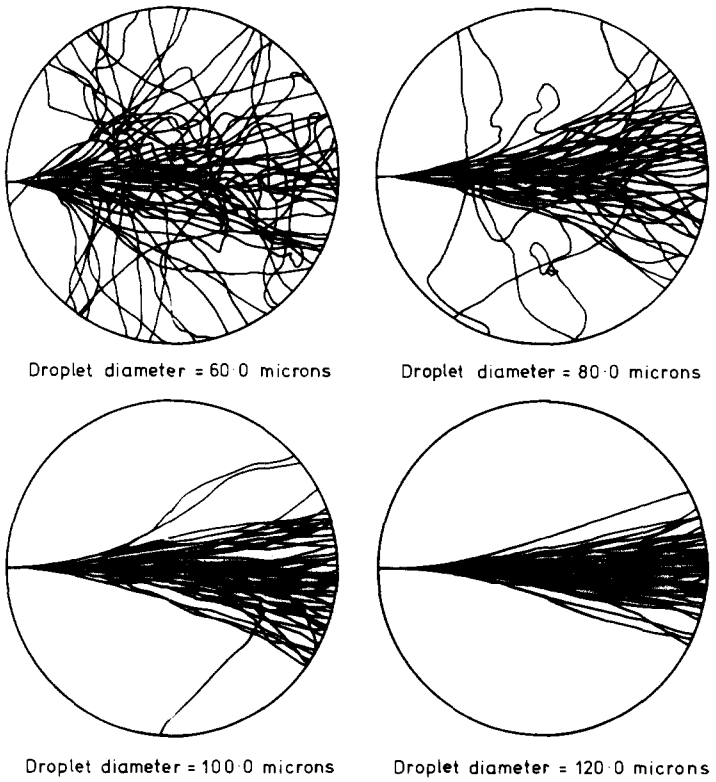

Figure 2. Simulation of droplet motion (initial velocity $v_i = 0.9$ m/sec for all cases).

The method developed in this paper for predicting the size of droplets at the transition between random-walk and trajectory motion is based on a dimensionless group which represents the ratio between the initial momentum of droplets in the plane normal to flow and the drag force exerted by turbulent eddies multiplied by the eddy characteristic time. The droplet initial momentum is

$$M_d = m_d v_i \qquad [38]$$

where v_i is the initial transverse velocity of the drop. For this analysis the mean values of this parameter are used even though experimental observation has shown velocities normally distributed about the mean with a standard deviation of 0.4 times the mean. However, as no correlation was found between the transverse velocity and the drop size it is assumed that this averaging is valid. The drag force can be set to be equal to

$$f_d = \frac{C_D}{2} \rho_G u^{*2} \frac{\pi D_d^2}{4} \qquad [39]$$

where D_d is the drop diameter and it has been assumed that the eddy velocity scales as the gas friction velocity. It has also been assumed that in [39] the eddy velocity represents the mean value of the velocity difference between droplet and eddy at the starting time of the eddy.

Following James *et al.* (1980) it is assumed that the eddy characteristic time, T_e, scales as the ratio between the characteristic length, L_e, and the gas friction velocity and L_e is proportional to the tube diameter D_t. A dimensionless group can then be defined as the momentum/drag ratio

$$\frac{M_d}{f_d \cdot T_e} \propto \frac{\rho_L v_i D_d}{C_D \rho_G u^* L_e}. \qquad [40]$$

The effect of eddies is also proportional to the number of interactions which occur before deposition. This number may be assumed to scale as the ratio between the tube diameter and the characteristic length. Finally, a dimensionless group, G, is obtained,

$$G = \frac{\rho_L v_i D_d}{C_D \rho_G u^* D_t} \qquad [41]$$

a critical value of which is the criterion which classifies the different types of motion, i.e. droplets which give values of G larger than this critical value to be determined from available data, will not be affected appreciably by turbulent eddies.

C_D, the drag coefficient, has been assumed to be given by the following equation, which correlates the measurements of Lapple & Shepherd (1940) in the so-called intermediate region ($0.3 < \mathrm{Re}_d < 1000$):

$$C_D = \frac{18.5}{\mathrm{Re}_d^{0.6}} \qquad [42]$$

with

$$\mathrm{Re}_d = \frac{\rho_G D_d u^*}{\mu_G}. \qquad [43]$$

An initial estimate of G can be obtained from the numerical experiment developed by James *et al.* (1980). The value of G is 0.51 when the criterion for transition is that half of the droplets deposit in the quadrant opposite the point of ejection and $G = 1.1$ when the criterion is that all droplets deposit in the quadrant opposite the point of ejection.

If in [41] the value of G is estimated as suggested above or calculated from experimental data and v_i is assumed to be given by [37], as determined from analysis of shadowgraph films, the droplet size at the transition can easily be determined. The drop size distribution and the average size can be predicted by the correlations recently developed by Azzopardi *et al.* (1980). These authors adopted the Rosin-Rammler distribution for correlating the cumulative volume distribution of droplet size.

This function has two adjustable parameters, which appear to be quite independent from flow conditions, and can be approximated as follows:

$$V = \exp\left[-[D_d/(\phi D_{32})]^N\right] \qquad [44]$$

where V is the volume fraction of liquid contained in droplets larger than D_d, D_{32} is the Sauter mean diameter and ϕ and N have both average values close to 2. According to Azzopardi *et al.* (1980) D_{32} is given by

$$\frac{D_{32}}{D_t} = 1.91 \frac{\mathrm{Re}_G^{0.1}}{\mathrm{We}^{0.6}} \left[\frac{\rho_G}{\rho_L}\right]^{0.6} + 0.4 \frac{\rho_G W_{LE}}{\rho_L W_G} \qquad [45]$$

By means of [41], [44] and [45] the fractions F_D and F_T of droplets which deposit by eddy diffusion or by direct impaction can easily be evaluated as $F_T = V$ when D_d is equal to the drop size at transition.

ANALYSIS OF DEPOSITION EXPERIMENTS

The equations described above have been used to analyse data from unidirectional deposition experiments (Cousins & Hewitt 1968) and interchange.

Cousins & Hewitt (1968) measured the deposition rate for upward annular flow of air–water by removing the liquid film and measuring the flow rate of the new film at a number of distances from the original film removal point. Most of their experiments were carried out in a tube of 9.5 mm diameter, a minority of the data were from a 32 mm diameter tube. Measurements from at least three deposition distances are necessary to determine the slope of the $\log_e W_{LE}$ vs x. However, at high values of deposition length, experimental measurements may be in error because re-entrainment occurs or flow conditions change appreciably along the tube. In other experiments the entrained liquid flow rate is very low and hence the measurements of this quantity are not reliable. Data which suffered from any of these defects were rejected, these were mainly from high gas flow rate runs. As already mentioned, Farmer *et al.* (1970) have shown that, for droplets which are entirely in a diffusion-like manner, log (W_{LE}) data are linear in distance. When the data of Cousins & Hewitt are plotted in this manner it is seen that data only lies on a straight line if the axial distance is greater than 0.15 m (for the 9.5 mm diameter tube) and 0.8 m (for the 32 mm diameter tube). However the sinter extraction unit was of finite size and if entrainment is assumed to cease half-way along the sinter a correction to the distances given above is necessary. The corrected maximum deposition length are 0.2 m (9.5 mm tube) and 0.9 m (32 mm tube).

Motion in a given time in the transverse direction is governed by v_i which is proportional to u_G whilst that in the axial direction is controlled by u_G. Therefore the motion will be geometrically similar for different gas velocities. That is the deposition length are almost independent of u_G.

A one-dimensional drag equation has been integrated twice to yield axial distances travelled by drops. The residence of the drops in the gas was determined from the transverse velocity and the average transverse distance travelled by drops. The distances travelled by droplets with diameters equivalent to the volume median diameter of the direct impaction droplets are in the ranges 0.08–0.1 m (9.5 mm tube) and 0.36–0.4 m (32 mm tube). The calculations assumed an average ejection velocity and were carried out over the ranges of gas velocities employed by Cousins and Hewitt. The mean distances are about half the maximum deposition lengths obtained from the deposition experiments (see above) and are in good agreement with the correlations and the other experimental data presented in this paper.

In figure 3 the intercept of the straight line through data with the ordinate at $x = 0$ represents the fraction of diffusing droplets after the sinter. However a more relevant value would be that at which entrainment ceased, this was assumed to be half-way along the sinter section. The extrapolation is fairly simple for the 32 mm tube data, as the first measurement length is within the region where both mechanisms of deposition are encountered. Data relative to the 9.5 mm tube have all been extrapolated as shown in figure 4 that is by tracing a straight line in semi-log paper through the point (0,1) which crosses the line through data at $x = 0.075$ m. This procedure is consistent with measurements and with the calculated values of axial distances reported above.

Values of F_D are determined as

$$F_D = \frac{F'_D}{F'_D + F'_T} \qquad [46]$$

where F'_D and F'_T are defined in figure 4.

The fraction F_D of droplets which deposit by a diffusion-like mechanism is plotted versus u_G in figure 5. It is seen that F_D is independent of gas density, but increases with gas velocity and tube size. From the value of F_D, the droplet size at the transition between diffusion and direct impaction can be determined by [44] with D_{32} calculated by [45]. The

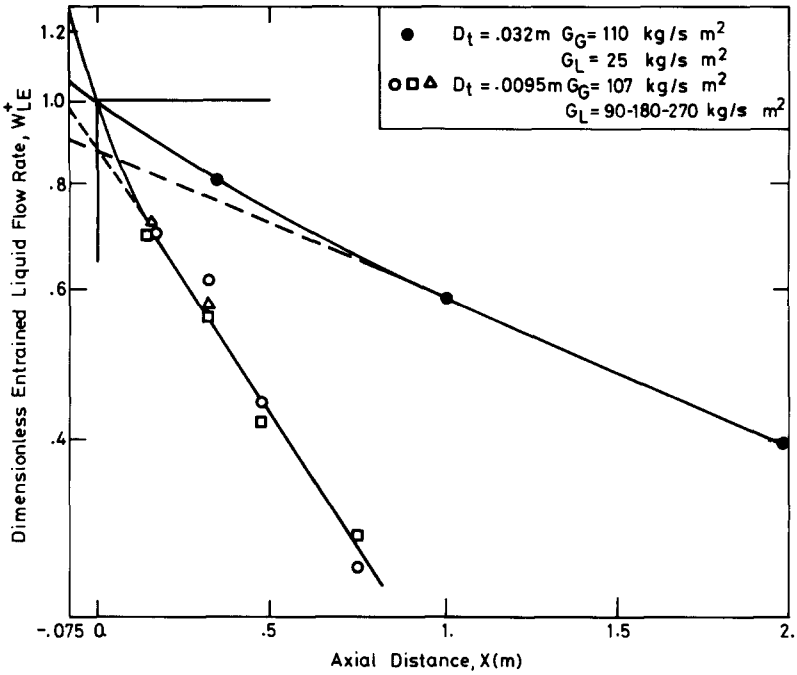

Figure 3. Measurement of rate of deposition deposition data (Cousins & Hewitt 1968).

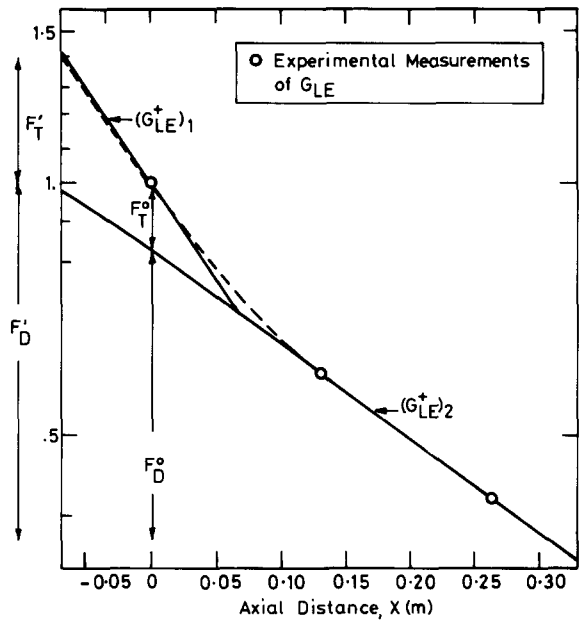

Figure 4. Extrapolation of measured entrained liquid flowrate.

dimensionless group G for the flow conditions analysed in the present work has been calculated from these values of v_i and D_d. Values of G were deduced and are plotted vs u_G in figure 6. As predicted by the theoretical analysis of droplet motion, values of G are, with a good approximation, constant and about equal to 0.7. This is within the limits determined by the numerical simulation of droplet motion.

ANALYSIS OF INTERCHANGE EXPERIMENTS

Interchange experiments have been conducted in a vertical 24 mm ID tube with air-water in downflow. Details of the flow loop and the other measurements made in this

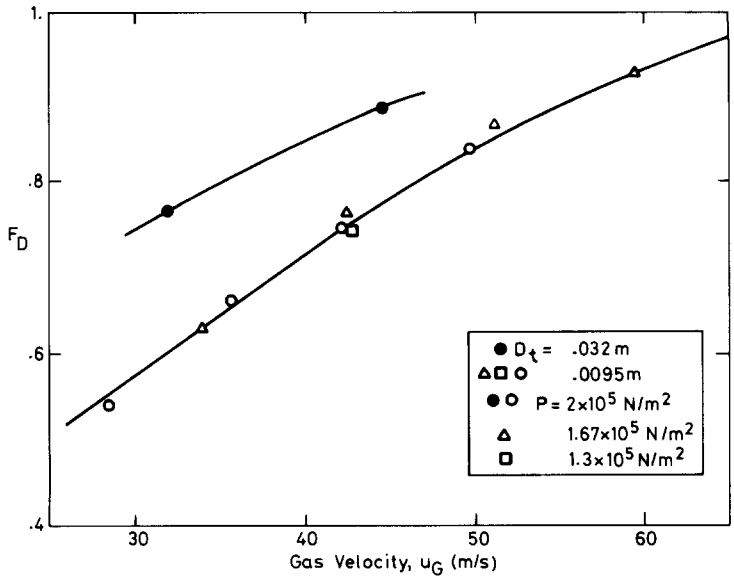

Figure 5. Measured fraction of droplets which deposit by a diffusion like mechanism.

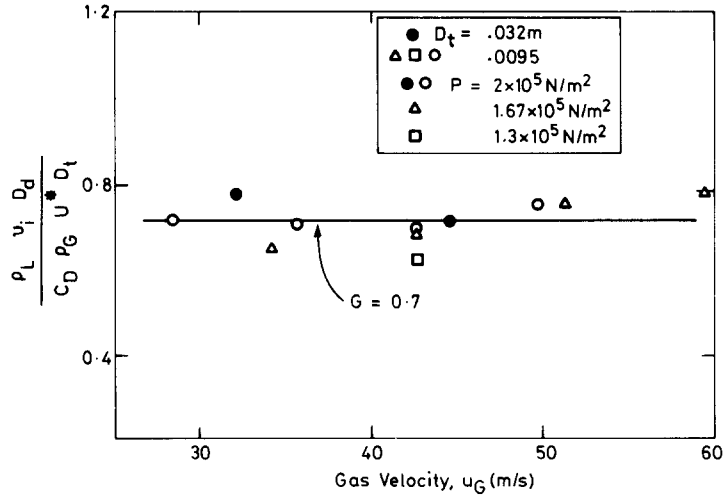

Figure 6. Correlation of droplet diameter at the transition between diffusion and direct impaction mechanisms of deosition.

apparatus can be found elsewhere (Andreussi & Zanelli 1979). The tracer, a concentrated solution of NaCl, was injected about 1.1 m before a liquid film extraction station at the outlet of the tube. The tracer concentration was determined by electrical conductivity measurements. Entrained liquid flowrates were also measured by measuring the liquid film flow rate at the film extraction station. Further details of the experimental technique, and a complete analysis of the results, will be presented in a forthcoming paper. The effect of gas flowrate on the dimensionless tracer concentration profiles is shown in figure 7 for a low value of the liquid Reynolds number. Although these data show some scatter, for $u_G = 37$ m/s the curve changes appreciably in slope at $x = 0.45$ m. The ratio between the slopes in the initial and final portion of this curve is about 1.6. This value and the value of x at which the $\ln(q)$ curve changes in slope are in a very good agreement with the predictions of the correlations developed in this paper. At higher gas velocities F_D increases and it is very close to one for the other cases shown in figure 7. The effect of liquid flowrate is shown in figure 8, at increasing values of Re_L the slope of $\ln(q)$ decreases both in the initial and final portions of the curves. The ratio of slopes is approximately constant or

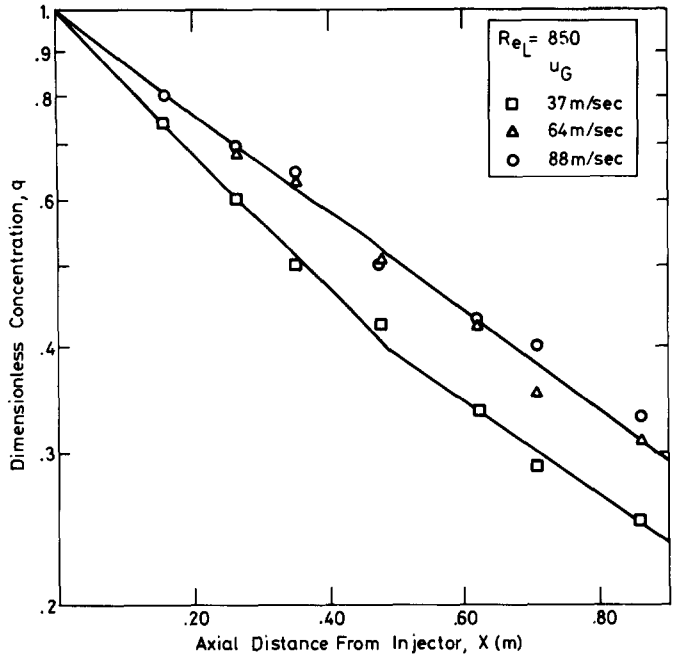

Figure 7. Tracer concentration in the liquid film; effect of gas velocity.

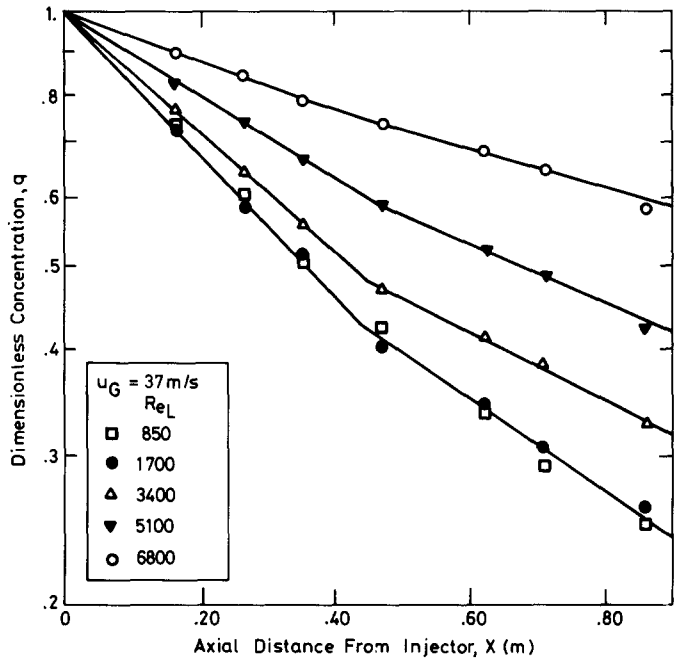

Figure 8. Tracer concentration in the liquid film; effect of liquid flowrate.

decreases with increasing liquid flowrate. According to the model for droplet motion developed here, F_D should increase with increasing average value of drop size. At a given gas velocity, experimental measurements of drop size reported by Andreussi *et al.* (1978) and by Azzopardi *et al.* (1980) show that drop size increases with droplet concentration in the gas core. This behaviour has been attributed by Azzopardi *et al.* (1980) to droplet coalescence. Coalescence may also affect the motion of large droplets and it seems likely that the theoretical analysis developed here only applies to low values of droplet concentration, less than $0.35\,\rho_G$. This limitation is also suggested by the fact that droplet concentration may also affect pressure drops and the structure of turbulence in the gas

core. Therefore further theoretical and experimental work seems to be needed for a more general model of droplet motion in annular flows.

CONCLUSIONS

The following conclusions may be drawn from the work described above.

(1) Measurements of droplet deposition and interchange show that droplet motion in annular flows can be significantly affected by the transverse velocity at which droplets are ejected by the liquid film. The influence of initial velocity decreases with gas velocity and tube diameter.

(2) The results obtained by the simple model developed by James *et al.* (1980) for describing droplet motion through the gas core are in good agreement with the analysis of deposition and interchange measurements reported in the present paper. It has also been shown that droplet size at the transition between eddy diffusion and direct impaction mechanisms can be predicted by means of a dimensionless number representing the ratio between droplet initial momentum in the plane normal to flow and the gas drag multiplied by the eddy characteristic time.

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