

## DROPLET TRANSFER IN TWO-PHASE ANNULAR FLOW

PAOLO ANDREUSSI

Departmento di Ingegneria Chimica, University of Pisa, Italy

(Received 15 April 1982; in revised form 23 March 1983)

**Abstract**—The tracer method, adopted by a number of experimenters for the measurement of the rate of liquid interchange between the wall layer and the dispersed phase in a fully developed annular flow, has been critically examined on the basis of a more accurate description of the flow behavior of entrained droplets. The results obtained indicate that this experimental technique can give reliable results in a wide range of flow conditions when the experiments are properly designed.

Present measurements of the interchange show that the dimensionless deposition coefficient is a function of the concentration of entrained droplets. This effect appears to be due to the dependence on concentration of droplet size.

The rate of entrainment is proportional to the liquid film flowrate and to the gas velocity squared.

### INTRODUCTION

Annular two-phase flow is encountered in a wide range of industrial applications, including nuclear reactors, evaporators, condensers and pipelines. In this flow configuration, the liquid flows along the wall as a film, and at a sufficiently high gas velocity also as droplets entrained by the gas. In adiabatic flows, equilibrium can be attained between the rates of droplet entrainment and deposition. In this case, the entrained liquid flowrate and the other flow parameters become approximately constant along the channel.

Diabatic flows are more often encountered in engineering applications and are characterized by a continuous variation of the flow configuration along the channel. In these flows, liquid transport between the wall layer and the gas core can be of critical importance in modeling heat and mass transfer processes. In particular, liquid entrainment and deposition must be carefully considered in the analysis of dryout and burnout.

The rates of droplet deposition and entrainment have been measured under equilibrium or developing flow conditions. However, as shown by Hewitt (1979) in a recent review, there are no well established techniques for measuring these quantities and no general correlations based on a clear-cut physical reasoning.

In the analysis of liquid transport in annular flows, developing flow conditions are encountered when it is desired to avoid the simultaneous presence of deposition and entrainment. For example, the deposition flow can be measured after the separation of the wall layer (Cousins & Hewitt 1968), and the entrainment flow can be evaluated as the initial rate of entrainment following the smooth injection of the liquid film (Tatterson 1975). In both cases, the correlation of measurements as a function of main flow parameters can be complicated by the continuous variation of the flow configuration along the tube. This problem does not arise in the analysis of measurements of the equilibrium rate of interchange, because, in this case, relevant flow parameters such as the liquid film thickness or the pressure gradient become constant along the tube and can be determined by a number of well established experimental methods.

The interchange flowrate can be measured by means of the tracer technique first adopted by Quandt (1965). This method consists of the injection of a tracer in the wall layer and the determination of the tracer partition between the wall layer and the entrained liquid at a

number of axial locations. The analysis of measurements is based on the assumption of complete mixing of the tracer within the gas core and the wall layer at any cross sections.

The experimental results obtained by the authors who adopted this technique after Quandt (1965) (Cousins *et al.* 1965; Jagota *et al.* 1973) seemed to contradict these assumptions. However, Andreussi & Zanelli (1976) have recently shown that from a theoretical standpoint tracer mixing within a turbulent, thin liquid film should be almost instantaneous. On the other hand, Andreussi & Azzopardi (1981) found that perfect mixing within the gas core could only be assumed at high gas velocity, but they also showed that this assumption was not indispensable for the evaluation of the interchange flowrate.

In the present paper, a detailed analysis of the tracer technique is presented along with a set of experimental measurements. These data, which have been tentatively correlated by empirical equations, shows that the deposition coefficient depends on the gas friction velocity and the concentration of entrained droplets. The rate of entrainment is found to be proportional to the liquid film flowrate and to the gas velocity squared.

#### DESCRIPTION OF EQUIPMENT

The rig used for the experiments was a modified version of that employed by Andreussi & Zanelli (1979) for measuring pressure drop, film thickness, drop size and liquid entrainment. The air flow was supplied by a high pressure facility line. The water flow was supplied from a reservoir pressurized to a constant pressure of 3 bars. Both air and water streams were metered using calibrated rotameters. The rig operated at atmospheric pressure.

The test section consisted of a 24 mm ID extruded Plexiglass tube which was in sections joined by carefully matched flanges. The flow was directed downward and the tube was carefully mounted to ensure that it was vertical. The air stream entered at the top of the tube and passed through a 0.5 m calming section before reaching the liquid injector. At the outlet from the test section, the liquid film was separated from the air stream. Air and entrained droplets then flowed to a gas-liquid separator. The liquid film and the entrained liquid flowrates were measured at the outlet by determining the passage time of carefully measured volumes of liquid. Pressure drops were measured by water purged inverted U-tube manometers.

Two different test sections have been used for the experiments. First experiments were conducted in 5.0 m long test section. The liquid film was introduced and extracted through porous sinter sections 0.1 m long. The tracer was injected through 120 0.5 mm diameter holes spaced equally around the periphery of the tube and divided into three rows. The distance between rows was 5 mm and the tracer injection section was located 1.1 m from the outlet. Samples of the liquid film could be withdrawn at nine locations after the tracer injector. At each sampling location four 0.6 mm diameter holes were drilled on the tube wall and surrounded by an annular ring. The samples were withdrawn by allowing the liquid with some gas to leak rapidly from the tube. The entrained liquid could be sampled at four locations by pitot-type probes 2 mm OD and 50 mm long.

Most of the experimental results relative to the measurement of the interchange reported in this paper have been obtained with this test section. However, in order to obtain a better understanding of the phenomena under investigation, an improved version of the test section has also been built. A schematic diagram of the new test section and of the tracer line is shown in figure 1. As indicated in this figure, in the new apparatus samples of the liquid film were obtained at nine axial locations (*A*) and samples of the dispersed phase at four axial locations (*C*). It was also possible to withdraw local samples of the liquid film around the periphery of the tube at two axial positions (*B*). At each of these positions, local samples were extracted through eight holes 1.0 mm diameter equally spaced around the circumference. As shown in figure 1, it was possible to sample the entrained liquid by means of pitot tubes 50 mm long at the same sections at which local concentration of the liquid film

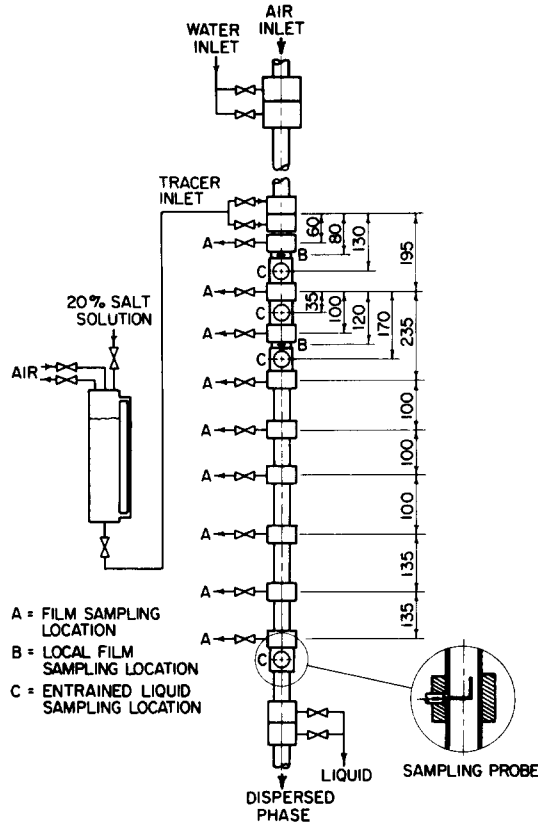


Figure 1. Schematic diagram of the experimental apparatus.

was determined. The extracted liquid was sampled along two traverses at  $90^\circ$ . The position of the sampling probe was determined by a micrometer head.

Sampling locations designated with *A* in figure 1 were made of 30 holes 0.7 mm diameter equally spaced in three rows. The distance between rows was 10 mm.

The new test section was 6.0 m long. The tracer injector was located 1300 mm from the outlet and consisted of two contiguous chambers, one with 100 holes 0.6 mm diameter, the other with 250 holes 0.7 mm diameter in seven rows. The distance between rows was 3 mm. In this apparatus, the liquid stream was introduced through 400 holes 2 mm diameter spaced in eleven rows and the liquid film was extracted through 1100 holes 1.2 mm diameter divided in two separate sections 50 mm and 40 mm long.

The tracer used in the experiments was a 20% sodium chloride solution. A constant flow of tracer was supplied to the test section from a reservoir filled with the salt solution and pressurized with air from a cylinder to a constant pressure of 3 bars. The tracer flowrate was determined by measuring at different times the level of the reservoir with a cathetometer. This flowrate was always less than 5% of the total water flow. Experiments conducted with tracer flowrates varying from 2% to 6% of the total liquid flow did not show any appreciable influence of this parameter. Tracer concentration was determined by measuring the electrical conductivity of small samples (2–5 ml) with a thermostated cell and a Metrohm 518 Conductometer.

#### EXPERIMENTAL METHOD

##### (a) Tracer mixing in the wall layer

The tracer method for determining the equilibrium rates of entrainment and deposition is based on the assumption that tracer injection leads to instantaneous mixing in the radial direction within the wall layer. In the experiments described by Cousins *et al.* (1965), tracer concentration measured close to the injection point was characterized by a large scatter and,

in a few cases, it was higher than the film concentration at the injection calculated assuming perfect mixing. In that work samples were withdrawn by allowing the liquid to leak slowly from four 0.8 mm diameter holes surrounded by an annular ring. The experimental results were explained by assuming a poor mixing of the tracer in the radial direction. It was believed that the samples were withdrawn preferably from the region close to the wall where the tracer concentration was higher.

The anomalous behavior of measurements taken close to the injection point can also be attributed to large variations of tracer concentration around the tube wall. Jagota *et al.* (1973), for instance, reported that they could not attain a uniform tracer concentration profile around the periphery of the tube. According to these authors, this was due to a non-uniform injection and to poor mixing around the circumference.

In the present work the effectiveness of tracer mixing in the radial direction close to the injection point has been tested with a simple experiment. At a given liquid flowrate the tracer was continuously added to the wall layer for increasing values of the gas velocity, but always within flow regimes which are characterized according to a previous paper by the author (Andreussi 1980) by negligible liquid entrainment. Samples of the liquid film were withdrawn 50 mm from the point of tracer injection through a local sampling hole or by means of a pitot tube 2 mm diameter positioned against the tube wall. Tracer concentration obtained by the two methods were about equal at any flowrate. The pitot tube was then placed at such a distance from the wall that only the tips of large disturbance waves could be sampled. In figure 2, we present the ratio  $\eta$  of tracer concentration within the large waves to mean local concentration as a function of the gas velocity. As can be seen, this ratio, which is a measure of the effectiveness of radial mixing within the liquid film, becomes very close to one at gas velocities below the value at which liquid entrainment is appreciable. It is concluded that the assumption of instantaneous mixing of the tracer in the radial direction can be safely adopted in the analysis of measurements, and also that local samples drawn from small holes drilled on the tube wall are representative of the average tracer-concentration at that circumferential position.

However, local samples withdrawn close to the tracer injection location confirmed the existence of appreciable concentration variations (within  $\pm 30\%$  of the average value) around the periphery of the tube previously noted by Jagota *et al.* (1973). It has also been noticed that at low gas velocity, with all the liquid flowing in the wall layer, the mean tracer concentration calculated as the arithmetic mean of local concentrations was different from the mean concentration at the outlet. As the injectors used in the present research seem to guarantee a uniform circumferential distribution of the tracer, we conclude that the observed concentration variations are due to non-uniform liquid flow around the tube wall.

Liquid film maldistribution around the periphery of the tube probably does not have relevant effects on the amount of liquid entrained by the gas. However, the experiments

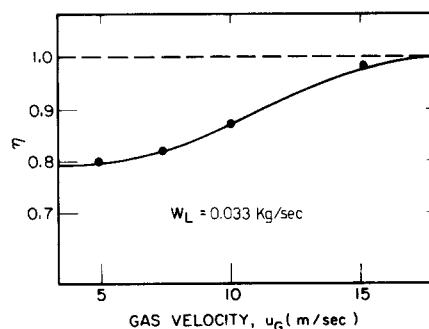


Figure 2. Effectiveness of tracer mixing in the wall layer.

described above show that the concentration of samples withdrawn close to the injector is not equal to the mean film concentration at that axial location. Therefore, we considered in the analysis of measurements only samples obtained at distances larger than 150 mm from the injector, as at these distances circumferential mixing appears to be reasonably complete.

(b) *Mathematical model of tracer transport*

Tracer concentration measurements have usually been correlated by means of the following mass balances relative to tracer transport in the wall layer and the dispersed phase:

$$W_F \frac{dC_F}{dz} = -\pi D_i R_l (C_F - C_l) \quad [1]$$

$$W_A \frac{dC_A}{Dz} = \pi D_i R_l (C_F - C_l). \quad [2]$$

In these equations  $W_F$  and  $W_A$ , are the liquid film and the entrained liquid flowrates,  $R_l$  is the rate of interchange,  $C_F$  is the tracer concentration in the wall layer and  $C_l$  is the mean tracer concentration of depositing droplets, which is usually assumed to be equal to  $C_A$ , the mean concentration across the gas core. In terms of the dimensionless concentration,  $q$ , defined as

$$q = \frac{C_F - C_e}{C_{F,0} - C_e} \quad [3]$$

where  $C_{F,0}$  and  $C_e$  are the tracer concentrations at the injection and the limiting or equilibrium concentration attained in a long tube, respectively, the concentration profile in the liquid film is given by

$$q(z) = \exp \left[ -\pi D_i R_l \left( \frac{1}{W_F} + \frac{1}{W_A} \right) z \right]. \quad [4]$$

Equations [1] and [2] are based on the assumptions of fully developed flow and uniform tracer concentration in the wall layer, and are integrated assuming perfect mixing of the dispersed phase at any cross sections. The assumption of perfect mixing of the entrained liquid does not imply that the mean tracer concentration,  $C_A$ , be a constant across the gas core, but only that  $C_A = C_l$ . In figure 3 we present tracer concentration profiles in the gas core determined at various axial distances from the injection point. As can be seen, close to the injection these profiles are far from being uniform. However, Andreussi & Azzopardi (1980) showed that at a sufficiently high gas velocity perfect mixing of the dispersed phase could safely be assumed. In particular, this is the case for the flow regime considered in figure 3.

In order to describe the flow behavior of entrained droplets Andreussi & Azzopardi (1981) introduced the cumulative residence length distribution function  $E(x)$ , which represents the fraction of droplets with residence length larger or equal to  $x$ . In terms of  $E(x)$  the amount of tracer entrained at an axial position between  $t$  and  $t + dt$  and not yet deposited back on the film at a position  $z$  downstream, is given by  $\pi D_i R_l C_F(t) E(z - t) dt$ . The tracer balance can be written as

$$W_F C_F(z) + W_A C_A(z) = W_F C_{F,0} \quad [5]$$

with

$$C_A(z) = \frac{\pi D_i R_l}{W_A} \int_0^z E(z - t) C_F(t) dt. \quad [6]$$

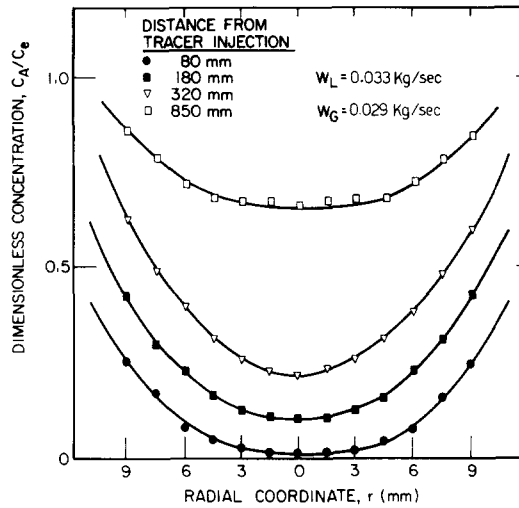


Figure 3. Tracer concentration profiles in the dispersed phase.

In terms of  $E(x)$ , the concentration of depositing droplets is equal to

$$C_A(z) = - \int_0^z C_F(t) \frac{dE(z-t)}{d(z-t)} dt. \quad [7]$$

Defining  $\alpha$  as

$$\alpha = \frac{\pi D_t R_t}{W_A} \quad [8]$$

we have from [6] and [7] that  $C_A(z) \equiv C_F(z)$  when

$$\int_0^z C_F(t) \left( \alpha E(z-t) + \frac{dE(z-t)}{d(z-t)} \right) dt \equiv 0. \quad [9]$$

As  $E(0) = 1$ , [9] is satisfied for

$$E(x) = \exp(-\alpha x). \quad [10]$$

The experimental measurements of droplet deposition reported by Farmer *et al.* (1970), by Cousins & Hewitt (1968) and the results of the present work suggest that the function  $E(x)$  evaluated for droplets which travel across the gas core by a diffusion like mechanism caused by a succession of interactions with gas phase eddies, is well approximated by [10]. When [10] is used for  $E(x)$ , [5] and [6] are the same as [1] and [2] with  $C_A = C_F$ . However, the use of the residence length distribution function allows a more general approach to the problem and an extended application for the tracer method.

The recent observations of droplet motion reported by James *et al.* (1980) show that larger droplets entrained by the gas travel across the gas core at about their initial velocity in a constant direction until they are deposited. This mechanism of deposition has been called direct impaction. Andreussi & Azzopardi (1981) showed that the effect of initial momentum on droplet motion becomes negligible at increasing values of the gas velocity and, at high gas velocity, the eddy diffusion mechanism of deposition prevails.

When both mechanisms are important, Andreussi & Azzopardi (1981) showed that the mean residence length of droplets depositing by direct impaction is lower than the residence length of droplets entrained by turbulent eddies and that the assumption of complete mixing of the dispersed phase is not valid.

In this case, [5], differentiated with respect to  $z$ , can be written in terms of the dimensionless concentration  $q(z)$  as

$$\frac{dq}{dz} + \pi D_t R_t \left( \frac{1}{W_A} + \frac{1}{W_F} \right) E + \frac{\pi D_t R_t}{W_F} \int_0^z E(z-t) \frac{dq(t)}{dt} dt = 0. \quad [11]$$

Close to the tracer injection point  $E \simeq q$  and the integral in [11] assumes a small value. Equation [11] can then be approximated by

$$\frac{dq}{dz} + \pi D_t R_t \left( \frac{1}{W_A} + \frac{1}{W_F} \right) q = 0 \quad [12]$$

from which it can be seen that the initial slope of  $\ln [q(z)]$  vs  $z$  has the same value as for the well-mixed case. Andreussi & Azzopardi (1981) showed that the initial slope of  $\ln [q(z)]$  is approximately constant for  $z < 20 D_t$ , then it decreases to a lower value. According to this analysis, the change in slope of  $\ln (q)$  vs  $z$  occurs at about the maximum residence length of droplets depositing by direct impaction. This length has been found to be independent of flow conditions.

The experimental evaluation of the initial and terminal slopes of  $\ln (q)$  can give a measure of the effect of the ejection velocity on droplet motion and valuable information about the residence length distribution function. Unfortunately, the evaluation of the terminal slope of  $\ln (q)$  can be subject to appreciable errors since  $q$  is proportional to the difference between the tracer concentration in the film and the equilibrium concentration and, at large distances from tracer injection,  $C_F$  approaches  $C_e$ . It is then possible that small absolute errors in the measurement of these concentrations will introduce large errors in the values of  $q$ .

It is concluded that reliable information about liquid transport through the gas core should be mainly derived from the initial slope of  $\ln (q)$  vs  $z$ . This slope is proportional to the rate of interchange of liquid between the film and the dispersed phase and it is not affected by deviations from the assumption of perfect mixing of the entrained droplets.

It has already been pointed out that samples withdrawn close to the tracer injection may be non-representative of the mean film concentration at that location. The further requirement that in the range of low gas velocities, only measurements for  $z < 20 D_t$  should be considered, limits the number of data which can be used for the evaluation of the slope of  $\ln (q)$  vs  $z$ . In the present work we could only use four measurements of concentration. These measurements were relative to axial distances from tracer injection of 0.16, 0.265, 0.33 and 0.47 m, for the first test section and 0.195, 0.295, 0.430 and 0.53 m for the test section shown in figure 1. In a single experiment the scatter of  $R_t$  values estimated from any couple of tracer concentrations was appreciable (up to  $\pm 20\%$  of the mean value). However, the reproducibility of measurements of  $R_t$  obtained from a least square fit of four measurements of concentration was  $\pm 10\%$  in the range of low gas velocities and less than 5% at high gas, low liquid flowrates.

#### CORRELATION OF EXPERIMENTAL MEASUREMENTS

##### (a) *Reduction of tracer concentration measurements*

Tracer concentration profiles along the tube for one liquid flowrate and six gas flows are

shown in figure 4. In this figure the dimensionless concentration  $p(z)$  is defined as

$$p(z) = \frac{C_F(z) - C_e}{C_e} \quad [13]$$

This concentration can be more easily evaluated from experimental measurements than  $q(z)$ , as it is not necessary to measure or calculate  $C_{F,0}$ . The equilibrium concentration,  $C_e$ , is related to  $C_{F,0}$  through the mass balance,

$$C_{F,0}W_F = C_e(W_F + W_A), \quad [14]$$

or

$$C_{F,0}(1-f) = C_e, \quad [15]$$

where  $f$  is the fraction of liquid entrained by the gas. Using the dimensionless concentration  $p$  and [15] for  $C_{F,0}$  [4] becomes

$$p(z) = \frac{f}{1-f} \exp \left\{ -\pi D_i R_i \left( \frac{1}{W_F} + \frac{1}{W_A} \right) z \right\}. \quad [16]$$

From this equation, it can be seen that the intercept of  $p(z)$  at  $z = 0$  is equal to  $f/(1-f)$ .

In the present work, the initial slope  $\beta$  and the intercept at  $z = 0$  of the straight lines fitting experimental values of  $\ln [p(z)]$  plotted vs  $z$  were determined by a linear least square method. From [16], we obtain that the slope is related to the rate of interchange by

$$\beta = \frac{\pi D_i R_i}{W_L} \frac{1}{f(1-f)}, \quad [17]$$

where  $W_L$  is the liquid flowrate. From [17] one can see that in the range of values of  $f$  encountered in this work the slope is not very sensitive to possible errors in the evaluation or measurement of  $f$ . We then calculated  $R_i$  from the values of  $\beta$  and  $f$  derived from the least square fit of the experimental measurements of concentration.

The fraction of entrained liquid measured at the outlet from the apparatus has been found to be very close to the value determined from tracer experiments, although system-

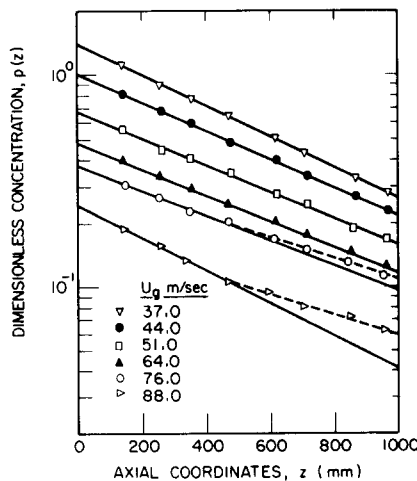


Figure 4. Dimensionless tracer concentration along the tube.



atically lower. This may be partly due to experimental errors, but also to non-fully developed flow conditions.

For a developing flow, the liquid and tracer balances in the wall layer can be written as

$$\frac{dW_F}{dz} = \pi D_i (R_D - R_E) \quad [18]$$

$$\frac{d(W_F C_F)}{dz} = \pi D_i [R_D C_A - R_E C_F] \quad [19]$$

where  $R_E$  and  $R_D$  are the rates of entrainment and deposition. These equations can be combined to give

$$W_F \frac{dC_F}{dz} = -\pi D_i R_D (C_F - C_A). \quad [20]$$

From this equation, it can be seen that for short distances after the tracer injection the variation along the tube of the tracer concentration in the wall layer is proportional to the rate of deposition. Therefore, in the case of developing flow conditions  $R_i$  in [4] should be assumed equal to  $R_D$ , while  $R_E$  can be evaluated by means of [18]. In the present experiments  $dW_F/dz$  appears to be small and we neglected the effect of developing flow conditions. However, it is useful to remark that measured values of  $R_i$  are a better approximation of  $R_D$  than  $R_E$ . By considering the tracer balance in the dispersed phase, it can be easily shown that  $R_E$  can be directly evaluated from tracer concentration measurements relative to the entrained liquid. The values of the rate of interchange, of the fractions of liquid entrainment measured at the outlet,  $f$ , and determined from tracer concentration measurements,  $f'$ , and of the pressure gradient for a number of flow regimes are reported in table 1.

#### (b) Rate of deposition

Droplet transport through the gas core is a complex phenomenon, in particular when the effect of the ejection velocity is appreciable. In order to deal with simple equations, it has generally been assumed that the rate of deposition is proportional through a deposition coefficient  $k$  to the concentration  $c_E$  of entrained droplets

$$R_D = k c_E \quad [21]$$

$c_E$  can be defined as

$$c_E = \rho_G \frac{W_A}{W_G}. \quad [22]$$

The terms  $R_D$  and  $k$  can also be expressed as functions of the rate of entrainment and the residence length distribution function. In a developing flow following the smooth injection of the liquid film, we have

$$R_D = -R_E \int_0^z \frac{dE(t)}{dt} dt = R_E [1 - E(z)] \quad [23]$$

$$k = \frac{u_G D_i}{4} \frac{1 - E(z)}{\int_0^z E(t) dt}. \quad [24]$$

Table 1. Experimental results

$W_L \times 10^3$ kg/sec	$W_G \times 10^3$ kg/sec	$f'$	$f$	$R_T \times 10^3$ kg/m <sup>2</sup> sec	dp/dz kN/m <sup>3</sup>
9.73	24.8	0.13	0.14	16.7	1.42
9.73	28.6	0.15	0.16	18.3	1.89
9.83	36.2	0.20	0.22	23.2	2.53
9.67	50.3	0.25	0.27	26.0	3.79
16.7	21.0	0.16	0.16	47.3	1.73
16.6	24.8	0.18	0.20	44.8	2.21
16.8	28.6	0.22	0.24	48.0	2.76
16.8	36.2	0.34	0.37	64.8	3.87
17.2	50.3	0.42	0.48	73.1	5.52
33.3	21.0	0.19	0.19	108.5	2.37
33.3	24.8	0.27	0.28	110.2	3.21
33.3	28.6	0.32	0.34	135.3	3.72
33.3	36.2	0.41	0.44	150.2	4.61
33.3	43.1	0.50	0.51	170.1	6.22
32.7	50.3	0.58	0.59	177.6	7.05
50.0	21.0	0.20	0.23	136.8	3.15
50.0	28.6	0.36	0.41	180.1	4.26
50.0	36.2	0.48	0.54	216.1	6.07
50.0	43.1	0.56	0.60	225.8	8.37
65.8	17.5	0.13	0.13	153.6	3.17
65.8	21.0	0.22	0.24	194.5	3.71
66.6	28.6	0.39	0.40	187.1	5.40
63.9	36.2	0.51	0.54	230.0	7.19
99.3	17.5	0.13	0.13	218.1	3.52
101.2	21.0	0.22	0.24	222.6	4.49
100.8	24.8	0.35	0.36	216.5	5.60
97.5	36.2	0.49	0.55	269.1	8.50
130.7	21.0	0.31	0.32	250.2	5.50
131.2	24.8	0.38	0.39	242.1	6.52
130.0	36.2	0.60	0.62	284.5	10.31
199.5	21.0	0.29	0.29	294.1	7.45
199.7	24.8	0.35	0.36	292.5	8.48
200.3	36.2	0.48	0.50	299.1	11.65

In these equations it has been assumed that  $R_E$  and  $E(t)$  do not depend on the position along the tube. From [24] it can be seen that  $k$  is, in general, a function of  $z$  and it is constant when  $E(t)$  can be approximated by an exponential decay function. In a fully developed flow  $k$  tends to a constant value, which can be obtained from [24] by letting  $z \rightarrow \infty$

$$k = \frac{u_G D_t}{4} \frac{1}{\int_0^\infty E(t) dt} \quad [25]$$

When the flow behavior of entrained droplets is described by an eddy diffusivity model,

[25] relates the one-parameter  $E(t)$  function, [10], to the deposition coefficient,  $k$ , which can be properly considered as a mass transfer coefficient. At low gas velocity this model is not adequate and [21] cannot be used with confidence for the analysis of developing flow conditions, or transient heat or mass transfer in a developed annular flow. For these reasons, we calculated  $k$  from measurements of  $R_l$  and  $c_E$  for all flow conditions, but we used an interpretation based on eddy diffusivity model only for the high gas velocity measurements.

In figure 5 we represent  $k$  as a function of  $u_G$  for all flow conditions. From this figure it can be noticed that—at low liquid flowrates  $k$  is only a function of  $u_G$ —at low gas velocities and intermediate liquid flows  $k$  is a decreasing function of  $u_G$ , as already noticed by Jagota *et al.* (1973)—at high liquid flows  $k$  is about constant.

According to the analysis of droplet motion developed by Andreussi & Azzopardi (1981), the effect of the ejection velocity for the present experimental conditions becomes negligible at  $u_G > 40$  m/sec. Present experiments confirm that above this velocity dimensionless tracer concentration profiles are well correlated by an exponential decay law, as also shown in figure 4. Furthermore, from figure 5 it can be noticed that for  $u_G > 40$  m/sec,  $k$  becomes a more regular, increasing function of  $u_G$  for all liquid flowrates, as expected for a mass transfer coefficient.

At lower gas velocity direct impaction seems to provide a more rapid mechanism of deposition than turbulent diffusion. However, the effect of the initial velocity decreases with increasing gas velocity and this can justify the presence of a minimum in the plot of  $k$  vs  $u_G$  for  $W_L = 0.066$  kg/sec and 0.10 kg/sec.

As suggested by the experiments reported by Farmer *et al.* (1970)  $k$  appears to be a function of droplet size at any gas velocity and the dependence of  $k$  on the liquid flowrate can be related to droplet size. In particular, Azzopardi *et al.* (1979) noticed that at low gas velocities, the mean droplet diameter is a decreasing function of the liquid flow. They also found that at high droplet concentrations, coalescence could cause the droplet size to increase with the liquid flowrate. Droplet coalescence can, as well, be an important mechanism of momentum transfer in random directions and this can possibly explain the low values of  $k$  in the range of high liquid flowrates.

At high gas velocity, droplet transport through the gas core appears to be mainly related to the interactions between droplets and turbulent eddies. The deposition coefficient is an increasing function of the gas velocity, but also in this case the experimental measurements of  $k$  are appreciably affected by the liquid flowrate. In figure 6 we plot the ratio between the deposition coefficient and the gas friction velocity,  $u^*$ , as a function of

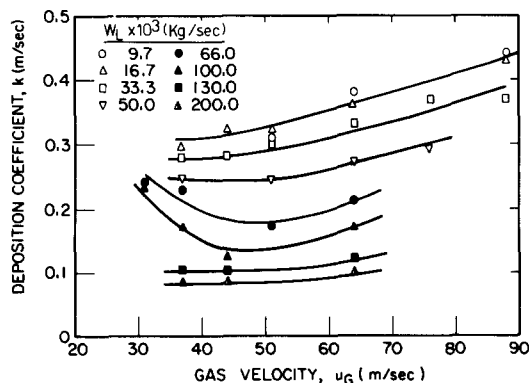


Figure 5. Deposition coefficient as a function of gas velocity.

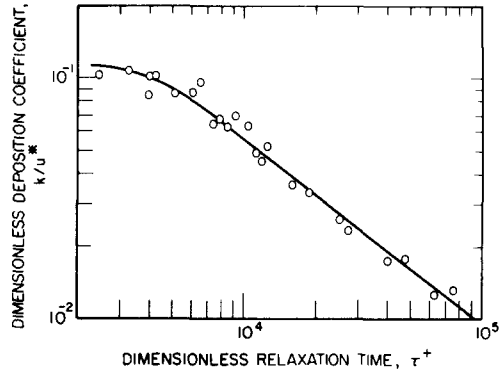


Figure 6. Effect of droplet concentration on deposition coefficient.

the concentration of entrained droplets. The gas friction velocity has been evaluated as  $u^* = \sqrt{\tau_i/\rho_G}$ , with  $\tau_i \simeq (dP/dx) D_i/4$ . As can be seen, this figure suggests that when the eddy diffusivity mechanism controls droplet transport, the dimensionless coefficient  $k/u^*$  is only a function of the liquid flowrate through the concentration of the entrained liquid. As shown in this figure, the empirical equation

$$\frac{k}{u^*} = \frac{0.115}{1 + 2.3 c_E/\rho_G} \quad [26]$$

gives a satisfactory fit to data for  $u_G > 40$  m/sec.

The dependence of  $k$  on droplet concentration was first observed by Namie & Ueda (1972) who attributed it to the effect of concentration on the state of turbulence in the gas core. As already mentioned, droplet coalescence can also explain the correlation between the deposition coefficient and the droplet concentration. In fact, at high  $u_G$  Azzopardi *et al.* (1979) found that the mean droplet size was a function of  $u_G$  and  $c_E$ . They correlated the experimental measurements of the Sauter mean diameter  $D_{32}$  by the equation

$$\frac{D_{32}}{D_t} = 1.91 \frac{\text{Re}_G^{0.1}}{\text{We}^{0.6}} \left( \frac{\rho_G}{\rho_L} \right)^{0.6} + 0.4 \frac{c_E}{\rho_G} \quad [27]$$

where the Weber number  $We$  is defined as

$$\text{We} = \frac{\rho_G u_G^2 D_t}{\sigma} \quad [28]$$

Equation [27] also gave a satisfactory fit to the measurements of droplet size obtained in this laboratory by Andreussi *et al.* (1978).

The experimental results reported by Farmer *et al.* (1970) show that the deposition coefficient varies inversely with particle diameter. This behavior is also predicted by the theoretical model developed by Hutchinson *et al.* (1971). More recent experimental and theoretical work on this subject (Liu & Agarwal 1974; Reeks & Skyrme 1976; McCoy & Hanratty 1977) considered the correlation between the dimensionless deposition coefficient,  $k/u^*$ , and the dimensionless relaxation time,  $\tau^+$ , defined as

$$\tau^+ = \frac{D_p^2 u^{*2} \rho_G \rho_L}{18 \mu_G^2} \quad [29]$$

Also in this case, it has been found that in the range of size encountered in annular flows,  $k$  decreases with particle size.

In order to calculate  $\tau^+$  and compare present measurements with available data, it is necessary to determine the droplet size for which the dimensionless relaxation time should be evaluated. If  $k$  is a function of particle size, the size distribution of droplets detaching from the wall layer or depositing back is different from the distribution which can actually be observed at any cross section. In a fully developed flow it can be assumed that the size spectrum of the deposition flow is equal to the spectrum of the entrainment flow. Let  $(dv/dD_p)_I dD_p$  be the volume fraction of the interchange flow due to droplets with diameter between  $D_p$  and  $D_p + dD_p$ . Let also  $(dv/dD_p)_A$  be the volume distribution relative to the droplets encountered at any cross section in a fully developed flow. These volume distributions are related by the equation

$$R_I \left( \frac{dv}{dD_p} \right)_I = k'(D_p) \frac{W_A}{W_G} \rho_G \left( \frac{dv}{dD_p} \right)_A \quad [30]$$

where  $k'(D_p)$  is the mass transfer coefficient which relates the deposition of droplets with diameter between  $D_p$  and  $D_p + dD_p$  to their concentration in the gas phase. From [30] and the definition of the average deposition coefficient,  $k$ , [21] it follows that

$$k = \int_0^\infty k'(D_p) \left( \frac{dv}{dD_p} \right)_A dD_p. \quad [31]$$

If we assume  $k(D_p) = a/D_p$  as suggested by the experiments reported by Farmer *et al.* (1970) and we use for  $(dv/dD_p)_A$  the volume distribution determined by Azzopardi *et al.* (1979), which can be written as

$$(v)_A = \exp \left\{ - \left( \frac{D_p}{2D_{32}} \right)^2 \right\}, \quad [32]$$

we obtain from [31]

$$k = \frac{\sqrt{\pi}}{2} \frac{a}{D_{32}}. \quad [33]$$

It follows that the characteristic size which should be adopted for the correlation of experimental measurements of  $k$  is

$$\bar{D}_p = \frac{2D_{32}}{\sqrt{\pi}}. \quad [34]$$

In the present work, as the measurements of droplet size reported by Andreussi *et al.* (1978) do not cover the complete range of flow conditions, we used [27] for predicting  $D_{32}$ . In figure 7 we plotted  $k/u^*$  as a function of  $\tau^+$ . As can be seen, the scatter of experimental data is appreciable; nevertheless, the effect of droplet concentration on the deposition coefficient seems to be explained.

The main features of the correlation between  $k/u^*$  and  $\tau^+$  shown in figure 7 agree fairly well with available data relative to deposition on a smooth wall. In particular, Liu & Agarwal (1974) found that for  $10 < \tau^+ < 10^3$ ,  $k/u^*$  was about constant and equal to 0.11. The deposition measurements of Cousins & Hewitt (1968) as analyzed by McCoy & Hanratty (1977) and by Andreussi & Azzopardi (1981) show that  $k/u^*$  ranges from 0.095

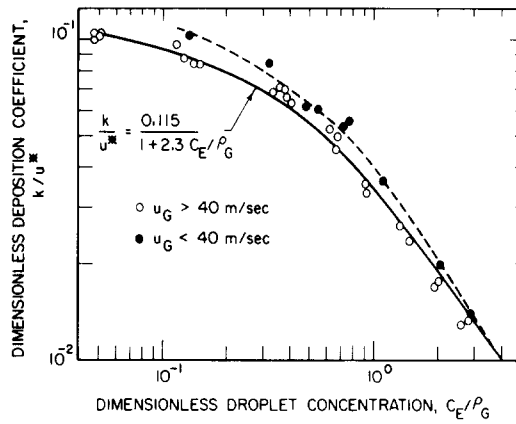


Figure 7. Correlation of deposition coefficient by dimensionless relaxation time.

to 0.06 for  $10^3 < \tau^+ < 10^4$ . A more detailed comparison of present results with experimental measurements of deposition on a smooth wall is questionable, because wall conditions may have some influence on the rate of deposition. The comparison is also difficult because in most cases mean droplet size and/or pressure drops have not been measured and the extension of available correlations for predicting these parameters does not seem to be safe. In particular, [27] has been developed from data obtained under similar conditions and requires a careful check on the effects of tube diameter and fluid properties.

(c) *Rate of entrainment*

Taylor (1940) and more recently Woodmansee & Hanratty (1969), have suggested that droplet entrainment by the gas is caused by pressure variations in the gas flow over atomizing wavelets which overcome the restoring force due to surface tension. According to the theory developed by Taylor (1940) the dimensionless rate of entrainment  $R_E$  defined as

$$R_E^* = \frac{R_E}{\rho_G^{1/2} \rho_L^{1/2} u_G} \tag{35}$$

assumes a constant value for the atomization of thick layers of low viscosity liquids.

Tatterson (1975) and Tatterson *et al.* (1977) proposed that for thin liquid films the wavelength of atomizing waves scales as the mean film thickness. In this case, the dimensionless rate of entrainment becomes a function of the Weber number,  $We'$ , defined as

$$We' = \rho_G \frac{u_G^2}{\sigma} m_F \tag{36}$$

where  $m_f$  is the mean film thickness. Tatterson *et al.* (1977) also suggested that for a better representation of the velocity field close to atomizing wavelets, the gas phase friction velocity,  $u^*$ , should replace  $u_G$  in [36]

Andreussi & Zanelli (1979) correlated their measurements of the entrained liquid flowrate by means of the fully developed flow relation

$$R_E = R_D. \tag{37}$$

They assumed  $k/u^* = \text{constant}$  and found that  $R_E^*$  was approximately proportional to the Weber number based on the gas friction velocity.

The equations proposed by Tatterson (1975) and by Andreussi & Zanelli (1979) are

complicated as it is required to predict or to measure the gas friction velocity and the mean film thickness. Besides, the choice of the characteristic dimension and velocity is questionable and requires further investigations. For practical purposes it can then be more convenient to represent  $R_E$  as a function of dimensionless groups which can be more easily evaluated than the Weber number relative to atomizing wavelets. Following this approach, Dallman & Hanratty (1979) proposed the following empirical correlation for  $R_E$

$$R_E = m((\Gamma - \Gamma')u_G^2 \rho_G^{1/2} \rho_L^{1/2}) \quad [38]$$

where  $m$  is an empirical coefficient,  $\Gamma$  is the liquid flowrate per unit perimeter, and  $\Gamma'$  is the value of  $\Gamma$  at the onset of entrainment. These authors found that for upward flow of air and water [38] with  $m = 6.7 \times 10^6 \text{ sec}^2/\text{kg}$  and  $\Gamma' = 0.046 \text{ kg/m sec}$  correlated the measurements of the initial rate of atomization reported by Cousins *et al.* (1965) and by Hinkle (1967) and also the fully developed rate of atomization estimated from deposition measurements of Cousins & Hewitt (1968).

As shown in figure 8, [38] also correlates present measurements of  $R_E$  in the range of low liquid flowrates. At high values of  $\Gamma$ , figure 8 suggests that  $R_E$  tends to become a linear function of the gas velocity. The same behavior is shown by the measurements of Hinkle (1967) analyzed by Dallman & Hanratty (1979). The value of  $m$  which gives the best fit to present measurements is equal to the value calculated by Dallman & Hanratty (1979).  $\Gamma'$  assumes the value determined by Andreussi (1980) for the same flow system used in the present work ( $\Gamma' = 0.08 \text{ kg/m sec}$ ).

This result is rather unexpected, considering the important differences in tube size, gas density, and flow orientation between the various groups of data. We then recommend further experimental and theoretical work before [38] can be used with confidence. It is also useful to notice that as the film thickness is roughly proportional to the ratio between liquid film and gas flowrates, the empirical correlation developed by Dallman & Hanratty (1979) is quite similar to the equations proposed for  $R_E$  by Tatterson (1975) and Andreussi & Zanelli (1979) on the basis of a physical model of the process.

## CONCLUSIONS

The tracer method can be very effective for determining the rate of interchange in a fully developed annular flow provided that the experiments are properly designed. This technique can eventually be adopted also for determining the rates of deposition or entrainment under

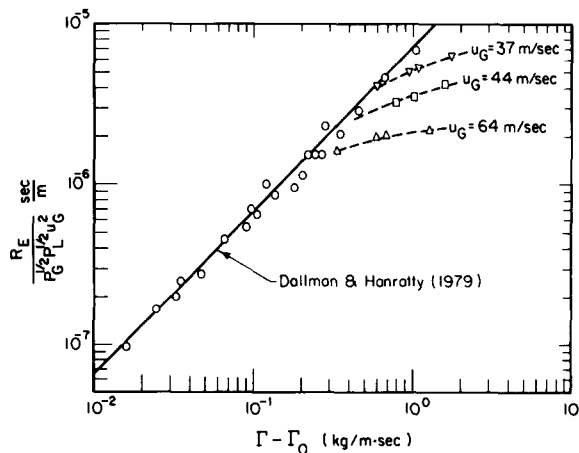


Figure 8. Correlation of the rate of entrainment as a function of the liquid flowrate per unit perimeter.

developing flow conditions. The rate of interchange is proportional to the initial slope of tracer concentration measurements plotted as  $\ln q$  vs  $z$ . In order to calculate this slope, only data relative to axial distances  $z < 20 D_i$  should be used.

The experimental results relative to the measurement of the deposition coefficient show that  $k$  is a function of the gas friction velocity and the droplet concentration. The dependence on  $c_E$  appears to be due to the effect of the concentration of the entrained liquid on the mean droplet size.

The rates of entrainment measured in the present work can be fairly well correlated by the empirical equation proposed by Dallman & Hanratty (1979). According to this equation,  $R_E$  is proportional to the liquid film flowrate and to the gas velocity squared. At high liquid flows,  $R_E$  appears to be proportional to the gas velocity.

The correlations proposed for the deposition coefficient [26] and for the rate of entrainment [38] can be combined to predict the entrained liquid flowrate. The resulting equation appears to be, at this stage, purely an empirical fit of the present measurements.

*Acknowledgements*—This work was supported by “Comitato Nazionale delle Ricerche”. Italy.

The experiments were carried out with the assistance of Mr. G. Sabatini, Mr. L. Guarracino and Mr. L. Tognotti.

The author wishes to thank Dr. B. J. Azzopardi, Dr. T. J. Hanratty, Dr. G. F. Hewitt, and Dr. S. Zanelli for useful discussions and advice during the course of the research reported here.

#### REFERENCES

- ANDREUSSI, P. & ZANELLI, S. 1976 Liquid phase mass transfer in annular two-phase flow. *Ing. Chim. Ital.* **12**, 132–136.
- ANDREUSSI, P., ROMANO, G. & ZANELLI, S. 1978 Drop size distribution in annular mist flows. Proc. from the First International Conference on Liquid Atomization and Spray Systems, Tokyo, August.
- ANDREUSSI, P. & ZANELLI, S. 1979 Downward annular and annular-mist flow of air–water mixtures. *Momentum Heat and Mass Transfer in Chemical, Process and Energy Engineering Systems* (Edited by F. DURST, G. V. TSIKLARI and N. AFGAN). Hemisphere McGraw-Hill, Washington, DC.
- ANDREUSSI, P. 1980 The onset of droplet entrainment in annular downward flows. *Can. J. Chem. Engng* **58**, 267–270.
- ANDREUSSI, P. & AZZOPARDI, B. J. 1981 Droplet deposition and interchange in annular gas–liquid flow. Atomic Energy Research Establishment, R10147, Harwell, England.
- AZZOPARDI, B. J., FREEMAN, G. & KING, D. J. 1979 Drop sizes and deposition in annular two phase flow. Atomic Energy Research Establishment, R9634, Harwell, England.
- COUSINS, L. B., DENTON, W. H. & HEWITT, C. F. 1965 Liquid mass transfer in annular two-phase flow. Atomic Energy Research Establishment, R6426, Harwell, England.
- COUSINS, L. B. & HEWITT, C. F. 1968 Liquid phase mass transfer in annular two-phase flow: droplet deposition and liquid entrainment. Atomic Energy Research Establishment, R5657, Harwell, England.
- DALLMAN, J. C. & HANRATTY, T. J. 1979 Interpretation of entrainment measurements in annular gas–liquid flows. *Momentum Heat and Mass Transfer in Chemical Process and Energy Engineering Systems* (Edited by F. DURST, G. V. TSIKLARI and N. AFGAN). Hemisphere McGraw-Hill, Washington, DC.
- FARMER, R., GRIFFITH, P. & ROHSENOW, W. M. 1970 Liquid droplet deposition in two-phase flow. *J. Heat Transfer* **92**, 587–597.
- HEWITT, C. F. 1979 Liquid mass transport in annular two-phase flow. *Momentum Heat and*



- Mass Transfer in Chemical Process and Energy Engineering Systems* (Edited by F. DURST, C. G. TSIKLARI and N. AFGAN), Hemisphere McGraw-Hill, Washington, DC.
- HINKLE, W. D. 1967 A study of liquid mass transport in annular air-water flow. Ph.D. Thesis, Massachusetts Institute of Technology, Cambridge, Mass.
- HUTCHINSON, P., HEWITT, G. F. & DUKLER, A. E. 1971 Deposition of liquid or solid dispersions from turbulent gas streams: a stochastic model. *Chem. Engng Sci.* **26**, 419-439.
- JAGOTA, A. K., RHODES, E. & SCOTT, D. S. 1973 Tracer measurements in two-phase annular flow to obtain interchange and entrainment. *Can. J. Chem. Engng* **51**, 139-147.
- JAMES, P. W., HEWITT, G. F. & WHALLEY, P. B. 1980 Droplet motion in two-phase flow. Atomic Energy Research Establishment, R9711, Harwell, U.K.
- LIU, B. Y. H. & AGARWAL, J. K. 1974 Experimental observations of aerosol deposition in turbulent flow. *J. Aerosol Sci.* **5**, 145-155.
- MCCOY, D. D. & HANRATTY, T. J. 1977 Rate deposition of droplets in annular two-phase flow. *Int. J. Multiphase Flow* **3**, 319-331.
- NAMIE, S. & UEDA, T. 1972 Droplet transfer in two-phase annular mist flow. *Bull. J.S.M.E.* **15**, 1568-1580.
- QUANDT, E. R. 1965 Measurement of some basic parameters in two-phase annular flow. *AIChE J.* **11**, 311-318.
- REEKS, M. W. & SKYRME, G. 1976 The dependence of particle deposition velocity on particle inertia in turbulent pipe flow. *J. Aerosol Sci.* **7**, 485-495.
- TATTERSON, D. F. 1975 Rate of atomization and drop size in annular two-phase flow. Ph.D. Thesis, University of Illinois, Urbana, IL.
- TATTERSON, D. F., DALLMAN, J. C. & HANRATTY, T. J. 1977 Drop size in annular gas liquid flows. *AIChE J.* **23**, 68-76.
- TAYLOR, G. I. 1940 Generation of ripples by wind blowing over a viscous fluid. Chemical Defense Research Department, Ministry of Supply, U.K., reprinted in *The Scientific Papers of G. I. Taylor* (Edited by G. K. BATCHELOR). University Press, Cambridge 1963.
- WOODMANSEE, D. E. & HANRATTY, T. J. 1969 Mechanism for the removal of droplets from a liquid surface by a parallel air flow. *Chem. Engng Sci.* **24**, 299-307.