Int. J. Multiphase Flow, Vol. 3, pp. 319–331. Pergamon/Elsevier, 1977. Printed in Great Britain.

RATE OF DEPOSITION OF DROPLETS IN ANNULAR TWO-PHASE FLOW

D. D. McCoy and T. J. Hanratty

Department of Chemical Engineering, University of Illinois at Urbana, Urbana, IL 61801, U.S.A.

(*Received* 10 *September* 1975)

Abstract—Droplets in a turbulent flow will deposit on surrounding surfaces. This paper critically examines measurements of the deposition rate with the goal of outlining our present ability to predict this quantity for annular two-phase flows.

1. INTRODUCTION

Liquid or solid particles carried by a turbulent air stream will deposit on bounding walls. Considerable attention has been given to the prediction of the mass rate of deposition per unit area, N_o, because of its importance in a number of engineering problems. This paper critically examines experimental measurements of N_o with the goal of outlining our present ability to predict this quantity for annular two-phase flows. The viewpoint that is taken has been greatly influenced by studies recently completed in our laboratory by Tatterson (1975) on drop size and by McCoy (1975) on deposition rates in horizontal flows. The results obtained by McCoy (1975) are considered in detail in this paper.

In the annular regime a liquid layer flows along the walls and a high velocity gas stream flows concurrently. The liquid layer has an agitated wavy surface and it can be entrained into the gas. This entrained liquid is carried by the gas as droplets with a large range of diameters. These droplets deposit on the wall layer so that the amount of entrained liquid depends on the relative rates of deposition and entrainment. Consequently, a knowledge of N_o is needed to predict the amount of entrained liquid and to develop improved methods to analyze the performance of two-phase systems (Hewitt & Hall-Taylor 1970).

Two types of equations have been used to correlate data on N_o. For one of these the mass flow rate of the dispersed droplets, W_{LE}, is the driving force

$$N_o = k_D' \frac{W_{LE}}{P} \quad [1]$$

where P is the perimeter equal to πd_t for a pipe. The rate constant k_D' then has an interpretation as the percent of the liquid deposited in a unit length of pipe. The more common procedure is to use the concentration of the entrained liquid in the units of mass/volume as the driving force

$$N_0 = k_D c_D. \quad [2]$$

The rate equation used in this study is

$$N_o = k_D \frac{\rho_G W_{LE}}{W_G} \quad [3]$$

where W_G is the mass flow rate of the gas and ρ_G is the density of the gas. It is equivalent to [2] if the droplets are moving at the same velocity as the gas and if $(W_G/\rho_G) \gg (W_{LE}/\rho_L)$, where ρ_L is the density of the liquid. It can be seen that the rate constant k_D is related to that appearing in [1] by the equation

$$k_D' = \frac{4k_D}{u_G d_t}, \quad [4]$$

where u_G is the bulk averaged gas velocity and d_t is the hydraulic diameter. The usefulness of rate equation [3] depends on the degree of independence of k_D from $(\rho_G W_{LE}/W_G) \approx c_D$. There is contradictory evidence in the literature regarding this point (Namie & Ueda 1972; Cousins & Hewitt 1968*a*) and it is a question which deserves clarification.

Three types of experimental investigations are pertinent to the estimation of k_D: the first of these involves the injection of particles or droplets into a gas stream and studying their deposition on the wall and include the data of Friedlanger & Johnstone (1957), Wells & Chamberlin (1967), Sehmel (1968), Schwediman & Postma (1961), Alexander & Coldren (1951), Farmer (1969), Forney & Spielman (1974) and Liu & Agarwal (1974). The range of particle diameters covered in these studies was 0.65–260 μm. Since, for annular two-phase flow most of the entrained liquid consists of droplets with diameters greater than 10 μm the work on larger sized particles by the last four of these investigators is of particular interest.

In the second type of study the liquid is introduced as an annulus through a porous section of the wall of a duct. The droplets are entrained in the air stream by the gas flowing over the liquid film on the wall. Cousins & Hewitt (1968*a*), Namie & Ueda (1972), Anderson & Russell (1970), and McCoy (1975) removed all of the wall layer or enough of it to stop atomization and determined the rate of deposition by measuring build up of the wall layer downstream of the suck off unit.

In the third, Quandt (1965), Cousins *et al.* (1965), and Jagota *et al.* (1973) determined deposition rates by injecting dye into the wall layer and measuring the change of the dye concentration because of the deposition of droplets. All of these studies covered about the same range of flow conditions and produced results consistent with one another. However, Jagota *et al.* (1973) seem to have given more attention to avoiding some of the errors associated with the use of this technique. Consequently, only their results will be considered.

Relations for k_D for annular flows have been suggested by Paleev & Filippovich (1966), Cousins & Hewitt (1968*a*), Jagota *et al.* (1973), Namie & Ueda (1972), Farmer (1969) and by Anderson & Russell (1970). No general agreement is noted. In addition to the usual problems associated with experimental accuracy and the limited range of variables covered in a single investigation, two other factors could be causing the apparently contradictory results. One of these has been the difficulty in using results obtained by injecting particles of controlled size into a gas stream because correlations for drop size in annular flows have been unavailable. The other, recently pointed out by McCoy (1975), is that results in horizontal flows could be quite different from those for vertical flows because of the influence of gravitational settling.

Consequently, in examining measurements of deposition rates we first consider vertical ducts in which particles of known size are injected into the air flow. The influence of gravity in such systems is examined by comparing results obtained in horizontal flows with those obtained in vertical flows. Results for vertical annular flows are compared with experiments using controlled particle size with the help of the recent correlation by Tatterson (1975) for drop size in annular flows. The influence of gravity on deposition rates in horizontal annular flows is considered by using a method of correlation found convenient in treating deposition data for injected particles.

2. CORRELATION OF DEPOSITION DATA

(a) *Vertical flows*

If one considers the fully developed turbulent flow of a gaseous suspension of spherical particles in a vertical pipe under conditions that entrance effects, electrical effects, and wall roughness are not important the following functionality can be written for the deposition rate constant, k_D, for dilute concentrations of particles:

$$k_D = f(\rho_P, \rho_G, u^*, d_P, \mu_G, d_t, D), \qquad [5]$$

where ρ_P is the density of the particle, d_P, the diameter of the particle, μ_G, the viscosity of the gas, d_t, the diameter of the pipe, D, the diffusivity due to Brownian motion (Davies 1966), and u^*, the friction velocity calculated using the Fanning friction factor.

The following dimensionless groups appear to be the most convenient for correlation data:

$$\frac{k_D}{u^*} = f\left(\tau^+, Sc, Re, \frac{\rho_P}{\rho_G}\right). \quad [6]$$

Here Sc is the Schmidt number, $\mu_G/\rho_G D$, Re, the Reynolds number, $d_t u_G \rho_G/\mu_G$, and τ^+, a particle relaxation time made dimensionless with respect to wall parameters, u^* and μ_G/ρ_G.

$$\tau^+ = \frac{d_P^2 u^{*2} \rho_G^2}{18 \mu_G^2} \frac{\rho_P}{\rho_G}. \quad [7]$$

A small particle that is experiencing a Stokesian resistance during motion through a fluid from an initial velocity u_{PO} will stop after a distance given as $S = (m_P u_{PO}/3\pi\mu_f d_P)$, where μ_f is the viscosity of the fluid and m_P is the mass of the particle, $(\pi d_P^3/6)\rho_P$. Since the relaxation time, τ, is defined as S/u_{PO}, it is seen that τ^+ can also be interpreted as a dimensionless stopping distance, S^+, where $u_{PO} = u^*$. In fact Friedlander & Johnstone (1957) in the analysis of their data used u_{PO} approximately equal to the turbulent velocity fluctuations outside the viscous wall region, i.e. $u_{PO} = 0.9u^*$. Their stopping distance characterizing the particle motion is thus equal to $0.9\tau^+$.

For particles in the submicron range τ^+ is small and the particles follow the streamlines of the fluid motion. The particles impinge on the wall by Brownian motion and the deposition constant is given by the relation defining mass-transfer of molecular species at large Schmidt numbers (Shaw 1975).

$$\frac{k_D}{u^*} = 0.0889\, Sc^{-0.704}. \quad [8]$$

For this limiting case only two dimensionless groups are needed.

Particles characterized by $\tau^+ > ca.\ 0.15$ do not follow the streamlines and impinge on the wall by an inertial mechanism. In this case the deposition constant is independent of the Brownian diffusion-coefficient and the Schmidt number is not an important dimensionless group. Starting with Friedlander & Johnstone (1957) a number of investigators (Owen 1960; Davies 1966; Sehmel 1968) have presented theoretical arguments which suggest that there is a range of τ^+ for which k_D/u^* is independent of Sc and depends primarily on τ^+.

No theoretical guidance is presently available to suggest the dominant dimensionless groups for $\tau^+ > ca. 20$. Consequently, we present measurements of k_D/u^* for very large particles as a function of τ^+ only as a matter of convenience.

(b) *Influence of gravity, wall roughness and electrical effects*

Factors not considered in the previous section are static electricity, wall roughness and gravity. A thesis by Montgomery (1970) and the papers by Wells & Chamberlin (1967), Lovett & Musgrove (1973) and Soo (1971) show that static electrical effects and wall roughness can cause large increases in the deposition constant of small diameter particles. However insufficient data are presently available to define these two effects quantitatively.

Gravity will be particularly important in horizontal system. The dimensionless group usually employed to characterize the influence of gravity is the ratio of the terminal settling velocity to the gas velocity, V_t/u_G.

3. MEASUREMENTS OF DEPOSITION RATES FOR VERTICAL FLOWS

(a) *Injected particles*

The measurements of the deposition constant for particles injected into vertical flows are summarized in table 1 and figure 1. All of these investigators used uniformly sized particles and directly measured the deposition rate to the wall.

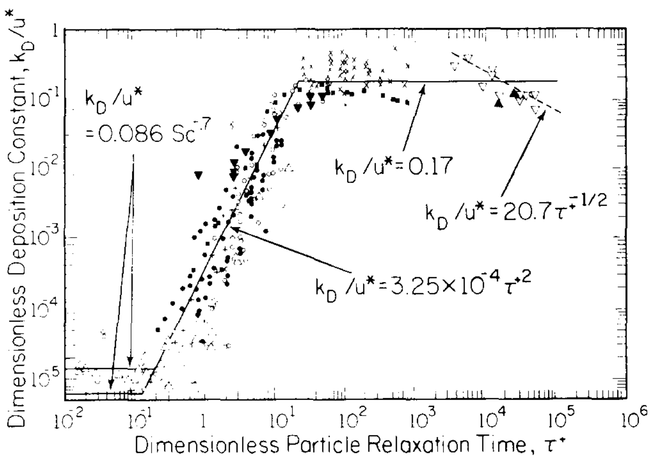

Figure 1. Summary of literature deposition data in vertical flow systems.

In figure 1 deposition data are plotted as k_D/u^* versus τ^+. According to [8] one would expect k_D/u^* to depend only on Sc or the particle diameter at very low τ^+. Consistent with this, it is found that k_D/u^* for a given particle size becomes independent of τ^+ as $\tau^+\to 0$. Data taken in the range where [8] is applicable were with particles with diameters from 0.65 to 2.1 μm or for Sc of 3.3×10^5 to 1.2×10^6.

From figure 1 we see that there is a range of τ^+, *ca.* $0.2<\tau^+<22.9$, for which k_D/u^* varies approx as τ^{+2}, as suggested Kneen & Straus (1969) and by Liu & Agarwal (1974). The equation used by Liu & Agarwal is nearly the same as the following equation that is the best average fit of all the data:

$$\frac{k_D}{u^*}=3.25\times10^{-4}\tau^{+2}. \qquad [9]$$

In the range where [9] is applicable there is a remarkably strong increase of k_D with air velocity and with particles diameter, $k_D\sim u^{*5}$ and $k_D\sim d_P{}^4$.

A detailed examination suggests that the neglect of the direct influence of ρ_P/ρ_G and of Re is not the major cause of the spread of the data. Consequently, more accurate measurements or a critical evaluation of presently available data would be needed to see the direct influence of these two dimensionless groups.

It is of interest to note that the analysis of Friedlander & Johnstone (1957) gives an equation of the same form as [9] for $\tau^+<$ *ca.* 1.0. According to their theory the exponent on τ^+ for small τ^+ is associated with the limiting behavior of the eddy diffusion coefficient, ϵ, at very small distances from the wall; i.e. the power on y^+ in the relation $(\epsilon/\nu_G)=by^{+n}$ for $y^+\to 0$. Here y^+ is the distance from the wall made dimensionless using the friction velocity, u^*, and the kinematic viscosity, ν_G.

At $\tau^+>$ *ca.* 22.9 the value of k_D/u^* appears to be relatively insensitive to the particle diameter or to the fluid velocity. In fact figure 1 suggests that the best average fit of all the data for $\tau^+>$ *ca.* 22.9 is

$$\frac{k_D}{u^*}=0.17. \qquad [10]$$

A number of theories, Friedlander & Johnstone (1957), Owen (1960), Davies (1966), Hutchinson *et al.* (1971) involving the use of an analogy between particle diffusion and turbulent momentum diffusion in the fluid seem consistent with this result in that they suggest that k_D/u^* is a weak function of Re and independent of particle diameter or particle density. However, none of these theories, with the exception of those of Hutchinson, Hewitt & Dukler (1971), and of Davies

Table 1. Work in vertical systems

Worker	Symbol	Flow direction	Particle size range	Particle material	ρ_P/ρ_G	Gas velocity range (m/sec)	Pipe diameter (m)
Farmer (1969)	▽	Downward	100–260 μm	Water	775	16.8–33.8	0.0127
			Vol. mean dia.				
Forney & Spielman (1974)	×	Downward	19.5 μm	Ragweed pollen	372	14.8–45.0	0.013–0.0365
			30.9 μm	Lycopodium spores	504	7.24–30.2	0.013–0.044
			32 μm	Polystyrene	852	11.8–23.6	0.0254
			48.5 μm	Pecan pollen	728	11.8–24.1	0.0178–0.0254
Friedlander	●	Upward	0.8–2.63 μm	Iron	6100	7.3–54.9	0.0054–0.0254
& Johnstone (1957)			Uniform dia.	Aluminum	2093	0.404–1.44	0.0138
Harwell (1968)	▲	Upward	70–204 μm	Water	775	23.8–41.5	0.00953
			Vol. median dia.				
Ilori (1971)	▼	Upward	6–9 μm	80% Methylene	1317	2.40–24.0	0.0298
			Uniform dia.	20% Uranine			
Liu & Agarwal (1974)	■	Downward	1.4–21 μm	Olive oil	713	11.0–55.0	0.0127
			Uniform dia.	Uranine			
Schwendiman	+	Upward	2–4 μm	Zinc sulfide	2744	2.16–17.7	0.0191
& Postma (1961)			Uniform dia.				
Sehmel (1968)	○	Upward	1–8 μm	Uranine	1163	5.09–36.6	0.0053–0.02926
			Uniform dia.	Methylene blue			
Wells & Chamberlin (1967)	△	Upward	0.65–2.1 μm	Tricresol	915	1.76–29.9	0.0127
			Uniform dia.	Phosphate			
Jagota *et al.* (1973)		Upward	—	Polystyrene	775	7.6–27.6	0.0127
				Water	775	11.8–30.5	0.0254

(1966) take into account the fact that as the particle size increases it will become less sensitive to turbulent velocity fluctuations in the fluid and that for infinitely large particles the mass-transfer–momentum-transfer analogy is not appropriate.

An examination of only the data of Farmer (1969), in fact, gives the more reasonable result that k_D varies inversely with the particle diameter for very large particles. Farmer examined too narrow a range of gas velocities to discover the exact influence of gas velocity, although his data show that k_D is either independent of or increases slightly with increasing u_G. The influence of particle diameter is more clearly indicated. We have chosen an equation of the form $k_D \sim d_P^{-1}$ as the best representation of his results. If it is assumed (perhaps incorrectly) that in the range of particle sizes covered by Farmer that k_D/u^* is primarily a function of τ^+, as has been found for smaller particles, then Farmer's data would be represented by the equation

$$\frac{k_D}{u^*} = 20.7\, \tau^{+[-(1/2)]}. \qquad [11]$$

This relation is indicated by the dashed line in figure 1. Consequently, his results suggests that a rather broad maximum in the deposition rate occurs at τ^+ approximately equal to 10^3.

It would appear that deposition rates of large diameter particles cannot be regarded as well defined. More measurements are clearly needed.

(b) *The study by Cousins & Hewitt of droplets originating from a wall layer*

A number of difficulties are encountered in the application of the deposition results just outlined to annular flow. Since the liquid droplets in annular flow originate from the wall layer, they could have an initial trajectory which dominates or greatly influences the motion of the droplets and they can have a distribution in the gas flow different from that for injected drops. The diameter of the droplets that occur in annular flows cannot be controlled independently. Waves on the liquid layer make the deposition surface a rough one and consequently could cause increased deposition rates over those encountered for smooth surfaces.

The most extensive set of measurements of deposition rates of droplets originating from a wall layer are those of Cousins & Hewitt (1968*a*) for vertical upflow of air and water in a 0.00953 m and in a 0.0318 m pipe. The liquid layer was removed by sucking it through a 0.0762 m length of porous wall. The mass flow rate of droplets at the beginning of the deposition test section, W_{LE_1}, was determined as the difference between the total liquid flow rate, W_L, and the flow rate of liquid from the suck off unit. The amount of liquid deposited over different pipe lengths was determined by sucking off the wall layer at positions from 0.152 to 1.98 m from the position at the inlet to the deposition test section where entrainment was stopped. The deposition constants determined from the first suck off unit were about 60% higher than the average of the other five. Cousins & Hewitt suggest that this is due to a more rapid deposition of the larger droplets. However, this seems at variance with the results of Farmer (1969) which indicate a decrease in the deposition rate with increasing particle size. Therefore, we have chosen to ignore the results obtained from the first suck off units. Averages of the values of the deposition constants measured by Cousins & Hewitt at the last five suck off units of their 0.00953 m pipe are summarized in table 2. The test section was located 2.19 m downstream where the flow was fully developed. Values of k_D are tabulated for different mass flow rates of gas, W_G, and of liquid, W_L. The flow rate of the wall layer, W_{LF_1}, and of the entrained liquid, W_{LE_1}, upstream of the suck off unit are also given. The ratio of the height of the wall layer to the pipe diameter upstream of the suck off unit was calculated from a correlation recently developed by Henstock (1975). It is noted that k_D shows only a narrow range of variation, but it does appear to increase with increasing gas velocity. We also present k_D in the dimensionless form of k_D/u_s^*, where u_s^* is the friction velocity calculated from the superficial gas velocity and the friction factor for a smooth pipe, $f = 0.0791\, Re^{-0.25}$. The

Table 2. Cousins and Hewitt's data for a 0.00953 m pipe

W_G	W_L	Inlet pressure	W_{LE_1}	W_{LF_1}	m/d_t	k_D	k_D/u_s^*
(kg/m)	(kg/m)	(Newtons/m^2 $\times 10^{-5}$)	(kg/m)	(kg/m)	—	(m/sec)	—
18.2	22.7	1.96	1.50	21.2	0.020	0.15	0.095
	34.1	2.02	2.55	31.5	0.024	0.18	0.115
	45.5	2.05	3.41	42.0	0.027	0.19	0.119
22.7	22.7	2.03	2.86	19.9	0.017	0.17	0.090
	34.1	2.10	4.55	29.5	0.020	0.19	0.103
	45.5	2.16	6.27	39.2	0.023	0.19	0.100
27.3	22.7	2.08	4.36	18.4	0.015	0.19	0.083
	34.1	2.15	7.14	27.0	0.018	0.19	0.086
	45.5	2.20	9.64	35.8	0.020	0.19	0.085
31.8	22.7	2.14	5.77	16.9	0.013	0.20	0.078
	34.1	2.20	10.55	23.5	0.015	0.22	0.087

The above deposition constants, k_D, were measured after a 2.19 m approach to equilibrium and averaged from deposition constants measured with deposition lengths of 0.305, 0.457, 0.610, 0.762 and 1.07 m.

Cousins and Hewitt's data for a 0.0318 m pipe

W_G	W_L	Inlet pressure	W_{LE_1}	W_{LF_1}	m/d_t	k_D	k_D/u_s^*
(kg/m)	(kg/m)	(Newtons/m^2)	(kg/m)	(kg/m)	—	(m/sec)	—
227	79.5	2.25	26.0	53.6	—	0.14	0.0689
318	72.7	2.40	28.8	44.0	—	0.18	0.0675

The deposition constants, k_D, were measured after a 9.75 m approach to equilibrium entrainment with a deposition length of 1.94 m.

W_{LE_1} is the mass flow rate of entrained droplets where the film is initially thinned.
W_{LF_1} is defined as $W_L - W_{LE_1}$.
u_s^* is the friction velocity based on the superficial gas velocity and a smooth tube friction factor f defined as $f = 0.0791/Re_G^{1/4}$.

data appear to be represented by either of the relations

$$k_D \simeq 0.18 \text{ m/sec} \quad [12]$$

or

$$\frac{k_D}{u_s^*} \simeq 0.095. \quad [13]$$

The runs by Cousins & Hewitt in a 0.0318 m pipe were less extensive than those in a 0.00953 m pipe. They were done with the initial suck off unit 9.75 m downstream of the entry and with deposition lengths of 0.343 m, 1.03 m, 1.94 m and 2.86 m. The deposition rates in these experiments were fast enough than in many of the runs the wall layer grew to such an extent that atomization occurred. This was the case for all of the data obtained at 2.86 m. In table 2 we summarized the results obtained by Cousins & Hewitt at the 1.94 m measuring unit under conditions that reentrainment was not occurring. These results for a 0.0318 m pipe again indicate that k_D increases with increasing gas velocity. They are represented by the equation

$$\frac{k_D}{u_s^*} = 0.068. \quad [14]$$

A comparison of [14] with [13] would indicate that k_D/u_s^* decreases with increasing pipe diameter. This is consistent with the theoretical result of Tatterson (1975) that larger droplets would be produced in a larger diameter pipe.

Cousins & Hewitt (1968*a*) used photographic techniques to determine the distribution of drop sizes that existed in their experiments in a 0.00953 m pipe. Consequently, it is possible to compare their measurements of deposition with measurements made with injected particles. We have chosen to characterize the size of the droplets by the volume median diameter, defined such that the droplets with diameters greater than this carry 50% of the volume. A value of $d_{v\mu} = 165\ \mu m$ was calculated from distributions measured at $W_G = 18.2$ kg/hr and a value of $d_{v\mu} \simeq 80\ \mu m$ for $W_G = 31.8$ kg/hr. Values of τ^+ calculated for these two gas flow rates are greater than 10^4 and consequently correspond to particle sizes much greater than those for which [9] is applicable. Using these values of τ^+ we have plotted in figure 1 the average values of k_D/u_s^* in table 2 for gas rates of 18.2 kg/hr and 31.8 kg/hr. The results of Cousins & Hewitt are found to agree approximately with those of Farmer. This is quite surprising considering the possible differences that could exist in the droplet distribution and in the droplet motion.

Consequently, we can expect that some guidance in correlating data for annular flows can be obtained from experiments using injected particles. For example, Tatterson (1975) has recently shown that for annular flows $(d_{v\mu}^2\rho_G u_s^{*2})/(d_t\sigma)$ is approximately constant where σ designates the surface tension. If we use a value of $(d_{v\mu}^2\rho_G u_s^{*2})/(d_t\sigma) = 2.56 \times 10^{-4}$ suggested by the drop size measurments of Cousins & Hewitt to eliminate particle diameter from [11] a relation for k_D/u_s^* is derived which is in reasonable agreement with the deposition rates measured by Cousins & Hewitt:

$$\frac{k_D}{u_S^*} \sim \left(\frac{\mu_g^2}{d_t\sigma\rho_P}\right)^{1/2}. \qquad [15]$$

Of course, the above relation must be regarded as speculative since [11] is based on too narrow a range of experimental results to be accepted as generally valid.

(c) *The study by Jagota, Rhodes & Scott of annular flows*

The measurements by Jagota *et al.* (1973) of deposition rates with dye tracers were done for fully developed up flow of air and water in 0.0254 m diameter pipe. The calculation of deposition constants for these experiments depended on the assumption that the dye is uniformly dispersed but at different concentrations in the wall layer and in the entrained liquid. The latter assumption cannot be correct (Cousins & Hewitt 1968*b*) and therefore the measurements are subject to some unknown error.

Values of k_D of 0.0125–0.0223 m/sec determined by Jagota *et al.* (1973) are of the same magnitude as those obtained by Cousins & Hewitt (1968*a*) in a 0.0318 m pipe. However, they show a tendency to increase in value with decreasing gas velocity, contrary to what would be expected from deposition studies of Farmer (1969) and of Cousins & Hewitt (1968*a*) and from the drop size correlation developed by Tatterson (1975). Consequently it is not clear whether the influence of gas velocity noted in these experiments is real or a consequence of experimental error.

Droplet deposition in the experiments of Jagota *et al.* was occurring under different conditions from those of Cousins & Hewitt. The liquid layer was not withdrawn from the wall so that the droplets were depositing on a highly agitated wavy surface rather than on a smooth liquid film. Consequently in analyzing their data the values of k_D were made dimensionless using both friction velocities calculated for a smooth surface, $u_s^* = u_G\sqrt{(f/2)}$ and for conditions that exist in annular flow, $u^* = u_G\sqrt{(f_i/2)}$ where u_G is the actual velocity in the gas core. The friction factor f_i for flow over a wavy surface was calculated using a relation recently developed by Henstock (1975). Values of k_D/u^* of 0.03–0.08 determined from the data of Jagota *et al.* are close to those obtained from the measurements of Cousins & Hewitt in a 0.0318 m pipe while the values of

k_D/u_s^* (0.08–0.30) are considerably larger. This would suggest that the influence of the waves on the deposition surface can be taken into account through their influences on the friction velocity.

4. MEASUREMENTS OF DEPOSITION RATES FOR HORIZONTAL FLOWS

(a) *Injected particles*

Experimental results on deposition rates determined by injecting particles into a horizontal flow has been obtained by two investigators, Sehmel (1973) and Alexander & Coldren (1951). The data obtained by Sehmel (1973) for very small particles clearly show the strong influence of gravity in that deposition rates to the floor are an order of magnitude larger than those to the roof.

The results of Alexander & Coldren (1951) were for much larger particles than studies by Sehmel (1973). They injected droplets by means of an atomizer at the inlet of a horizontal 0.0472 m diameter duct through which air was flowing, and measured droplet flux at different radial positions by withdrawing samples from the air stream. The total flow rate of droplets at any given cross section was obtained by integrating these profiles and the rate of deposition on the walls was calculated from the change in droplet flow rate. The Sauter mean diameter of the droplets injected into the duct was calculated to be 27 μm from the correlation developed by Nukiyama & Tanasawa (1938) for the atomizer used. In the early part of the duct the droplets redistributed themselves until an essentially flat concentration profile was obtained. Rate constants were calculated only from the data obtained in the region where the droplet profile is fully developed. A surprising aspect of this study is that rate constants in the undeveloped region were larger than for the developed region. Some possible explanations are that the droplets were initally given a radial trajectory because of the method of injection, that there was an elutriation of droplets, or that the particles had not fully accelerated to the free stream velocity (Farmer 1969).

It is noted that the rate constants determined from the data of Alexander & Coldren (1951) are in good agreement ($k_D/u_s^* = 0.17$) with eqn [10] for large particles in vertical flow. However these data are a perimeter average of the rate constant, since deposition rates to the top and bottom of the pipe were not measured separately. This masks any effect of gravity. However concentration profiles presented by them are symmetric. This suggests that gravity did not have a strong effect on their measured deposition rates.

The range of variables studies in horizontal flows is summarized in table 3. As in the case of vertical flows, we have chosen to represent the drop diameter by $d_{v\mu}$ when a spray was depositing on the wall because this would be characteristic of the droplets carrying most of the mass.

Since Alexander & Coldren (1951) reported a Sauter mean diameter without giving sufficient information to allow the calculation of the volume median diameter the Sauter mean diameter was used in subsequent calculations for this data set. However, the use of the Sauter mean diameter will result in a more conservative estimate of the quantities of interest in figure 2, k_{Dv}/V_t and k_D/V_t, as the Sauter mean diameter is less than the volume median diameter.

We have found it convenient to correlate the results of these studies by comparing them to the results obtained in vertical flows, as is done in figure 2. Here k_D/V_t is plotted against k_{DV}/V_t, where k_{DV} is the deposition constant for a vertical flow defined by [9] and [10] and V_t is the terminal velocity, calculated using $d_{V\mu}$ when a spray is being considered. For large values of k_{DV}/V_t the effect of gravity should be negligible and $(k_D/V_t) = (k_{DV}/V_t)$.

It is interesting to note that for values of $(k_{DV}/V_t) <$ *ca.* 0.1 the data of Sehmel (1973) indicate that $k_D \simeq V_t$ for deposition to a floor and that $k_D \to 0$ for deposition to a roof as suggested by Owen (1960). This would indicate that in cases where gravity dominates the particles deposit by a free fall mechanism.

The data of Sehmel (1973) and of Alexander & Coldren (1951) indicate that values of $(k_{DV}/V_t) >$ *ca.* 10.0 are needed for gravity to play a secondary role in determining deposition.

(b) *Droplets originating from a wall layer*

Investigations of the deposition of droplets that originated from a water layer flowing along

Table 3. Work done in horizontal systems

Worker	Symbol	Particle size range	Particle material	ρ_P/ρ_G	Gas velocity range (m/sec)	Pipe diameter (m)	k_D/u_s^*
Alexander & Coldren (1951)	×	27 μm Sauter mean diameter	Water	775	24.7–89.9	0.0472	0.13–0.244
Anderson & Russell (1970)		—	Water	775	29.9–44.2	0.0254	0.78–0.95†
Montgomery (1970)		0.44–2.16 μm Uniform diameter	Uranine Methylene blue	1162	6.4–45.4	0.152	—
Namie & Ueda (1972)	Δ	187–267 μm Est. vol. med. diam. 131–187 μm Sauter mean diameter	Water	775	34–63	0.06 × 0.01 m rect. channel	0.034–0.116†
Sehmel (1973)		0.1–28 μm Uniform diameter	Uranine	1162	2.2–13.4	0.60 × 0.60 m channel	
	O	Deposition floor					3.0×10^{-5}–0.18
	●	Deposition roof					
McCoy (1975)	Δ	530–1000 μm Vol. med. diameter 413–617 μm Sauter mean diameter	Water	775	24.4–48.8	0.305 × 0.0254 m channel	0.24–0.66†

u_s^* is the friction velocity based on a superficial gas velocity and a smooth tube a friction factor defined as $f = 0.0791/Re^{1/4}$.
†Deposition constant calculations are based on the assumption that all deposition is occurring on the bottom wall. Remaining deposition constants are calculated based on total surface area available for deposition.

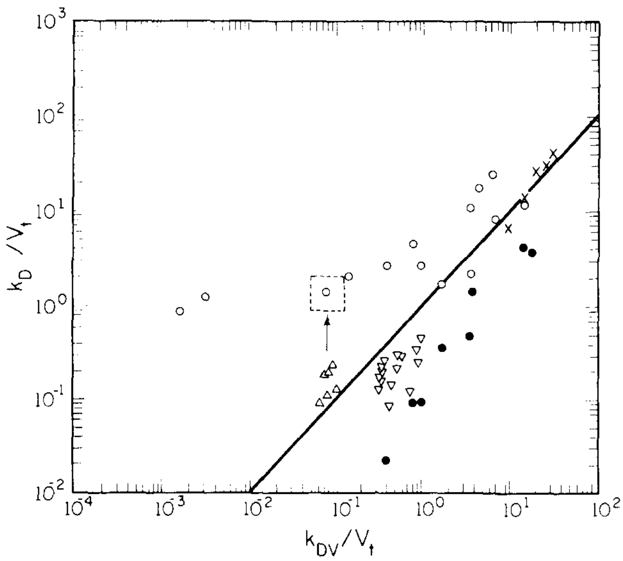

Figure 2. Deposition in horizontal systems.

the walls of a horizontal duct were carried out by Anderson & Russell (1970) in a 0.0254 m pipe, Namie & Ueda (1972) in a 0.06 × 0.01 m channel and by McCoy (1975) in a 0.305 × 0.0254 m channel. At the beginning of the deposition section the wall layers were thinned out sufficiently that no further atomization was occurring. Namie & Ueda (1972) and McCoy (1975) determined the change of the droplet concentration at different distances downstream in the deposition test section by directly sampling the gas flow. Anderson & Russell (1970) measured, at increasing distances downstream of the film thinner, the liquid film height at the bottom of the pipe and the film velocity with a salt tracer technique. This allowed the determination of a deposition constant k_D. However, small errors in the film flow rate can lead to significantly larger errors in the deposition constant because the film flow rates are large compared to the amount of liquid deposited.

The range of variables covered in these investigations is summarized in table 3. Namie & Ueda (1972) caught liquid droplets in an oil film and measured their diameters, They reported number mean diameters, d_{10}, in the range of 27–40 μm. The volume median diameter given in table 3 for their runs was calculated using the result that $(d_{V\mu}/d_{10}) \simeq 6.67$ obtained by Tatterson (1975) from his analysis of droplet distributions for annular two-phase flow. We note that the size of the droplets in the experiments of Namie & Ueda (1972) are five to seven times larger than those in the experiments of Alexander & Coldren (1951). The values of $d_{V\mu}$ reported in table 3 for the experiments of McCoy (1975) were measured using an electric probe developed by Tatterson (1975). They are about five times greater than those found in the experiments of Namie & Ueda (1972). Anderson & Russell (1970) did not measure drop sizes, but they should be of the same magnitude as experienced by McCoy (1975) since the two experiments were carried out under similar conditions.

Of considerable importance in interpreting the results of these experiments are the measurements of the concentration profiles of the droplets. As mentioned earlier, those measured by Alexander & Coldren (1951) were symmetric. Namie & Ueda (1972) found asymmetric droplet profiles with the maximum closer to the wall. In McCoy's experiments the profiles were even more asymmetric than was found by Namie & Ueda (1972). These measurements as well as visual observations suggest that in the experiments of McCoy (1975) almost all of the deposition was occurring on the bottom wall. This is consistent with the results of Anderson & Russell (1970) who measured the circumferential variation of interchange and found that nearly 90% of the interchange occurred on the bottom half of the pipe. Clearly gravity was playing an important role in the experiments of Namie & Ueda (1972), Anderson & Russell (1970), and McCoy (1975). Consequently, the deposition constants given in table 3 for the measurements of McCoy and Namie & Ueda were calculated by assuming all of the liquid was depositing on the bottom wall. These results suggest that the motion of very large droplets in annular flow can be described by a series of trajectories which originate at the liquid layer on the bottom wall. A few of these will intercept the top wall but most will end at the bottom wall some distance downstream.

The deposition experiments by Namie & Ueda (1972) and by McCoy (1975) in horizontal annular flow are compared with measurements of the deposition rates of injected droplets in figure 2, where V_t has been calculated using the volume median drop diameter. This comparison suggests that the experiments of McCoy (1975) were in a range of drop sizes where gravitational settling completely controlled the deposition rate and that the experiments of Namie & Ueda (1972) were in an intermediate range between a turbulence and a gravitational controlled deposition.

Yet, in apparent contradiction of the results with injected drops, McCoy's experiments yield values of $(k_D/V_t) \ll 1$. An explanation can be obtained by calculating the vertical velocity, V_w, with which a particle placed in the center of the channel with zero velocity would arrive at the wall. The only forces acting on the particle are assumed to be gravity and the hydrodynamic drag of the fluid. For small particles or for high channels the particles arrive at the wall with the terminal velocity, V_t. However, this type of calculation would indicate that the large sized droplets that existed in McCoy's experiments never would reach terminal velocity if they were settling on to the wall. In fact, if the measurements of McCoy (1975) are plotted as k_D/V_w, represented in figure 2 by the area enclosed by the dotted line, rather than as k_D/V_t the results appear in much closer agreement with data obtained by Sehmel (1973) using injected droplets.

5. SYNOPSIS

The sizes of the droplets carrying most of the mass of the dispersed liquid in annular two-phase flow are much larger than have been employed in experimental studies of particle deposition from a turbulent stream. They are characterized by stopping distances which are of the order of the dimensions of the duct. The only study of the deposition of particles injected in a vertical air flow which covers a range of particle size of interest in annular flows is that by Farmer

(1969). These indicate that the deposition constant varies inversely with particle diameter. More experiments are needed both to confirm the results of Farmer (1969) and to extend the range of variables covered. For the present, estimates of the deposition constant for vertical annular flows must be based on the experiments of Cousins & Hewitt (1968*a*) from which we get

$$\frac{k_D}{u_s^*} = 0.095 \qquad [16]$$

for a 0.00953 m pipe and

$$\frac{k_D}{u_s^*} = 0.068 \qquad [17]$$

for a 0.0318 m pipe. By using drop size measurements made by Cousins & Hewitt (1968*a*) for their experiments in a 0.00953 m pipe we find close agreement between the deposition rates measured by Cousins & Hewitt (1968*a*) and by Farmer (1969). This would suggest that further studies of the deposition of injected particles along with the development of methods to predict drop size in annular flow should lead to the development of a more reliable estimate of deposition constants for annular flow. This approach is illustrated by the development of the tentative equation

$$\frac{k_D}{u^*} \sim \left(\frac{\mu_G^2}{d_t \sigma \rho_P}\right)^{1/2} \qquad [18]$$

from the deposition studies of Farmer (1969) and the drop size correlation of Tatterson (1975).

Gravitational settling can have an important effect in horizontal annular flows. Consequently, the mechanism of deposition can be different from that for vertical flows. Correlations obtained from experiments with vertical flows can be used to estimate deposition rates for horizontal flows only if $(k_{DV}/V_t) > ca.$ 10.0. For $(k_{DV}/V_t) < ca.$ 0.1 gravitational settling is controlling. For small particles $(k_{DV}/V_t) \sim 1$ when $(k_{DV}/V_t) < ca.$ 0.1. However for conditions that exist in annular flows the particle sizes are large enough that settling particles might not have sufficient time to reach the terminal velocity and consequently measured values of k_D can be much smaller than V_t when gravitational settling is controlling.

Acknowledgement—This work has been supported by the National Science Foundation under grant NSF ENG 71-02367 and by the American Institute of Chemical Engineers through its Design Institute for Multiphase Processing.

REFERENCES

ALEXANDER, L. G. & COLDREN, C. L. 1951 Droplet transfer from suspending air to duct walls. *Ind. Engng Chem.* **43**, 1325–1331.

ANDERSON, R. J. & RUSSELL, T. W. F. 1970 Film formation in two-phase annular flow. *A.I.Ch.E.Jl* **16**, 626–633.

COUSINS, L. B., DENTON, W. H. & HEWITT, G. F. 1965 Liquid mass transfer in annular two-phase flow. Proceedings of the Symposium on Two-Phase Flow, Exeter, June, Paper C4.

COUSINS, L. B. & HEWITT, G. F. 1968*a* Liquid phase mass transfer in annular two-phase flow: droplet deposition and liquid entrainment. AERE-R 5657.

COUSINS, L. B. & HEWITT, G. F. 1968*b* Liquid phase mass transfer in annular two-phase flow: radial liquid mixing. AERE-R 5693.

DAVIES, C. N. 1966 Deposition of aerosols from turbulent flow through pipes. *Proc. R. Soc. A.* **289**, 235–246.

FARMER, R. A. 1969 Liquid droplet trajectories in two-phase flow. Ph.D. thesis, Massachusetts Institute of Technology.

FORNEY, L. J. & SPIELMAN, L. A. 1974 Deposition of course aerosols from turbulent flow. *J. Aerosol Sci.* **5**, 257–271.

FRIEDLANDER, S. K. & JOHNSTONE, H. F. 1957 Deposition of suspended particles from turbulent gas streams. *Ind. Engng Chem.* **49**, 1151–1156.

HENSTOCK, W. H. 1976 The interfacial drag and the height of the wall layer in annular flows. *A.I.Ch.E.Jl* **22**, 990–1000.

HEWITT, G. F. & HALL-TAYLOR, N. S. 1970 *Annular Two-Phase Flow.* Pergamon Press, Oxford.

HUTCHINSON, P., HEWITT, G. F. & DUKLER, A. E. 1970 Deposition of liquid or solid dispersons from turbulent gas steams: a stochastic model. AERE-R 6637.

ILORI, T. A. 1971 Turbulent deposition of aerosol particles inside pipes. PhD. thesis, University of Minnesota.

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**, 139–148.

KNEEN, T. & STRAUSS, W. 1969 Deposition of dust from turbulent gas streams. *Atm. Env.* **3**, 55–67.

LIU, B. Y. H. & AGARWAL, J. K. 1974 Experimental observations of aerosol deposition in turbulent flow. *J. Aerosol Sci.* **5**, 145–155.

LOVETT, C. D. & MUSGROVE, P. J. 1973 An electrical method for measuring the particle deposition velocity from a turbulent gas flow. Proceedings from Pneumotransport 2, second international conference on the pneumatic transport of solids in pipes, Guildford, September, Paper D5.

MCCOY, D. D. 1975 Particle deposition in turbulent flow. M. S. thesis, University of Illinois.

MONTGOMERY, T. L. 1970 Aerosol deposition in a pipe with turbulent air flow. PhD. thesis, University of Pittsburgh.

NAMIE, S. & UEDA, T. 1972 Droplet transfer in two-phase annular mist flow. *Bull. J.S.M.E.* **15**, 1568.

NUKIYAMA, S. & TANASAWA, Y. 1938 An experiment on the atomization of liquid by means of an air stream. Report 2, *Trans. J.S.M.E.* **4**, 524–526.

OWEN, P. R. 1960 Dust deposition from a turbulent airstream. Int. J. Air–Water Poll. **3**, 3–25.

PALEEV, I. I. & FILIPPOVICH, B. S. 1966 Phenomena of liquid transfer in two-phase dispersed annular flow. *Int. J. Heat Mass Transfer* **9**, 1089.

QUANDT, E. R. 1965 Measurement of some basic parameters in two-phase annular flow. *A.I.Ch.E.Jl* **11**, 311.

SCHWENDIMAN, L. C. & POSTMA, A. K. 1961 Turbulent deposition in sampling lines. USAEC Rpt. HW-65309.

SEHMEL, G. A. 1968 Aerosol deposition from turbulent air streams in vertical conduits. AEC Res & Dev. Rpt. BNWL-579.

SEHMEL, G. A. 1973 Particle eddy diffusivities and deposition velocities for isothermal flow and smooth surfaces. *J. Aerosol Sci.* **4**, 125–138.

SHAW, D. A. 1976 Mechanism of turbulent mass transfer to a pipe wall at high Schmidt number, Ph.D. thesis, University of Illinois.

SOO, S. L. 1971 A review on electrical effects in pneumatic conveying. Proceedings of Pneumotransport 1, September, Paper R1.

TATTERSON, D. F. 1975 Rates of atomization and drop size in annular two-phase flow. Ph.D. thesis, University of Illinois.

WELLS, A. C. & CHAMBERLIN, A. C. 1967 Transport of small particles to vertical surfaces. *Br. J. Appl. Phys.* **18**, 1793–1799.