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DROPLET CONCENTRATION DISTRIBUTION IN HORIZONTAL ANNULAR FLOW

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1. INTRODUCTION

This presentation is based on an ongoing research effort, aimed at improving our understanding and modeling horizontal annular two-phase flow. Detailed data are presented on droplet concentration distribution in the pipe cross-section. A considerable amount of similar information is available for vertical annular flow. However, for horizontal pipes the recent work of Williams (1986) in a 4 inch pipe-loop appears to be the only other study where such detailed data have been obtained.

Local fluxes of droplets entrained in the gas core have been measured isokinetically, with a pitot-type sampling tube. By interpolating and integrating the data throughout the pipe cross-section, the mean liquid entrainment has been obtained over a fairly wide range of air and water flow rates. Under the same flow conditions, time records have been collected of the local film thickness around the pipe circumference, using "parallel wire" conductance probes. The data on liquid film thickness distribution, as well as other statistical characteristics of the gas/liquid interface, are reported elsewhere (Paras and Karabelas, 1990) and provide essential information for the modeling effort.

In this presentation the experimental procedures are outlined and the main characteristics of local and average fluxes are briefly discussed. Furthermore, a relatively simple model is proposed as a first step in the direction of predicting concentration distribution in the pipe cross-section and droplet deposition rates to the wall. Comparisons are finally made between model predictions and data from this investigation as well as from Williams' (1986) thesis.

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2. DROPLET ENTRAINMENT IN THE GAS

2.1 Flow Loop

The experiments are carried out in a recently constructed horizontal flow loop. Two-phase flow develops in a 16 m long straight section, of 50 mm ID. The mixing section for the two phases is a simple tee with the liquid phase introduced in the branch and the gas phase in the run. The test section is positioned about 300 diameters downstream of the mixing section of the two phases and the flow is considered to be fully developed at this location.

2.2 Experimental Techniques

One of the most reliable experimental methods to obtain the fraction of liquid dispersed in the gas core is to measure the droplet flux distribution in the pipe cross-section and to integrate it. This liquid fraction is usually referred to as entrainment. Entrained water is collected in a small plexiglas separator. Adjustment of sampling conditions is possible by alternately using two separators connected in parallel. The sampling set-up is shown in Figure 1. To obtain roughly isokinetic conditions, suction is applied to the top of the separators and the air flow from the sampling tube is measured with a rotameter. However, it is observed that exactly isokinetic conditions are not necessary, as is also pointed out by Romano et al (1978).

Measurements are made for horizontal annular flow, under various gas and liquid flow rates. An L-shaped sampling tube is inserted into the pipe to collect droplets entrained in the gas core. The tube has a 3.17 mm ID and a 0.22 mm wall thickness at the end facing the flow, and extends 10 cm upstream to minimize any local disturbances caused by the vertical tube section entering the pipe. A traversing mechanism is employed to move the pitot tube along the inside diameter of the pipe, whereas the circumferential orientation of the probe is adjusted by rotating the entire test section. With this arrangement it is possible to sample droplets within approximately 3 mm of the pipe wall.

Droplet fluxes are measured by collecting the liquid in a separator, for a specified time period. The flow from the pitot tube is switched from one of the separators to the other by simultaneously activating two electrically operated 3-way valves, i.e. one for the suction and one for the feed from the pitot tube. The

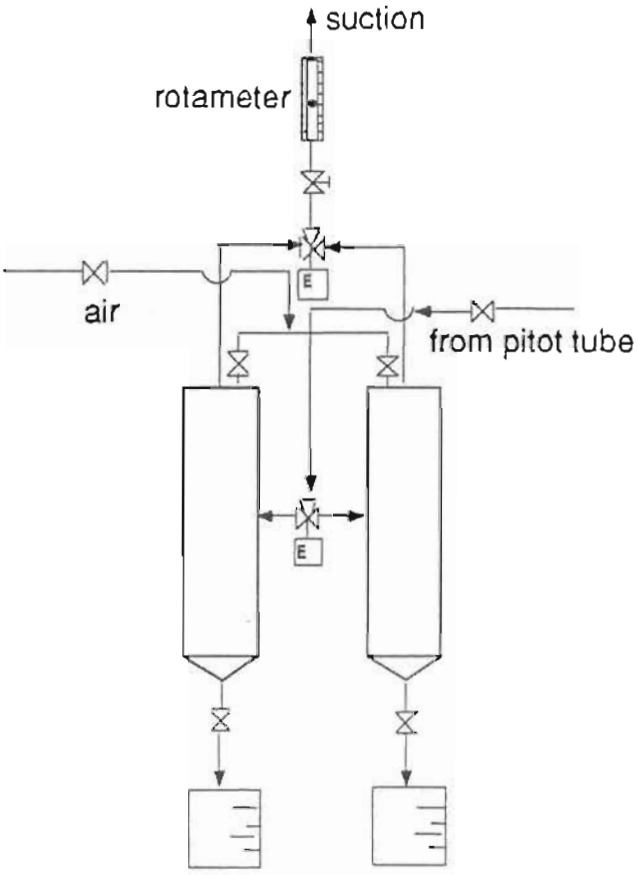

Figure 1. Set-up for entrainment measurements.

amount of liquid entrained in the gas is calculated by using the horizontal and vertical droplet flux profiles and by integrating throughout the pipe cross-section. Detailed measurements of film thickness at the pipe wall (Paras and Karabelas, 1990) are employed to exclude areas in the pipe cross-section occupied by the liquid film. Moreover, the statistical characteristics of the liquid film (mainly the rms values) are necessary in order to compute the equivalent wall roughness and to estimate the gas velocity distribution U(r).

2.3 Data on Droplet Fluxes and Entrainment

The ranges of superficial velocities, covered in the tests, were U_L= 2 to 20 cm/s for water and U_G= 31 to 66 m/s for air. Seventeen data sets were collected. Three droplet flux profiles were measured for each set of flow rates, i.e. a vertical, a horizontal and one at 45°. Measured vertical and horizontal droplet flux profiles are shown in Figure 2 . The horizontal profiles display an almost uniform distribution across the pipe diameter. The strong influence of gravity is evident in the vertical profiles.

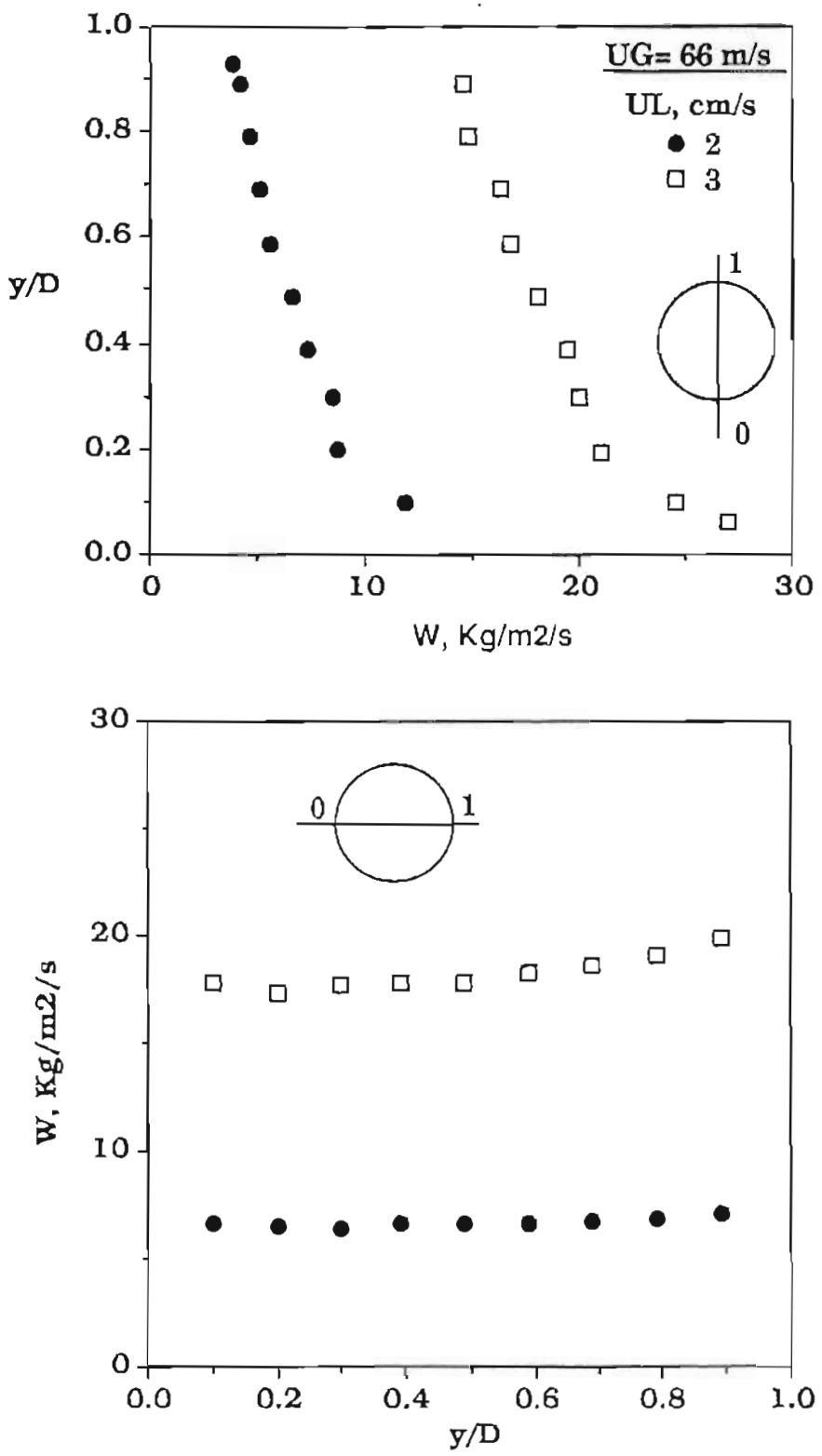

Figure 2. Vertical and horizontal droplet flux profiles for high gas flow rate.

(U_G=66m/s; U_L= 2 and 3 cm/s)

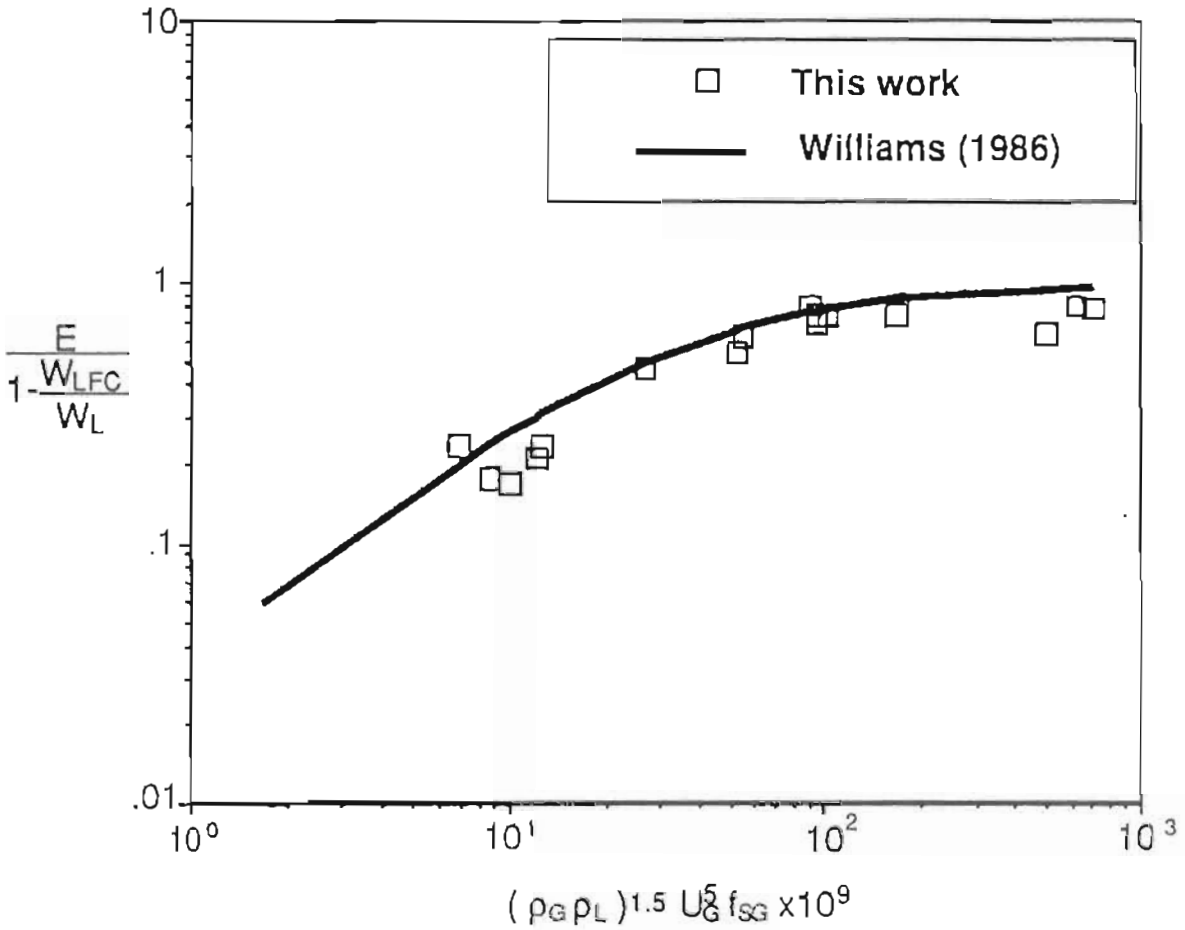

Figure 3. Comparison of entrainment data with correlation developed by Williams (1986).

Liquid entrainment , E, defined as

$$E=\frac{\dot{W}_{LE}}{W_L}=\frac{W_L-W_{LF}}{W_L}$$

where W_L is the total liquid flow rate, W_{LE} is the entrained liquid flow rate and W_{LF} is the flow rate of the liquid film. The entrainment data from this study are compared with an entrainment correlation developed by Williams (1986) and are in very good agreement, as shown in Figure 3. This expression correlates satisfactorily data for air/water and pipe diameters in the range D=25 to 100 mm. Thus it may be useful for practical applications, after testing with fluids of various properties.

3. A MODEL FOR PREDICTING DROPLET CONCENTRATION DISTRIBUTION

Such a model is useful for practical applications; e.g., for developing sampling or measuring procedures of the dispersed phase. It can be also helpful in the general effort for modeling annular flow.

Local concentrations seem to be more convenient than fluxes for modeling and data interpretation. Concentration of droplets C_i (Kg/m^3) at a point is given by $C_i=W_i/U_i$ where W_i is the local flux (Kg/m^{2}s) and U_i the local gas velocity (m/s). Local velocities are calculated by the well-known logarithmic expression (Schlichting, 1960) assuming that a symmetrical velocity profile exists in the pipe. The data by Tsuji and Morikawa (1982) taken under similar conditions, for flow of a dilute particle/gas mixture in horizontal pipes, lend support to this assumption. Another assumption is that a uniform wall roughness exists, due to the wavy liquid film, in the entire pipe circumference regardless of the film asymmetry. Rms data (Paras and Karabelas, 1990) show that this assumption may be realistic only at relatively high gas velocities, i.e. at U_G>40 m/s.

The proposed model, based on turbulent diffusion, can be used for predicting the concentration distribution in the pipe cross-section as well as the deposition rates of the entrained droplets onto the film. The basic assumptions of the model are as follows:

i. The entrained liquid in the gas phase is a dispersion of uniform droplet sizes.

ii. The droplets are small so that gas and particle diffusivities are equal, i.e., $\varepsilon=\varepsilon_p$.

iii. The concentration distribution is two-dimensional with respect to the vertical axis.

Using these assumptions a simplified diffusion equation is obtained (Karabelas,1977) :

$$\varepsilon \frac{dC}{dy} + wC(1-C) = a \qquad (1)$$

Considering that the droplets in the gas core are in dilute dispersion, $C(1 - C) \approx C$. Indeed, in our tests the mean droplet concentration is less than 0.003 m^3 water/m^3 air. Thus Eq.(1) reduces to :

$$\varepsilon \frac{dC}{dy} + wC = a \qquad (2)$$

where $\varepsilon \frac{dC}{dy}$ = droplet flux due to turbulent diffusion (Kg/m^2s) and

wC = droplet flux due to gravity (Kg/m^2s)

a = a constant flux (Kg/m^2s)

w = droplet settling velocity (m/s)

y = vertical coordinate (m).

ε = eddy diffusivity given by the expression $\varepsilon = \zeta.R.U^*$,

ζ = dimensionless droplet diffusivity (taken equal to 0.1)

R = pipe inside radius (m)

U^* = friction velocity (m/s).

One might interpret the constant flux **a** as follows. It has been observed by James et al (1980) in vertical annular flow that, in the process of liquid atomization, some droplets break off the film surface with large initial velocities. The nearly straight trajectories of such particles may not be significantly influenced either by turbulence or by gravity. In horizontal annular flow such trajectories may result in a net droplet flux which (at relatively low gas velocities) is in a direction opposite to gravity.

By defining $\alpha \equiv \frac{a}{w}$ and $k \equiv \frac{wR}{\varepsilon} = \frac{w}{0.1U^*}$, Eq. (2) becomes

$$\frac{dC}{C-\alpha} = -\frac{k}{R}\,dy \qquad (3)$$

The solution of Eq.(3) leads to

$$C = \alpha + \beta \cdot \exp\left[-k\frac{y}{R}\right] \qquad (4)$$

where β is the constant of integration. This constant (B) can be evaluated by applying the condition that, at steady state, the mean concentration $\langle C \rangle$ of the droplets in the pipe cross section (of area $\mathcal{A}$) is obtained as follows:

$$\langle C \rangle = \int_{\mathcal{A}} C \, d\mathcal{A}$$

Thus the final solution for the distribution becomes :

$$\frac{C}{\langle C \rangle} = A + \frac{1 - A}{E(k)} \exp[-k\frac{y}{R}] \qquad (5)$$

where $A = \frac{\alpha}{\langle C \rangle} = \frac{a}{w \langle C \rangle}$ is a dimensionless flux and E(k) is an expression involving Bessel functions. For small values of k it can be shown (Karabelas,1977) that

$$E(k) = 1 + \frac{k^2}{8}\left[1 + \frac{k^2}{24}\right] + O(k^6)$$

It is evident that only one parameter (the dimensionless flux A) is introduced in the model, in addition to the usual droplet settling velocity and diffusivity.

Predictions based on this model are in fairly good agreement with the data of this investigation and with the limited data in the open literature (Williams, 1986), as shown in Figures 4 and 5. The greatest discrepancy between measured and predicted profiles is observed in a few cases at high liquid rates. In those cases, the measured concentration profiles display an inflection point and a local maximum above the pipe centerline. However, before any definite conclusions are drawn, it must be recalled that there is an uncertainty regarding the true velocity distribution which is required to transform local fluxes to concentrations.

A value of the dimensionless flux A is obtained from the best fit of each profile. It is somewhat surprising that these values cannot be correlated with the gas flow rate. However, the parameter values from our work and from the profiles measured by Williams (1986) show an interesting correlation when plotted against the reduced liquid superficial velocity $U_R = U_{Ls}/U_{Lcr}$ (Figure 6). The critical liquid film velocity, (U_{Lcr}), corresponds to the film thickness, attained for high gas flow rates, below which there is no atomization. Values of the critical film flow rate W_{LFcr} from which U_{Lcr} is computed, for D = 50 mm and 100 mm, are given by Laurinat (1982) and Williams (1986). Figure 6 suggests that there is a correlation between A and U_{Ls}/U_{Lcr}. Furthermore, an obvious physical significance can be attached to the fact that as the liquid velocity

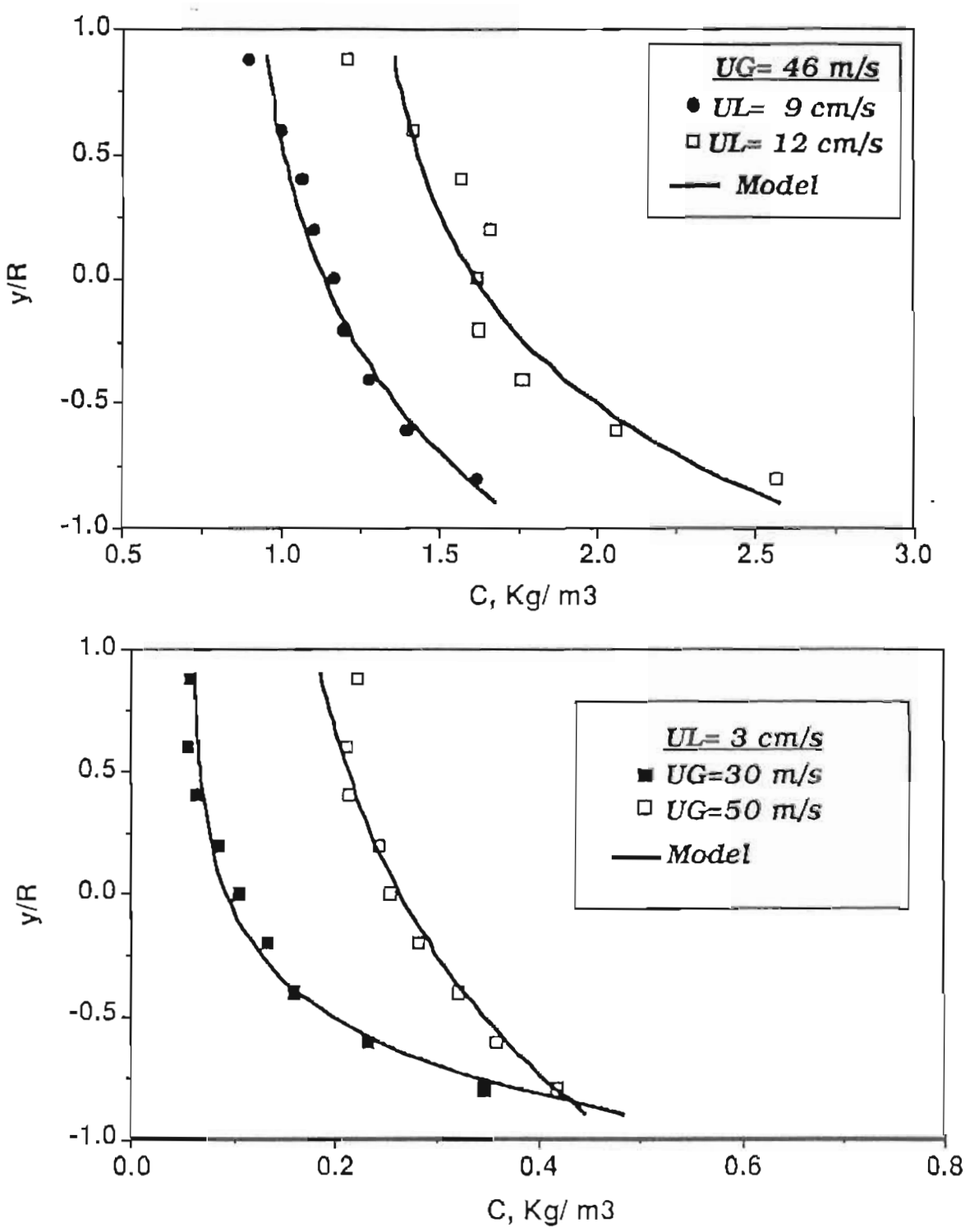

Figure 4. Comparison between our data and model predictions .

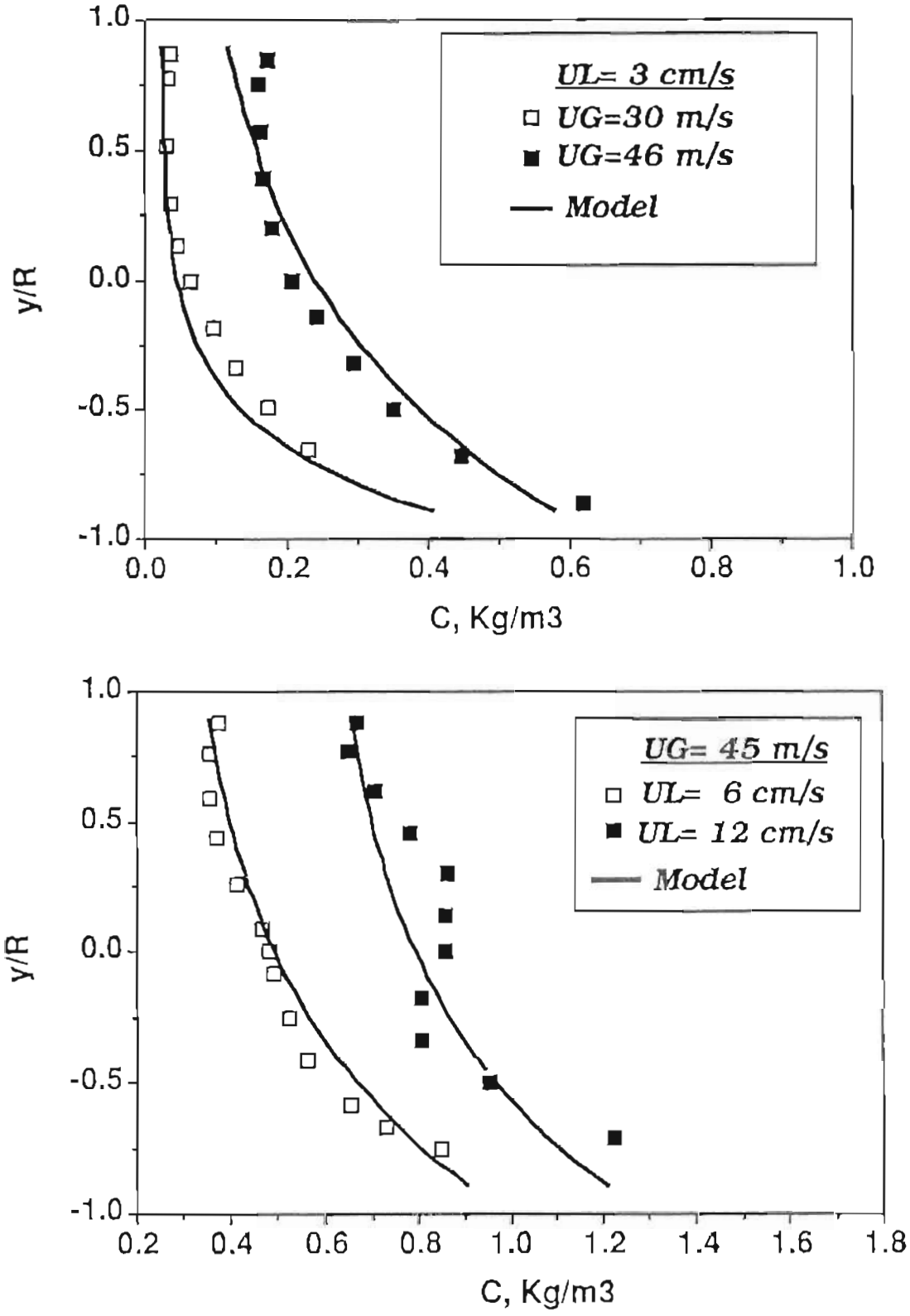

Figure 5. Comparison between data by Williams (1986) and model predictions.

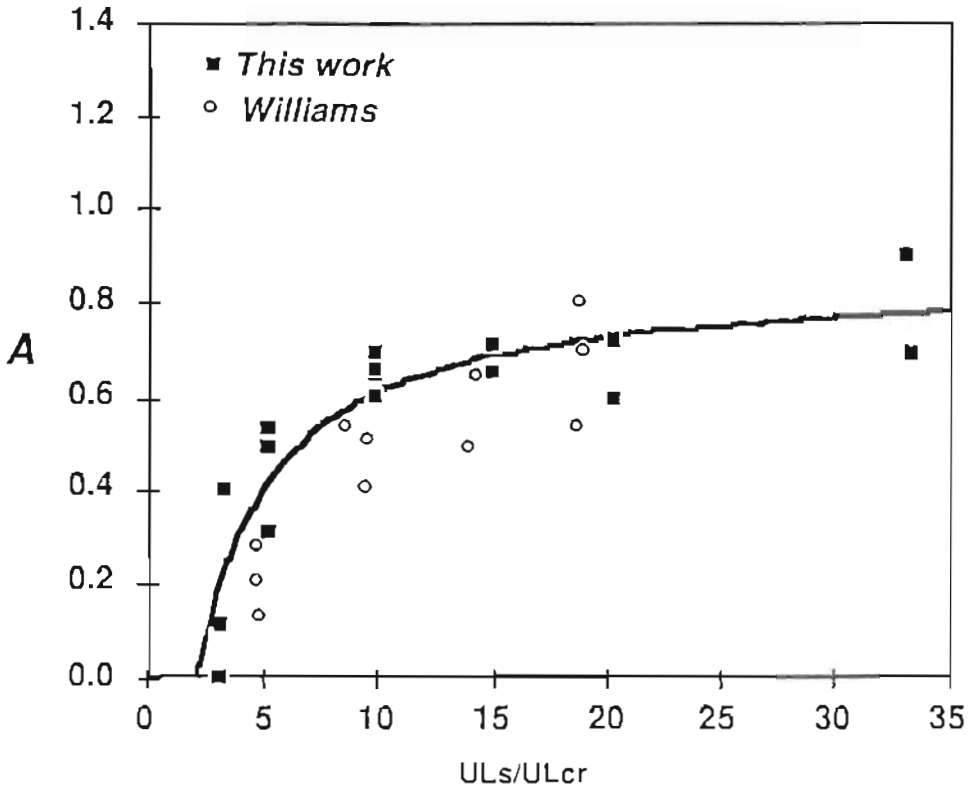

Figure 6. Influence of reduced liquid velocity on model parameter A.

approaches its critical value, i.e. as (U_L/U_{Lcr}) tends to 1.0 at which no atomization can occur, the flux A tends to zero. This observation lends support to the interpretation offered for A.

4. COMPUTATION OF DROPLET DEPOSITION RATES

Deposition rates of droplets (R_D) were calculated from our experimental data by assuming that deposition is a diffusion-like process, so that

$$R_D = k_D * C_i$$

where R_D is the droplet deposition flux (Kg/m^2s), k_D is the deposition rate constant (m/s) and C_i (Kg/m^3) the local droplet concentration close to the pipe wall. Droplet concentration on the film surface is considered to be zero.

McCoy and Hanratty (1977), among others, suggested that k_D/U^* should be a function of a droplet relaxation time τ^+:

$$\tau^+ = \frac{d_p^2 (U^*)^2 \rho_G^2}{18\,\mu_G^2} \frac{\rho_p}{\rho_G}$$

where U^* is the friction velocity, d_p and ϱ_p are the particle diameter and density and μ_G, ϱ_G are the gas viscosity and density, respectively. For $\tau^+ > 3000$, which applies to our data, McCoy and Hanratty suggested that

$$k_D = 20.7 \cdot U^* \cdot (\tau^+)^{-0.5}$$

In order to calculate τ^+ and k_D one needs reliable estimates of a mean droplet diameter. Azzopardi (1988) proposed an empirical relation for the Sauter mean diameter, d_{32}, based on available experimental and theoretical studies on drop sizes in vertical annular flow :

$$\frac{d_{32}}{\lambda} = \frac{15.4}{We^{0.58}} + \frac{3.5 G_{LE}}{\rho_L U_G} \qquad (6)$$

where $\lambda = \sqrt{\frac{\sigma}{\rho_L g}}$, $We = \frac{\rho_L U_G^2 \lambda}{\sigma}$ and G_{LE} = entrained liquid mass flux.

A respresentative mean diameter for our work is $d_p = d_{32}/0.7$ (Tatterson et al, 1977), with d_{32} obtained from Equation (6).

Following the above procedure, local deposition rates in the pipe circumference were computed from our data. Upon integration the average deposition rate R_d was obtained. These rates are plotted in Figure 7 versus the corresponding values of the parameter (flux) A. The trend in these data is similar to that in Figure 6, showing a steep increase of A with increasing R_d at small deposition rates, and a nearly constant A value at high R_d. Considering that under steady state conditions the average rate of atomization R_a is equal to average rate of deposition, i.e. $R_a = R_d$, this trend is consistent with the physical interpretation given for the flux A.

The values of R_d (or R_a) obtained here are currently studied and compared with other data from the literature. A method of correlation relating R_a to the liquid film characteristics, which was proposed very recently by Schadel and Hanratty (1989), appears to give satisfactory results. Moreover, our data on R_d appear to be in general agreement with the R_a values obtained by Schadel and Hanratty (1989), for <u>vertical</u> annular flow.

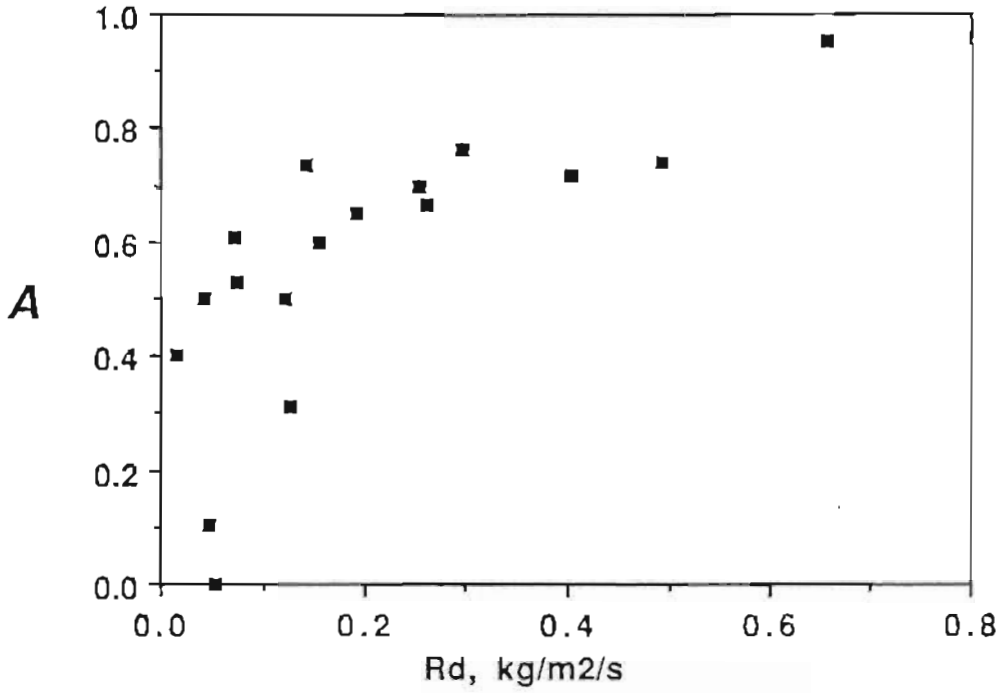

Figure 7. Correlation of dimensionless flux A with deposition rate

5. CONCLUSIONS AND COMMENTS

Local flux measurements, made with a direct sampling technique, allow the determination of mean liquid entrainment E, and of local deposition rate in the pipe circumference. The data on entrainment are satisfactorily correlated by an expression available in the literature (Williams, 1986). The computed local and average deposition rates appear to be quite reliable. However, comparisons with other data and correlations are required to better assess their accuracy.

A relatively simple model is proposed for predicting the droplet concentration distribution in the pipe cross-section and the circumferential variation of deposition rate. Terms representing fluxes due to turbulent diffusion and gravitational settling are used in the model. An additional dimensionless flux A is introduced and is considered to be related to droplets gaining high inertia upon atomization. Attempts are currently made to develop a reliable correlation for obtaining independent estimates of A.

The aforementioned model, although quite convenient for practical calculations and modeling, relies on several physical quantities which have

not been firmly established for the system under study. Such quantities are the mean droplet size (or size distribution), a characteristic droplet settling velocity (or slip velocity in the direction of gravity) - not to mention the droplet and gas velocity distributions. Work related to some of these quantities is in progress in our Laboratory.

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