Pitch-to-Diameter Effect on Two-Phase Flow Across an In-Line Tube Bundle

Void fraction and friction pressure drop measurements were made for an adiabatic, vertical two-phase flow of air-water across two horizontal, in-line, 5 × 20 rod bundles, one with a pitch-to-diameter ratio, P/D, of 1.3, the other 1.75. For both bundles the average void fraction showed a strong mass velocity effect and values were significantly less than those predicted by a homogeneous flow model. All void fraction data were found to be well correlated, with no P/D effect, using the dimensionless gas velocity, j^*_g. The two-phase friction multiplier data exhibited a strong effect of P/D and mass velocity, however, the data for both bundles could be well correlated with the Martinelli parameter for $G > 200\ kg/m^2s$. The correlations developed for void fraction and two-phase friction multiplier were successfully tested in predicting the total pressure drop in boiling freon experiments.

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Introduction

It has been estimated that more than 50% of all heat exchangers employed in process industries are used to boil fluids and involve two-phase flow on the shell side. The hydrodynamics and boiling characteristics are, however, only poorly understood at present for vertical flows across a tube bundle (Bell, 1983).

In a photographic study of boiling two-phase flow in a horizontal kettle reboiler, Cornwell et al. (1980) observed strong variations in voidage across the bundle, and two-dimensional flow with considerable inflow from the sides and predominantly upward flow in high voidage regions near the center and top of the bundle. This would suggest that results from forced cross-flow experiments conducted with known mass velocity, G, and quality, x, could be used to analyze the kettle and internal boiler performance. Many authors (Brisbane et al., 1980; Palen and Yang, 1983; Whalley and Butterworth, 1983; Fair and Klip; 1983) have presented circulation boiling models in which average void fraction in the bundle and two-phase friction pressure drop were estimated using either in-tube correlations, a homogeneous flow model, or other methods, none of which has been verified to be generally valid for a bundle geometry. More recently, Jensen (1988) proposed a one-dimensional model for the recirculating flow in a kettle reboiler by predicting the pressure drop between various points in the reboiler using the void fraction and two-phase friction multiplier correlations proposed by Schrage et al. (1988).

The amount of experimental data on two-phase flow across a bundle obtained under well-defined boundary conditions is also limited. Grant and Chisholm (1979) measured pressure drop on the shell side of a segmentally baffled heat exchanger under horizontal and vertical upward and downward flows. Without any measurement of actual void fraction, they assumed that the hydrostatic components of pressure drop in the upward and downward flow sections would cancel. Kondo and Nakajima (1980) made the first indirect measurements of void fraction in vertical air-water upflow across a tube bank. They found that the void fraction was dependent on superficial gas velocity, not on liquid velocity, although the study was performed at very low mass velocities ($G < 5\ kg/m^2s$). Chan and Shoukri (1987) also measured void fraction profiles in heated 3 × 3 and 3 × 9 bundles using R-113 under pool boiling conditions.

Recently, Schrage et al. (1988) made void fraction and pressure drop measurements in an in-line 4 × 27 bundle using air and water under controlled inlet conditions. They used a pair of quick-shutoff plate valves to measure bundle average void fraction and found that the homogeneous flow model overpredicted their data. They noted that there was a strong mass velocity effect on void fraction and proposed a void fraction correlation in terms of a Froude number and quality. Until now, however, there have been few studies that investigated the effect of pitch-to-diameter ratio on void fraction and two-phase friction pressure drop in a bundle (Kondo and Nakajima, 1980; Hsu and Jensen, 1988).

In the present work, void fraction profiles and two-phase pressure drop have been measured in two different in-line

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bundles with pitch-to-diameter ratios, P/D, of 1.3 and 1.75, in order to elucidate the two-phase flow characteristics of air-water mixtures under adiabatic vertical crossflow conditions. The data are compared with the results of previous work and new correlations for bundle average void fraction and two-phase friction mutiplier are proposed. They are then tested by predicting the total pressure drop in boiling two-phase flow experiments.

Experimental Apparatus and Procedure

The air-water test loop used is shown in Figure 1. Air from the building supply was metered using a bank of rotameters and injected into the entrance of the test section through a 25.4 mm OD porous tube that extended across the test section normal to the rod axis. The air flow rate was controlled by a valve located at the inlet of the rotameter and the air pressure was monitored with a Bourdon gauge. Deionized water stored in a tank was pumped through an orifice or turbine flow meter into a nozzle at the bottom of the test section. The two-phase mixture traveled up through a rectangular test section containing the rod bundle and exited through a 50 mm ID nozzle into an open-top separator tank.

Each bundle consisted of 20 rows of four full rods and two half-rods each 80 mm long, arranged in a sqaure array as shown in Figure 2. One bundle was fabricated with 19 mm OD rods and $P/D = 1.3$, the other with 12.7 mm OD rods and $P/D = 1.75$. The two half-rods were attached to the side walls in order to minimize bypass leakage. The test section walls and rods were made of clear acrylic. Flow straighteners in the form of parallel plates 210 mm long were located 65 mm above the air injection nozzle and 65 mm from the bottom of the rod bundle. Flow visualization revealed that the air-water flow was well mixed before reaching the first row of rods. The top of the rod bundle was located approximately 400 mm below the exit nozzle and diffuser plates were placed before the exit to obtain smooth flow out of the bundle.

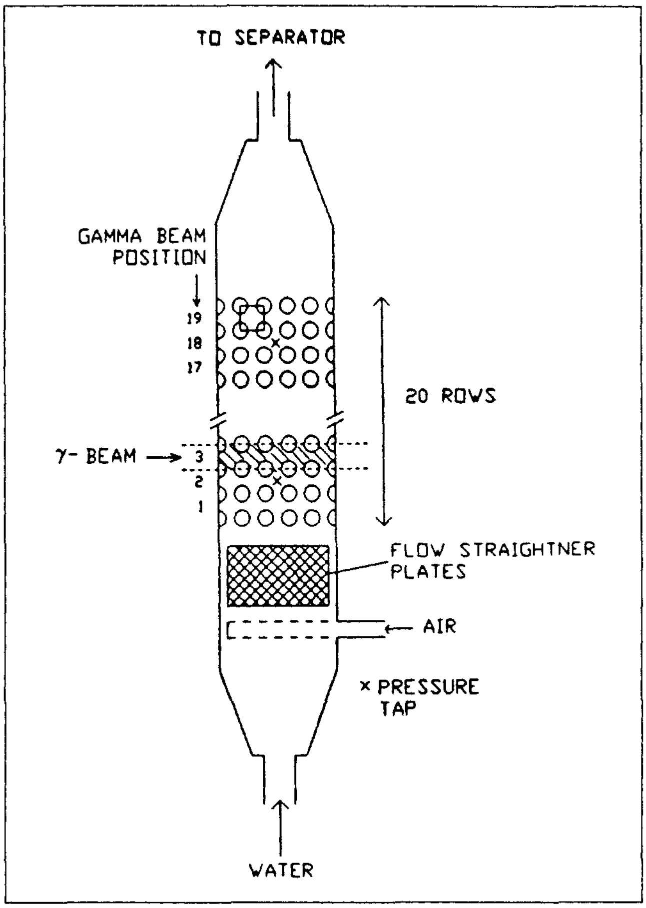

Figure 2. Test section.

In order to determine the two-phase friction pressure drop, an accurate measurement of the bundle average void fraction is required to estimate the gravitational pressure drop which is then subtracted from the measured total pressure drop. If the pressure taps are located within the bundle to minimize the end

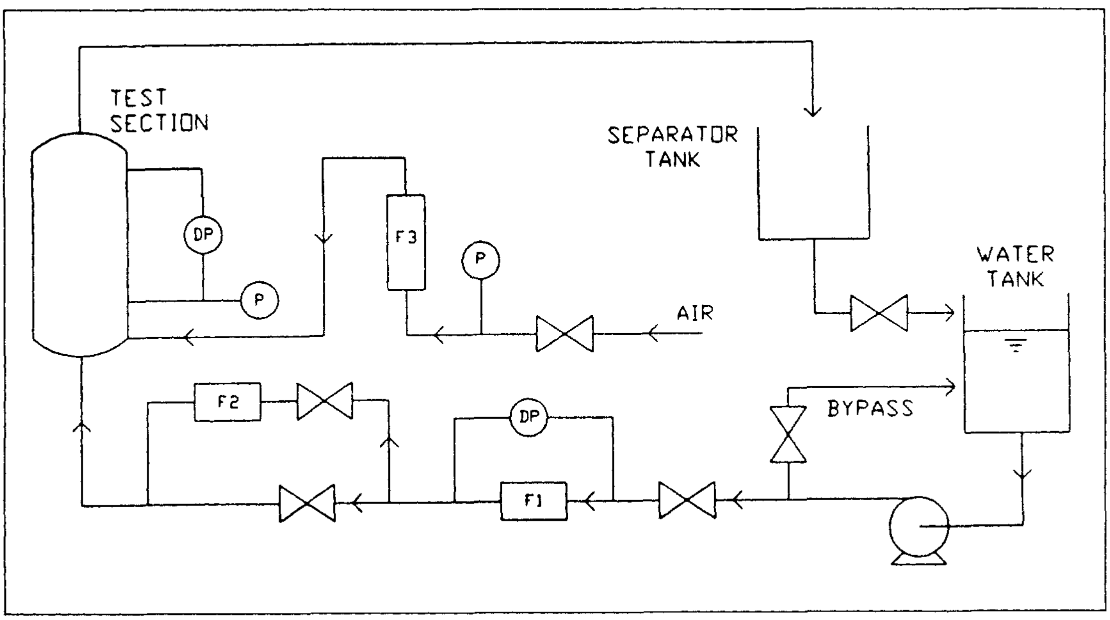

Figure 1. Experimental flow loop.

effects at the inlet and exit of the bundle as in the present experiment, the bundle average void fraction should be measured over the section of the bundle bounded by those pressure taps. For this reason, a gamma densitometer was used to measure the local void fractions at various elevations between the pressure taps, from which an average void fraction would readily be calculated.

A single-beam gamma densitometer with a well-collimated beam (24 mm high × 50 mm wide) was used to measure the void fraction at various elevations in the bundle, Figure 2. The gamma source used was Co-57 which emits 122 and 135 keV gamma rays and gives good sensitivity for air-water flows even in small flow channels. A detailed description of the gamma densitometer and calibration procedure used is given by Chan and Banerjee (1981). The gamma source and detector were mounted on a vertically traversible platform with a digital position readout device to accurately position the beam and assure reproducibility of the measurements.

For each void fraction measurement at a specific beam position, at least seven readings of the attenuated gamma beam intensity were obtained for each of the three conditions: test section full of water, I_l, full of air, I_g, and with a two-phase mixture, $I_{2\phi}$. The chordal average void fraction, α_c, was then evaluated from the average counts using the log-interpolation method:

$$\alpha_c = \frac{\ln (I_{2\phi}/I_l)}{\ln (I_g/I_l)} \tag{1}$$

The bundle average void fraction, α, was then calculated by averaging the chordal average void fractions obtained at differential positions between the pressure taps.

In order to obtain pressure drop data free of end effects, a differential pressure transducer, Rosemount model 1151DP with a range of −2.49 to 6.225 kPa, was connected to the pressure taps located above the second row and below the nineteenth row, respectively. Pressure transducers were also connected to the lower tap and above the tenth row in order to measure the absolute test section pressure. All pressure sensing lines were purged free of air bubbles before each test run and the signals from all transducers were recorded on a chart recorder.

For a given two-phase flow run with the specified water and air flow rates, a gravitational pressure drop was calculated using the bundle average void fraction and it was subtracted from the measured pressure drop to obtain the frictional two-phase pressure drop, $\Delta P^F_{2\phi}$. The acceleration pressure drop was neglected in the present analysis. The two-phase friction multiplier was then computed based on the pressure drop for the liquid phase flowing alone, ΔP^F_l, with the same liquid flow rate as in the two-phase flow run.

$$\phi_l^2 = \frac{\Delta P^F_{2\phi}}{\Delta P^F_l} \tag{2}$$

The experiments were conducted at near atmospheric pressures (101–180 kPa) under the following conditions:

$P/D = 1.3$	$P/D = 1.75$
$27 \le G \le 818$ kg/m²s	$90 \le G \le 542$ kg/m²s
$0 \le x \le 0.33$	$0 \le x \le 0.08$
$520 \le Re_l \le 15{,}700$	$1151 \le Re_l \le 6{,}932$
$20 \le \Delta P^F_{2\phi} \le 8{,}540$ Pa	$70 \le \Delta P^F_{2\phi} \le 3{,}300$ Pa

The liquid Reynolds number Re_l above is defined in terms of the rod diameter and maximum velocity, V_{max}, between the adjacent tubes. For a series of runs with a given mass velocity, the liquid flow rate was set at the specified value and the single-phase pressure drop was first measured. The air was then introduced as specified and after reaching steady conditions, the two-phase pressure drop and test section pressure were recorded. The void fraction measurements were then made by traversing the gamma densitometer vertically along the bundle.

The flow behavior was also recorded using a video camera with an electronic shutter speed of 1/1000 s.

Results and Analysis

Single-phase friction factor

The pressure drop data for single-phase liquid runs were used to compute the bundle friction factor as follows:

$$\Delta P^F_l = \frac{Nf[G(1-x)]^2}{2\rho_l} \tag{3}$$

The computed friction factors for both in-line bundles with $P/D = 1.3$ and 1.75 are shown in Figure 3. A Blasius-type correlation was fitted to the data for the higher Reynolds number region. The friction factor data are seen to compare well with the correlation of Zukauskas (1972). For greater accuracy, however, the measured single-phase pressure drop was directly used in the calculation of friction multipliers.

Void fraction

The void fraction profiles measured across and above the bundle for four different runs are shown in Figure 4. Although the void fraction can change quite abruptly at the outlet of the bundle, the measurements at beam positions 1 through 19 inside the bundle show a relatively uniform profile in every run. These relatively uniform void profiles were also observed for the other test bundle with $P/D = 1.75$. Consequently, only four beam positions (positions 3, 8, 12, and 17) were used to obtain the bundle average void fraction in most of the tests.

The above void measurements were made with the 50 mm wide gamma beam placed at the middle of the bundle along the rod axis. Uniformity of void fraction along the rod axis was also checked by making measurements with the beam moved close to

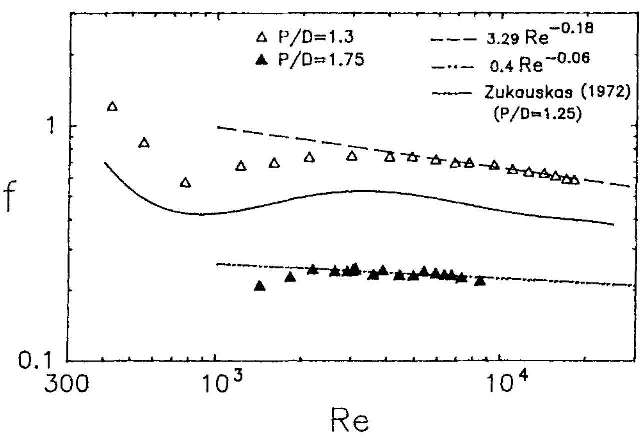

Figure 3. Comparison of single-phase friction factor data.

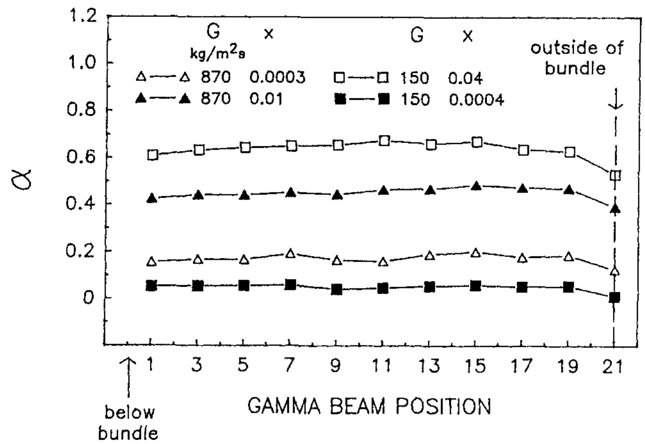

Figure 4. Test section void fraction profiles, P/D = 1.3.

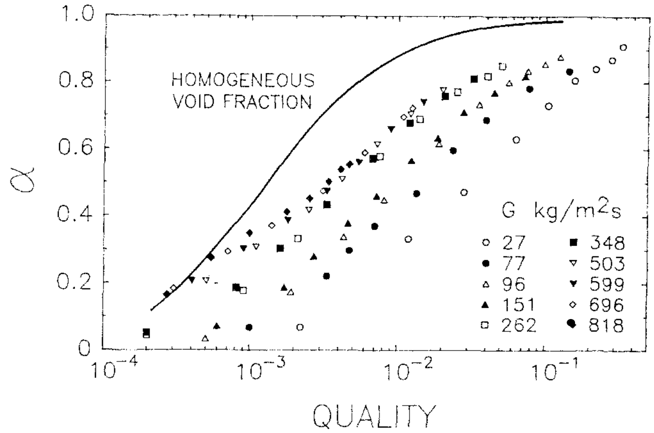

Figure 5a. Void fraction data and mass velocity effect, P/D = 1.3.

the ends of the rods. For both low and high mass velocity flows with various qualities, the maximum difference in void fractions measured at the center and near the wall was 3%.

The bundle average void fraction data are plotted against quality in Figures 5a and 5b for the bundles with P/D = 1.3 and 1.75, respectively. Comparison of the data from both bundles shows that the effect of P/D appears to be negligible for given values of mass velocity. The present data also show a strong mass velocity effect, where for a given quality, higher void fraction is obtained with increasing mass velocity.

Compared to the predictions by the homogeneous flow model, the measured void fractions are significantly lower for both bundles. The discrepancy between the homogenous flow model and the present data is seen to increase with decreasing mass velocity and quality, as has been observed in two-phase flow in vertical tubes. This is because the homogeneous flow model assumes no slip between the phases and the validity of this assumption depends on the degree of mixing achieved by the two phases. At low mass velocities, the effect of buoyancy is significant, especially at low qualities. The flow visualization using a video camera revealed that at low mass velocities, the air bubbles tend to flow as a jet between the neighboring rods in the vertically connected channels. There is little crossflow of air bubbles from one vertically connected channel to another. Similar observations were made by Cornwell et al. (1980) in a kettle reboiler tube bundle. At high mass velocities, the turbulence in the liquid phase helps in mixing the two phases and a more homogeneous mixture is obtained.

The mass velocity effect is also evident when the bundle average void fractions are plotted against the Martinelli parameter as shown in Figure 6 for the bundle with P/D = 1.3. The Martinelli parameter assuming turbulent regimes for both gas and liquid is given by

$$\chi_{tt}^2 = \left(\frac{1-x}{x}\right)^{2-m} (\rho_g/\rho_l)\,(\mu_l/\mu_g)^m \tag{4}$$

where the value of m is derived from the Blasius-type friction factor correlation as shown in Figure 3. It is noted, however,that a value of 0.2 was used for m in Eq. 4 for both bundles, since this resulted in slightly less scatter than when m was allowed to vary according to the measured Reynolds number dependence. Similar results have also been reported by Ishihara et al. (1979) and Schrage et al. (1988). Because of the strong mass velocity effect, a Lockhart-Martinelli type correlation, used often for in-tube flows, is not applicable here. Similar results were also obtained for the bundle with P/D = 1.75.

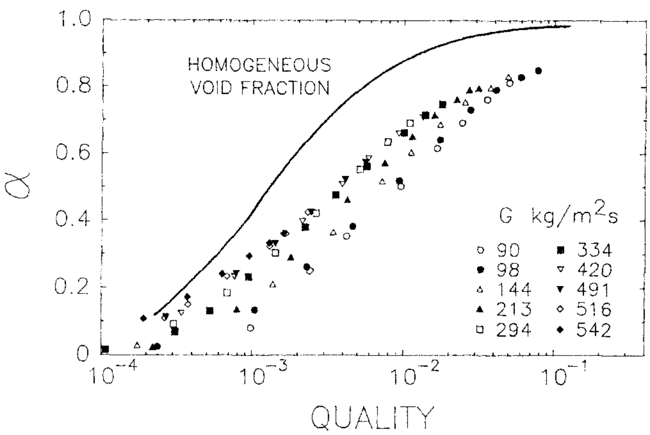

Figure 5b. Void fraction data and mass velocity effect, P/D = 1.75.

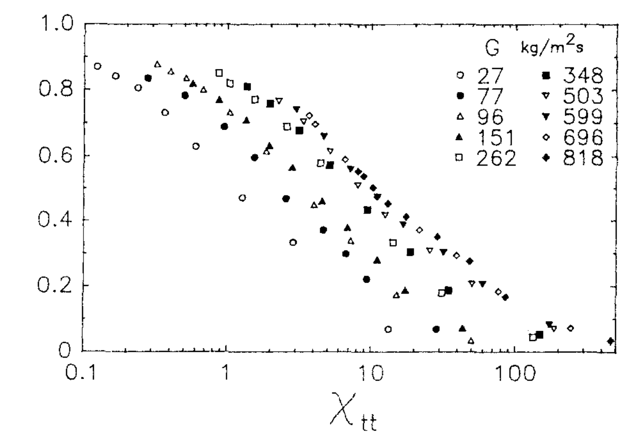

Figure 6. Variation of void fraction with Martinelli parameter, P/D = 1.3.

For vertical two-phase flows, buoyancy is expected to be important. If the mass velocity effect is assumed to be caused primarily by the balance of buoyancy and inertial forces, then a parameter representing these forces should be used instead of the Martinelli parameter. One such parameter is the following dimensionless gas velocity (Wallis, 1969).

$$j_g^* = \frac{\rho_g^{1/2} j_g}{\sqrt{gD(\rho_l - \rho_g)}} \tag{5}$$

where the superficial gas velocity, j_g, is defined in terms of the minimum flow area, the rod diameter is used for D, and the gas density is evaluated at the average test section pressure. This parameter has been shown by Dowlati et al. (1988) to correlate fairly well the void fraction data obtained in a bundle with $P/D = 1.26$ and 10 rows (half the number of rows used in the present study). As shown in Figure 7, the void fraction data from the present study are correlated quite well with the Wallis parameter above using the following correlation,

$$\alpha = 1 - \frac{1}{(1 + C_1 j_g^* + C_2 j_g^{*2})^{1/2}} \tag{6}$$

It should be noted that Eq. 6 is similar in form to the correlation in terms of the Martinelli parameter suggested by Chisholm and Laird (1958) for two-phase flow in circular tubes. The constants, C_1 and C_2, set equal to 35 and 1, respectively, gave the best overall fit, as shown in Figure 7, for void fraction data from both bundles, with an average deviation of 10%. In order to improve the correlation of Eq. 6 to predict void fraction for $j_g^* \geq 0.2$, it is suggested that C_2 be set equal to 30.

We note that no pitch-to-diameter ratio effect on void fraction is observed for the present bundles with $P/D = 1.3$ and 1.75. Kondo and Nakajima's results (1980) also showed no pitch-to-diameter ratio effect on void fraction in bundles with greater than 13 rows over the P/D range from 1.08 to 1.4, although the mass velocity was limited to less than 5 kg/m²s in their work. The void fraction data for the shorter bundle tested by Dowlati et al. (1988) ($P/D = 1.26$, 10 rows) fell slightly below those shown in Figure 7 and this is perhaps due to the inlet and exit effects present in shorter bundles, as also reported by Kondo and Nakajima (1980).

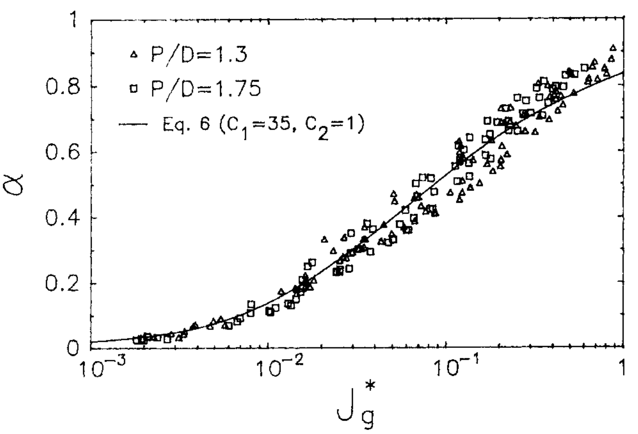

Figure 7. Correlation of void fraction data with dimensionless gas velocity.

The success of the use of j_g^* in eliminating the mass velocity effect may be explained by the fact that it contains both the total mass velocity and quality. This is clear when j_g^* is expressed in an alternate form,

$$j_g^* = \frac{Gx}{\sqrt{\rho_g gD(\rho_l - \rho_g)}} \tag{7}$$

In comparison with Eq. 6, the correlation of Schrage et al. (1988), based on their void fraction data obtained in a $P/D = 1.3$ bundle using a pair of quick-closing plate valves, was found to underestimate the present data by as much as 50% of the measured value.

Two-phase friction multiplier

A model proposed by Lockhart and Martinelli (1949) to calculate the two-phase friction multiplier in horizontal tube flow and represented by a simple equation due to Chisholm and Laird (1958),

$$\phi_l^2 = 1 + \frac{C}{\chi_{tt}} + \frac{1}{\chi_{tt}^2} \tag{8}$$

has also been applied by various investigators for shell-side two-phase flow (Ishihara et al., 1979; Schrage et al., 1988). Thus, the two-phase friction multiplier data were plotted against the Martinelli parameter for both bundles with $P/D = 1.3$ and 1.75 as shown in Figures 8a and 8b, respectively. Again, it is noted that a value of 0.2 for m was used in Eq. 4 for both bundles.

For $P/D = 1.3$, Figure 8a, the data for $G \geq 260$ kg/m²s follow the same trend suggested by Eq. 8 and the best fit curve is given by $C = 8$. For $G < 260$ kg/m²s, however, a strong mass velocity effect is observed. In a review of both two-phase friction multiplier data and various models, Ishihara et al. (1979) also found that a C value of 8 best fit Eq. 8 to their data base, although large scatter was seen for $\chi_{tt} > 0.2$, and the void fraction correlation used to compute their friction multiplier values was not specified. Schrage et al. (1988), on the other hand, found that a C value of 8 overpredicted their friction pressure drop data by an average of 17% and furthermore suggested dependence of C on flow patterns. They calculated C values for all data points and found that values of C lower than 8 were associated mostly with low mass velocity, high-quality flows (i.e., $\chi_{tt} < 1$) such as in spray flow. This was not the case in our experiment, as shown by the values of ϕ_l^2 lying above the $C = 8$ curve in Figure 8a. Underprediction of the two-phase friction multiplier would result if lower than actual void fraction were used to estimate the gravitational pressure drop, which is subtracted from the measured total pressure drop to determine the friction pressure drop. The void fraction correlation proposed by Schrage et al. (1988) for the $P/D = 1.3$ bundle was found to significantly underpredict our void fraction data, as previously noted.

For Martinelli parameter greater than 1, the C values computed from the data of Schrage et al. were mostly greater than 8 and ranged up to 30. This is consistent with our experimental results.

For the bundle with $P/D = 1.75$, it is clear that a value of $C =$ 8 in Eq. 8 can no longer predict the two-phase friction multiplier

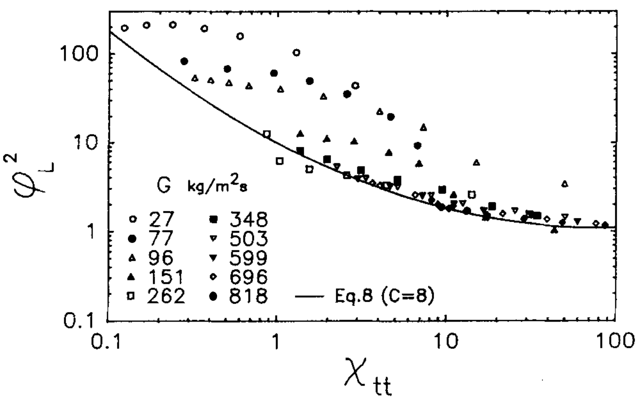

Figure 8a. Liquid-only two-phase friction multiplier data and Martinelli parameter, $P/D = 1.3$.

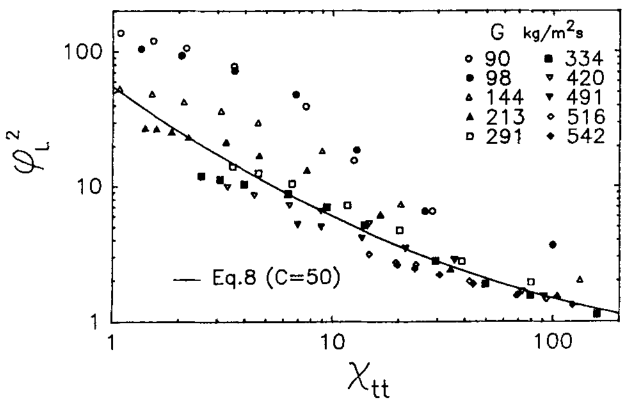

Figure 8b. Liquid-only two-phase friction multiplier data and Martinelli parameter, $P/D = 1.75$.

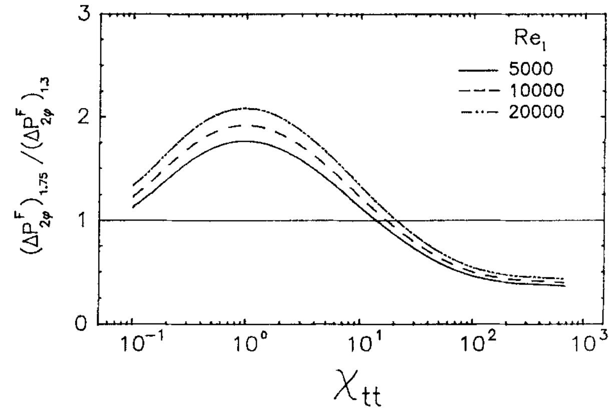

Figure 9. Ratio of two-phase friction pressure drops for bundles with $P/D = 1.3$ and 1.75.

data, Figure 8b. Instead, $C \approx 50$ gave a fairly good correlation for $G > 200$ kg/m²s. For $G < 200$ kg/m²s, a strong mass velocity effect was again observed as in the previous bundle.

In comparing Figures 8a and 8b, it is clear that the two-phase friction multiplier, for a given value of χ_{tt}, is found to be greater for the larger P/D bundle. The P/D effect observed tends to increase as χ_{tt} decreases. This does not necessarily mean that the two-phase frictional pressure drop in a bundle with larger pitch will always be greater than that in a smaller pitch bundle. In order to obtain a clearer picture of the effect of increasing pitch for an in-line bundle with a given tube diameter, the two-phase friction pressure drops were calculated for the two bundles with $P/D = 1.75$ and 1.3, using the two-phase friction multiplier given by Eq. 6 and the single-phase friction factor correlations shown in Figure 3. The ratio of the friction pressure drops for the larger to smaller pitch bundle was then plotted, as shown in Figure 9, against the Martinelli parameter for several values of Re_l, which correspond approximately to $G > 200$ kg/m² · s.

For sufficiently large Martinelli parameters ($\chi_{tt} > 50$), the bundle average void fraction is still small ($\alpha < 0.2$), the friction pressure drop for the larger pitch bundle is smaller than that for the smaller pitch bundle, and the ratio of the pressure drops is well below unity. This is not surprising since at low void fractions the flow behaves similar to a single-phase liquid flow and the single-phase friction factors are smaller for the larger pitch bundle as discussed previously. Therefore, as χ_{tt} increases above 50, the ratio of the two-phase friction pressure drops for a specified Reynolds number will asymptotically approach the ratio of the single-phase friction factors.

As the Martinelli parameter decreases to values below about 50 and the void fraction increases to above 0.2, the effect of increased friction multiplier for the larger pitch bundle starts to dominate and the ratio of pressure drops increases to values greater than unity. The larger pitch bundle now has greater friction pressure drop. The ratio continues to increase until it reaches a peak at about $\chi_{tt} = 1$, where the void fraction is about 0.8. Although the void fraction and pressure drop data are not available for the $P/D = 1.75$ bundle for $\chi_{tt} < 1$, further calculations, assuming that Eq. 8 is still applicable, indicate decreasing ratio of the friction pressure drops for smaller χ_{tt} and void fractions greater than about 0.8.

At sufficiently low values of the Martinelli parameter (less than about 0.05), the ratio is expected to fall below unity and the friction pressure drop for the larger pitch bundle to become smaller than that for the smaller pitch bundle. At these small values of χ_{tt}, the void fraction would be significantly greater than 0.8 and the flow is expected to be in the dispersed droplet flow regime and to behave similar to single-phase gas flow. The drop in the ratio to values below unity would again be consistent with the lower friction factors for the larger pitch bundle in single-phase gas flow. Thus, as χ_{tt} further decreases, the ratio of the two-phase friction pressure drops should asymptotically approach that of the single-phase friction factors.

The above results also apply to the total pressure drop, since the void fractions at the same mass flux and quality were seen to be very similar for the in-line bundles with different P/D ratios and gravitational pressure drop as well as accelerational pressure drop for boiling flow situations would be equal.

In order to explain why the two-phase friction pressure drop is greater for the larger pitch bundle compared to the smaller pitch bundle over a wide range of void fractions, it is speculated that with a larger pitch the two-phase wake behind each tube becomes more developed since the downstream tube is no longer hindering its formation. Considered as an energy sink, the more developed wake can lead to an increase in frictional pressure loss as the two-phase mixture flows past the tubes. In fact, video pictures revealed that at relatively high qualities the wake became less clearly defined for the $P/D = 1.3$ bundle compared to $P/D = 1.75$.

Pressure drop in boiling crossflow

In order to test the applicability of the friction multiplier and void fraction correlations, Eqs. 6 and 8, to a boiling bundle, the total pressure drop data of Hsu (1987) were analyzed. Hsu used an in-line 5 × 27 tube bundle with 7.94 mm OD tubes and $P/D = 1.3$, to boil freon R-113 at pressures of 200–500 kPa, with a mass velocity range of 50–700 kg/m²s. The static pressure profile in the bundle was obtained from several pressure drop measurements using inclined manometers connected to pressure taps located along the bundle height. To test the present correlations, the pressure drop between two elevations sufficiently far apart was computed from Hsu's data and compared with the predictions of the total pressure drop due to gravity, friction, and acceleration.

The gravitational pressure drop between the two elevations was calculated by evaluating the local void fraction from quality using the void fraction correlation, Eq. 6, and numerically integrating the results. Since Hsu did not measure the void fraction in his experiments, a direct comparison between the measured and predicted void fractions could not be made.

The two-phase friction pressure drop was calculated using Eq. 2, where Hsu's single-phase friction factor correlation was used to determine the single-phase friction pressure drop, and the two-phase friction multiplier was computed using Eq. 8. The two-phase frictional pressure drop was then numerically integrated between the two quality values at the given elevations. For all runs with $G \geq 260$ kg/m²s, the value of $C = 8$ was used in Eq. 8 as suggested in Figure 8a. For $260 > G > 90$ kg/m²s and $G < 90$ kg/m²s, respective C values of 35 and 70 obtained from our data for $\chi_{tt} > 1.0$ were used to predict the two-phase friction multiplier for the remainder of Hsu's data.

Due to the large variation in quality across the bundle, the accelerational pressure drop, ΔP_{acc}, was also taken into account. This was obtained from the separated flow model (Collier, 1982),

$$\Delta P_{\text{acc}} = G^2 (v_2' - v_1') \tag{9}$$

where

$$v' = \frac{(1 - x)^2}{(1 - \alpha)\rho_l} + \frac{x^2}{\alpha \rho_g} \tag{10}$$

The void fraction was again evaluated from quality using Eq. 6.

Figure 10 shows the ratio of the experimental overall pressure drop, as determined from Hsu's data, to the predicted pressure drop plotted against the average quality between the two elevations considered. Our correlations are successful in predicting 90% of the data points considered, within ±20%. Although the present correlations have been derived from experiments performed with air and water at near atmospheric pressures, it is clear from Figure 10 that they also work well in predicting the total pressure drop for freon at the elevated pressures.

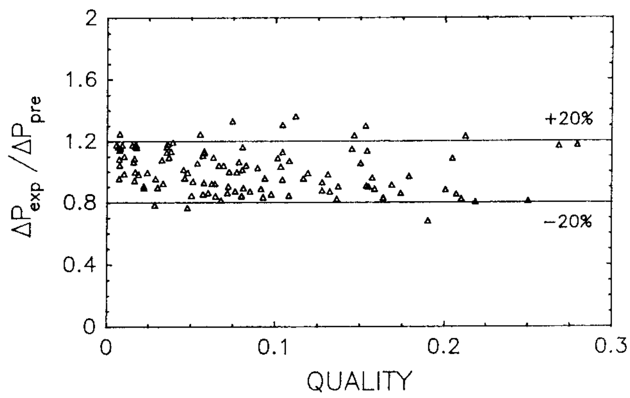

Figure 10. Prediction of overall pressure drop data obtained by Hsu (1987) in freon boiling experiments.

Conclusions

Void fraction and friction pressure drop data were obtained for vertical two-phase flow of air-water across two horizontal, in-line 5 × 20 rod bundles, one with $P/D = 1.3$ and the other $P/D = 1.75$. The experimental mass velocity ranged from 27 to 818 kg/m²s and quality between 0.0002 and 0.33. Analysis of the data revealed the following.

1. The void fraction profiles were relatively uniform inside the bundles over the entire range of flow rates covered.
2. The bundle average void fractions were much smaller than the values predicted by a homogeneous flow model and showed a strong mass velocity effect. A new void fraction correlation was proposed in terms of a dimensionless gas velocity.
3. The void fraction data when plotted against the dimensionless gas velocity showed no mass velocity and pitch-to-diameter ratio effect over the range covered.
4. The two-phase friction multiplier data could be correlated well in terms of a Martinelli parameter, however, strong mass velocity effects were observed for low mass velocities ($G < 200$ kg/m²s) in both bundles.
5. Increasing the pitch in an in-line bundle with a given tube diameter can lead to a higher two-phase pressure drop over a wide range of void fractions.
6. The void fraction and two-phase friction multiplier correlations presented here were reasonably successful in predicting the total pressure drop across an in-line bundle with boiling freon at pressures up to 500 kPa.

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Notation

C = C factor, Eq. 8
C_1, C_2 = coefficients, Eq. 6
D = tube diameter, m
f = single-phase friction factor, Eq. 3
G = mass velocity based on minimum flow area, kg/m²s
g = gravitational constant, 9.806 m/s²
I = attenuated gamma beam count, Eq. 1
j_g = Gx/ρ_g, superficial gas velocity, m/s
j_g^* = dimensionless gas velocity, Eq. 5
m = exponent in Blasius-type equation
N = number of tube rows between pressure taps
ΔP = pressure drop, kPa
P = pitch, m
Re_l = $(\rho_l D V_{max})/\mu$, liquid Reynolds number

V_{max} = maximum liquid velocity between adjacent tubes, m/s
v' = weighted specific volume, Eq. 10, m^3/kg
x = quality

Greek letters

α = void fraction
μ = dynamic viscosity, kg/m · s
ρ = density, kg/m^3
ϕ^2 = two-phase friction multiplier
χ_{tt} = Martinelli parameter, Eq. 4

Subscripts

g = gas phase only
l = liquid phase only
2ϕ = two-phase

Superscripts

F = friction

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