Lin, C. S., E. B. Denton, H. S. Gaskill, and G. L. Putnam, "Diffusion-Controlled Electrode Reactions," *Ind. Eng. Chem.*, **43**, 2136 (1951).
Linton, W. H., and T. K. Sherwood, "Mass Transfer From Solid Shapes to Water in Streamline and Turbulent Flow," *Chem. Eng. Progr.*, **46**, 258 (1950).
McConaghy, G. A., "The Effect of Drag Reducing Polymers on Turbulent Mass Transfer," Ph.D. thesis, Univ. Ill., Urbana (1974).
Mizushina, T., F. Ogino, Y. Oka, and H. Fukuda, "Turbulent Heat and Mass Transfer Between Wall and Fluid Streams of Large Prandtl and Schmidt Numbers," *Intern. J. Heat Mass Transfer*, **14**, 1705 (1971).
Monin, A. S., and A. M. Yaglom, *Statistical Fluid Mechanics; Mechanics of Turbulence*, The M.I.T. Press, Cambridge, Mass. (1965).
Newson, J. D., and A. C. Riddiford, "Limiting Currents for the Reduction of the Tri-iodide Ion at a Rotating Platinum Disk Cathode," *J. Electrochemical Soc.*, **108**, 695 (1961).
Schutz, G., "Untersuchung des Stoffaustausch-Anlaufgebietes in einem Rohr bei Vollausgebildeter Hydrodynamischer Strömung mit einer Elektrochemischen Methode," *Intern. J. Heat Mass Transfer Ser.*, **7**, 1077 (1964).
Shaw, D. A., "Mechanism of Turbulent Mass Transfer to a Pipe Wall at High Schmidt Number," Ph.D. thesis, Univ. Ill., Urbana (1976).
———, and T. J. Hanratty, *AIChE J.*, **23**, No. 2 (1977).
Shaw, D. A., "Measurement of Frequency Spectra of Turbulent Mass Transfer Fluctuations at a Pipe Wall," M.S. thesis, Univ. Ill., Urbana (1973).
Sirkar, K. K., and T. J. Hanratty, "Relation of Turbulent Mass Transfer to a Wall at High Schmidt Numbers to the Velocity Field," *J. Fluid Mech.*, **44**, 589 (1970).
Son, J. S., "Limiting Relation for the Eddy Diffusivity Coefficient Close to a Wall," M.S. thesis, Univ. Ill., Urbana (1965).
———, and T. J. Hanratty, "Limiting Relation for the Eddy Diffusivity Close to a Wall," *AIChE J.*, **13**, 689 (1967).
Theory of the Flowrator, Catalog Section 98-A, Fisher and Porter Co., Hatboro, Pa. (1947).
Van Shaw, P., "A Study of the Fluctuations and the Time Average of the Rate of Turbulent Mass Transfer to a Pipe Wall," Ph.D. thesis, Univ. Ill., Urbana (1963).
Vulterin, J., and J. Zyka, "Investigation of Some Hydrazine Derivatives as Reductimetric Titrants," *Talanta*, **10**, 891 (1963).

Manuscript received May 27, 1976; revision received October 6 and accepted October 8, 1976.

Vortex Free Downflow in Vertical Drains

NORTON G. MCDUFFIE

Chemical Engineering Department
The University of Calgary
Calgary, Alberta T2N 1N4
Canada

Performance of flush drains in gas-liquid separators has been evaluated under conditions of variable gas flow and pressure, liquid height and flow rate, and drain diameter. For the two-phase flow regime, liquid flow rate is found to be independent of vessel pressure or gas flow rate. Transition from weir control to orifice control of liquid flow rate occurs over the range of H/D ratios from 0.4 to 1.5. In this range, simple orifice equations do not apply. Revised design equations for use to determine minimum liquid height for maintenance of single-phase downflow in gas-liquid separators are presented.

SCOPE

Downflow of liquid from a process vessel presents several design problems as outlined by Simpson (1968). The whole downcomer system should be designed to minimize cavitation problems, both in free flow and pumped systems (Anderson et al., 1971; Simpson, 1968). This consideration generally determines dependent values of drain diameter and required head in the process vessel, whether it be predominantly pressure or hydraulic head. A consequent requirement in most processes is that the height of the lower phase (phase 1, Figure 1) be sufficient to repress coning and subsequent mixing of phases (phases 1 and 2, Figure 1); if the second phase is gas, this is especially crucial for vessels supplying liquid to centrifugal pump suctions. Vortex flow and subsequent entrainment can be restricted by installation of cross baffles or vanes in the downcomer mouth. After this precaution is taken, the minimum height H of the lower phase is determined by requirements for stable, vortex free downflow. The study discussed herein was undertaken to attempt to verify and extend bases suggested by Simpson (1968) for determination of minimum lower-phase levels for vessel and drain design.

CONCLUSIONS AND SIGNIFICANCE

This study shows that for a pressurized gas-liquid separator drain system, the maximum attainable steady state liquid superficial velocity V is a function of H/D alone, even in the two-phase flow region (low viscosity liquid). An increase in gas pressure and gas downflow rate does not affect V in the two-phase region. At H/D values from 0.4 to 3, the transition from full liquid flow to two-phase flow occurs at Froude numbers given by the relationship $1.6\,(H/D)^2 \leq Fr \leq 5.1(H/D)^2$. The same relationship applies to liquid superficial Froude numbers attained in two-phase downflow. The transition relationship essentially corroborates Simpson's assumption (1968) utilizing Kalinske's equation (1941) for free downflow with a constant discharge coefficient. The Harleman theoretical

equation proposed by Simpson (1968) as a design equation for minimum liquid height for gas-liquid separators is unsatisfactory for that use. It is recommended that the relationship $Fr \leq 1.6(H/D)^2$ be used for this determination instead. In the range of H/D values from 0.4 to 1.5 and maximum single-phase liquid velocities, weir flow exists at the lowest H/D ratio, and there is a transition to an orifice type of flow with increasing ratios. Within the 0.4 to 1.5 H/D range, simple orifice flow relationships do not apply.

There are several different flow regimes to be considered in development of mathematical models for vortex free downflow. If we consider the case of a gas-liquid containing vessel with a short flush drain under conditions of free hydraulic flow, these regimes can be classed with increasing values of H/D, where D is the diameter of the drain. As reviewed by Simpson (1968), the initial regime is essentially a self-venting weir flow in which liquid flows over the perimeter of the drain and down its inner surface without any appreciable concurrent air flow. Souders et al. (1938) successfully applied a modification of the Francis equation for circular downcomers in bubble trays. The equation for values of H/D less than 0.25, in terms of the Froude number, is

$$Fr = 2.36(H/D)^{1.5} \tag{1}$$

This differs only slightly from the equation used by Anderson et al. (1971) in related hydraulic studies. Their equation for circular weir flow in the same form becomes

$$Fr = 2.31(H/D)^{1.5} \tag{2}$$

If liquid downflow is restricted to a lower rate than that given by Equation (1) or (2), the downcomer will begin to fill with liquid.

At values of H/D greater than about 0.25, with Froude numbers over 0.3 or so (up to Froude numbers of 0.55 in some instances), the second flow regime begins (Simpson, 1968). In this regime, gas is sucked into the flowing liquid stream through a cone formed in the drain (Kalinske, 1941). For a given drain diameter and length with free flow, increasing values of H are accompanied by increasing gas intake rates which peak and then decrease to zero as the liquid heights become sufficient to stop gas inspiration. The limiting curve for this situation of full liquid flow has been plotted from Kalinske's free downflow data by Simpson (1968). If this curve is analyzed, it can be seen to follow a simple equation;

$$Fr = 4.4(H/D)^2 \tag{3}$$

Simpson (1968) used the theoretical equation developed by Harleman et al. (1959); namely

$$Fr = 3.2(H/D)^{2.5} \tag{4}$$

for unstable downflow limits for stratified downflow as the theoretical expression of the above limit. Since the curve for the equation of Harleman et al. fell below Kalinske's experimental curve in the range plotted, Simpson (1968) suggested that the Harleman theoretical equation could be used as a conservative equation for drain sizing in gas-liquid separators. Plots of Equations (3) and (4) approach each other at values of H/D over 1.0 and intersect at a value of 1.89 for H/D. Thus, it would appear preferable to use some other equation for design of gas-liquid separators, especially for values of H/D greater than 1.0. In addition, such an equation should apply to flow under pressure as well as to free downflow.

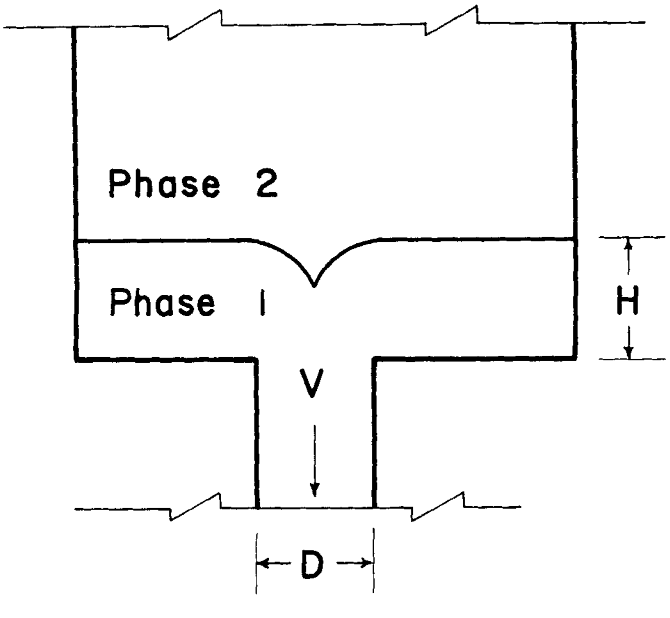

Fig. 1. Typical downflow system.

EXPERIMENTAL

Two different apparatus setups were used to test the equations for downflow. The first consisted of a flanged Perspex tube of 7.6 cm I. D. closed at both ends. Air pressure was controlled at a top inlet and measured by a calibrated oil manometer. Air flow rates were monitored by use of a rotameter. Water flow was measured with a calibrated rotameter. The water was introduced at the top of the liquid in the vessel in four quadrants separated by thin baffles extending to the bottom to inhibit vortex formation. The drain consisted of a 6.0 mm diameter hole in the bottom 12.7 mm thick cap of the vessel. This produced a drain consisting of a very short tube. With this apparatus, water flow rates along with corresponding vessel pressures and liquid heights were recorded at the points of incipient air inspiration into the drain. Further readings were taken in the two-phase flow regions at increasing air flow rates and pressures.

The second setup was devised to test flows through a 3.0 mm diameter drain in a 7.6 cm diameter vessel. Just as in the first setup, the drain length was 12.7 mm. Baffles were not used in this setup, but procedure was adapted to eliminate drain vortex effects. The procedure was begun by stoppering the drain and then partially filling the air pressurized vessel with water. The water was allowed to stand until motion had subsided. A drop of dye was added to one side to indicate any vortices; then the drain plug was removed, allowing the water to flow into a 2.5 cm I. D. receiver while the air pressure in the vessel was maintained by three-stage pressure control. Total water flow was measured by use of a transducer-storage oscilloscope arrangement producing a photographic record of receiver bottom static pressure. Water flow rates as well as heights of water remaining in the separator vessel were obtained from differential and integral readings of these records. For the conditions used the water flow rates were very constant up to the air inspiration points, whereupon they dropped decidedly. Within the second phase (with air inspiration), the photographic records of liquid height vs. time followed an identical trace after transition from full liquid flow had occurred. From the records it was

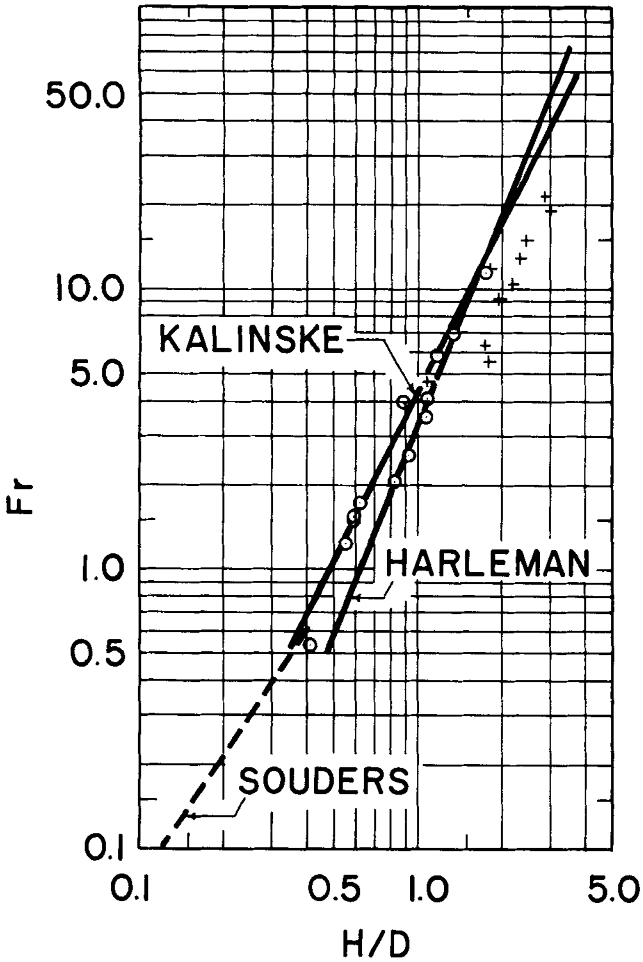

Fig. 2. Liquid superficial Froude numbers for steady state vortex free downflow for weir flow (Souders equation) and incipient inspiration (Kalinske and Harleman equations) as functions of *H/D* ratios at drain, points from this study, + for *D* = 3.0 mm and ○ for *D* = 6.0 mm.

possible to obtain quite accurate measurements of separator liquid heights at incipient air inspiration as well as liquid flow rates immediately preceding breakthrough. With sufficient care, drain vortices did not form (as indicated by dye streamlines); runs in which vortices did form were not used. In cases of higher flow rates a receiver water jet impact correction was necessary, but it was easily evaluated because of the constancy of flow rate.

RESULTS AND CONCLUSIONS

Results of this study are represented in Figures 2 and 3. The weir flow regime was found to persist up to values of *H/D* around 0.41. At and below this value of *H/D*, liquid flow was self-venting and air was not pulled down. Just above this value, air was inspired with the liquid to produce a vacuum in the vessel. At higher values of *H/D*, positive pressure was required to force air into the drain. In Figure 2, data points are plotted for liquid Froude numbers just at the point of incipient air inspiration for the *H/D* range beyond weir flow. In Figure 3, the pressure and equivalent total head required to maintain liquid flow and level at steady state just below the point of air inspiration are shown. At pressures below the points shown, full liquid flow persisted in the drain. Above, two-phase flow persisted. In addition, for a given liquid flow rate and drain diameter, *H* was fixed at all air flow rates used and independent of increasing vessel pressures so long as two-phase flow occurred.

In Figure 2, lines are drawn for the Souders equation, the Harleman theoretical equation, and a form of Kalinske's equation. The latter form is derived from Kalinske's equation for his experimental free flow data

$$Q = C(gD)^{1/2}H^2 \tag{5}$$

by setting the discharge coefficient C to his lowest experimental value of 3.36 (Kalinske, 1941). From Equation (5), the following conservative equation is obtained for the liquid Froude number:

$$Fr = 4.28(H/D)^2 \tag{6}$$

This equation is represented as Kalinske's equation in Figure 2. It can be seen that many of the experimental points for forced flow fall below the lines for both the Kalinske and the Harleman equations. A study by Lubin and Springer (1967) for free downflow in the water-air system with *H/D* values from about 0.6 to 1.3 gave results for initial dip formation which fell along a line that can be shown to represent Harleman's (1959) theoretical equation [Equation (4) herein]. In Lubin and Springer's plot, the experimental results are included together with those for various liquid-liquid systems. Their experimental results differ from those of Harleman's (1959), which are best represented by

$$Fr = 2.05(H/D)^2 \tag{7}$$

The present study extends the air-water system data up to an *H/D* value of 3.0 with a pressurized flow system instead of a free flow system. The relationship encompassing experimental points herein also encompasses Lubin and Springer's (1967) points for the free flow air-water system; namely

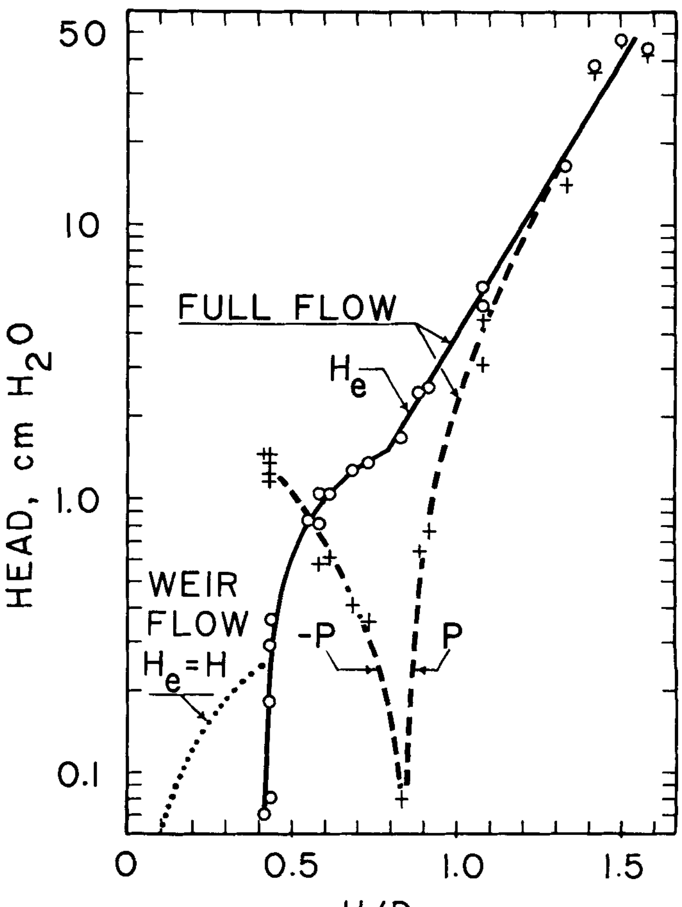

Fig. 3. Pressures associated with steady state water downflow through a 6.0 mm diameter by 12.7 mm length drain. P = static pressure above liquid surface and H_e = controlling pressure drop across drain.

$$1.6(H/D)^2 \leq Fr \leq 5.1(H/D)^2 \quad (8)$$

The Harleman theoretical equation proposed by Simpson (1968) as a design equation for minimum liquid height for gas-liquid separators is not satisfactory for that use. Instead, it is recommended that the following relationships be used for conservative design in critical applications:

$$Fr \leq 1.6(H/D)^2 \quad (9)$$

$$Q \leq 1.26(g'D)^{1/2}H^2 \quad (10)$$

The equivalent head H_e represented in Figure 3 is effectively the total static and pressure head required to maintain full liquid flow at steady state. The curve for weir flow, where the static head of liquid height H is the total driving force, is shown for comparison. Maximum steady state vacuum is attained in sucking flow just at the transition from weir flow to full flow. At this point, flow pressure drop (H_e) is only about 0.1 cm of the total available head represented by the length of the downcomer (1.27 cm) plus H (about 0.25 cm). Liquid Euler numbers decrease from a maximum of about 0.9 in the H/D range of 0.5 to 1.0 to about 0.6 at H/D values near 1.5. Thus, true orifice entrance control is realized only at the higher flow velocities. Inferences normally given for use of standard orifice equations with decreasing discharge coefficients at lower velocities for efflux of liquid from a vessel are thus incorrect for values of H/D between 0.4 and 1.5. For the weir flow regime, superficial Euler numbers decrease to zero with H/D, since from Equation (1)

$$Eu = \frac{2.36}{\sqrt{2}}(H/D) \quad (11)$$

This experimental study has been directed to gas-liquid systems; the experimental equation of Harleman et al. (1959) should still be used for design related to selective removal of stratified liquid or liquid from lower phases of decanters, etc. Converted for volumetric flow, this equation is

$$Q = 1.61\sqrt{g'}\,H^{2.5} \quad (12)$$

It is interesting to note that the diameter of the drain does not appear in this equation. The only geometric factor is the lower phase height. This is a consequence of the assumption of a point sink.

NOTATION

C = dimensionless discharge coefficient
D = inside diameter of drain, cm
Eu = Euler number = $V/\sqrt{2g'H_e}$
Fr = Froude number = $V/\sqrt{g'D}$
g = acceleration of gravity, cm/s^2
g' = $g(\rho_1 - \rho_2)/\rho_1$
H = height of lower liquid phase above bottom of vessel, cm
H_e = equivalent head across drain = $H + P + L$ for full pipe flow or = H for weir flow, cm H_2O
L = length of drain, cm
P = static pressure head above lower liquid phase, cm H_2O
Q = liquid volumetric flow rate, cm^3/s
V = liquid superficial velocity in drain, cm/s
ρ = density, g/cm^3

Subscripts

1, 2 = designation of phase or stratified layer number from bottom of vessel

LITERATURE CITED

Anderson, A. G., P. P. Vaidyaraman, and C. S. Chu, "Hydraulics of Long Vertical Conduits and Associated Cavitation," *U.S. E.P.A. Water Pollution Control Research Series 11034 FLU 06/71* (1971).
Harleman, D. R. F., R. L. Morgan, and R. A. Purple, "Selective Withdrawal from a Vertically Stratified Fluid," *Reprint p. 10-C-1*, 8th Congress, International Association for Hydraulic Research, Montreal, Canada (Aug., 1959).
Kalinske, A. A., "Hydraulics of Vertical Drain and Overflow Pipes," in "Investigations of the Iowa Institute of Hydraulic Research 1939-1940," J. W. Howe, ed., *University of Iowa Studies in Engineering*, Bull. No. 26, p. 26 (1941).
Lubin, B. T., and G. S. Springer, "The Formation of a Dip on the Surface of a Liquid Draining from a Tank," *J. Fluid Mech.*, **29**, 385 (1967).
Simpson, L. L., "Sizing Piping for Process Plants," *Chem. Eng.*, **75**, No. 13, 192 (1968).
Souders, M., Jr., R. L. Huntington, H. G. Corneil, and F. L. Emert, "Performance of Bubble-Plate Columns, Froth Heights and Pressure Differentials," *Ind. Eng. Chem.*, **30**, 86 (1938).

Manuscript received September 2, and accepted November 8, 1976.

The Characterization of Fixed Beds of Porous Solids from Pulse Response

The response of a fixed bed of porous particles through which fluid flows to a pulse of tracer gas has been analyzed by a number of methods. It is found that analysis of the Fourier transformed and the Laplace transformed response curves for small values of the parameters of the integrals provides accurate estimates of intraparticle diffusivity and Peclet group for axial dispersion when additional information provided by the experiments is used to resolve interaction between the process coefficients in the dynamic response.

HASEEB BASHI
and
DONALD GUNN
Department of Chemical Engineering
University College
Singleton Park, Swansea, Wales

SCOPE

The response of a fixed bed of porous particles through which a gas flows to an injected pulse of tracer gas may be used in theory to provide estimates of the process coefficients, the coefficient of axial dispersion, the intraparticle diffusivity, the particle to fluid mass transfer coefficient, and the interstitial fluid velocity. The response

Correspondence concerning this paper should be addressed to Donald Gunn.