CHAPTER I
INTRODUCTION

Two-phase gas/liquid flow is important in a variety of chemical engineering applications, such as the simultaneous transport of gas and oil in horizontal pipelines or in vertical wells, condensate return lines flashing into steam, vapor-liquid feed lines entering distillation columns and refrigerant-return lines, that must maintain a specific vapor-liquid ratio for efficient operation. The major complexity in two-phase flows results from the growth and collapse of the gas-liquid interfaces that can give rise to various flow regimes. The linear velocities of gas and liquid phases in each flow regime are dictated by the system's thermohydraulic behavior. In fact, the liquid in a two-phase flow can be accelerated to velocities approaching or exceeding vapor velocities. Such high velocities can cause "erosion corrosion" in equipment and piping systems. In this study, bubble and slug flow patterns were produced by varying the inlet air and water flow rates. The superficial gas velocities ranged from 0.0029 to 0.7042 m/s, while the superficial liquid velocity was varied from 0 to 0.1470 m/s have been tested. The resulting flow types were observed and filmed with a camcorder. Once a sufficient range of inlet conditions has been observed, flow pattern maps of bubble and slug flow could be created for the vertical tube systems. Moreover this thesis also investigated relation between rise velocity of single slug and the slug length, void fraction at different air and water flow rate, rise velocity of continuously generated slugs, and air-lift pump operation within slug flow.

The followings are some background about two-phase flow in a vertical tube.

1.1 Flow Regimes

For vertical pipes, there are four main regimes, shown in Figure 1.1, and which occur successively at ever-increasing gas flow rates:

(a) *Bubble* flow: There is a continuous liquid phase, and the gas phase is dispersed as bubble within the liquid continuum. The bubble travel with a complex motion within the flow may be coalescing and are generally of non-uniform size.

(b) *Slug* flow: This flow regime occurs when the bubble size tends toward that of the channel diameter, and characteristic bullet-shaped bubbles are formed.

(c) *Annular* flow: This configuration is characterized by liquid travailing as a film on the Channel walls, and gases flowing through the center. Part of the liquid can be carried as droplets in the central gas core.

(d) *Mist* flow: in which the velocity of the continuous gas phase is so high that it reaches as far as the tube wall and entrains the liquid in the form of droplets.

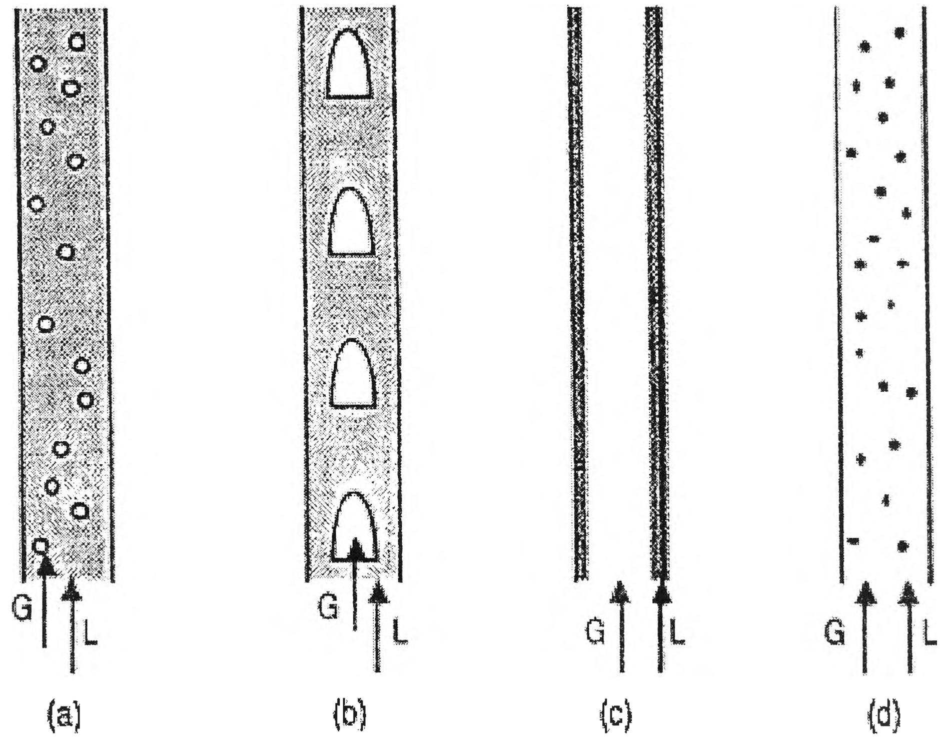

Figure 1.1 Two-phase flow regimes in a vertical tube: (a) bubble, (b) slug, (c) annular, and (d) mist flow. In each case, the gas is shown in white, and the liquid is shaded or black.

1.2 Bubble Flow

Figure 1.2 shows gas bubbles and liquid in upwards co-current flow. Consider the plane A–A, drawn so that it lies entirely in the liquid. If $\overline{U}_l$ is the mean upwards liquid velocity across A–A, continuity requires that:

$$\overline{U_l} = \frac{G + L}{A} \tag{1}$$

(Over a relatively shot vertical span, the pressure varies little and the gas bubbles have an essentially constant volume.) Thus, the gas bubbles just below A–A are rising relative to a liquid that is already moving at a velocity $\overline{U}_l$, so that the velocity of the gas bubbles is:

$$v_g = \overline{U_l} + \overline{U}_b = \frac{G + L}{A} + \overline{U}_b \tag{2}$$

in which U_b is the bubble velocity rising into a stagnant liquid. But the total volumetric flow rate of gas is:

$$G = \varepsilon A v_g \tag{3}$$

So that the void fraction is given by:

$$\frac{G}{\varepsilon A} = \frac{G+L}{A} + \overline{U}_b \qquad \text{or} \qquad \varepsilon = \frac{G}{G+L+U_b A} \tag{4}$$

This relation for the void fraction holds within certain limits for G and/or L *negative* – that is, for *downflow* of one or both phases.

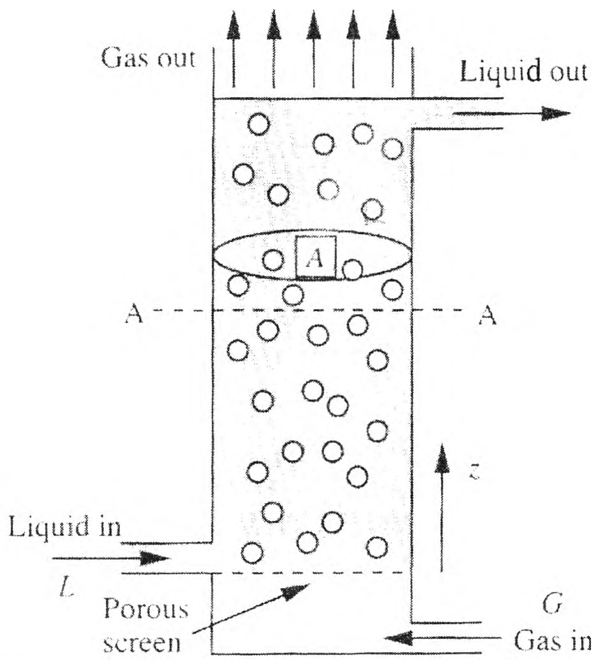

Figure 1.2 Bubble flow.

Concerning the pressure gradient in the upwards vertical direction, note that the density of the liquid, which occupies a fraction (1-ε) of the total volume, is much greater than that of the gas. Also, for the relatively low liquid velocities likely to be encountered in the bubble-flow regime, friction is negligible. Therefore, the pressure gradient is given fairy accurately by considering only the hydrostatic effect:

$$\frac{dp}{dz} = -\rho_l g(1-\varepsilon) \tag{5}$$

1.3 Slug Flow

The general situation is shown in Figure 1.3(a), in which the gas (air) and liquid (water) are traveling upwards together at individual volumetric flow rates G and *L* respectively, in a tube of internal diameter *D*. In general, there will be an upward liquid velocity u_{Lm} across a plane A—A just ahead of a gas slug. By applying continuity and considering the gas to be incompressible over short distances, the total upwards volumetric flow rate of liquid across A—A must be the *combined* gas and liquid flow rates entering at the bottom, namely, $G + L$. The *mean* liquid velocity at A-A is therefore $U_{Lm} = (G + L)/A$, where A is the cross-sectional area of the tube.

Next, consider Figure 1.3(b), which shows a somewhat different situation —that of a single bubble, which is moving steadily upwards with a rise velocity U_b in an otherwise *stagnant* liquid. For liquids such as water and light oils that are not very viscous, the situation is one of *potential* flow in the liquid. Under these circumstances, Davies and Taylor [2] used an *approximate* analytical solution that gave a specific value for c in the equation:

$$U_b = c\sqrt{gD} \tag{6}$$

in which g is the gravitational acceleration and c = 0.33. Experimental evidence shown that constant should be c = 0.35. The situation of Figure 1.3 (a) is now shown enlarged, in Figure 1.3(c). The slug is no longer rising in a stagnant liquid, as in Figure 1.3(b), but in a liquid whose mean velocity just ahead of it is U_{Lm}. Further, near the "nose" O of the slug—at the center of the tube, where the velocity is the highest—the liquid velocity will be somewhat larger, namely, about $1.2U_{Lm}$, as shown by Nicklin, et al. [1], provided that the Reynolds number between slugs exceeds 8,000. Therefore, the actual rise velocity of the slug will be

$$U_s = 1.2\frac{G+L}{A} + U_b = 1.2\frac{G+L}{A} + c\sqrt{gD} \tag{7}$$

Now, by conservation of the gas, we must have

$$G = U_s A \varepsilon \tag{8}$$

in which ε is the *void* fraction (the fraction of the total volume that is occupied by the gas). Hence, eliminating U_s between equations (7) and (8), we obtain

$$\frac{G}{\varepsilon A} = 1.2\frac{G + L}{A} + c\sqrt{gD} \tag{9}$$

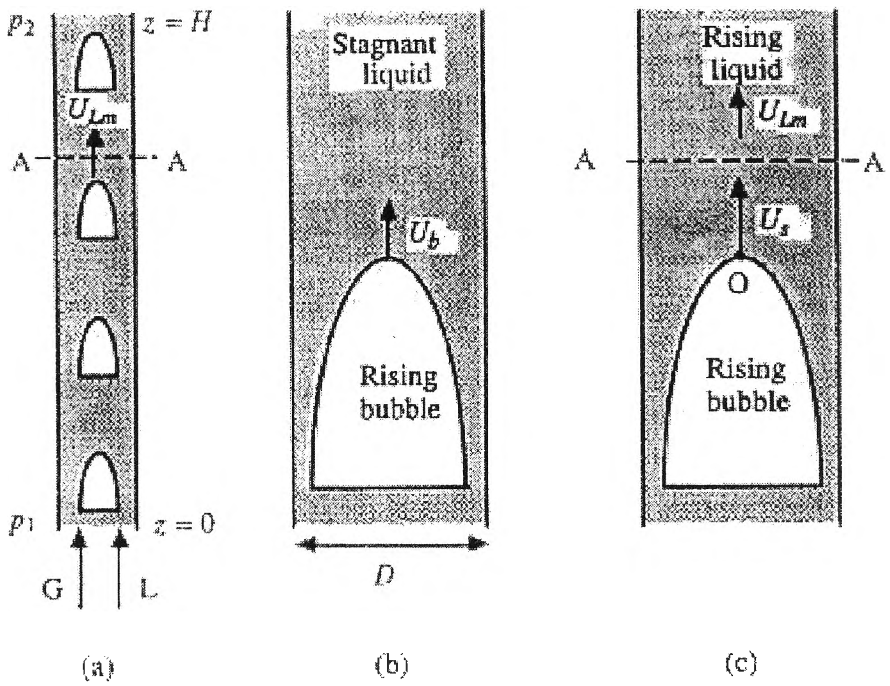

Figure 1.3 Two-phase flow in a vertical tube: (a) gas and liquid ascending, (b) bubble rising in stagnant liquid, (c) bubble rising in moving liquid.

If G and L are known, equation (9) gives the void fraction, which is very important in determining the pressure drop in a tube of height H. Also note that the weight of the liquid, which occupies a fraction (1 — ε) of the total volume, is much greater than that of the gas. Therefore, the pressure drop is given to a first approximation by

$$p_1 - p_2 = p_L g H(1 - \varepsilon) \tag{10}$$

A secondary correction to (10) would include the wall friction on the liquid "pistons" between successive gas slugs.

1.4 Gas-Lift Pump Operation

Figure 1.4 shows a gas-lift pump, in which the buoyant action of a volumetric flow rate G of gas serves to lift a volumetric flow rate L of liquid from a height H_0 in a reservoir to a height H in a vertical pipe of diameter D and cross-sectional area A, in which slug flow may be assumed. Neglect liquid friction in both the supply pipe and in the vertical pipe.

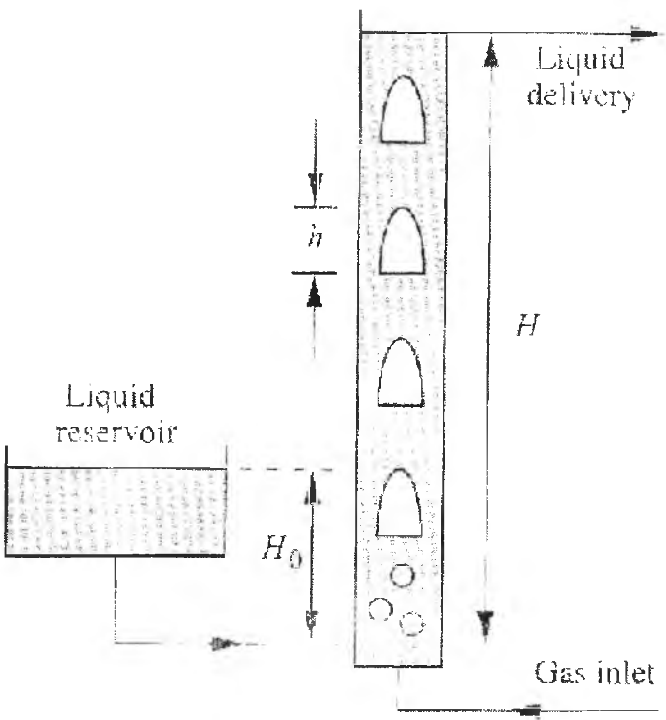

Figure 1.4 Gas-lift pump.

From hydrostatics, the pressure at the base of the column is obtained in two ways:

$$\rho_l g H_0 = \rho_l g H (1 - \varepsilon)$$

which gives:

$$\varepsilon = 1 - \frac{H_0}{H} \tag{11}$$

From equation (6) and (7) with $L = 0$:

$$\frac{G}{\varepsilon A} = 1.2\frac{G}{A} + U_b = 1.2\frac{G}{A} + c\sqrt{gD}$$

$$\frac{G}{A}\left(\frac{1}{\varepsilon} - 1.2\right) = c\sqrt{gD}$$

$$\frac{G}{A} = \frac{c\sqrt{gD}}{\left(\frac{1}{\varepsilon} - 1.2\right)} \tag{12}$$

Substitute equation (11) in equation (12). It gives superficial velocity of gas, which then suffices to achieve the desired liquid flow rate.