Chapter - 2

Literature Survey
2.1 Introduction

A detailed literature survey of two-phase flow which includes the basic definition of two-phase flow, various models like slip, mixture, two-fluid etc. and their advantages, different problem areas in the modelling of two-phase like interfacial transfer terms, interfacial area etc., various flow regimes, maps and transition criteria have been presented in this chapter.

Two-phase flow is a term covering the interacting flow of two phases (gas, liquid or solid) where the interface between the phases is influenced by their motion. This chapter is concerned only with gas-liquid and vapour-liquid two-phase flows, which cover bubbly to churn to slug flow.

Two-phase flow takes place in a wide range of industrial plants, boilers and nuclear reactors. The requirement for economic designs, optimisation of the operating conditions and assessment of safety factor leads to the need for quantitative information. Increase in rating, competition in capital and operating costs and the importance of reliability and safety have accentuated this need in the last few decades. The solution of these types of problems may involve a large number of experimental investigations of a particular geometry, two-phase flow regimes and various optimisation techniques.

Two-phase flows can be broadly classified as vertical and horizontal flow. Among the vertical flows, we can have co-current upward, co-current downward, counter current flow with the liquid flow downward and gas flow upward. The other possibility of liquid flow upward and gas flow downward is not possible.

In general, inclined flows are analysed as combination of vertical and horizontal flow. In this thesis, analysis and review is carried out for co-current upward flow.
2.2 Models for Two-Phase Flow

Besides the different averaging methods [60-61] for developing a model, there are two fundamentally different formulations of the macroscopic balance equations for two-phase flow systems. These are the mixture (or diffusion) model and two-fluid model. The mixture model is formulated by considering the motion of a two-phase mixture as a whole in terms of mixture momentum equation. The relative motion between phases is taken into account by a closure equation for relative velocity. The most important assumption associated with the mixture model is that a strong coupling exists between the motions of two-phases [62]. This implies that the mixture model is an accurate approximation to the two-fluid model whenever there is a sufficient interaction time or length for two phases. This model may not appropriately describe certain two-phase problems involving a large acceleration of one phase, with respect to the other phase or a flow in very short tube. In these cases inertia terms of each phase should be considered separately by the use of two-fluid model.

2.2.1 Mixture (or Diffusion) Models

Depending on the form of the closure equation for the relative velocity and on the treatment of the thermal non-equilibrium between phases, a number of different mixture models have been proposed, i.e., the homogeneous flow model, slip flow model and drift-flux models. In the case of homogeneous model, the phase velocities are assumed same i.e. slip \((u_s/u_i)\) is 1.0. However, it is an unreasonable assumption. In the case of vertical upward flows, due to buoyancy, slip is always greater than 1.0 and more so in the case of high density liquid metals. However at high fluid velocity slip tends toward 1.0.
In the case of slip flow models, in addition to solving continuity equation of phases, combined momentum and energy equations, empirical relations based on experiment have to be provided with either for $\alpha$ or for slip. Depending upon the geometry, flow rates, fluids used, there are many empirical relations available in the literature [63-70]. Most of them are confined to steam-water or air-water flows. The main disadvantages of these models are, they are valid for specific system and likely to be invalid, when applied to different flows, geometries or operating at different ranges. Very little work has been carried out for the case of liquid metal flows.

J. Dounan et al. (1985) relation is based on air-water flows in the rectangular channels and they have developed models for both bubble and slug flow [71]. Serizawa and Michiyoshi relation is derived from the experimental data from light density liquid metals (sodium and potassium) steam-water and air-water [72-73]. The empirical relations developed by them for slug flow predicted very low void fraction. Unger et al. (1986-87) have extended Serizawa and Michiyoshi (1973) and developed empirical relation for $\alpha$ as a function of fluid density ratio, Froude number of liquid metal flows and quality for steam-mercury upward cocurrent flow [74-76].

El-Boher et al. (1988) have developed an empirical relations based on experiments conducted in air-water, steam-mercury and steam-alloy of lead and bismuth and void fraction measured by them using quick closing valve method. The void fraction was expressed as a function of volumetric flow rate ratio, Froude number, viscosity ratio, liquid superficial Reynolds number and Weber number [77].

In the case of drift flux model, void fraction is expressed as a function of vapour and mixture volumetric flux, drift velocity and distribution parameters [78-83]. The main advantage of this model is, cross-sectional variation in the phase parameters like velocity, void distributions etc. can be accounted explicitly. Some of the well-known drift
flux models are developed by Chexal et al. (1996) for steam-water, air-water, hydrocarbon etc. for wide range of pressures, flows and void fraction. This correlation not only covers co-current upward flow but co-current downward flow, counter current and also flows inclined to horizontal plane [84-86]. However, there are no drift flux models developed for liquid metals. Since the thermal and transport parameters of the liquid metals differ widely from that of water (density, viscosity, surface tension wet-ability, conductivity etc.), we cannot use directly empirical relations based on water-steam or air-water.

The main disadvantages of all these models are, the equations do not take into consideration explicitly the various flow regimes. They also do not account for different initial conditions (bubble size etc.).

2.2.2 Two-Fluid Models

Two-fluid models [60,79,87-93] are formulated by considering each phase separately in terms of two sets of conservation equations governing the balance of mass, momentum and energy of each phase. Since the macroscopic field is not independent of the other phases, the interaction terms, which couples the transport of mass, momentum and energy of each phase across the interface, appear in the field equation. For liquid metal two-phase flows, Mond and Sukoriansky (1984), Eckert et al. (1993) have studied the flow by two-fluid model with multibubble assumption [94-95].

General three-dimensional two-fluid model is obtained by Ishii and Delhaye et al. by using temporal or statistical averaging [60-61,87,93]. The model is expressed in terms of two sets of conservation equations governing the balance of mass, momentum and energy in each phase. Since the average fields of one phase are not independent of the other phase, the interaction terms appear in the field equations as source terms. These are too complex for analysis. Hence based on these equations, one-
dimensional two-fluid equations are derived.

Previous studies have indicated that unless phase momentum interaction terms are accurately modelled, the advantage of the two-fluid model over the mixture model disappears and numerical instabilities arise. At least two transient forces, i.e., the virtual mass and Basset forces exist in addition to drag and interfacial shear forces [90,96-103]. However, these transient momentum interaction terms are important under rapid transient conditions and for numerical-stability problems [62,71]. In spite of existing shortcomings for interfacial transfer terms, there is no substitute available for modelling accurately two-phase phenomena where two-phases are weakly coupled. The two-fluid model can more easily analyse the non-equilibrium effects by using two sets of balanced equations. This model is even more useful in studying the mechanical non-equilibrium effects, e.g., sudden mixing of two phases, transient flooding and flow reversal and rapid transient flow. Thermal non-equilibrium between phases can be analysed accurately by the inclusion of two-energy equations [60,91].

Another main advantage of two-fluid model is, it can take the flow structure explicitly. In the riser of the LMMHD systems, the flow consists of bubbly, churn and slug as the void fraction varies from 0.1 to 0.8. Fig. 2.1 shows the typical flow structure of the flow regimes discussed above. At higher velocities, annular flow occurs. In view of relatively low velocity, this flow does not occur in LMMHD system. Near the mixer, the flow is typically bubbly until the void fraction increases to around 0.25 to 0.3 [104-114], where bubble coalesces. In the bubbly flow, the vapour or gas phase is distributed as discrete bubbles in a continuous liquid flow. The bubbles may be small and spherical at one extreme and relatively large non-spherical shape at the other extreme. With the increase in gas flow rate, the bubble density increases and a point is reached where the dispersed bubbles become so closely packed that many collision occur and the rate of
agglomeration to larger bubbles increases sharply. This results in a transition to slug or churn flow.

Depending upon the distance from the mixer and flow rate, the flow either become slug or churn. In the slug flow, the size of the vapour or gas bubbles is approximately the diameter of the pipe and is called Taylor bubble. The nose of the bubble has a characteristic spherical cap and the gas in the bubble is separated from the pipe wall by a slowly descending liquid film. The liquid flow is contained in liquid slugs, which separate successive gas bubble. The length of the main gas bubble can vary considerably. The Taylor bubbles and liquid slug propagates at constant speed.

The process of developing a stable slug near the entrance section can be described as follows: At the mixer inlet the gas and liquid metal introduced form short liquid slugs and Taylor bubbles as shown in Fig. 2.1. A short liquid slug is known to be unstable and it falls back and merges with the liquid slug coming from below causing it to approximately double its length. In this process, the Taylor bubble following the liquid slug overtakes the leading Taylor bubble and coalesces with it as the slug between the two bubble collapses. This process repeats itself and the length of liquid slug as well as the length of Taylor bubbles increase as they move upward in the riser until the liquid slug is large enough to be stable and forms a competent bridge between two consecutive Taylor bubbles. Between the inlet and the position at which a stable slug is formed, the liquid slug alternately rises and falls and this is precisely the condition for churn flow. As the bubble goes up, the length of the entrance region increases to the extent that it can occupy the entire length of the of any test section. Thus, one should think of churn flow as an entrance phenomenon. Since in practice the pipes are of finite length, it would be useful to provide some estimates of the lengths over which churn flow is in predominant mode. With this objective, Taitel et al. (1980), Duckler and
Fig. 2.1: Schematic of Bubbly, Churn and Slug Flow
Taitel (1987) have developed a method for calculating the entry length required to develop stable slug flow [109,112]. The distance from the entrance to that length will be observed to be in churn flow pattern. From this, we find that this dimensionless entrance length for churning depends on the void fraction and flow velocities [108-111].

When the distance from the mixer is less than certain length called 'entrance length', the flow does not stabilize as slug instead becomes churn flow. This flow is formed by the breakdown of the large vapour bubbles in the slug flow. The vapour or gas flows in more or less chaotic manner through the liquid, which is mainly displaced to the channel wall. The flow has an oscillatory or time varying character; hence, the descriptive name, 'Churn' flow is given. These regions are also sometimes referred to as semi-annular or slug annular flow. In the churn flow, the liquid slug is too short to support a stable liquid bridge between two Taylor bubbles. The falling film around the bubble penetrates deeply into the liquid slug, creating a highly agitated aerated mixture at which point the liquid slug is seen to disintegrate and to fall in a rather chaotic fashion. The liquid re-accumulates at a lower level at the next slug, where the continuity is restored and the slug then resumes its upward motion. Thus, we observe the oscillatory motion of the liquid which, we consider the characteristic identification of churn flow.

One of the important links in the two-fluid model formulation is the closure equation for interfacial transfer terms. The difficulties arise due to the complicated motion and geometry of interfaces in a general two-phase flow. Furthermore, these equations should be expressed by the macroscopic variables based on proper averaging. As has been shown in detail [60,90-91], the interfacial transfer terms in a two-fluid model appear as averaging of local instant transfers of mass, momentum and energy. Because these terms appear as source terms in the basic equations, proper averaging alone is not sufficient to develop these basic equations. It is
essential to clarify different physical mechanism controlling these interfacial transfers as well as to identify important controlling parameters, which govern them.

The interfacial transfer terms are one of the important aspects of the two-fluid formulation modelling. However, there exist considerable difficulties in this area both in terms of experiments and modelling. These terms in two-fluid models specify the rate of phase change; momentum transfer and heat transfer at the corresponding interfaces between phases. The interfacial transfer terms are strongly related to the interfacial area and to the local transfer mechanisms, such as degree of turbulence near the interfaces. The interfacial transport terms are proportional to the interfacial area concentration and to the driving force [60,90-91,114-117].

The interfacial area concentration is defined as the interfacial area per unit volume of the mixture and characterises the first order geometrical effect. Thus, it must be related to the structure of the two-phase flow field. Since the interfacial area concentration is a parameter, which characterises the structure of a flow, its modelling should be based on the geometrical factor, void fraction and flow. The approach applied by Ishii and Mishima (1984) tells that the interfacial area concentration is related to the geometry of the flow such as the bubble sizes, bubble number density, bubble shape factor, roughness of the interfaces and hydraulic diameter, as well as hydrodynamic parameters such as the amount of droplet entrainment and void fraction.

When the flow transition takes place from bubble to slug or churn, the interfacial area is modified due to (1) coalescing of the bubble and (2) distortion of spherical bubble. Coalescing decreases the area whereas distortion increases. However, in general there is a net decrease in the interfacial area per unit volume.

A number of experiments have been carried out to estimate the interfacial area concentration. Among them are chemical absorption
technique [118-119] based on a pseudo-first order chemical reaction, light attenuation method [120-122] and photographic method [123-124]. The light attenuation method and photographic method require a flow channel with transparent walls. The optical Method have been reviewed and applied to obtain the data for bubbly flow by Veteau (1981) and Viteau-Charlot (1980-1981). The local measurement of the interfacial area concentration is possible using a local probe method [125-126]. A double sensored probe [127-129] method has also been suggested.

A further experimental study utilising these methods for measuring the interfacial area concentration and careful analysis of data and measurement techniques is highly desirable. Such detailed measurements of the local quantities of two-phase flow will greatly increase the understanding of interfacial transport phenomena, structure of two phases and regimes. For a churn flow the interfaces around the large bubbles become very irregular due to turbulent motions. Ishii-Mishima (1984) obtained the interfacial area concentration in bubbly and slug flow with the estimated small bubble size of 2.0 mm. It indicates that there is a drastic reduction in the interfacial area density when flow transitions take place between bubbly flow to slug or churn. In the case of the slug flow, interfacial area density depends also on void fraction in the slug [130-131]. Based on this analysis, Ishii, Zuber, Mishima and others have developed appropriate time averaged relations for drag force, virtual mass force etc. [90,96,100].

Thus, in two-phase flow systems, the void fraction and interfacial area concentration are two of the most important geometrical parameters. Since these two parameters appear in the balance equations as additional variables, two empirical relations should be supplied for void fraction and interfacial area. For the interfacial area concentration the relation may be given by functional relation or by a transport equation [60,91]. Thus the knowledge of interfacial area concentration is indispensable in the
modelling of two-fluid.

In this thesis, the model developed by Ishii-Mishima (1984) for interfacial drag and virtual mass force has been used [91]. Taitel-Bornea- Duckler (1980) criteria for flow transition are assumed [109].