CHAPTER 2

LITERATURE SURVEY

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A fundamental requirement for the systematic study of any chemical unit operation is the knowledge of its hydrodynamic characteristics and for bubble column reactor, the equilibrium size of bubble, hold-up, specific gas-liquid surface area, dispersion coefficients and mass transfer coefficient are important parameters for establishing the degree of chemical conversion. Bubble velocity and its residence time in the column are controlled by the properties of the liquid or slurry phase in addition to its size and configuration. These phase properties are in turn also influenced by the operating conditions such as temperature, pressure, catalyst particle size, and size range. Thus it is obvious that the selection of all these interacting design parameters for a real reacting system is a difficult task.

In the following sections of this chapter some aspects of hydrodynamics and important design parameters for slurry bubble column will be presented. This includes the following topics:

2.1. Flow Regime.

2.2. Bubble Dynamics.

2.3. Gas hold-up.

2.4. Liquid phase back mixing or dispersion coefficient.

2-1- Flow Regime

In bubble column reactors the hydrodynamics, transport and mixing properties such as pressure drop, holdup of various phases, fluid-fluid interfacial areas and inter-phase mass and heat transfer coefficients depend on the prevailing flow regime [Shah et al., 1982]. These can be categorized as:

2-1-1-Homogenous Regime or Bubbly Flow

This regime is characterized by almost uniformly sized bubbles with equal radial distribution. This regime occurs if the superficial velocity is less than 0.05 m/s and the rise velocity of the bubbles ≤ 0.3 m/s. The bubble size distribution is narrow and a roughly uniform bubble size, generally in the range 1 - 7 mm is found [Shah et al., 1982].

2-1-2-Heterogeneous or Churn -Turbulent Regime

When the superficial gas velocity U_G reaches the value U_{Trans}, coalescence of the bubbles takes place to produce the first fast rising "large bubbles". This changes the hydrodynamic picture dramatically. The hydrodynamic regime is commonly referred to as the heterogeneous flow regime.

In the heterogeneous regime, small bubbles combine in clusters to form large bubbles in the size range 20-70 mm. The large bubbles are non spherical and of varying form with very mobile and flexible surfaces, for instance, spherical caps [Deckwer et al., 1981]. They travel up through the column at high velocities (1-2 m/s) in a more or less plug flow manner. These large bubbles have the effect of churning up the liquid phase. The large bubbles are mainly responsible for the throughput of gas through the reactor at high velocities. Small bubbles, which co-exist with large bubbles in the churn-turbulent regime, are "entrained" in the liquid phase, and as a good approximation have the same back mixing characteristics of the liquid phase. The heterogeneous or churn-turbulent regime is most commonly encountered in industrial bubble columns where the two bubble classes with significantly differing sizes and rise velocities can be observed [Krishna and Sie, 2000]. However the homogeneous flow regime is encountered as well, particularly in pressurized BCSRs [Deckwer and Schumpe, 1993].

2-1-3-Slug Flow

In small diameter columns at high gas flow rates, large bubbles are stabilized by the column wall leading to the formation of bubble slugs. Bubble slugs can be observed in columns of diameters up to 0.15 m [Shah et al., 1982].

The type of sparger used, physico-chemical properties of liquid, and the liquid velocity can affect the transition between the flow regimes [Shah et al., 1982]. The dependence of flow regime on column diameter and gas velocity can be roughly estimated from Figure (2-1).
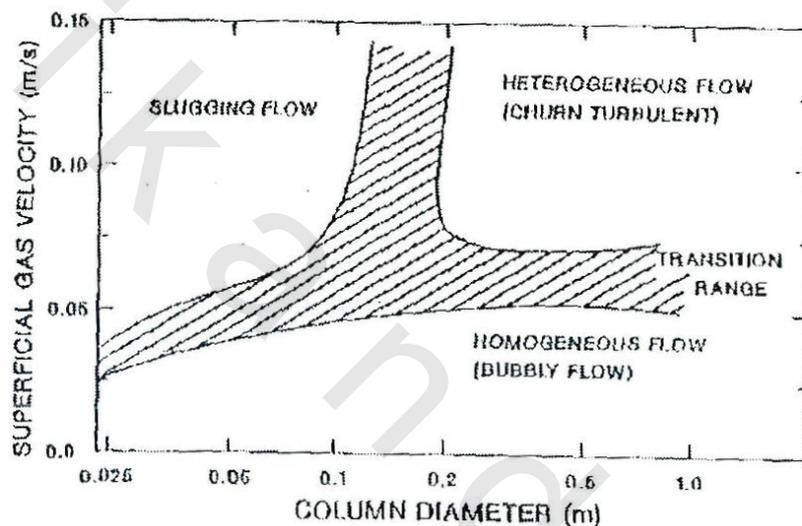

Figure (2-1): Types of Flow Regimes [Shah et al., 1982].

2-2-Bubble Dynamics

Bubble size, bubble rise velocity, bubble size distribution and liquid or bubble velocity profiles have a direct bearing on the performance of bubble columns. The majority of bubble column reactors used in industry are operated at high pressures. Studied systems showed that up to pressures of 1.6 MPa, there is no effect of pressure on the gas holdup and mean bubble diameter if the gas velocity is corrected to take into account the pressure in the column [Shah et al., 1982].

The mean rise velocity of the bubbles in a bubble swarm is equivalent to the interstitial gas velocity V_G^* which follows from the superficial gas velocity U_G [Niclin, 1962]

$$V_G^* = U_G / \varepsilon_G \qquad (2\text{-}1)$$

In the bubbly flow regime the rise velocity may vary from 0.03 to 0.2 m/s. If the flow is churn-turbulent, the rise velocities are considerably higher ≥ 0.8 m/s. The transition from bubbly to churn-turbulent flow is usually accompanied by a sharp increase in V_G^* [Shah et al., 1982].

2-3-Gas Holdup

Gas holdup is one of the most important parameters characterizing the hydrodynamics of bubble columns. It can be defined as the percentage by volume of the gas in the two or three phase mixture in the column. It has two fold applications. On one hand, the gas holdup in two phase systems gives the volume fraction of the phases present in the reactor and hence their residence times. On the other hand, the gas holdup in conjunction with the knowledge of mean bubble diameter d_b, allows the determination of the interfacial area and thus leads to the mass transfer rates between the gas and liquid phase. Gas holdup depends mainly on the superficial gas velocity and often is very sensitive to the physical properties of the liquid and gas [Shah et al., 1982]. The relative gas holdup varies locally in axial and radial direction. It largely depends on the flow regime, if it is homogeneous (bubbly) or heterogeneous (churn-turbulent).The dependence of the gas holdup on gas velocity is generally of the form, $\varepsilon_G \quad \alpha \quad U_G^n$

The value of the exponent, n, depends on the flow regime. In homogeneous bubbly flow regime the value of n varies from 0.7 to 1.2, while for churn-turbulent or the transition regime, the effect of U_G is less

pronounced and the exponent n takes values from 0.4 to 0.7 [Smith et al. (1977)].

A large number of correlations for gas holdup have been proposed in the literature [Shah et al., (1982)]; however, the large scatter in the reported data does not allow a single correlation.

2-3-1-Factors Affecting Gas Holdup

i-Sparger And Column Geometry

The gas hold up is virtually independent of column dimensions and sparger layout for low as well as high pressures [Smith et al. (1977)] provided the following three criteria are fulfilled [Wilkinson, et al., (1992)]:

1- The column diameter has to be larger than 0.15 m.

2- The column height to diameter ratio has to be in excess of 5.

3- The hole diameter of the sparger has to be larger than 1-2 mm.

ii- Gas Density

It is an important factor that afects the gas hold-up. Ozturk, et al., (1987) studied gas hold-up in different gas-liquid system (Gases: H_2, He, N_2, Air, Co_2) at atmospheric pressure and applied the correlations reported in literature to fit the data. They recommended the correlations of Akita and Yoshida (1973) which gave best fit with a mean error of only 11%. Also, other authors [Reilly, et al., 1986, Idogawa, et al., 1982;1987, De Bruijn et al.,1987, Zov, et al., 1988, Oyevaar 1989] reported careful experimental studies in which a significant effect of gas density on gas hold-up, either due to increasing gas molecular mass (different gases) or operating pressure was observed .

All the previous work on the effect of gas density and elevated pressure on gas hold-up in bubble column reactors has been critically reviewed and analyzed by Wilkinson and Von Dierendonek (1990) and

Wilkinson (1991). These authors concluded that the gas hold-up increases with gas density due to reduction in bubble size.

Wilkinson and Dierndonck (1990) have demonstrated that a higher gas density increases the rate of bubble break-up especially for large bubbles. The large bubbles rise fast through the column virtually in plug flow ,while the small bubbles display a wide residence time distribution [Krishna ,et al (1990)].This means that increasing in the density or the pressure of a gas, leads to a delay in reaching to U_{trans} or starting the heterogeneous regime ,therefore, the total gas hold-up increases. It is important to point out that the results and interpretations given by Wilkinson and Von Dierndonck also explain the high gas hold-up observed in coal liquefaction process [Tarny, et al., 1984] and Fischer-Tropsch synthesis in slurry phase [Deckwer, et al., 1980;1981]. Based on the results of Wilkinson and Von Dierndonck and additional measurements with the dynamic gas disengagement technique of Krishna (1981), Krishna, et al. (1991) proposed a model for the gas hold-up BCRs which incorporates the influence of gas density on flow regime transition.

iii- Liquid Properties

Liquid properties such as viscosity and surface tension have an influence on the regime transition and hence gas holdup. A higher liquid viscosity promotes coalescence of large bubbles. Consequently, large bubbles occur especially in bubble columns with high viscosity liquids. Furthermore, because of the high-rise velocity of these large bubbles, the gas holdup in viscous liquids is expected to be low, whereas the transition to the churn turbulent regime occurs at very low gas velocity. The surface tension also has a pronounced influence on bubble break up and thus holdup. The occurrence of large bubbles is minimal due to bubble break up especially in those liquids that combine a low surface tension with a low viscosity. As a result, relatively high gas holdup values are to be expected

for such liquids, whereas the transition to the churn-turbulent regime due to the formation of large bubbles is delayed to relatively high gas holdup values [Wilkinson et al., 1992]

iv- Temperature

It has been demonstrated in general, that increasing in temperature leads to a higher gas holdup. The change in temperature can have an influence on gas holdup, due to the influence of temperature on the physical properties of the liquid, and on the vapor pressure. Both the surface tension and the liquid viscosity decrease if the temperature is increased, and this will lead to higher gas holdup [Wilkinson et al., 1992].

v- Suspended Solids

From many publications it has become clear that the addition of solids to a bubble column will lead to a small decrease in gas holdup (Reilly et al., 1986) and the formation of large bubbles. An exception occurs for very small particles $(0 - 100)$ μm at low weight fractions $(< 4\%)$ [Wilkinson et al., 1992]

According to Deckwer (1980): the addition of solids decrease the gas hold-up slightly at low superficial velocity $U_G < 4cm/s$, Shah et al.(1982) stated that the presence of solids does not affect the gas holdup significantly at high gas velocities > 0.1 m/s , but according to Krishna (1997):

o   The total gas hold-up ε_G is significantly decreased with increasing slurry concentration.

o   The dense phase gas hold-up ε_{DF} (small bubble hold-up) in the heterogeneous regime of operation is practically independent of the operating gas velocity and is significantly reduced with increasing solids concentration.

o   The dilute-phase gas hold-up ε_{LB} (large bubble hold-up) is virtually independent of slurry concentration, but is a significant function of the column diameter.

VI-Liquid Velocity

The liquid velocity in a bubble column is usually low, and consequently its influence on gas holdup is often claimed to be negligible. In principal, however, liquid flowing co currently up ward will lower gas holdup, while a counter current liquid flow will increase gas holdup [Wilkinson et al.,1992]

2-3-2-Correlations for Calculating the Gas Holdup

Shah et al., [1982] listed 14 different correlations for gas holdup in two- and three phase systems together with the system employed and the range of operating parameters, only few of them are based on numerous experimented data e.g.

Akita and Yoshida, [1973]

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4}=0.2\left[\frac{D_R^2\,\rho_L\,G}{\sigma}\right]^{1/8}\left[\frac{g\,D_R^3\,\rho_L^2}{\mu_L^2}\right]^{1/12}\left[\frac{U_G}{\sqrt{g D_R}}\right] \tag{2-2}$$

And Hikita et al. [1980]

$$\varepsilon_G = 0.672\left[\frac{U_G\,\mu_L}{\sigma}\right]^{0.578}\left[\frac{(\mu_L^4 g)}{\rho_L\sigma^3}\right]^{-0.131}\left[\frac{\rho_G}{\rho_L}\right]^{0.062}\left[\frac{\mu_G}{\mu_L}\right]^{0.107} \tag{2-3}$$

Wilkinson et al., [1992] listed some other equations incorporating the gas phase properties and accounts for the influence of gas velocity on gas holdup in the churn-turbulent regime by assuming that gas holdup increases less than linear proportionally [$\varepsilon_G \approx U_G^{0.4-0.6}$ or $\frac{\varepsilon_G}{1-\varepsilon_G} \approx U_G^{0.6-0.9}$] to the superficial gas velocity U_G, examples are :

Hammer et al., [1984]

$$\frac{\varepsilon_G}{1-\varepsilon_G} = 0.4 \quad U_G^{0.87} \ \rho_L^{0.1} \ \rho_G^{0.17} \ \sigma^{-0.06} \ \mu_L^{-0.21} \ g^{-0.27} \qquad (2\text{-}4)$$

Idogawa et al., [1985]

$$\varepsilon_G = I/(I+1) \qquad\qquad I = 1.44 \quad U_G^{0.58} \ \rho_G^{0.12} \ (\sigma)^{(-0.16\,\exp(-P))} \qquad (2\text{-}5)$$

P pressure in MPa ; σ surface tension in mN/m

Reilly et al., [1986]

$$\varepsilon_G = 296 \quad U_G^{0.44} \ \rho_L^{-0.98} \ \rho_G^{0.19} \ \sigma^{-0.16} + 0.09 \qquad (2\text{-}6)$$

Idogawa et al., [1987]

$$\frac{\varepsilon_G}{1-\varepsilon_G} = 0.059 \, U_G^{0.8} \ \rho_G^{0.17} \left[\frac{\sigma}{72}\right]^{-0.22\,\exp(-P)} \qquad (2\text{-}7)$$

P: pres in MPa ; σ : surface tension in mN/m and U_G in cm/s

In the heterogeneous flow Krishna et al., [1991] and Wilkinson, [1991] and Krishna and Sie, [2000] assume that the small bubbles holdup is constant and equal to that at the end of the homogeneous flow regime (ε_{trans}). Increasing the gas velocity beyond the transition velocity U_{trans} from homogeneous to heterogeneous flow increases only the large bubble holdup and the following relations are assumed to apply:

Krishna et al. [1991]

$$\varepsilon_{G\,hom} = 4\,U_G$$

$$\varepsilon_{LB} = A\,(U_G - U_{trans})^n \qquad U_G \geq U_{trans}$$

$$\varepsilon_{hetro} = \varepsilon_{trans} + A\,(U_G - U_{trans})^n$$

$$(2\text{-}8)$$

$$\varepsilon_{trans} = 4\,U_{trans}$$

With A$=1$, $n = 0.8$ and U_{trans} taken from experiments could describe gas holdup data up to 0.5 with striking agreement.[Krishna et al. 1991].

Wilkinson, [1991]

$$\varepsilon_G = U_G / V_{DF} \qquad\qquad U_G \leq U_{trans}$$

$$\varepsilon_G = \frac{U_{trans}}{V_{DF}} + \frac{U_G - U_{trans}}{V_{LB}} \qquad\qquad U_G > U_{trans}$$

$$\frac{V_{DF}\mu_L}{\sigma} = 2.25 \left[\frac{\sigma^3 \rho_L}{g\,\mu_L^4}\right]^{-.273} \left[\frac{\rho_L}{\rho_G}\right]^{0.03} \qquad (2\text{-}9)$$

$$\frac{V_{LB}\,\mu_L}{\sigma} = \frac{V_{DF}\,\mu_L}{\sigma} + 2.4 \left[\frac{\mu_L\left(U_G - U_{trans}\right)}{\sigma}\right]^{0.757} \left[\frac{\sigma^3 \rho_L}{g\mu_L^4}\right]^{-0.077} \left[\frac{\rho_L}{\rho_G}\right]^{.077} \frac{U_{trans}}{V_{DF}}$$

$$\varepsilon_{trans} = 0.5 \exp\left(-193\,\rho_G^{-0.61}\mu_L^{0.5}\,\sigma^{0.11}\right) \qquad (2\text{-}10)$$

The correlation is valid for the range:

$\sigma = 0.02 - 0.073$ Nm^{-1}

$\mu_L = 0.0004 - 0.055$ Pa s

$\rho_L = 683 - 2960$ Kg m^{-3}

$\rho_G = 0.09 - 38$ Kg m^{-3}

Krishna and Sie, [2000] further assumed that the rise velocity of small bubbles will increase with increased solids holdup due to enhanced coalescence according to :

$$V_{DF} = V_0 \left(1 + 0.8\ \varepsilon_s / V_0\right) \qquad\qquad (2\text{-}11)$$

$$\varepsilon_{DF} = \varepsilon_{DO} \left(\frac{\rho_G}{\rho_{GO}}\right)^{0.48} \left(1 - 0.7\ \varepsilon_s / \varepsilon_{DO}\right) \qquad\qquad (2\text{-}12)$$

$$U_{DF} = \varepsilon_{DF}\ V_{DF} \qquad\qquad (2\text{-}13)$$

$$U_{LB} = U_G - U_{DF} \qquad\qquad (2\text{-}14)$$

$$\varepsilon_{LB} = 0.3\ U_{LB}^{0.58} \left(\frac{\rho_G}{\rho_{GO}}\right)^{0.5} \qquad\qquad (2\text{-}15)$$

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$$\varepsilon_G = \varepsilon_{LB} + \varepsilon_{DF}\left(1 - \varepsilon_{LB}\right) \qquad\qquad (2\text{-}16)$$

With: $V_0 = 0.095$ m/s $\rho_{G0} = 1.3$ $\varepsilon_{DF} = 0.27$ D>1m

$\varepsilon_s > 0.16$

Stern [1985]

$$\varepsilon_G = .03\ U_G / (1 + .08\ U_G) \qquad\qquad (2\text{-}17)$$

Deckwer & Schumpe[1993]

$$\varepsilon_G = .2\ (Bo)^{-0.13} * (Ga)^{0.11} * (Fr)^{0.54} \qquad\qquad (2\text{-}18)$$

$$Bo = g * (d_b)^2 * (\text{Þ}_L / \sigma_L)$$

$$Ga = g * (d_b)^3 * (\text{Þ}_L / \mu_L)^2$$

$$Fr = U_G / (g * d_b^{0.5})$$

2-4- Liquid Dispersion Coefficient

In a bubble column reactor, the gas flow generates significant backmixing in the liquid phase. At a thorough analysis it can be noticed that the backmixing of the dense phase is caused by the eddies which derive their energy from the large, fast-rising bubbles. The maximum size of the eddies, for vessels with $H/D_T > 1$, is limited by the column diameter, D_T. The induced circulation patterns have a profound effect on mass transfer and productivity of these systems and they are especially important in eliminating concentration gradients within the vessel. The design of a bubble column reactor always requires consideration of the backmixing in the liquid phase. Krishna (2000).

Mixing of the liquid phase in the bubble column reactor has to be attributed to various phenomena such as turbulent vortices, liquid entrainment in the wakes of rising bubbles, liquid circulation, radial exchange flows, etc. All these phenomena are obviously interrelated and

are primarily dependent on the gas holdup structure, i.e. local holdup variations, bubble size and rise velocity distributions.

The degree of mixing within any reaction vessel will invariably lie between the hypothetical limits of ideal plug flow and perfectly mixed flow. Many types of models were developed to characterize non ideal behavior .some relay on the analogy between mixing in the actual flow and diffusion process. These are called dispersion model. Krishna (2000).

The general form of the dispersion model expresses the dispersion coefficient and velocity as a function of position [El Temtamy1982].

Since the axial dispersion model is so simple and characterizes the backmixing by only a single parameter, it is the most widely used in the representation of the non ideal mixing behavior for each phase in bubble column reactors. The dispersion flux can be expressed in analogy with Fick's diffusion law:

$$\frac{\partial C_L}{\partial t} = D_{ax,L}\frac{\partial^2 C_L}{\partial z^2} \qquad (2\text{-}19)$$

Where the parameter $D_{ax,L}$, called the axial dispersion coefficient is the unique parameter which characterizes the degree of back mixing during process. The term 'axial' is used in order to mark clearly the difference between the mixing in the direction of flow and the mixing in the lateral or radial direction. In bubble columns these two quantities are quite different in magnitude, so that the axial dispersion coefficient is significantly exceeding the radial dispersion coefficient.[El-Temtamy et al.(1979)]

Axial dispersion coefficients of the liquid phase in vertical gas-liquid contactors have been reviewed by Shah et al. (1978). Table 2.1 gathers the conditions of the work on axial dispersion coefficients from the available literature.

The dispersion coefficient is expressed in the dimensionless form as the Peclet number, its value denoting the degree of back mixing in the column. If $Pe = 0$ bacmixing is complete and if $Pe = \infty$ plug flow prevails.

Most of reported empirical correlations indicate the dispersion coefficient to be dependent on the gas velocity and column diameter. A significant influence of the flow direction (i.e. co-current or counter-current) has not been pointed out. In the range of liquid velocities used in industrial operation, the superficial liquid velocity, U_L appears to have no influence on liquid-phase dispersion. Steady state measurements by Deckwer *et al.* (1973), Badura *et al.* (1974) and Towell and Ackerman (1972) confirmed this fact. It is usually assumed that the dispersion coefficient does not depend on the column height. However, studies of Schügerl (1967) and Deckwer et al. (1973) show that dispersion coefficients may differ along the column height, decreasing from top to bottom.

The degree of axial dispersion is also affected by vessel internals and surface-active agents that delay the coalescence. Konig et al. (1978) have demonstrated the effects of surfactants and sparger type by experimenting with weak alcohol solutions using three different porous spargers. They clearly indicated that the interaction of surfactants and sparger can be very complex. Surfactants can produce either much more or much less backmixing than surfactant-free systems, depending on the bubble size, which, in turn, depends on the sparger used.

Although industrial bubble columns usually contain cooling tubes, only two studies have tackled this point, Forret et al.(2000), and Chen et al. (1999).

Chen et al. (1999) presented the results obtained for the gas hold-up profiles, time-averaged liquid velocity profiles, turbulent stresses and eddy diffusivities (radial and axial) obtained in a 18" (44cm) diameter column

without and with internals similar to those used in industrial scale units (e.g. heat exchanger tubes) using both air/water and air/drakeoil 10 (viscosity ≈ 30 cP) at gas superficial velocities of 2, 5 and 10 cm/sec. At high superficial velocity (10 cm/s), the turbulent stress and eddy diffusivities in air-water are higher than in air-drakeoil. Turbulent stress and eddy diffusivities were lower in the column with internals due to the fact that they physically reduce the length scales of turbulence. The difference was more pronounced in the radial direction than in the axial direction.

Forret et al. (2000) have undertaken a comprehensive study of the hydrodynamics (liquid velocity profile, axial dispersion) in 1m diameter bubble column with and without cooling tubes. Although the value of axial dispersion coefficient was identical for the two cases, the ratio of axial to radial dispersion coefficient changed significantly .This change was attributed to the existence of ordered circulation pattern in presence of the cooling tubes.

The influence of the physical properties of the liquid on the dispersion coefficient has been investigated by several authors (Aoyama et al., 1968; Hikita and Hikukawa, 1974; Riquarts, 1981; Walter and Blanch, 1983; Bernemann, 1989) but little effect has been observed.

Some authors also underlined the disastrous effect of small deviations from true verticality, such as Tinge and Drinkenburg (1986) and Rice & Littlefield (1987). The scatter in the literature data was attributed to originate from very slight and unreported (or unmeasured) deviations from verticality.

For reactor scale-up purposes, the most useful relation for vertical cylindrical vessels with pure liquids without coalescence inhibitors is considered to be the one of Baird and Rice (1975). Making use of

Kolmogoroff's theory of isotropic turbulence, they proposed the following equation:

$$D_{ax,L} = 0.35 g^{1/3} D_T^{4/3} U_G^{1/3}$$

(2-20)

Furthermore, the authors extended the correlation to fluidized beds and liquid-liquid spray columns, which proved to be partially successful.

The dependency of $D_{ax,L}$ on D_T and U_G expressed by Eq. (2.20) has been confirmed by many studies and can be also transcript in dimensionless terms of Pe and Fr criteria:

$$\frac{D_{ax,L}}{D_T U_G} = K \left(\frac{g D_T}{U_G^2} \right)^{1/3}$$

(2-21)

The constant K has different values for different authors, as can be seen in Table.2.2.

In 1996 Degaleesan et al. developed a phenomenological model for liquid mixing, mainly based on the monitored flow circulation and turbulent eddy diffusion. The model was called "recirculation and cross flow with dispersion" (RCFD) and accounts for two mixing mechanisms. The first one is the convective recirculation, due to the large scale liquid motion. In this context, the column can be simply considered as a big recirculation cell, with liquid moving upwards in the central core zone and descending along the annular region, near the walls of the column. Besides this, fluctuated motion of the liquid elements superimposes, due to the bubble turbulence. This second mixing contribution takes into account the complex motion of the gas bubbles, and therefore the random motion of fluid elements in axial and radial direction caused by the wakes of fast-rising large bubbles. The upflow and downflow, characterized by average interstitial velocities are connected to each end by two regions which are

assumed well-mixed. The turbulent axial mixing is accounted for by an axial dispersion coefficient in each section; the radial mixing is incorporated into an exchange coefficient between the two sections (i.e. the upflow region and the downflow region). The input parameters of the model (such as local liquid velocities and void fraction) were obtained using Computer Automated Particle Tracking (CARPT) and Computed Tomography (CT) as experimental techniques.

The model was recently further developed and extended also for bubble columns slurry reactors by Degaleesan et al. (1997). The two-compartment convective-diffusion model has been formulated by considering a finite volume discretization of the two-dimensional axis-symmetric convective-diffusion model developed for bubble columns flows. The same experimental techniques were used for the data base, together with data from literature. The existing data-base was limited to air-water systems, certain column sizes and superficial gas velocities. The methodology aimed to extrapolate the data giving the following cross sectional averaged eddy diffusivities $\overline{D_{zz}}$ and $\overline{D_{rr}}$:

$$\overline{D_{zz}} = -\frac{0.00584}{D_T^{0.8}} + 0.1689(D_T U_G)^{0.3}$$

(2-22)

$$\overline{D_{rr}} = -\frac{0.000879}{D_T^{0.8}} + 0.0206(D_T U_G)^{0.3}$$

(2-23)

for $U_G \geq 0.05$ m/s.

The authors limited the applicability of the above developed correlations only to air-water systems in churn turbulent regime, at atmospheric pressure. When the axial term is compared with experimental data from this work, they show a good agreement for low gas velocity only. This can not be attributed only to the fact that the radial contribution

is also taken into account, as long as this term is much smaller, but to the currently limited data base, as is also mentioned by the authors. They advised that for higher gas velocities the equations presented above have to be verified in compliance with experimental data.

For bubble columns, as long as the one-dimensional description is accepted, the axial dispersion coefficient in the liquid phase is linking all mixing phenomena as mentioned before.

Table (2-1): Experimental Studies on Liquid Phase Mixing In Bubble Columns

Reference	System	Flow rate, m/sec.	Tracer system	Column characteristics mx (1000)	Dispersion studied
Kato and Nishiwake (1972)	Air-water concurrent	$U_L = 0.005-0.015$ $U_G = 0 - 0.3$	1-1.5N KCl pulse electrical conductivity cell	$Dt = 66.122.214$ $L_c=2000,20102,4050$ $d_o = 1, 1.4, 2, 3$ $nh = 7, 19, 31, 37, 97, 336.$	Axial
Eissa, et al (1975)	Air-water concurrent	$U_L = 0.0075-0.013$ $U_G = 0.014-0.078$	Salt solution steady state point source conductivity meter	$Dt = 50$ $Lc = 420$ Multioriface plate distributor.	Axial and Radial
Reith, et al. (1968)	Air-water concurrent and countercurrent	$U_L = 0 - 0.02$ $U_G = 0.02-0.45$	NaCl solution p-sulfobenzeneazo-B-naphthol, conductivity cell & spectrometer	$Dt = 50.8, 140, 290$ $L_c=152, 200, 353, 380,$ axial d_o = one hole of 0.002m diam $104m^2$	Axial
Towell and Ackerman (1972)	Air-water concurrent and counter current	$U_L = 0.0024-0.014$ $U_G = 0-0.152$	Methylene blue dye Freon step & pulse spectrophotometer. Ionization cell (ionization induced by emmison from tritum source. Ionization current measured by electrometer.	$Dt = 406, 1067$ $L_c=2743, 5182$ $d_o = 6.35$ $nh = 19, 41.$	Axial

Cont. Table (2-1)

Reference	System	Flow rate, m/sec.	Tracer system	Column characteristics mx (1000)	Dispersion studied
Eissa and Schugerl (1975)	Air-water concurrent	U_L = 0.0035-0.014 U_G = 0.002-0.06	NaCl Steady state injection Conductivity meter.	Dt = 159 L_c= 3900 d_o = 2 nh =200	Axial
Cova (1966)	N_2-water -acetone -CCl_4 -EtOH -Aq EtOH -Aq Sugar soln-Cyclohexanol concurrent	U_L = 0.002-0.0087 U_G = 0.0051-0.173	Heat Steady state Thermocouples	Dt = 19.1, 45.7 L_c= 3.8, 4ft sintered, metal disk, single 1.32mm diam orifice 13x0.483mm diam.	Axial
Argo and Cova (1965)	N_2-water NH$_3$-water counter concurrent	U_L = 0.0038-0.016 U_G = 0.0041-0.203	KCl and NaCl Steady state point source Potentiometric titration Electrical conductivity.	Dt = 45.7, 102, 448 L_c= 914, 1219, 8230 18x3.18mm perforated plate	Axial
Ohki and Inoue (1970)	Air-water	U_L = batch U_G = 0-0.25	4N aqueous KCl pulse Pt electrical conductivity cell.	Dt = 40, 80, 160 L_c= 200, 250, 300 d_o = 0.4, 0.7, 1, 2, 3 nh = 2, 5, 7, 13, 19, 55, 91	Axial
El-Temtamy et al.(1979 a)	Air-water & Air- water glass beads (0.45and 0.96mm diameter.)	U_L = 0 -0.14 Ug = 0 – 0.07	NaCl tracer, steady state point source tracer Electrical conductivity cell	Lc = 1200 D_T= 50	Axial

Reference	System	Flow rate, m/sec.	Tracer system	Column characteristics mx (1000)	Dispersion studied
El-Temtamy et al.(1979 b)	Air-water & Air- water-glass beads (0.45,0.96,2&3 mm diameter.)	$U_L = 0$ -0.14 $Ug = 0 – 0.07$	NaCl tracer, steady state point source tracer, electrical conductivity cell	Lc = 1200 D_T= 50	Axial & Radial
El-Temtamy. Et. al. (1985)	Air-water-yeast suspension	U_L batch U_G 0-0.06 m/s	NaCl tracer pulse conductivity prob.	L_C = 2 m D_T = 0.15 m	Axial
Hikita and Kikukawa (1974)	Air-water Air-aq. MeOH solution	U_L = batch U_G = 0.043-0.338	Aq. KCl Pulse Pt electrical conductivity cell	Dt =100, 190 L_c= 1500, 2400 d_o =13.1, 20.6, 36.2 nh = 1	Axial
Deckwer, et al., (1974)	Air-water	U_L = batch U_G = 0-0.05	Dye, heat, electrolytes Diroc pulse.	Dt = 150, 200 L_c= 4400, 7230 cross of 56, nozzle 1mm diam, sintered porous plate of 150 μm	Axial
Aoyama, et al.,(1968)	Air-water-Aq. Glycerin Air-Aq tween20 concurrent	U_L = 0.0018-0.011 U_G = 0.0031-0.078	Heat, 3NKCl Steady state pulse Thermocouples Electrical conductivity cell.	Dt = 50, 100, 200 L_c=1000, 1900, 2020 porous plate distrusted 0.068–0.13 perforated plate d_o = 0.5, 1 nh = 19, 31, 33, 85	Axial

Cont. Table (2-1)

Reference	System	Flow rate, m/sec.	Tracer system	Column characteristics mx (1000)	Dispersion studied
Alexander and Shah (1976)	Air-Aq. Glycerine solution Air-Aq. Triton DF 12 solution	U_L = batch U_G = 0-0.07	H_2SO_4, pulse Electrical conductivity.	Dt = 60.3, 152.4, 76.2x229 rectangular L_c= 1240, 3080, 1120 1x6.4mm diam orifice, 13x0.79mm diam porous plate	Axial

Table (2-2): Literature Correlations for Axial Dispersion Coefficient

	Author	Axial dispersion coefficient
1.	Reith et al. (1968)	$Pe_L = \dfrac{U_r D_T}{D_{ax,L}} = 3 \pm 0.3$ $U_r = 2U_G + u_{b\infty}$
2.	Ohki and Inoue (1970)	for $0 \le U_G \le 0.07$ m/s (bubble flow regime) $D_{ax,L} = 75.4\, D_T^2 U_G^{1.2} + 17000 d_0$ for $U_G \ge 0.1$ m/s (coalesced bubble slug flow regime) $D_{ax,L} = \dfrac{0.14\, D_T}{(1-\bar{\varepsilon})^2}$
3.	Kato and Nishiwaki (1971, 1972)	$Pe = \dfrac{13\, Fr_G}{1 + 6.5(Fr_G)^{0.8}}$ $D_{ax,L} = \dfrac{g^{0.5}\, D_T^{1.5}\left[1 + 6.5\left(\dfrac{U_G}{\sqrt{gD_T}}\right)^{0.8}\right]}{13}$
4.	Towell and Ackerman (1972)	$D_{ax,L} = 1.23\, D_T^{1.5} U_G^{0.5}$

5.	Akita and Yoshida (1973)	$D_{ax,L} = U_G D_T \dfrac{0.06 + 0.55\left(\dfrac{U_G}{\sqrt{gD_T}}\right)^{0.7}}{\left(\dfrac{U_G}{\sqrt{gD_T}}\right)}$
6.	Badura (1974)	$D_{ax,L} = 0.692\, D_T^{1.4} U_G^{0.33}$
7.	Deckwer (1974)	$D_{ax,L} = 0.678\, D_T^{1.4} U_G^{0.33}$
8.	Hikita and Kikukawa (1974)	$D_{ax,L} = \left(0.15 + 0.69 U_G^{0.77}\right) D_T^{1.25} \left(\dfrac{10^{-3}}{\eta_L}\right)^{0.12}$
9.	Baird and Rice (1975)	$D_{ax,L} = 0.35\, g^{1/3} D_T^{4/3} U_G^{1/3}$
10.	Field and Davidson (1980)	$D_{ax,L} = 0.9\, D_T^{1.5} \left[H(U_G - \bar{\varepsilon} V_b) \right]^{1/3}$ $V_s = V_b (1 - \bar{\varepsilon})^{1.39} (1 + 2.55\bar{\varepsilon}^3)$; $V_b = 0.164\,\text{m/s}$
11.	Joshi (1980)	$D_{ax,L} = 0.33 D_T V_c$ $V_c = 1.31 \left[gD_T (U_G - \bar{\varepsilon} u_{b\infty}) \right]^{1/3}$
12.	Miyauchi et al. (1981)	$D_{ax,L} = U_G^{1/4} D_T^{3/2} \left(0.291 + \dfrac{0.341}{\left(1 - 0.54 U_G^{1/2}\right)^2} \right)$
13.	Riquarts (1981)	$D_{ax,L} = 0.068\, g^{3/8} D_T^{3/2} U_G^{3/8} \nu_L^{-1/8}$

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14.	Zehner (1982)	$D_{ax,L} = 0.368 g^{1/3} D_T^{4/3} U_G^{1/3}$
15.	Walter and Blanch (1983)	$D_{ax,L} = 0.606 D_T V_1 \text{ or } \quad Pe = \dfrac{V_1 D_T}{D_{ax,L}} = 1.65$ $V_1 = \left(\dfrac{U_G g H}{4 + \dfrac{70}{Re^{0.25}} \dfrac{H}{D_T}} \right)^{1/3}$ for $Re > 1000$ and $\dfrac{H}{D} > 3$; $\quad Re = \dfrac{V_1 D_T \rho_L}{\eta}$
16.	Bukur et al. (1983)	$D_{ax,L} = 3.676 \, (U_G)^{0.32} * (D_T)^{1.34}$
17.	Stern (1985)	$D_{ax,L} = 0.68 * (D_T^{1.4}) * (U_G^{1/3})$
18.	Kawase and Moo-Young (1986)	$D_{ax,L} = 0.343 n^{-8/3} g^{1/3} D_T^{4/3} U_G^{1/3}$
19.	Deckwer & Schumpe (1993)	$D_{ax,L} = 0.35 * D_T * U_G \, [g * D_T / U_G^2]^{1/3}$
20.	Deckwer et al.(1973)	$D_{ax,L} = 2.7 * (D_T^{1.4}) * (U_G^{0.3})$
21.	Kantak (1994)	$D_{ax,L} = 0.632 D_T^{1.25} \dfrac{U_G}{\bar{\varepsilon}}$
22.	Krishna (2000)	$D_{ax,L} = V_L(o) * D_T$ $V_L(o) = 0.21 * (g*D_T)^{0.4} (U_G^3 / g*10^{-6})^{0.2125}$