Effect of volumetric flow rate on axial dispersion coefficient of fluid in liquid-solid bed with RTD curves studies

J. Behin*, F. Aligoli
Department of Chemical Engineering, Faculty of Eng., Razi University, Kermanshah, IRAN

Abstract
Liquid phase RTD curves were investigated in classical fixed and fluidized bed regimes with high density particles. Using an impulse tracer technique in a column of 5cm in diameter and height of 1.2m, liquid's RTD, mean residence time (MRT) and axial dispersion coefficient (ADC) were determined. The effect of liquid velocity and solid's density were studied on hydrodynamics of bed. ADC varied from 1.62 to 8.23 for the particle Reynolds number of 43.18 to 279.41. ADC increases with increasing liquid superficial velocity. As difference between solid and liquid density increases, ADC decreases in an identical Reynolds number.

Keywords: liquid-solid fluidized bed, axial dispersion coefficient (ADC), residence time distribution (RTD) curves, hydrodynamics

Introduction

Liquid-solid fixed (packed) and fluidized beds are encountered in many processes in industrial operations. Numerous examples can be found in the chemical, petrochemical, pharmaceutical, agricultural, biochemical, hydrometallurgical, electrical power-generation, and food industries. Because of the low efficiency of the conventional fixed bed system, much attention has been paid to fluidized bed system [1]. Many catalyzed reactions as fluidized catalyst cracking of petroleum occur in fluidized bed reactors [2]. These are common and important reactors in process engineering because of easy handling, good mixing and transportation, good mass-heat transfer rates between the fluid and the particles and also between the particles and the sidewall. The operations such as combustion, water treatment, crystallization, ion exchange, and adsorption are the other examples of liquid-solid fluidized beds [3-5]. The efficient-successful design, analysis, operation, scale-up, and prediction of the hydrodynamic behavior of this beds require detailed knowledge of pressure drop ,axial dispersion coefficient (ADC), and hold-up (voidage) of liquid phase [6,7]. One of the main factors which influence the hydrodynamics of liquid-solid fluidized beds is the interaction between the solid and liquid phases. The effect of liquid-solid interaction is generally quantified in term of the liquid's ADC, which provides information about the degree of fluid mixing existing in the bed [8]. Axial liquid phase mixing greatly affects the interphase heat and mass transfer rates, the reactant concentration distribution, and ultimately reactant conversions. Information on axial liquid mixing is therefore crucial to reactor design and process optimization [9]. For example, an increase in axial mixing reduces the driving force of the transport processes and reduces the level of conversion. Hence, a quantitative estimation of mixing of phases is a key factor [4]. During recent years, considerable progress has been made in exploring and understanding the hydrodynamics of fluidized beds. Bed expansion is one of the important factors to be considered in the design of fluidized beds and it has been the subject of many experimental and theoretical studies [10]. Residense time distribution (RTD) studies have proved to be an indispensable tool in this respect where a tracer is injected at some locations of the bed and its concentration is monitored at a point in down stream. The concentration distribution data is processed to extract quantitative

* Corresponding author. Address: Department of Chemical Engineering, Faculty of Engineering, Baghe Abrisham, Kermanshah, Iran. Tel.: +98-831-8250167; Fax: +98-831-8250167; E-mail: Behin@razi.ac.ir

information about the dispersion characeristics of the bed [8]. Among the studies on ADC in liquid-solid fluidized beds, the most extensive work was reported by Chung and Wen (1968). They proposed a generalized correlation for ADC in both packed and fluidized beds. Kikuchi et al. (1984) reported that ADC in a fluidized bed of polystyrene particles deviated significantly from literature correlation equations, including that obtained by Chung and Wen (1968). Moreover their data for beds of glass beads also showed poor agreement with those predicted by the Chung and Wen's equation. Therefore, the significant deviation in Kikuchi's data for low density particles may result from operating or design variable effects, in addition to solid's density effect [9]. Tang and Fan (1990) measured ADC of low density particles (polystyrene, acrylic, nylon, etc.) in liquid-solid fluidized beds. They used the plexiglass column of 7.62cm inner diameter and 130cm height. A 750µm wire screen was placed on top of column to prevent elutriation of particles from the system. A distributor consisting 30micrometers porous plate atop a packed bed claming section was used to provide a uniform liquid velocity profile. The particles have densities and diameters ranging 1.05-1.3g/cm^3 and 1-2.5mm respectively. A potassium chloride solution (3N) was used as the tracer. The tracer injection was controlled by a switch-to-open solenoid valve. The concentration of tracer was detected by a conductivity probe connected to a YSI (Yellow Spring Instrument Co., Inc.) conductivity-meter. They showed that the axial dispersion coefficients determined at r=(2/3)R differed only by 7% from those determined at the axis. They analyzed dispersion of tracer based on the axial dispersion model and showed that liquid's ADC (D_l) is proportional to the liquid superficial velocity (U_o) to a power of 1.28 for low density particles [9]:

$$D_l = 0.575\,U_o^{1.28} \tag{1}$$

To compare the axial liquid mixing behavior's results in beds of low density particles with most of the literature data for heavy particles, an experimental run for a fluidized bed of heavy particles, 1.0mm glass beads, was performed. They obtained the axial dispersion coefficient of 7.64cm^2/s in the 1.0mm glass beads at the superficial liquid velocity of 5.0cm/s and a bed voidage of 0.634 which was compared to 8.06cm^2/s calculated from the Chung and Wen (1968) correlation equation. Their results indicates that difference, if one exists, between the axial liquid mixing behavior in beds of low density particles and that reported in the literature for beds of heavy particles can not be attributed to the measuring method applied and the column and distributor design used in their study. However, they showed that significant scattering between the data and the curve based on the single parameter correlation equation, indicating that axial dispersion coefficient depends on some other parameters. They also studied the relationship between ADC and the bed voidage (ϵ). The ADC for all the low density particles increases monotonically with increasing ϵ. It appears that for the same bed voidage, particles with higher terminal velocity have higher value of ADC. The results are expected since a higher relative liquid velocity is required to maintain a given bed voidage for particles with a higher terminal velocity. Consequently, higher turbulence will be generated, which in turn leads to higher extent of mixing. A maximum ADC at a bed voidage of 0.75 was not observed in their system (which uses particles no smaller than 1mm), as reported by Kikuchi et al. (1984) for beads of polystyrene spheres with a diameter smaller than 1mm. However, the energy dissipation rates per unit liquid mass (e), for all particles used exhibit a maximum at $e \approx 0.7$. For a liquid-solid fluidized bed, energy dissipation rates per unit liquid mass is given as:

$$e = (\rho_s - \rho_l)(1-\varepsilon)\,U_o\,g/(\varepsilon\rho_l) \tag{2}$$

In which U_o is liquids superficial velocity, ρ_s and ρ_l are solid particle and liquid densities, respectively. The specific dissipation rate is equal to the rate of energy per unit liquid mass supplied to overcome the drag of particles, with the assumption due to wall friction is negligible. The energy dissipated by the particles is thus related to mixing in particle wakes.

Consequently, there may exist a maximum axial mixing at $e \approx 0.7$ due to wake dispersion, as postulated in the wake dispersion model by Letan and Elgin (1972). A maximum axial mixing at $e \approx 0.7$ however may not be observed due to a sharp increase in the tube dispersion and a development of a specific particle circulation flow with increases in ϵ after $e \approx 0.7$. This equation is valid only for fluidized beds of particles with densities between 1.04 and 1.3 g/cm^3 and sizes between 1.0 and 2.5 mm, and a bed voidage between 0.5 and 0.92 [9].

Asif et al. (1992) have been studied effect of liquid velocity on the ADC with negligible distributor effect. The fluidized bed was a Plexiglas column of 7.68cm in diameter. They determined ADC of 2-12cm^2/s for cylindrical propylene particles (ρ_s: 1.61g/cm^3, Dp: 3.0mm, Lp: 3.0mm, U_o: 2-10cm/s), 4-35 cm^2/s for spherical glass particles (ρ_s: 2.46g/cm^3 Dp: 2.0mm & 3.0mm, U_o: 4-16cm/s) respectively. The tracer, methylene blue, was kept air pressurized in a one-liter tank. Time and duration of tracer injection was controlled with an on-off valve operated with Taurus computer data acquisition and control system interfaced with an IBM personal computer. For their experiments, the pulse time was kept around 120ms. The amount of tracer injected was controlled with the help of the air pressure in the tracer tank (320-200kPa) so that its concentration stayed within the measurement range of the measuring equipment. The concentration measurement system consisted of Brinkmann probe colorimeter (Model 700). It was based on fiber optic technology to measure the color of the solution and utilized a fiber-light guide that transmitted phase shifted, AC modulated light to the probe tip. One-half of the fiber-optic bundles transmitted light into the test solution. The other half returned the light to the instrument where it passed through an interference filter prior to impinging on the silicon detector cell. The light was returned by the probe tip, which had a mirror set at a fixed distance from the end of the fiber optics. The effective path length in the fluid was twice that distance, since the light traveled through the fluid two times-first to the mirror, and then back to the fiber-optic assembly. A glass window prevented fluid from coming into contact with fiber optic, assuring that only the probe tip comes into contact with the test solution. The electronic circuitry of the instrument was designed in such a way that extraneous or ambient light can not affect its readings. Their results have good agreement with Tang & Fan correlation [8].

Liquid phase residence time distribution studies are reported in 2-phase inverse fluidized bed for the first time by Renganathan and Krishnaiah (2004). They have been studied the mixing characteristics of the liquid phase in liquid-solid inverse fluidized bed (LSIFB) for very low density particles of 0.897g/cm^3. They showed that the liquid phase ADC increases with increase in liquid velocity and Archimedes number and is independent of static bed height. They determined liquid phase ADC and also showed that it increases with increase in liquid velocity. The liquid phase ADC is reported to increase with increase in liquid velocity and reach a maximum at a velocity corresponding to a voidage around 0.7 and decrease with further increase in liquid velocity. However, it is also observed by Chung and Wen and Tang and Fan, a monotonic increase of dispersion coefficient with increase in liquid velocity even beyond voidage of 0.7. The increase and decrease in the trend of dispersion coefficient with liquid velocity is usually observed for particle sizes of less than 1000micrometers, whereas the monotonic increase in dispersion coefficient is reported for particle size of greater than 1000micrometers. The particles with size greater than 1000micrometers used in their study followed this reported monotonic increase in dispersion coefficient. This may be due to a sharp increase in the tube dispersion and a development of a specific particle circulation flow with increase in voidage beyond 0.7. In their study, a particle of size 180micrometers was also used. Even these particles showed a monotonic increase in dispersion coefficient with liquid velocity which is against the reported trend in their study for small particles in classical fluidized beds. In spite of repeated experiments, the same trend is observed. This particle has the lowest "Archimedes number (Ar)" of 17.6 in their study, the next being 12947. To get better understanding of the variation of dispersion coefficient with liquid velocity in IFB, experiments with particles of *Ar* between 17.6 and 12947 may be necessary. The column was made of acrylic with an inner diameter of 89mm and an outer diameter of 97mm using multiple sections. The total height of the column was 2.8m and that of test section was 1.86m. A packed bed of berl saddles was used at the top of the column to provide uniform distribution of liquid. Water was pumped from a storage tank through calibrated rotameters and admitted at the top of the column through distributor. Different rotameters were used to cover wide range of liquid flow. The water exited from the bottom of the column and recirculated back to the storage tank through an overflow weir. The overflow weir helps to maintain a constant water level in the column. Provisions were made for online measurement of tracer by the conductivity method. To measure the conductivity, SS 314 electrodes were fixed with the inside surface flush with the wall of the column. This arrangement would not hinder the free movement of fluidized particles in the column. A pair of electrodes was placed diametrically opposite with the longer side perpendicular to the axis of the column. Eighteen sets of such electrodes were placed at an equal distance of 10 cm along the length of the column. To record the online conductivity measurements, the electronic circuit consisting of a switching circuit, a conductivity meter and an A/D card was used. The switching circuit used to switch between the 18 electrodes can be operated automatically using a software trigger from the computer or manually. A pulse of 3ml of 5N NaCl solution was injected using a syringe through the septum at the top of the column. The time taken for the tracer injection in all the experiments was less than one-third of a second. They have considered the effect of "liquids superficial velocity to minimum fluidization velocity (U_o/U_{mf})", and "Archimedes number (Ar)" on ADC and proposed the following correlation [6]:

$$D_l = 1.48\times10^{-4}\,Ar^{0.66}\left(Re_p/Re_{mf}\right)^{1.73} \tag{3}$$

Where Re_{mf} is the minimum fluidization Reynolds:

$$Re_{mf} = \sqrt{(33.7)^2 + 0.0408Ar} - 33.7 \tag{4}$$

RTD curves

Information on flow behavior can be obtained from residence time distribution (RTD) measurements [10]. RTD is very important to many chemical processes at least in the following two main branches: for designing and controlling the process. It is connected with the knowledge of contact and reaction time between fluids or solid fluid [11]. It gives information on how long the various elements have been in the vessel. The knowledge of the liquid RTD is important for a number of reasons allowing an accurate kinetic modeling of the system, and help reactor design to achieve or preserve a desired flow pattern [12]. The measurement and analysis of RTD are important tools in the study of continuous flow system. It gives an idea of flow pattern. The principle of a tracer experiment consists of a common impulse-response method: injection of a tracer at the inlet of a system and recording the concentration-time curve at the outlet [11]. RTD theory, which uses the results of the stimulus response method, is one of the important tools to understand the complex phenomena taking place inside the process system. The RTD can be directly measured by injecting a unit impulse (unit amount of labeled particles or tracer particles) in to the system's inlet. Two techniques, namely the transient and the steady state tracer injection methods, are frequently used to determined axial liquid dispersion coefficient. The transient technique introduces a specific from of tracer input into the system and detects the tracer residence time distribution at a location downstream. The steady state method measures the tracer axial concentration profile upstream from a plane tracer source. To use the steady state method ,a high degree of axial mixing is needed such that a longitudinal tracer concentrations upstream from a plane tracer source can be detected accurately. Therefore, the axial dispersion coefficients were determined based on the tracer residence time distribution obtained from a pulse tracer injection design provided an approximately uniform tracer concentration across the column at the time of injection [8]. Supposing that the injection of tracer is made instantaneously at t=0 (Dirac's impulse). If k(t) representing the conductivity function of tracer at the outlet stream of the system, the exit age distribution is defined:

$$E_t = E(t) = \frac{k(t)}{\int_0^\infty k(t).dt} \cong \frac{k(t_i)}{\sum_{i=1}^{n} k(t_i).\Delta t_i} \quad (5)$$

The mean residence time (t_m) can be obtained from the following equation:

$$t_m = \frac{\int_0^\infty t.k(t).dt}{\int_0^\infty k(t).dt} \cong \frac{\sum_{i=1}^{n} t_i.k(t_i).\Delta t_i}{\sum_{i=1}^{n} k(t_i).\Delta t_i} \quad (6)$$

In which Δt_i is time step between two measurments. The time dependant vaiance of the RTD curves is defined by [11]:

$$\sigma_t^2 = \frac{\int_0^\infty (t-t_m)^2.k(t).dt}{\int_0^\infty k(t).dt} \cong \frac{\sum_{i=1}^{n} (t_i-t_m)^2.k(t_i).\Delta t_i}{\sum_{i=1}^{n} k(t_i).\Delta t_i} \quad (7)$$

Therefore, the dimensionless variance:

$$\sigma_\theta^2 = \frac{\sigma_t^2}{t_m^2} \quad (8)$$

Dispersion model

The concentration distribution in the fluid phase of a liquid fluidized bed containing inert, nonadsorbing componet and nonporous solid is given by the following dispersion equation:

$$\frac{\partial C}{\partial t} = D_l \frac{\partial^2 C}{\partial z^2} - U_{act} \frac{\partial C}{\partial z} \quad (9)$$

Where D_l is the axial dispersion coefficient describing the degree of dispersion (by diffusion and liquid mixing) in the direction of flow (z) through the column and U_{act} is the liquid interstitial (actual) velocity ($U_{act}=U_o/\varepsilon$) in which U_o and ε are liquid superficial velocity and average void fraction of bed respectively. This equation can be written in dimensionless form of:

$$\frac{\partial C}{\partial \theta} = \left(\frac{D_l}{U_{act} \cdot L}\right) \frac{\partial^2 C}{\partial z^2} - \frac{\partial C}{\partial z} \quad (10)$$

with

$$\theta = t \cdot U_{act} / L \quad (11)$$

In which L (or H) is the length (height) of bed. The location of the tracer injection and measurement points has important implications on the choice of the boundary conditions to solve Equation (7). The dispersion model uses a single parameter to describe mixing, which is termed the vessel dispersion number (approximately inverse of mass Peclet number),

$$N_D = D_l / U_{act} \cdot L \quad (12)$$

The vessel dispersion number relates convective transport of liquid to dispersion and is used to describe the influence of axial mixing on the performance of the whole fluidized bed setup.

Using a an impulse input of tracer, a short-cut method can be used with the dispersion model when the vessel dispersion number is low (i.e., N_D<0.01). At such small extents of dispersion, the boundary conditions (i.e., open or closed) imposed on the vessel have little impact on the shape of the tracer curve and a Gaussian approximation for the RTD can be assumed. From the dispersion model and RTD curve which described in detail by levenspiel [11]:

$$\sigma_\theta^2 = 2\frac{D_l}{U_{act} \cdot L} \tag{13}$$

The dimensionless variance can be calculated from Equation (8) and used in Equation (13) to determine the vessel dispersion number. If there is a large deviation from plug flow (i.e., N_D>0.01), the tracer response is broad and spreads as it slowly passes the measurement point giving an asymmetric response, the assumptions made in the use of short-cut method are no longer valid [12].

Experiment

Experiments were carried out in a classical plexiglas liquid-solid fluidized bed of 5cm in diameter and height of 1.2m with tap water as the liquid phase and heavy metalic shots as the solid particles (Figure 1).The density and diameter of the paticles were 10.73g/cm^3 and 2.45mm respectively. The temperature was approximately 20˚C and water flow rate was kept approximately constant in each run of experiments. It varied in the range of 33-220cm^3/s.A 3N sodium chloride solution was used as the tracer. Delta function (impulse) input method was used in this work to measure the RTD of liquid phase. The concentration (conductivity) of tracer was determined by a conductivity meter (Jenway; England). The three different tests were run; one in the column filed with only liquid phase without solid phase, and two tests in the fixed and fluidized bed regimes whose results were presented here. The tracer (1cm^3) was injected as impulse input about 5cm away from the distributor so that the response was free from any kind of flow non-uniformities existing in the distributor region and reflected only the effect of the liquid-solid interaction in the bed itself. Experiments were designed in such a way that the distance between the tracer injection and the tracer measurement points were the same for different tests. The location of the tracer detection point was always at least 5cm below the top of the fluidized bed to keep the measurement free from boundary effects.

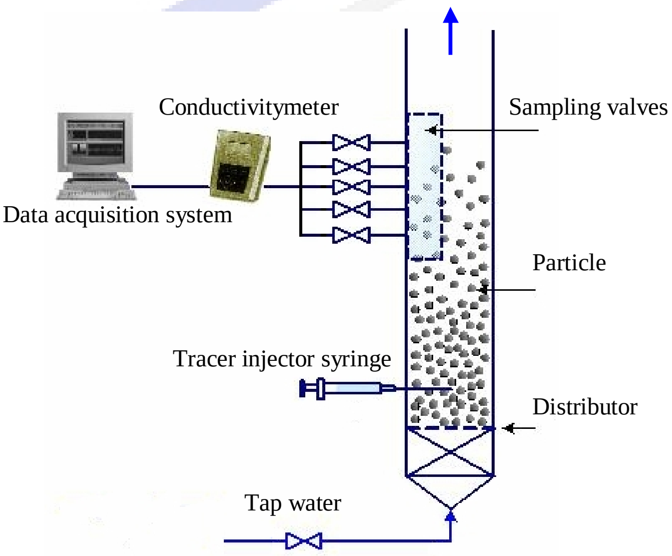

Figure 1: Schematic diagram of experimental setup

The values of the bed voidage were determined from the measured height of expanded bed according to the relationship:

$$\varepsilon = 1 - \frac{M_S}{\rho_P \cdot A \cdot H} \tag{14}$$

Where,

M_S: the total mass of solid in the bed

A: the cross-sectional area of bed

H: height of bed

ρ_P: particle density

Results and discussion

Figure 2 shows the effect of liquid flow rate on the RTD of liquid phase in the fixed bed. In this curves the effect of input RTD (actual impulse injection) was considered. The variances of RTD curves and liquid MRT decrease with increasing volumetric flow rate of liquid.

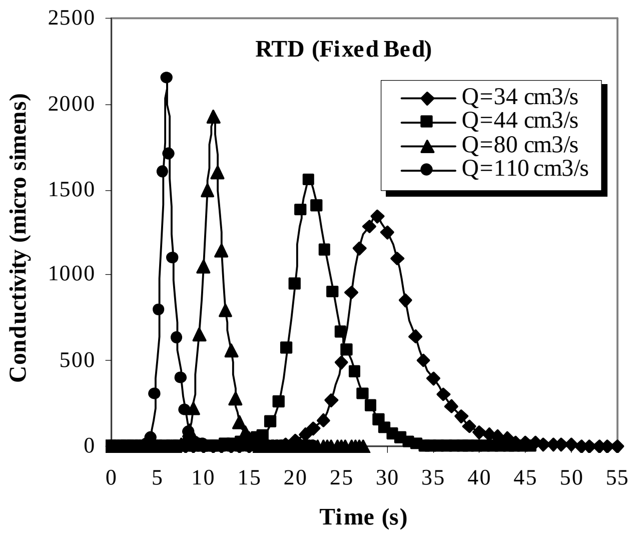

Figure 2: RTD curves of liquid at different liquid flow rate (fixed bed)

With increasing the liquid flow rate more than 110cm^3/s the bed started to become fluidized, and more than 220cm^3/s the particle entrainment was occurred. The results (Figures 3) show that the variance of obtained RTD also decreases with increasing the liquid flow rate, and apparently, it seems that the mixing regime of liquid is become nearly plug flow with liquid velocity.

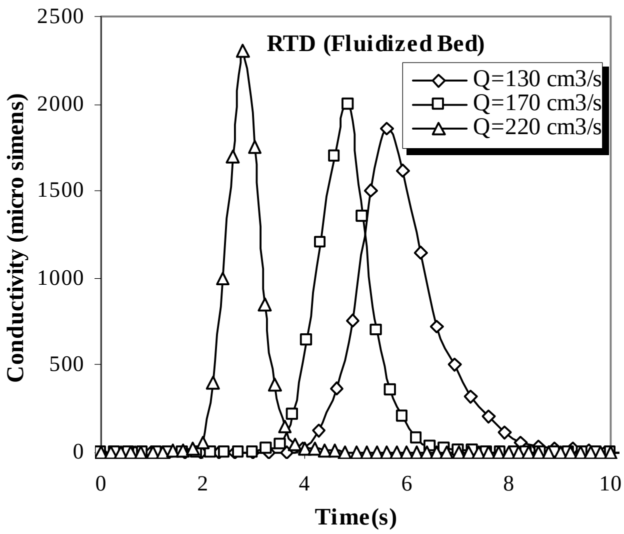

Figure 3: RTD curves of liquid at different liquid flow rate (fluidized bed)

In fact, the calculated ADC (as presented in Table 1) shows the opposite meaning. In the other hand, the effect of liquid velocity, which was not considered in variance, was included in the ADC. Therefore we observed that ADC of liquid increases with increasing the liquid flow rate. The ADC was varied between 1.62 & 4.32cm^2/s in the fixed bed regime and between 5.17 & 8.23cm^2/s in the fluidized bed regime.

Table 1: Effect of liquid superficial velocity on ADC of liquid

Q	Bed's regime	H	ε	U_o	$U_{act.}$	Re_P	σ^2	t_m	$t_{hyd.}$	$D_1^{exp.}$
cm^3/s		cm	-	cm/s	cm^3/s	-	-	s	s	cm^2/s
34	Fixed bed	73	0.269	1.73	6.44	43.18	18.82	30.15	13.97	1.62
44				2.24	8.34	55.88	8.58	22.66	10.80	1.69
80				4.08	15.15	101.61	1.51	11.24	5.94	2.19
110				5.61	20.84	139.71	0.66	6.24	4.32	4.32
130	Fluidized bed	74	0.274	6.62	24.18	165.11	0.67	5.91	3.72	5.17
170		76	0.293	8.66	29.55	215.91	0.38	4.87	3.05	6.22
220		82	0.344	11.21	32.51	279.41	0.13	2.81	2.77	8.23

In which,

Q: liquid flow rate
H: height of bed
ε: average void fraction of bed
$U_{act.}$: actual liquid velocity
Re_P: particle's Reynolds number
σ^2: variance of RTD curves
t_m: MRT of liquid
$t_{hyd.}$: apparent (hydraulics) MRT
$D_l^{exp.}$: experimental ADC

Effect of liquid velocity: The variation in ADC of liquid phase with particle Reynold's number is shown in Figure 4. It can be seen that the ADC increase with increasing Reynold's number. The movement of particles intensifies with increasing liquid velocity, and thus the liquid in the bed is subjected to increasingly vigorous turbulence, resulting in the increase of mixing of the liquid phase. The increase in axial dispersion coefficient with liquid velocity is also observed by Chung and Wen, Krishnaswamy and Shemilt and Tang and Fan for classical liquid-solid fluidized bed [6]. The ADC in fixed bed is smaller than the fluidized bed. Because of the low efficiency of the conventional fixed bed system, much attention has been paid to fluidized bed system.

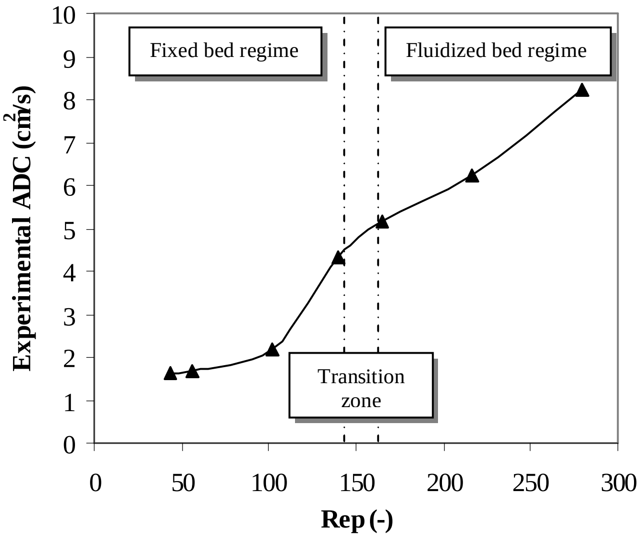

Figure 4: ADC of liquid phase at different particle Reynold's number

The mean residence time of liquid phase (t_m) obtained from the RTD data was compared with the apparent (hydraulics or hydrodynamics) mean residence time (t_{hyd}) calculated by actual liquid velocity and height of bed (Figure 5). In fludized bed regime, the close agreement between residence time (based on hydrodynamics) and residence time (based on RTD) particularly at high liquid flow rates, indicates that the bed behavior become nearly ideal pulg flow in this regime. The difference between t_m and t_{hyd} is due to the dead zones which at high liquid flow rate is smaller than at the low liquid flow rate. In dead zones, the liquid is traped between solid particles with minimum motion. The dead zones were removed by removal of apparent agglomeration of particles. The bed voidage, mixing of solid particles, and turbulent intensity increase with the liquid flow rate which results removeing of dead zone. It can be also cocluded that volume of dead zones decreases with prressure drop [8].

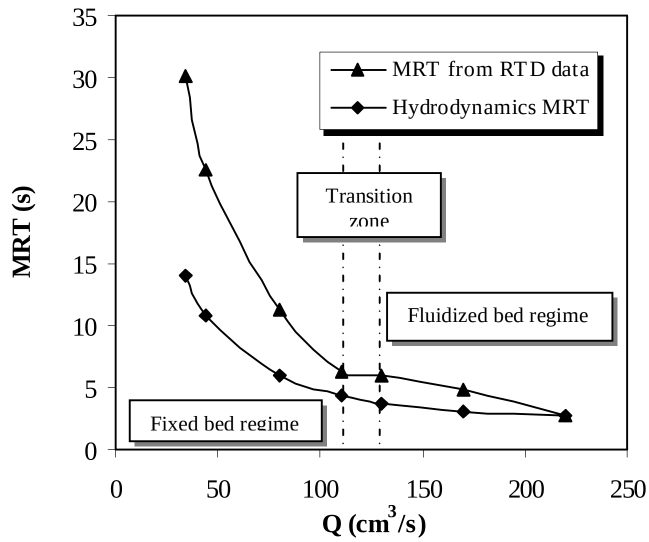

Figure 5: MRT from RTD data and hydrodynamics parameters at different flow rate

Effect of particle density: Our experimental results (Table 2) were compared with that predicted by Tang & Fan (low density particles) and Renganathan & Krishnaiah (very low density particles) for the identical particle's diameter of 0.245cm. ADC increased with increase in liquid superficial velocity (Reynold's number), in our experimental range of flow rate. As difference between solid and liquid density increases, ADC decreases in an identical Reynolds number. For high density particles, the ADC cannot be predicted certainly by existing published correlation equations which are developed for beds of low density particles, indicating that particle density affects axial liquid mixing behavior significantly. Also Chung & Wen have been demonstrated that ADC in a fluidized bed of low density particles behaves differently from a bed of heavy particles.

Table 2: ADC at different liquid superficial velocity for particle's diameter of 0.245cm

Correlation	d_C	ρ_P	$\lvert\rho_P - \rho_L\rvert$	$D_l(cm^2/s)$		
	cm	g/cm^3	g/cm^3	Rep=165	Rep=216	Rep=279
Renganathan et al. (2004)	8.90	0.917	0.080	19.18	30.56	47.59
		0.897	0.100	15.48	24.67	38.41
Tang & Fan (1990)	7.62	1.180	0.183	6.46	9.11	12.68
Our work	5.00	10.730	9.730	5.76	6.22	8.22

In which,
d_C, d_P: column and particle diameter ρ_P, ρ_L: particle and liquid density
Table 3 shows the effect of Rep/Re_{mf} on ADC at different Archimedes numbers. Archimedes number is defined as:

$$Ar = \frac{d_P^3 (\rho_l - \rho_P) \rho_l g}{\mu_l^2} \tag{15}$$

ADC increases with Rep/Re_{mf} in each constant Ar. With increase in Archimedes number, the minimum fluidization velocity increases and so the actual liquid velocity is more for a particle with a higher Archimedes number than for a particle with a lower Archimedes number. In fluidized bed regime, increased liquid velocity implies more axial dispersion coefficient as explained under the effect of liquid velocity. This explains the reason for the increasing trend of axial dispersion coefficient with Archimedes number which is also observed by Chung and Wen and Tang and Fan in classical liquid-solid fluidized beds. To understand the variation in liquid phase dispersion coefficient with liquid velocity for particles having *Ar* between 17.6 and 12947, more studies are required [6]. Literature studies on the axial liquid dispersion coefficients in liquid-solid fluidized beds show significant scatter. This scattering may be associated with the measuring techniques and specific system designs such as column size and distributor design.

Table 3: ADC of liquid phase at different Rep/Re_{mf} for particle's diameter of 0.245cm

Correlation	ρ_P	Ar	Re_{mf}	Rep/Re_{mf} ($D_l(cm^2/s)$)		
	g/cm^3	-	-	Rep=165	Rep=216	Rep=279
Renganathan et al. (2004)	0.917	11969	6.6	25.0 (19.2)	32.7 (30.6)	42.3 (47.6)
	0.897	14961	8.1	20.4 (15.5)	26.7 (24.7)	34.5 (38.4)
Tang & Fan (1990)	1.180	27379	13.8	12.0 (6.5)	15.69(9.1)	20.3 (12.7)
Our work	10.730	1456177	212.4	0.78 (5.8)	1.02 (6.2)	1.3 (8.2)

Figure 6 shows the results of achieved ADC (Table 3) versus Rep/Re_{mf} for different particles densities but the same diameter of 0.245cm. The values present a linear relationship. In a fixed Reynolds number and fluidized regime (i.e. fixed liquid velocity due to identical diameter of particles), the mixing of particles which have low density is more than that of heavy particles. Therefore ADC of light particles is more than that of heavy particle. Besides ADC is also depend to the column diameter and increase with column diameter (Table 2).

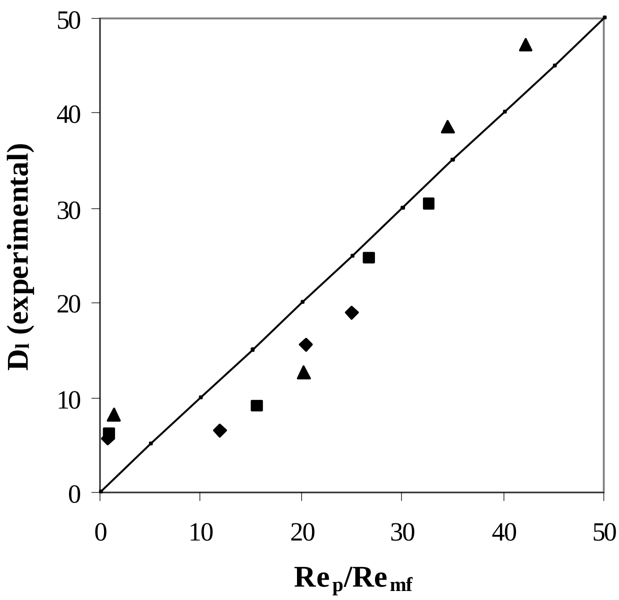

Figure 6: ADC at different Rep/Re_{mf} (fluidized bed)

Conclusion

Results show that axial dispersion coefficient (ADC) of liquid increased with increasing liquid flow rate (Reynold's number) for heavy particles. It observed that particle's density affects axial liquid mixing behavior significantly. The ADC cannot be predicted precisely by Tang et al. and Renganathan et al. correlations, because these literature correlation equations were obtained for light and very light particles. As difference between solid and liquid density increases, ADC decreases. This is due to the fact that it is very sensitive to liquid interstitial velocity.

References

1. Jin, G., and Liu, D., "Modeling and simulation of liquid pulsed particulate fluidized beds", *Powder Technology*, **154**, 138 (2005).
2. Wachem, B., and Almstedt, A., "Methods for multiphase computational fluid dynamics", *Chemical Engineering Journal*, **96**, 81 (2003).
3. Asif, M., and Petersen, J. N., "A dynamic model for the hydrodynamics of a liquid fluidized bed", *Ind. Eng. Chem .Res.*, **33**, 2151 (1994).
4. Kuwagi, K., Takano, K., Horio, M., "The effect of tangential lubrication by bridge liquid on the behavior of agglomerating fluidized beds", *Powder Technology*, **113**, 287 (2000).
5. Aghajani, M., Muller-Steinhagen, H., Jamialahmadi, M., "New design equations for liquid/solid fluidized bed heat exchangers", *International Journal of Heat and Mass Transfer*, **48**, 317 (2005).
6. Renganathan, T., and Krishnaiah, K., "Liquid phase mixing in 2-phase liquid-solid inverse fluidized bed", *Chemical Engineering Journal*, **98**, 213 (2004).
7. Limtrakul, S., Chen, J., Ramachandran, P., Dudukovic, M., "Solids motion and holdup profiles in liquid fluidized beds", *Chemical Engineering Science*, **60**, 1889 (2005).
8. Asif, M., Kalogerakis, N., Behie, L., "Hydrodynamics of liquid fluidized beds including the distributor region", *Chemical Engineering Science*, **47**, 4155 (1992).
9. Tang, W., and Fan, L., "Axial liquid mixing in liquid-solid and gas-liquid fluidized neds containing low density particles", *Chemical Engineering Science*, **45**, 543 (1990).
10. Miura, H., Takahashi, T., Ichikawa, J., Kawase Y., "Bed expansion in liquid-solid two-phase fluidized beds with Newtonian and non-Newtonian fluids over the wide range of Reynolds numbers", *Powder Technology*, **117**, 239 (2001).

11. Levenspiel, O. "Chemical Reaction Engineering", 3rd ed., John Wiley, New York, 295 (1999).
12. Theodossiou I., Elsner H. D., Thomas O. R. T., Hobley., T. J. "Fluidisation and dispersion behaviour of small high density pellicular expanded bed adsorbents", *Journal of Chromatography A*, **964**, 77 (2002).
13. Hartman, M. , Trnka, D., Havlin, V., "A relatoinship to estimate the porosity in liquid-solid fluidized beds", *Chemical Engineering Science*, **47**, 3162 (1992).