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Prediction of bed voidage in multi-phase fluidized bed using Air/Newtonian and non-Newtonian liquid systems

Senthilkumar Kandasamy[a,*], Sivakumar Venkatachalam[b]

[a]*Department of Chemical Engineering, Kongu Engineering College, Perundurai, Erode, Tamil Nadu, India, Mobile No. +91 9842844880; email: uksen2003@gmail.com*
[b]*Department of Chemical Engineering, A.C. College of Technology, Anna University, Chennai, Tamilnadu, India, Mobile No. +919842723454; email: drvsivakumar@gmail.com*

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ABSTRACT

In general, multiphase reactors with different configurations are used in various industries, a fluidized bed reactor has found extensive application in wastewater treatment and other biochemical processes. For the design and development of three-phase fluidized bed reactors, knowledge of the hydrodynamic parameters such as bed voidage is essential. In this paper, an attempt has been made using water, glycerol with different concentrations, and mono ethanolamine as Newtonian liquid and different concentrations of carboxy methyl cellulose as non-Newtonian liquids and seven different particles to study the effect of fundamental and operating variables on bed voidage in a three-phase fluidized bed. The dependency of the bed voidage on various parameters such as the gas and liquid flow rates, particle size and shape, and the physical and rheological properties of liquids are analyzed. The bed voidage increases with fluid flow rates and decreases with an increase in particle diameter and sphericity and it increases with an increase in the viscosity of Newtonian liquids and fluid consistency index of non-Newtonian liquids. On the basis of the experimental results, a generalized correlation has been developed to predict bed voidage in a fluidized bed using Newtonian and non-Newtonian liquids. The experimental results showed good agreement with those predicted according to the developed correlation, with a wide range of operating conditions.

Keywords: Bed voidage; Fluidized beds; Hydrodynamics; Multiphase reactors; non-Newtonian liquids

1. Introduction

The operating modes of three-phase contactors used in industry can be broadly classified under two categories viz., (i) the solid particles in a fixed state (packed bed), and (ii) the solid particles in a suspended state (fluidized bed). The choice of the position of solids depends mainly on the nature of the reaction system. Though three-phase contractors with varying configurations are used in industry, fluidized bed reactor is preferred for many chemical engineering operations, namely, catalytic hydrogenation, hydro-cracking, coal liquefaction, hydro-desulphurization of petroleum fraction, etc., because it offers high mass transfer rates as a result of good mixing [1–4]. Recently, it gained importance in the area of the biotechnological process such as fermentation and aerobic wastewater treatment applications, where bacteria or enzymes are entrapped within porous particles or immobilized on the surface of the inert solids [5]. The industrial effluent is fed into the fluidized bed reactor at a given superficial fluid velocity enough to suspend the support media. The suspended support media are powerfully agitated by the fluid passing through the bed, great mixing is obtained. The main use of the fluid-fluid distributor is to uniformly distribute the wastewater across the

* Corresponding author.

reactor bed. The fluidized media may be an immobilized catalyst in the advanced oxidation process and microorganisms in the biological treatment process. Sometimes, recycling of treated wastewater may be carried out in order to enhance the removal efficiency.

The applications of embedded anaerobic fluidized bed membrane bioreactor for effluent treatment and resulted in minimum energy consumption and lowered membrane fouling. Some of the excellent features of fluidized bed reactor (FBR) are low operating cost, high resistance to system upsets, excellent mixing, high mass transfer rates, and low sludge production, etc. [5,6]. Apart from this, if solid particles are used as catalysts, they can be easily added or withdrawn from the reactor for regeneration. Besides, wastewater which contains more organic matter can be treated by aerobic methods or by anaerobic digestion in fluidized bed reactors or up-flow anaerobic sludge blanket reactors (UASBR). The microbial electrochemical-fluidized bed reactor (ME-FBR) is a single compartment reactor and is comparatively quite easy to perform. In ME-FBRs, the anode behaves in a fluid-like state and it has some advantages such as good electron acceptor, high surface area and carrier for biomass growth which in turn minimize the cells wash-out, higher mass and temperature transport, along with good mixing within the fluidized bed reactor [7]. Various research aspects such as the movement of electrodes, stirred conductive granules, capacitive conductive granules, operating different modulated units in parallel or serial in order to improve treatment efficiency are currently performed. However, this type of reactor may not be cost-effective to treat large volumes of industrial effluent [7].

The challenges like exploiting the present infrastructures of effluent treatment plants (the fluidized bed configurations) in the design of the reactor can be removed by implementing bio-electrochemical effluent treatment along with significant investments. The recent developments in the treatment of industrial effluent are hybrid microbial electrochemical technologies, membrane bioreactor-microbial fuel cells, and hybrid systems based on the combination of electrocoagulation with various biological reactors [7]. The successful scale-up, design, and operation of fluidized beds mainly depend upon the accurate prediction of the behavior and features of the system such as bed voidage. The pioneering work was initiated by Richardson and Zaki [1] and then many investigators made significant contributions toward bed voidage in fluidized beds [8–10]. Further, it was also observed that most of the authors developed bed voidage correlations using Newtonian liquids only [11–16]. The availability of bed voidage correlations using non-Newtonian liquids is limited. Since many biochemical reaction fluids behave as power-law non-Newtonian liquids [17–25], there is a vital need to obtain data with a wide range of variables using non-Newtonian liquids and to develop generalized correlation for representing the data. Hence, an attempt has been made using water, various concentrations of glycerol and mono ethanol amine (MEA) as Newtonian liquid and different concentrations of carboxy methyl cellulose (CMC) as non-Newtonian liquids and different solid particles to study the influence of fundamental and operating variables on bed voidage in a three-phase fluidized bed. The main objectives of the present study is to analyze the dependency of bed voidage on various parameters such as gas and liquid flow rates, particle size and shape, and the physical and rheological properties of liquids in a three-phase fluidized bed. Also to develop a dimensionless correlation for the prediction of bed voidage using Newtonian and non-Newtonian liquids with a wide range of operating conditions and to analyze the applicability of the proposed correlation with the available literature's bed voidage data.

2. Experimental setup and procedure

All experiments were carried out in a Perspex column (0.15 m inner diameter and 1.8 m height) [15,17]. The experimental column had a provision to feed the gas and liquid at the base of the reactor. Using a centrifugal pump, liquid from the storage tank was pumped into the reactor through gas–liquid distributor. Calibrated rotameters were used for the measurement of both gas and liquid flow rates with an accuracy of ±2%. The liquid phase flowed through a calming section of 0.1 m height filled with 0.0048 m Raschig rings and entered the bed through a wire screen, supporting the particles. Compressed air was fed into the bottom of the column through a pressure regulating valve. A gas distributor was provided at the bottom of the fluidized column. The gas–liquid distributor's design details are found in previous works [15,17,19]. Water, different concentrations of glycerol, and MEA are Newtonian liquid systems and different concentrations of CMC, are non-Newtonian liquids and seven different particles were used. The details of the characteristics of different fluids and solids used in the present study are mentioned in Table 1. The bed voidage was determined by measuring the bed height [16]. The bed voidage (ε) is represented as:

$$\varepsilon = \left(\varepsilon_g + \varepsilon_l\right) = 1 - \varepsilon_s \quad (1)$$

where the mean hold-up of solids in the bed was calculated based on the weight of dry particles and height of the fluidized-bed column by:

$$\varepsilon_s = \frac{M_s}{\rho_s A h} \quad (2)$$

A minimum of 3–5 readings were taken and the average value was used for calculations and the reproducibility of the errors was found to be within ±2%.

3. Results and discussion

3.1. Effect of superficial fluid velocities on bed voidage

The bed voidage is found to vary with respect to the fundamental and operating variables. In three-phase fluidization process, it was found that bed voidage depends upon the gas and liquid flow rates. Fig. 1 shows the effect of superficial gas and liquid flow rates on bed voidage for spherical particles (d_p = 0.004 m). As evident from Fig. 1, for any given constant liquid flow rate, the bed voidage increases with an increase in gas flow rates. Similarly, for any fixed gas flow rate, when the liquid flow rate increases, the bed voidage increases. However, at low gas superficial velocities (0.009–0.043 m/s), the effect is not significant. This may have been due to the low preliminary value of the expansion of the bed. From the experimental results, we found that when gas was introduced, for the smaller size particles (1 and 2 mm glass beads) less significant initial contraction in bed voidage was observed which is in concurrence with the literature [8,16]. At high gas superficial velocities (0.06–0.095 m/s) there was a rapid increase in bed voidage. This is because of the larger drag forces applied to the solid particles by an increase in the liquid superficial velocity causing the solid bed to expand.

3.2. Effect of particle diameter and sphericity on bed voidage

The effect of particle diameter on bed voidage can be seen in Fig. 2. It may also be observed that the bed voidage decreases with increasing particle diameter and decreases with increasing sphericity of particle which is shown in Fig. 3, for given constant gas and liquid flow rate. Bed voidage decreases with increasing particle diameter as well as decreasing particle sphericity and mainly due to the bubble wakes of large bubbles present at the bottom of the particle which resists the solid particle from expanding.

3.3. Effect of physical and rheological properties of liquids on bed voidage

The dependency of the bed voidage on the liquid properties was analyzed using eight different liquid systems (water, 20% glycerol, 60% glycerol, 90% glycerol, MEA, 0.1% CMC, 0.5% CMC, and 1% CMC). The bed voidage increases with increasing viscosity of Newtonian liquids (Fig. 4) and fluid consistency index (k) of non-Newtonian liquids (Fig. 5). Increasing fluid consistency index of the CMC solutions increases shear force between the liquid–solid interfaces, thus leading to an increase in bed voidage.

3.4. Improved correlation for bed voidage

The predictive ability of the important available literature correlations (Table 2) was compared with the present data and literature data. The statistical analysis of the present experimental and literature data on bed voidage (Table 3) shows that most of the literature correlations are restricted to the individual author's own range of data. A few researchers have failed to consider the effect of particle characteristics and rheological properties of the fluids such as flow consistency index (k) on bed voidage [8,11] and hence those

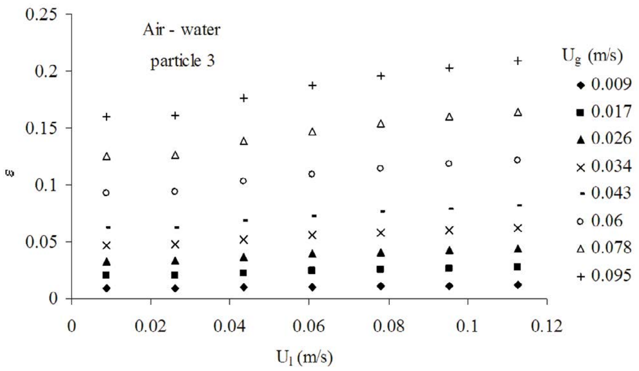

Fig. 1. Effect of liquid and gas velocities on bed voidage.

Table 1
Details of particles and liquid system used in the present work for bed voidage

Bed characteristics		d_p (m)	ρ_s (kg/m^3)	ϕ_s	d_c (m)
Particle 1	Spheres	0.001	2,480	1	0.15
Particle 2	Spheres	0.002	2,480	1	0.15
Particle 3	Spheres	0.004	2,480	1	0.15
Particle 4	Spheres	0.0055	2,480	1	0.15
Particle 5	Spheres	0.0072	2,480	1	0.15
Particle 6	Berl saddles	0.0048	2,050	0.33	0.15
Particle 7	Raschig rings	0.0051	2,480	0.58	0.15
Properties of fluids		Density (kg/m^3)	Viscosity (kg/m/s)		Surface tension (N/m)
			k (kg/m^{-1}/s^{n-2})	n	
System 1	Water	1,000	0.00085	1	0.072
System 2	20% Glycerol	1,010	0.002	1	0.069
System 3	60% Glycerol	1,020	0.006	1	0.069
System 4	90% Glycerol	1,040	0.01	1	0.068
System 5	MEA	1,050	0.015	1	0.045
System 6	0.1% CMC	1,020	0.00842	0.92	0.072
System 7	0.5% CMC	1,020	0.01838	0.88	0.072
System 8	1% CMC	1,020	0.0548	0.86	0.069

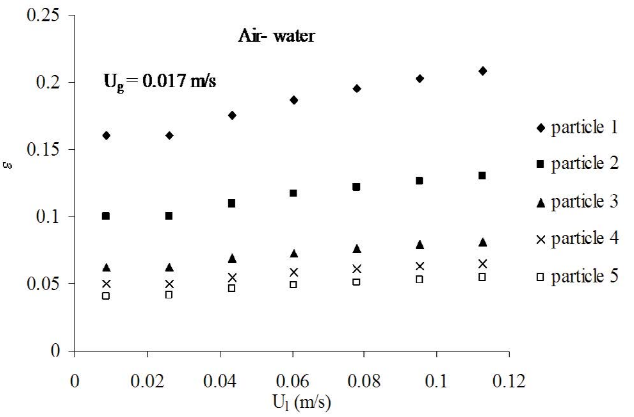

Fig. 2. Effect of particle diameter on bed voidage.

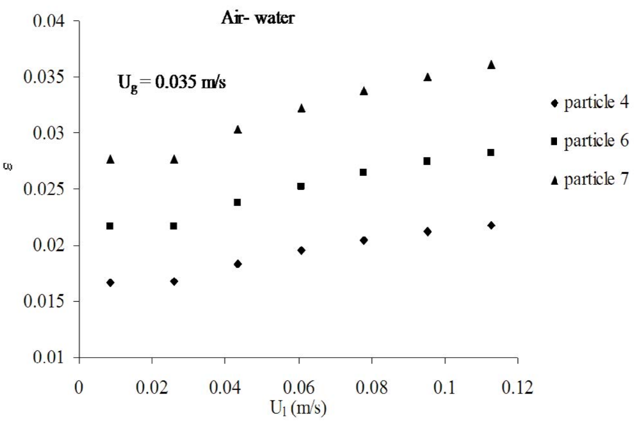

Fig. 3. Effect of sphericity of particle on bed voidage.

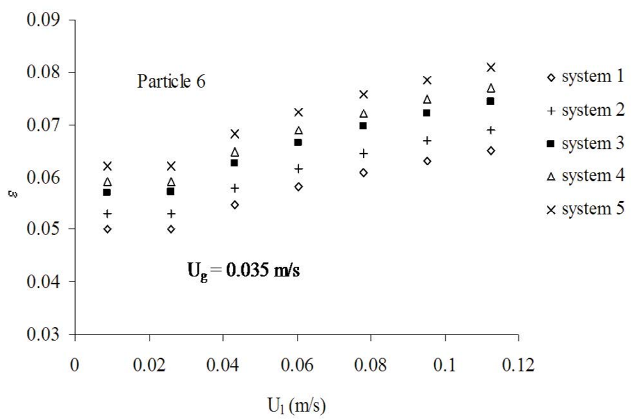

Fig. 4. Effect of physical properties of liquids on bed voidage.

correlations predict results with higher deviation for both present and literature data [8–10]. Graphical analysis of the present data shows that the variation of bed voidage can be attributed to the effect of all the above-said variables.

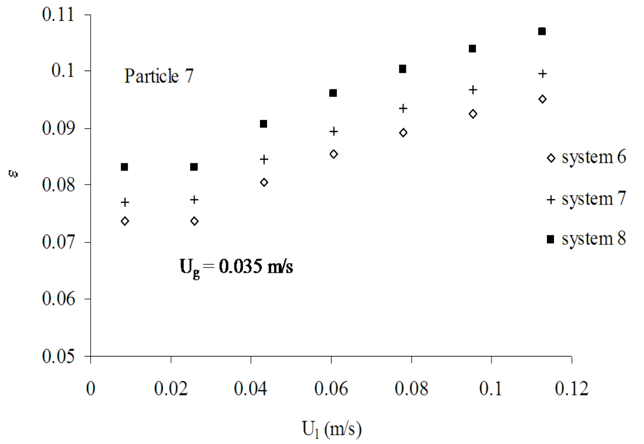

Fig. 5. Effect of rheological properties of non-Newtonian liquids on bed voidage

In this study, the approach of the dimensionless method was adopted for the establishment of bed voidage correlations. The combined effects of the liquid properties were accommodated using Modified Morton's number. Regression analysis of the experimental and available literature data (Table 3), obtained by using nineteen liquid systems with 25 different particles, gave the constants and indices for the bed voidage correlation as given below:

$$\varepsilon = 0.9\left(\mathrm{Fr}_l\right)^{0.09}\left(1+\mathrm{Fr}_g\right)^{0.6}\left(\mathrm{Mo}_{l,M}\right)^{0.017}\left(\frac{\rho_s}{\rho_l}\right)^{-0.03}\left(\frac{d_p}{d_c}\right)^{0.01}\left(\varphi_s\right)^{-0.37} \quad (3)$$

Using the proposed correlation (Eq. (3)), statistical error analysis has been performed and obtained an absolute average relative deviation (AARD) of ±8.8% for bed voidage indicating a satisfactory representation of the available data for air-Newtonian and air-non-Newtonian systems. The applicability of the present correlation has been tested with the available literature's bed voidage data [8–10], which shows a satisfactory agreement (Table 4).

4. Conclusion

In a three-phase fluidized bed, the dependency of bed voidage on various operating and fundamental variables with a wide range has been analyzed. A dimensionless correlation for the prediction of bed voidage has been developed and its applicability has been analyzed using the present experimental data along with those of published literature sources covering a wide range of variables. The statistical analysis confirmed that the predictive ability of the proposed correlation is good. Therefore, the proposed correlation can be confidently used for the estimation of the bed voidage, with the knowledge of the fundamental and operating variables.

Table 2
Details of literature data used for bed voidage analysis

Bed characteristics	d_p (m)	ρ_s (kg/m^3)	ϕ_s	d_c (m)	Reference
Glass ballotini	0.00028	2,960	1	0.1016	[8]
Glass ballotini	0.00058	2,940	1	0.1016	
Glass ballotini	0.0012	2,700	1	0.1016	
Glass ballotini	0.002	2,880	1	0.0508	
Glass ballotini	0.0022	2,500	1	0.1016	
Spheres	0.0013	2,700	1	0.056	[9]
Spheres	0.00013	1,100	1	0.056	
Spheres	0.00106	2,700	1	0.056	
Spheres	0.002235	2,710	1	0.056	
Spheres	0.003348	2,400	1	0.056	
Spheres	0.006844	2,400	1	0.056	
Spheres	0.00489	2,260	1	0.056	
Spheres	0.003	7,707	1	0.056	
Spheres	0.006	2,300	1	0.66	[10]
Spheres	0.001	2,950	1	0.66	
Properties of fluids	Density (kg/m^3)	Viscosity (kg/m/s)		Surface tension (N/m)	
		k (kg/m^{-1} /s^{n-2})	n		
Water	680	0.00036	1	0.018	[8]
Water	995	0.00085	1	0.0712	[9]
Kerosene	800	0.0017	1	0.026	
0.17 wt.% CMC	1,002	0.02	0.916	0.0733	[10]
0.35 wt.% CMC	1,001	0.07	0.914	0.0738	
25 wt.% Sugar	1,090	0.00237	1	0.0729	
42 wt.% Sugar	1,170	0.0076	1	0.0759	
40 vol.% Acetone	960	0.00143	1	0.0398	
Water	1,000	0.001	1	0.0728	
0.1 wt.% CMC	1,004	0.0063	0.971	0.0728	
0.17 wt.% CMC	1,002	0.02	0.916	0.0733	
36 wt.% Sugar	1,150	0.00464	1	0.0755	

Table 3
List of important literature correlations for bed voidage

Author	Correlations	System	Range of variables
Begovich and Watson [4]	$\varepsilon = 3.93\left(\mu_l\right)^{0.055}\left(U_l\right)^{0.271}\left(U_g\right)^{0.041}\left(\rho_s-\rho_l\right)^{-0.316} d_p^{-0.268}\left(d_c\right)^{-0.033}$	Air–water	d_p = 0.0046–0.0062 m ρ_s = 1,720–2,440 kg/m^3 U_g = 0–0.17 m/s U_l = 0–0.12 m/s d_c = 0.0762 and 0.152 m
Nikov et al. [11]	$\varepsilon = \left(2.5+13.2\mu_l^{0.64}\right)\left(U_l\right)^{0.271}\left(U_g\right)^{0.041}\left(\rho_s-\rho_l\right)^{-0.316} d_p^{-0.268}\left(d_c\right)^{-0.033}$	Air–mineral oils and mixtures of Vitrea oils with kerosene	d_p = 0.002–0.006 m ρ_s = 2,420–2,850 kg/m^3 U_g = 0–0.0.05 m/s U_l = 0.003–0.043 m/s μ_l = 10 to 120 MPa.s ρ_l = 850–870 kg/m^3

Table 4
Statistical comparison of bed voidage with present and literature data

Correlations	Present data		Kim et al. [10]		Dakshinamurthy et al. [9]		Ostergaard and Theisen [8]	
	AARD%	Bias	AARD%	Bias	AARD %	Bias	AARD %	Bias
Begovich and Watson [4]	20.37	1.43	10.26	1.09	25.34	0.87	21.86	1.28
Nikov et al. [11]	19.8	1.49	10.49	1.05	24.29	0.88	22.33	1.29
Present correlation (Eq. (3))	8.8	1.01	9.7	1.09	11.2	1.01	11.4	1.08

Symbols

A — Cross-sectional area of bed, m^2

AARD — $\frac{1}{N}\sum_{i=1}^{N}\frac{\left|\left(\text{experimental} - \text{calculated}\right)\right|}{\text{experimental}}$

Ar_l — Liquid Archimedes number, $\frac{g d_p^3 \rho_l^2}{\mu_l^2}$

Bias — $\exp\frac{1}{N}\sum_{i=1}^{N}\ln\frac{\text{experimental}}{\text{calculated}}$

Bo — Bond number, $\frac{g d_p^2 \rho_l}{\sigma_l}$

d_c — Column diameter, μ

d_p — Particle diameter, μ

Fr_g — Froude number of gas, $\frac{U_g^2}{g d_p}$

Fr_l — Froude number of liquid, $\frac{U_l^2}{g d_p}$

g — Acceleration due to gravity, m/s^2

h — Height of bed, m

k — Flow consistency index, kg/m^{-1}/s^{n-2}

Mo_l — Morton number of liquid, $\frac{\mu_l^4 g}{\rho_l \sigma_l^3}$

$\text{Mo}_{l,M}$ — Modified Morton number of liquid, $\frac{\text{We}^3}{\text{Fr}_l N_{\text{Re}\,l,M}^4}$

M_s — Total solid mass in bed, kg

N — Number of data points

n — Fluid behavior index

N_{Ga} — Galileo number, $\frac{g d_p^3 \rho_l \left(\rho_s - \rho_l\right)}{\mu_l^2}$

$N_{\text{Re}l}$ — Reynolds's number of liquid, $\frac{d_p U_l \rho_l}{\mu_l}$

$N_{\text{Re}l,M}$ — Modified Reynolds's number of liquid, $\frac{d_p^n U_l^{2-n} \rho_l}{k}$

U_g — Superficial gas velocity, m/s

U_l — Superficial liquid velocity, m/s

We_l — Weber number, $\frac{d_p u_\ell^2 \rho_\ell}{\sigma_\ell}$

Greek

ρ_l — Liquid density, kg/m^3

σ_l — Liquid surface tension, N/m

μ_l — Liquid viscosity, kg/m/s

ρ_s — Particle density, kg/m^3

ε_s — Solid holdup

ε — Voidage of bed

ϕ_s — Sphericity of particle

ρ_g — Gas density, kg/m^3

μ_g — Gas viscosity, kg/m/s

Abbreviations

AARD — Absolue average relative deviation

CMC — Carboxy methyl cellulose

MEA — Mono ethanol amine

UASBR — Upflow anaerobic sludge blanket reactor

ME-FBR — Microbial electrochemical-fluidized bed reactor

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